Process integration and economics evaluation of sugar beet pulp conversion into ethanol

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Abstract: The aim of the study was to evaluate, from an economic standpoint, the feasibility of using sugar beet pulp (SBP) as the feedstock in an existing sugar processing plant to ethanol. Two base cases were studied. Case 1 incorporated dilute sulfuric acid pretreatment, enzymatic hydrolysis, and fermentation using *S. cervisiae*. Case 2 neglected the pretreatment step and used a series of enzymes in Simultaneous Saccharification and Fermentation (SSF) with *S. cervisiae* yeast followed by *E.coli K011* fermentation. The ethanol production cost for each case was estimated to be $1.50 and $1.10 per gallon of ethanol for case 1 and case 2, respectively. Assuming a 10% discount rate, a minimum selling price of $2.35 per gallon was obtained for case 1 and $1.53 per gallon for case 2. These prices can be competitive with the increasing gasoline prices. However, base case 2 has higher potential to be feasible with the discovery of efficient microbial species.

Keywords: sugar beet pulp, fuel conversion, lignocellulosic, biomass, ethanol, feasibility, economics, pretreatment, enzymatic hydrolysis
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1 Introduction

Presently, most of the ethanol produced in the United States (U.S.) is derived from corn. However, corn is a valuable food source for humans and animals. Nonetheless, dedicating all the U.S. corn production to bioethanol would meet only 15% of the annual gasoline consumption[1]. The Energy Independence and Security Act of 2007 developed by the U.S. government require 36 billion gallons of renewable fuel by the year 2022[2]. To help alleviate this in part, there is a need to investigate an alternative feedstock for ethanol production. Ethanol from lignocellulosic material is advantageous because it is abundantly available with low cost. Sugar beet pulp (SBP) is an attractive feedstock for ethanol production, because it is a co-product from the table sugar industry. After sucrose extraction, the remaining plant fiber is SBP.

Sugar beets are farmed throughout the world in temperate climates; however, in the U.S, sugar beet farming is concentrated in the northern plains, North Dakota, South Dakota, Minnesota, Montana and Colorado. The SBP consists of 20%-24% cellulose, 25%-36% hemicellulose, 20%-25% pectin or uronic acids, 1%-2% lignin, and 7%-8% protein, all expressed as a percentage of dry weight of total solids[3]. The SBP addresses some of the logistical constraints most biomass encounter, such as feedstock harvest, feedstock prices, transportation, and storage. Beet harvesting equipment and transportation methods are well established to deliver

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the product to the sugar processing plant. After sucrose extraction, the remaining pulp is dried, pelletized, and sold to farmers as animal feed. Doran and Foster[4] estimated that between 30%-40% of the overall energy cost of sugar beet processing is devoted to dehydrating and pelletizing the pulp. Consequently, because SBP would not need to be dried, the use of SBP as an ethanol feedstock could be more profitable than processing for animal feed. The sugar beet industry already has many built-in production advantages that favor the production of high performance fuels at existing sugar beet processing plants. First, there are large quantities of SBP already concentrated with no additional transportation cost. Second, energy consumption would be minimal. In the past, much attention was paid to pretreatment[5-7], enzymatic hydrolysis[8-10], and fermentation[11-13] stages. This work combines all of the aforementioned stages.

The National Renewable Energy Resource Laboratory (NREL) has conducted a number of feasibility studies on lignocellulosic biomass materials to ethanol processes, primarily with corn stover[14,15]. Also NREL has investigated the feasibility of integrating lignocellulosic ethanol into existing corn based ethanol plants[16]. Nguyen and Saddle[17] used process simulation models to evaluate the technical and economic feasibility of a lignocellulosic to ethanol bioconversion process using aspen wood. Outlaw et al.[18] analyzed the feasibility of integrating ethanol production into the existing sugar mill that uses sugarcane juice as the feedstock for ethanol production.

