

## Article

# Techno-Economic Implications of Fed-Batch Enzymatic Hydrolysis

Ellen Argo and Deepak R. Keshwani \*

Department of Biological Systems Engineering, University of Nebraska, Lincoln, NE 68583, USA;  
eemanuel@nebraska.edu

\* Correspondence: dkeshwani2@unl.edu; Tel.: +1-402-472-6714

Received: 9 October 2019; Accepted: 7 November 2019; Published: 12 November 2019



**Abstract:** Fed-batch enzymatic hydrolysis has the potential to improve the overall process of converting cellulosic biomass into ethanol. This paper utilizes a process simulation approach to identify and quantify techno-economic differences between batch and fed-batch enzymatic hydrolysis in cellulosic ethanol production. The entire process of converting corn stover into ethanol was simulated using SuperPro Designer simulation software. The analysis was conducted for a plant capacity of 2000 metric tons of dry biomass per day. A literature review was used to identify baseline parameters for the process. The sensitivity of the ethanol production cost to changes in sugar conversion efficiency, plant capacity, biomass cost, power cost, labor cost, and enzyme cost was evaluated using the process simulation. For the base scenario, the ethanol unit production cost was approximately \$0.10/gallon lower for fed-batch hydrolysis. The greatest differences were seen in facilities costs, labor costs, and capital costs. Using a fed-batch operation decreased facilities costs by 41%, labor costs by 21%, and capital costs by 15%. The sensitivity analysis found that cost of biomass had the greatest effect on ethanol production cost, and in general, the results support the proposition that fed-batch enzymatic hydrolysis does improve the techno-economics of cellulosic ethanol production.

**Keywords:** fed-batch hydrolysis; techno-economic analysis; cellulosic ethanol; process simulation

## 1. Introduction

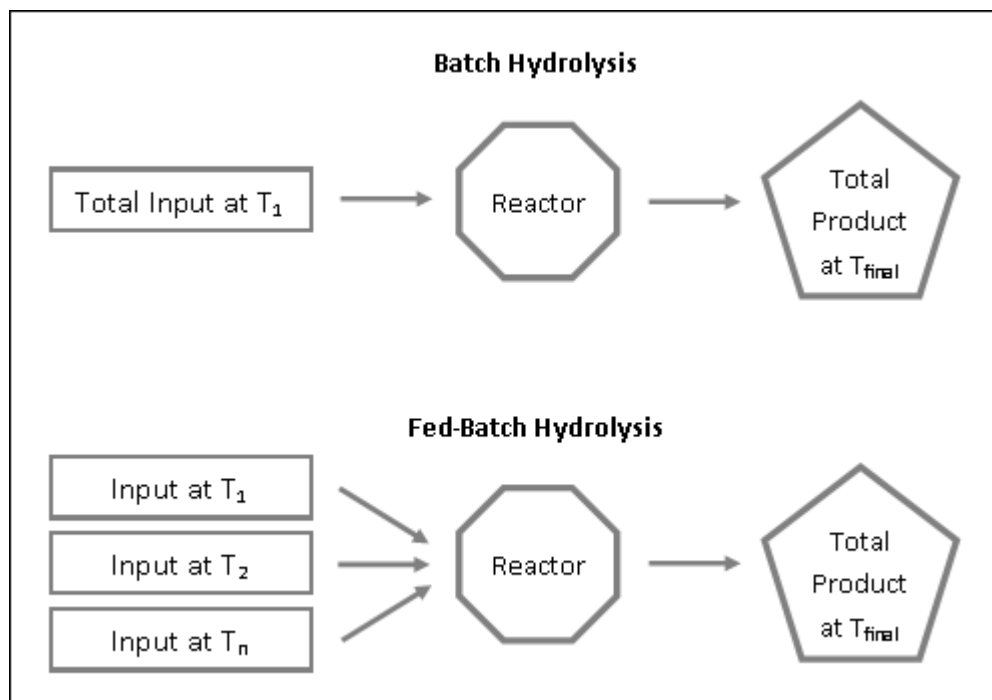
Ethanol production capacity in the United States has increased for seven consecutive years [1]. However, production of cellulosic ethanol continues to fall short of reaching target production goals established by the Energy Independence and Security Act of 2007, and the U.S. Environmental Protection Agency (EPA) has offered cellulosic waiver credits to reach Renewable Fuel Standard requirements since 2010 [2]. The EPA website for the waivers states, “Cellulosic fuels have not yet been produced in sufficient amounts to satisfy the volume levels listed in the Clean Air Act” [3]. Clearly there are many opportunities for optimizing the cellulosic ethanol production process.

This research aims to optimize the hydrolysis step in cellulosic ethanol production. Most cellulosic ethanol production uses a batch method for hydrolysis. However, previous work in this lab and other studies [4–7] has investigated the potential benefits of utilizing fed-batch enzymatic hydrolysis instead. The research suggests that using a fed-batch method will maintain a higher reaction rate [8–11], relieve mass transfer problems, and allow for higher total cumulative solids in the hydrolysis reactor [4,5].

Likely, a fed-batch process could also lead to improvements (reductions) in water, chemical, and energy use in cellulosic ethanol production. Kazi et al. [12] found that increasing solids’ consistency during pretreatment improved the product value because it could use a smaller reactor volume, which meant a lower process heating requirement. One would expect similar findings for fed-batch enzymatic hydrolysis.

Figure 1 illustrates the differences in batch versus fed-batch processes. In a batch process, all of the inputs are added to the reactor at the same time, and when the reaction is complete, all of the product is removed at the same time. In a fed-batch process, inputs are added to the reactor in small amounts over time intervals. When the reaction is complete, all of the product is removed at one time.

This paper identifies and quantifies techno-economic differences between cellulosic ethanol production using batch enzymatic hydrolysis versus fed-batch enzymatic hydrolysis. A sensitivity analysis is included for both the batch and fed-batch scenarios to observe the magnitude of effects caused by changing parameters. These effects are compared between the batch and fed-batch scenarios, as well as with their respective base case scenarios.



**Figure 1.** Visual comparison of batch versus fed-batch hydrolysis, where T represents time interval, n, and the total input in batch hydrolysis is equal to the sum of the inputs in fed-batch hydrolysis.

## 2. Materials and Methods

Two separate simulations were set up as base case scenarios: One using batch enzymatic hydrolysis and one using fed-batch enzymatic hydrolysis. The simulations differed only in their hydrolysis operation; they were exactly the same for every other process operation. Simulation results from these base cases were compared to identify the techno-economic effects of using a fed-batch operation instead of a batch operation.

This study used the SuperPro Designer (SPD) simulation software (Version 9.5, Intelligen, Scotch Plains, NJ, USA, 2015) [13], because it was designed specifically to model bioprocesses. SPD also has built-in economics calculations, which was a key component of this study. It is important to note there are three levels of complexity in an SPD simulation. “The simplest physico-chemical transformation step that can be modeled by SuperPro Designer [is a unit operation]. Operations are strung together to form a unit procedure and unit procedures are put together to make up a process (or a recipe)” [14]. An operation may be as simple as ‘Charge’ or ‘Mix,’ or it may be more complex, e.g., ‘Distill’ or ‘React.’ A procedure is “a sequence of actions representing the most elementary physico-chemical transformations supported by the software all assumed to take place within the same equipment resource” [14]. This paper uses the same naming convention for these steps.

SPD comes with an example process flow sheet for converting corn stover to ethanol. This process flow sheet was modified to fit the needs of this study. Appendix A provides a sample flowsheet from this study, for reference. The operating parameters for the simulation can be found in Table 1.

**Table 1.** Operating parameters for SuperPro Designer base case simulations. IRR indicates the internal rate of return.

	BATCH BASE CASE	FED-BATCH BASE CASE
PROCESS TYPE	Continuous	Continuous
HOURS OF OPERATION PER YEAR	7920	7920
PLANT CAPACITY	2000 metric tons/day	2000 metric tons/day
DEPRECIATION	10 years, straight-line method	10 years, straight-line method
% EQUITY FINANCED	100	100
PROJECT LIFE	20 years	20 years
IRR (AFTER TAX)	3.52%	6.33%
STARTUP PERIOD	4 months	4 months
CONSTRUCTION PERIOD	30 months	30 months
YEAR OF ANALYSIS	2013	2013
INFLATION	4.00%	4.00%

The plant in this study is assumed to be located in Ravenna, Nebraska due to the availability of corn stover, as this is a high corn-producing region of the state. The plant capacity is set to match the processing capacity of plant designs in other techno-economic analyses, such as the National Renewable Energy Laboratory (NREL) standard [15,16] and others [12,17–20].

Corn stover is the biomass in this study because it has shown promise as a lignocellulosic ethanol feedstock, and it is readily available in Nebraska. The study assumed the corn stover biomass would be transported 50 kilometers (km), and each shipment contained 20 metric tons (MT). Our overall operation would require nearly 66,000 shipments/year. The composition of the corn stover was assumed to be as follows (mass percentages given): 5.2% ash, 37.4% cellulose, 21.1% hemicellulose, 18% lignin, and 18.3% other solids [15]. After the feedstock arrives at the plant facility, it is first washed and ground to reduce particle size. The feedstock mixture for the ethanol production process consisted of 50% corn stover, 50% water (mass percentages given).

For pretreatment, our design uses thermal hydrolysis (hot steam). The thermal hydrolysis pretreatment will degrade the structure of the biomass and leave the cellulose more accessible to the enzyme in the upcoming enzymatic hydrolysis operation. Hot, high-pressure steam is fed into the reactor at a rate of 30 metric tons per hour, temperature of 200 degrees Celsius (°C), and pressure of 10 bar. The feedstock slurry enters the reactor at 215 metric tons per hour, 88 °C, and 10 bar. Within the reactor, the contents sit at 180 °C and 10 bar. The residence time is 30 minutes. During this time, some cellulose is broken down into glucose, and the majority of the hemicellulose is broken down into xylose. The conversion of cellulose to glucose is set to 10%. The conversion of hemicellulose to xylose is set to 70%. The pretreatment reaction is assumed to be adiabatic.