The objective of the present work was to investigate the feasibility of integrating an ethanol producing plant into an existing sugar processing plant that uses SBP as feedstock. The primary idea was after sucrose extraction, the remaining pulp at 75% (W/W) moisture would be transported to the integrated ethanol plant. The energy previously used to dry and pelletize the pulp would be diverted to operate the ethanol plant. The benefit of conducting this research is to show the feasibility of using SBP as feedstock over corn stover and ultimately to corn starch.

2 Materials and methods

The development of process flow diagrams is the first step in any conceptual process design. Because of the complexity of lignocellulosic ethanol production, there is a variety of processing conceptual designs presented in the literature. Two process flowsheets from the literature were used in this study. The first is by NREL[14] and the second is by Rorick[19]. The processes were modeled using ASPEN Plus simulation software. This software was used to solve the mass and energy balances and to calculate thermodynamic properties of the streams in the process. The component physical properties used in the simulation were obtained from Wooley & Putsche[20]. The properties not directly available were estimated from similar components. Material balances were used to size major equipment. Total project investments were determined for both cases. We used a discounted cash flow analysis to determine the production cost of ethanol when the net present value of the project was zero for both cases.

American Crystal Sugar, which is owned by approximately 3 000 shareholders, planted about 443 000 acres of sugar beets resulting in about 10 million tons of sugar beets in 2010[21]. Their processing plant in East Grand Forks, Minnesota, one of the five in the Red River Valley, handles an average of 2 300 000 tons of sugar beets annually[22]. For a typical sugar beet processing plant, 250 kg of pressed beet pulp with 75% moisture remain after the removal of sucrose from one ton of sugar beets, equivalent to about 62.5 kg of dry matter beet pulp material[10]. As a result, the plant is assumed to operate at 574 990 MT of SBP per year.

2.1 Experiment

2.1.1 Compositional analysis

The SBP was provided by American Crystal Sugar (East Grand Forks, MN 56721). Compositional analysis of the SBP was determined based on National Renewable Energy Laboratory Analytical Procedure (NREL/ TP-510-42619).

2.1.2 Pretreatment

There are many types of pretreatment processes but one unifying goal is to get the highest fermentable sugar

conversion possible after enzymatic hydrolysis. Dilute sulfuric acid pretreatment was used in this study because it has been widely used and relatively inexpensive and effective to treat different biomass species[23]. In addition, this will hydrolyze the amorphous polymeric hemicellulose into monomeric saccharides in the hydrolyzate and thus improves the cellulose accessibility to enzymes during enzymatic hydrolysis[24, 25].

Pretreatments were performed in a 300 mL internal volume batch reactor manufactured by Autoclave Engineers (Erie, PA 16509). The reactor was fabricated with Hastelloy ® C-276 because this alloy is highly resistant to acid corrosion in high temperature environment. Reactor was equipped with external jacket for heating and cooling. Saturated steam and cold water were used as heating and cooling elements. The average heating kinetics of the reactor was around 35℃/min. Details of the novel pretreatment reactor can be found at the work of Degenstein et al.[26]. The agitation was performed by magnetic motor and was maintained constant at 60 r/min throughout the reaction period. Steam was injected into the reactor from the boiler by operating a three-way valve manually. The reaction time was initiated when the desired temperature was reached in the reactor. After completion of the reaction pretreatment run, steam was shut off and cooling water was pumped into the external jacket of the reactor. Once the reactor was cooled down below 40℃, slurry samples were withdrawn from the reactor into polyethylene bottles for further analysis.

2.1.3 Enzymatic hydrolysis

Enzymatic hydrolysis of the pretreated solids was performed in a MaxQ 4000 thermal incubator manufactured by Thermo Scientific (Portsmouth, NH 03801) at 50℃ at 250 r/min for 72 h. Hydrolysis was performed with sodium citrate buffer with 50 mM L^{-1} concentration (pH of 4.8) and sodium azide with concentration of 20 g/mL. These reagents along with de-ionized (DI) water were added to the 1.5% dry substrate of glucan so that total volume of the batch was 10 mL. A commercially available enzyme GC220 (Genencor, Palo Alto CA) with loading of 20 mg of protein/g of cellulose was used to perform enzymatic hydrolysis. After hydrolysis, samples were filtered and analyzed in HPLC for fermentable sugar yields. The cellulose digestion was calculated by using the Equation (1). The value 0.9 was used in the equation as a correction factor for hydration.

$$\%\,\text{Digestion}=\frac{\text{Grams of cellulose digested}\times 0.9\times 100}{\text{Grams of cellulose added}} \quad (1)$$

2.1.4 Analytical procedure

Pretreated slurry samples were vacuum filtered and separated into Water Insoluble Solids (WIS) and Water Soluble Carbohydrates (WSC). The WSC were analyzed for monosaccharides and inhibitor products. This analysis was performed based on the NREL analytical procedures (NREL/TP-510-42623). The WIS were analyzed for glucan, xylan, and acid insoluble lignin (AIL) content. Quantitative analysis for determining monosaccharides present in WSC was performed by Agilent 1200 HPLC with Transgenomic CHO-Pb column. All samples were replicated and analyzed for 30 minutes in HPLC. The mobile phase used for analysis was DI water with a flow rate of 0.6 mL/min.

2.2 Case 1 process description

Aden et al.[15] developed a process flowsheet for conversion of biomass to ethanol. Figure 1 is a simplified version of their model, which was used as the base case for producing ethanol from SBP. Slight alterations of NREL's model were done as the feed handling section was eliminated because SBP does not require a preparation step prior to pretreatment. Also the cellulose production section was removed. We assumed that purchasing the enzymes from vendors would be economical. Furthermore, we assumed that the plant will produce only ethanol. In NREL's model, they dehydrated and burned the remaining slurry after distillation in a boiler to generate steam and electricity[23]. The slurry mainly consists of lignin. The benefits gained from producing electricity would be offset by the cost of dehydrating the slurry. Most biomass consists of 10%-30% lignin while SBP contains 1%-2% lignin[10]. Dehydrating and burning the slurry to produce electricity was not considered in this study. A waste water treatment section was also not considered in the ASPEN

plus model since the existing sugar processing plant had a water treatment facility in place and it was assumed to have the capacity to treat the waste output from the integrated plant.

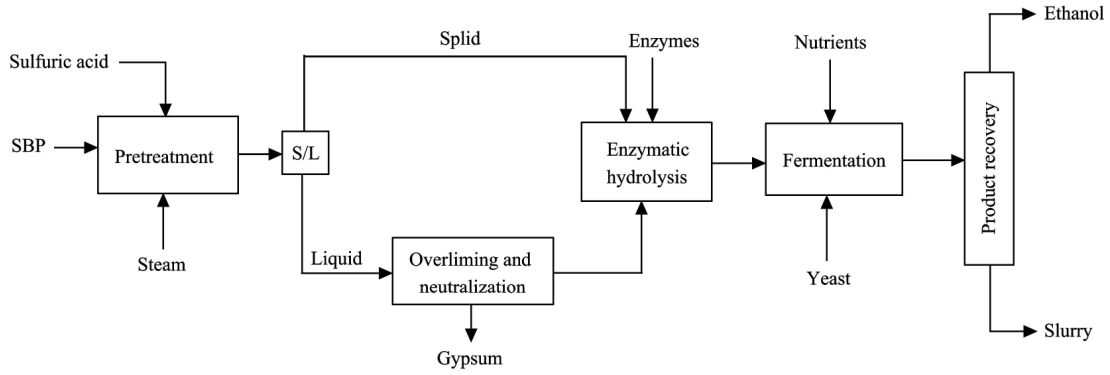

Figure 1 Process flowsheet diagram for base case 1 (SBP is sugar beet pulp as the feedstock , S/L is solid/liquid separation step)

The process flowsheet shown in Figure 1 was used in the ASPEN Plus simulation of the plant. After sucrose extraction, the remaining pulp is transferred to the pretreatment section. For a typical biomass to ethanol conversion process, pretreatment is used to combat the recalcitrant nature of the biomass[6]. The main objective of the dilute acid and high temperature pretreatment is to solubilize the hemicellulosic fraction of the biomass. Moreover, some of the lignin in the feedstock is also solubilized[7]. As a result, cellulose is more susceptible to enzymatic attack. The glucan in the hemicelluloses and a small portion of the cellulose also are converted to glucose. Degradation products such as: pentose sugars (primarily furfural), hexose sugars (primarily hydroxymethyl furfural), and acetic acid are formed during pretreatment[15].

The data generated from the 300 mL Hastelloy ® C-276 batch reactor was used in the ASPEN Plus simulation. The SBP pretreatment was conducted at 150℃ with 1.1% sulfuric acid and 10% solid loading for a total residence time of 12 minutes. The pretreatment temperature was selected based on the equipment capability and previous study[27]. The temperature used is lower than that would be used commercially. A longer residence time was used to offset the lower temperature. Five percent of glucan is converted to glucose, 60% of xylan, and 55% of arabinan are converted to their respective sugars under these conditions. A small residence time is preferred for the commercial plant because the pretreatment reactor is expensive. Therefore, we modeled the pretreatment based on recommendations made by Aden et al.[15]. The pulp was treated with 1.1% dilute sulfuric acid at 190℃ with a residence time of two minutes. We measured the yield using our lab scale equipment (300 mL reactor). We also assumed that yields similar to our experiments would be achieved at this elevated temperature with the shorter residence time at commercial scale. These yields from the experiments were used in the process simulation.

After pretreatment, the effluent is flash-cooled to remove the degradation products detrimental to downstream fermentation microorganisms. Some of acetic acid and most of 5-hydroxy-2-methylfurfural HMF and furfural are removed. In addition, the liquid is separated from the solid and treated by addition of lime. Overliming is a temperature and pH treatment designed to aid the conversion of hydrolyzed sugars to ethanol during fermentation[15]. For efficient organism sugar uptake, it is necessary for minimum concentrations of calcium. Overliming limits calcium concentration to tolerable levels[14]. The hydrolyzate is adjusted to pH 4.8 with sodium hydroxide and the resulting gypsum is separated and discarded. The conditioned liquid is mixed with the solid stream and fed to the enzymatic hydrolysis reactor.

In the hydrolysis step, water was added to bring the hydrolysis total solids to 20%. Though the lab scale pretreatment experiments were conducted at 10% solid due to equipment constrain. Due to the SBP's low bulk density, 20% was achievable in the experimental

apparatus. For this feasibility studies, we assumed similar results would be achieved at 20% solid loading at a commercial scale. This step was modeled as simultaneous saccharification and co-fermentation (SSCF). This configuration allows the hydrolysis step to be operated at the enzymes optimal operating temperature. The hydrolysis section was modeled using a continuous reactor at 50℃ with a total residence time of 36 hours. The cellulase enzyme (GC220) with a protein content of 212 mg protein/L was diluted with water and fed to the reactor with enzyme loading of 20 mg/g of cellulose. The data used in the simulation was obtained from this laboratory work. Enzymatic hydrolysis of SBP was conducted in 125 mL Erlenmeyer flasks with enzyme loading of 20 mg/g of cellulose and SBP solid loading of 1.5% for 72 hours. Fifty-five percent of the glucose hydrolyzed during the first 34 hours and 80% after 72 hours. We assumed that similar yields would be achieved at a larger scale. Therefore, we used the obtained yields in the simulation. Enzyme recycle was not considered in this study because it had the potential to build up degradation products in the process.

After 36 hours, the slurry was cooled to 31℃ and fed to the fermentation tank. The fermentation process was modeled with a continuous reactor at 31℃ with 36 hours of residence time. It was assumed that saccharification would continue during the fermentation step; however, at a slower rate. *Saccharomyces cerevisiae* was added to the reactor to ferment glucose to ethanol. This organism was selected for its ability to ferment glucose at industrious scale and for its high ethanol tolerance. *Saccharomyces cerevisiaes* achieve lower yield due to the organism inability to ferment arabinose and galacturonic acid from SBP[12]. In the simulation, 20% ethanol yield based on glucose was used. The yield used in the simulation was obtained from laboratory work done by Rorick[19]. In addition to ethanol, acetic acid, and CO_2 were also produced.

The product recovery section consists of two distillation columns and a molecular sieve. The first column removes most of the dissolved CO_2 and most of the water. The second column concentrates the ethanol to a near azeotropic composition. Molecular sieves further remove the water present in the azeotropic mixture. A water scrubber recovered most of the ethanol in the fermentation vents. The ethanol is cooled and sent to storage tank.

2.3 Case 2 process description

For the second base case, we assumed the same plant capacity (574 990 MT per year) as case 1. In case 2, the pretreatment section was removed from the process flowsheet. The main purpose of pretreatment is to remove the lignin barrier in biomass. However, SBP contains 1%-2% lignin. Therefore, it is economical to remove the pretreatment section[3]. The SBP contains 20%-25% pectin. The presence of pectin forms a barrier that reduces cellulose susceptibility to enzymatic attack. As a result, pectinase was added to hydrolyze the pectin portion. ASPEN Plus simulation software was also used to model this case. The simulation model was based on the flowsheet shown in Figure 2. The model consists of a simultaneous saccharification and fermentation (SSF) section, followed by another fermentation step, and then a product recovery section. The data used in the simulation were obtained from laboratory work conducted by Rorick[19]. Rorick[19] investigated maximizing ethanol titers and yields through enzymatic hydrolysis of SBP and fermentation of the five and six carbon sugars. Hemicellulose and pectin were also hydrolyzed and fermented separately from cellulose to increase ethanol titers and yields.

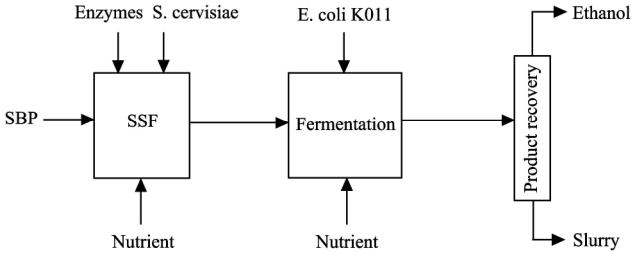

Figure 2 Process flowsheet diagram for base case 2 (SSF is simultaneous saccharification and fermentation, SBP is sugar beet pulp as the feedstock)

The SSF section was modeled with two continuous stirred reactors in series operating at 39℃. The SBP was treated with pectinases, cellulases, and *S. cerivisiae* organisms. Pectinase hydrolyzed the pectin and the

hemicellulose portion of the biomass to galacturonic acid and mostly arabinose. Cellulase hydrolyzed the cellulose portion to glucose. Cellubiases was added to hydrolyze the remaining cellulose in the SBP completely. *Saccharomyces cerivisiae* fermented the resulting glucose to ethanol. The enzymes and organisms have different optimal temperatures. We assumed that 20% yield based on glucose and 21% yield based on hemicellulose was achievable at the industrial scale based on the lab scale data from Rorick[19] and used it in the simulation. After a total residence time of 72 hours, the slurry was sent to the *E. coli* fermentation section. *Saccharomyces cerevisiae* organism performs well when converting glucose to ethanol. However, it cannot convert the arabinose and galacturonic acid portions, which accounts for 42% of SBP to ethanol. *Escherichia coli KO11* (*E. coli KO11*), on the other hand, can convert both arabinose and galacturonic acid to ethanol. Therefore, *E. coli KO11* organism was fed to two continuous stirred tanks in series at 37℃ for an additional 72 hours. Please note *E. coli* can survive in an ethanol solution generated by yeast. However, the yield may be lowered due to the presence of inhibitors and ethanol concentration.

2.4 Theory and calculation

The component stream balances were used to size the major process equipments. Most of the equipment costs used in this study were obtained from Humbird et al.[27]. The pretreatment reactor was estimated using Guthrie's purchase cost correlation[28]. Due to the presence of sulfuric acid and elevated temperatures in case 1, Hastelloy C Alloy was selected as the material of construction for the pretreatment reactor to combat corrosion effect.

$$New\ Cost = Original\ Cost \times \left(\frac{New\ Size}{Original\ Size} \right)^{\exp} \quad (2)$$

Equation (2) was used to adjust the equipment costs for other alternative sizes being evaluated. The scaling exponents (exp) were obtained from Ulrich[29] and Aden et al.[15]. The equipment installation factors used in this study were obtained from Peters & Timmerhaus[30]. Using the Chemical Engineering Purchased Equipment Index, the equipment costs were adjusted to the projected year. Accessible values for the index ranging from 1990 to 2008 were regressed to a simple equation. The generated equation was used to extrapolate to future years. Once the total installed equipment cost (TIEC) was determined, other costs associated with the project were added: such as warehouse, construction cost, and start-up cost. We assumed the cost of warehouse, construction and start-up to be 1.50%, 10%, and 8% of TIEC, respectively. Summation of the above costs gives the total project investment.

Raw materials quantities used in determining the operating cost were obtained from the ASPEN Plus model. The costs of the materials were obtained from NREL[15]. The costs of chemicals associated with the process were indexed to the projected year using the Industrial Inorganic Chemical Index. The fixed operating cost includes labor cost and various overhead items. Overhead items consist of general overhead, maintenance, insurance, and taxes. Applying guidelines presented by Douglas[28], it was assumed the general overhead to be 60% of the total salaries, maintenance to be 2% of installed equipment cost, insurance, and taxes to be 1.50% of the total installed equipment cost. The labor costs were indexed to the operating year using the labor index from the Bureau of Labor Statistics[31]. The available data were regressed to a simple equation. The regression equation was used to extrapolate to the assumed projected year.

Discounted cash flow analysis was used to determine the minimum selling price per gallon of ethanol produced for both cases based upon the total project investment, variable operating costs, and fixed operating cost. Microsoft Office Excel 2007 solver was used to iterate on the selling price of ethanol until the net present value of the project is zero.

The assumptions used to evaluate the discounted cash flow rate of return (DCFROR) analysis are listed in Table 1. Double declining balance (DDB) depreciation method was used for the income tax calculation. This was used because it provides greater returns at the early stages of the project. The discount rate for this analysis was set at 10%. The assumptions presented in Table 1 are based on the recommendations made by Short[32] on

how to perform economic evaluation of renewable energy technologies. He recommended using a discount rate of 10% in the absence of statistical data on discount rates used by industrial, transportation, and commercial investors for investments with risks similar to those of conservation and renewable energy investments.

Table 1 DCFROR assumptions for the two base cases

Plant life	20 years
Salvage Value	0
Depreciation Method	DDB
Depreciation Periods	7 years
Income Tax Rate	39%
Discount Rate	10%
Working Capital	10%

3 Results and discussion

3.1 Economic evaluation

We present results of an economic evaluation of integrating a lignocellulosic ethanol plant into an existing sugar processing plant using SBP as the feedstock. We assumed that the energy saved from omitting the SBP dehydrating and pelletizing steps would be used to operate the plant. As a result, utilities were not considered in this work. Two base cases were considered and their results are presented.

All of the costs reported in this study were in 2012 US dollars. The plant capacity for both base cases was 574 990 MT of SBP per year. The overall production capacity for base case 1 was 1.7 million gallons per year and 2.26 million gallons of ethanol per year for case 2. The observed difference in the production capacity is due to the yields achieved at the enzymatic hydrolysis and fermentation stages.

Table 2 summarizes the capital requirements and percentage breakdown of installed equipments costs for both examined cases. For case 1, the pretreatment and distillation sections were the largest contributors to the total direct cost and together represented 84% of the total direct cost. The total project investment was estimated to be \$11.4 million for base case 1 and \$6.4 million for case 2. The total capital investment (TPI) for case 2 is 50% less than that of case 1. This is not surprising, considering that the pretreatment section representing 55% of the total direct cost that was not considered in case 2.

The overall ethanol production cost for the base cases were estimated to be \$1.50 and \$1.10 per gallon of ethanol for case 1 and case 2, respectively. The production cost in case 1 is 37% higher than in case 2, because case 1 has a lower production capacity and higher capital investment.

Table 2 Total project investment (2012 dollars)

	Case 1	TDC /%	Case 2	TDC /%
Direct cost				
Pretreatment	\$5 238 500	55		
Enzymatic hydrolysis	\$53 000	7		
Fermentation	\$789 100	8	\$2 520 300	48
Distillation	\$2 774 200	29	\$2 774 200	52
Total direct cost (TDC)	\$9 453 900		\$5 294 500	
Indirect costs	\$1 101 400		\$616 800	
Other costs	\$844 400		\$472 900	
Total project investment	\$11 399 700		\$6 384 200	

The fixed and variable operating costs are presented in Table 3. The maintenance, taxes, and overhead costs are based on guidelines presented by Peters & Timmerhaus[30]. The Table presents total annual costs and cost per gallon of ethanol for both base cases. As expected, case 2 has a higher organisms cost than case 1. Case 2 incorporated three different enzymes in the SSF step, which led to a higher operating cost in this category.

Table 3 Operating cost breakdown for two base cases

Items	case 1 cents/gal	case 2 cents/gal
Organisms[a]	12.50	47.20
Chemicals	14.30	3.20
Labor	49.70	32.60
Waste disposal	11.50	-
Maintenance, taxes, overhead	49.50	27.80
Total	126.6	110.90

Note: [a] Enzymes cost are included in organisms cost.

In case 2, the organisms cost corresponds to 47.24 cents per gallon of ethanol. It would seem organisms recycling would be beneficial. However, recycling could increase the amount of contaminants in the process, thus, reducing the effectiveness of the downstream organism. Only gypsum was considered in the waste disposal cost. We assumed that the existing treatment

facility could handle the generated waste streams; as a result, it was not included in the cost analysis.

Once the total project cost and operating cost were determined, a discounted cash flow analysis was used to determine the minimum selling price per gallon of ethanol produced. Assuming 10% discount rate, Microsoft Office Excel 2007 solver was used to iterate on the selling price of ethanol until the net present value of the project was zero. For case 1, a minimum selling price of \$2.35 per gallon was obtained and \$1.53 per gallon for case 2. Case 1 has a higher selling price for ethanol than case 2 because it has a lower ethanol production capacity with high operating cost. NREL[27] assumed a discount rate of 10% and obtained a minimum selling price of \$2.15 (2012 dollars) per gallon for a standalone lignocellulosic ethanol plant using corn stover with a capacity of 2000 MT per day. The selling price for case 1 is higher than that of NREL and case 2 is lower than NREL model. Though the presented processes are undemonstrated technologies with uncertain yields at larger scale, case 2 has the potential to be feasible for a company to implement the process.

3.2 Sensitivity analysis

We identified the costs of the enzymes and the yield of organisms as the key variables that could greatly affect the economics of the process. A sensitivity analysis was carried out to investigate the effect of enzyme cost and yield of organisms on the minimum selling price of ethanol. We assumed that all other process input variables were constant. Variation in the cost of enzyme would have a direct effect on the ethanol selling price. For instance, if the cost of the enzyme was reduced by 50%, the minimum selling price reduces from \$2.35 to \$2.23 and from \$1.53 to \$1.28 for case 1 and case 2, respectively. As expected, reducing enzyme cost has a larger effect on case 2 than case 1 due to the high enzymes requirement.

However, if the cost of the enzymes increased by 50%, the minimum selling price would increase from \$2.35 to \$2.41 for case 1 and \$1.53 to \$1.75 for case 2. To combat this effect, the residence time in the saccharification tanks could be increased. Increasing the residence time could reduce the amount of enzymes required to achieve a specific yield. The total project investment (TPI) is expected to increase with increasing residence time. By doubling the residence time, TPI increased by \$0.65 million and \$0.66 million for case 1 and case 2. Also the ethanol selling price increased by \$0.09 per gal for case 1 and \$0.11 per gal for case 2. As a result, the benefits gained from doubling the residence time in the saccharification and fermentation tanks are offset by the increased in the TPI. The price of enzyme is expected to fall in the coming future as enzyme production technologies advance[33].

The effect of increasing the yield of enzymatic hydrolysis and fermentation stages on the selling price of ethanol was also analyzed. The yield at the enzymatic hydrolysis and fermentation stages were increased by 10%, 20%, and 30% for both cases. Figure 3 shows how the minimum selling price is affected by an increase in the organisms yield. As expected, increasing the yields reduced the minimum ethanol selling price for both cases. Increasing the yields by 30% resulted in selling price of \$1.40 and \$1.17 for case 1 and case 2, respectively.

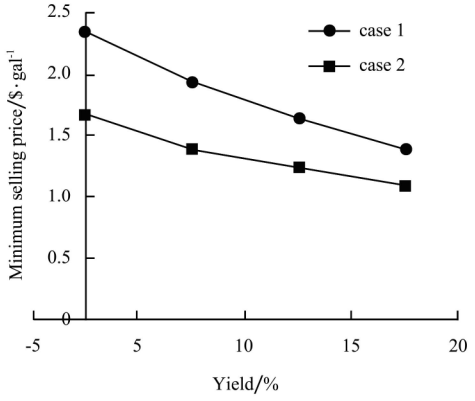

Figure 3 Effect of increasing the yield of enzyme and the organism during the hydrolysis and fermentation stages on the minimum selling price of ethanol

Case 1 only utilized the cellulose portion of the biomass. The SBP contains 20%-24% of cellulose[4]. Case 2, on the other hand, utilized the cellulose, pectin, and hemicellulose portion, which accounts for about 65% of the SBP. Case 2 has the potential to be cost competitive in today's ethanol market. However, it is suffering from low reaction yields because the enzyme mixtures and the organisms operate at different optimal

temperatures (50℃ and 35℃). The presence of inhibitors from the SSF step limits the effectiveness of the *E. coli* organisms in the second fermentation step. Rorick[19] suggested including a separate step to remove some of the acetic acid present in the slurry prior to second fermentation step. This would reduce the level of inhibitors and increase the fermentation yield for case 2; however, it could introduce additional cost.

NREL[15] investigated the minimum selling price of ethanol as a function of plant sizes and suggested that plant sizes below 2000 MT per day have higher costs. A rule of thumb in plant economics is the cost reduces with increasing plant size. The plant size in this study was 1643 MT per day because it is limited by the SBP output from the sugar processing plant. North Dakota has five sugar processing plants. Rorick[19] suggested building a centralized lignocellulosic ethanol plant that would utilize the SBP produced by the sugar processing plants. This would increase the plant size from 1643 MT to 7143 MT per day and reduce the minimum ethanol selling price. However, the plant will have a higher energy cost with additional transportation cost.

4 Conclusions

The minimum selling ethanol prices obtained in both cases are comparable to NREL study[15, 27] for production of ethanol from corn stover. However, the yields of SBP conversion to ethanol presented have not been proven at a pilot scale. The obtained cost of $2.35 for case 1 is probably too high for the project to be feasible with the current gasoline price. The obtained cost of $1.53 for case 2 has good potential for the project to be feasible and competitive with the gasoline price. Developing organisms that can ferment the cellulose, pectin, and hemicelluloses portions efficiently will make SBP an attractive feedstock for lignocellulosic ethanol. The onset of low cost and high efficiency enzymes will also make this process economically attractive. It is beneficial for U.S. to be energy independent. However, the economics today are not viable for commercialization. Further studies are needed to address some of the process uncertainties.

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