Entering the pretreatment reactor, the feedstock slurry has the following composition (approximate mass percentages given): 15% cellulose, 9% hemicellulose, 48% water, 13% lignin, 3% glucose, 1% xylose, 11% other. Leaving the pretreatment reactor, the slurry composition changes to become (approximate mass percentages given) 12% cellulose, 2% hemicellulose, 53% water, 11% lignin, 4% glucose, 7% xylose, 11% other.

After the thermal hydrolysis, the slurry is flash cooled. Some excess water is removed and some of the xylose is filtered out of the slurry. After cooling and filtration, the slurry has the following composition (mass percentages given): 16% cellulose, 3% hemicellulose, 45% water, 15% lignin, 3% glucose, 6% xylose, 12% other.

Next, the hydrolase enzyme is mixed into the slurry for the enzymatic hydrolysis operation at a rate of 13 metric tons per hour, 25 °C, and 1 bar. After mixing the enzyme into the slurry stream, the hydrolase comprises just 0.2% mass composition of the stream. This study assumed the hydrolase is

purchased from an external source at \$11.40/kg protein. This price factors out to about \$0.50/gallon (gal) of ethanol produced.

For both simulations (batch and fed-batch), the hydrolase enzyme is mixed into the stream before the slurry enters the hydrolysis reactor. The batch enzymatic hydrolysis reaction uses 2123 metric tons of hydrolase enzyme per year, which comes to 0.268 metric tons per hour. The batch enzymatic hydrolysis reaction was assumed to be adiabatic. The contents of the reactor were recorded at about 45 °C and a pressure of about 10 bar. The cellulose to glucose reaction was assumed to run to 90% completion, and the hemicellulose to xylose reaction was assumed to run to 70% completion. When the batch enzymatic hydrolysis is complete, the slurry stream composition is as follows (approximate mass composition percentages given): 2% cellulose, 1% hemicellulose, 48% water, 14% lignin, 18% glucose, 7% xylose, 10% other.

The simulated fed-batch enzymatic hydrolysis reaction uses 2091 metric tons of hydrolase enzyme per year, which comes to 0.264 metric tons per hour. The fed-batch enzymatic hydrolysis was also assumed to be adiabatic. The contents of the reactor were recorded at about 45 °C and a pressure of about 10 bar. The reaction was assumed to run to full completion due to the nature of a fed-batch operation within a continuous process. When the fed-batch enzymatic hydrolysis is complete, the slurry stream composition is as follows (approximate mass composition percentages given): 0% cellulose, 3% hemicellulose, 47% water, 14% lignin, 19% glucose, 6% xylose, and 11% other.

After hydrolysis, the hydrolysate slurry is filtered. The stream containing mostly glucose and water is sent on to fermentation. The stream containing mostly lignin, ash, and water is further processed. Most of the lignin is sent to be burned in the utilities section of the plant to generate power.

In the fermentation section, some of the slurry is used in seed fermentation tanks to grow the yeast cells. The whole slurry is fermented into ethanol. Our process used four 2220 m<sup>3</sup> fermentation tanks with a temperature of 37 °C and a cycle time of 48 h. The slurry stream then enters a storage holding tank until it can be distilled to a higher percentage of ethanol.

The slurry stream leaves the storage holding tank and enters a heat exchanger to facilitate the distillation process. Leaving the heat exchanger, the stream has a temperature of 47 °C. The stream is now 9% ethanol and 80% water (approximate mass percentages given) when it begins the distillation process. The distillation columns operate at a temperature of 106 °C. Leaving distillation, the stream is 90% ethanol, 9% water (approximate mass percentages given). Next, an adsorption operation further dehydrates the stream, removing the little water remaining, such that the ethanol product reaches 99.9% purity.

The utilities section burns lignin obtained from hydrolysis to generate power. The generated power is sold back to the grid; it is not used within the plant. Selling the power generates additional profit for the production plant facility. The utilities section also recycles water for continued use within the plant.

Using the simulation, the sensitivity of the ethanol production cost to different process parameters can be tested. A review of the literature yielded differing values for basic parameters, so each parameter value was altered while the other parameters were held constant. Only one parameter was altered at a time, and the change in ethanol production cost relative to the base case scenario was monitored.

Table 2 gives an economic summary of some basic economic parameters for our simulated SPD base cases. Comparing the cash flow analyses, the batch process had higher capital investments, lower gross profits, greater depreciation, and lower net cash flows. Hence, the batch process had a lower net present value and lower internal rate of return (IRR) compared to the fed-batch process. All economic values for this study are reported in US 2013 dollars.

**Table 2.** Economic summary of SuperPro Designer base cases. ROI indicates return on investment; IRR indicates internal rate of return; gallon is abbreviated as gal.

	Batch	Fed-Batch
Total capital investment (\$)	196,487,072	167,194,736
Annual operation cost (\$/year)	117,650,740	111,650,689
Annual ethanol revenue (\$/year)	116,565,239	115,657,243
Ethanol unit production cost (\$/gal)	2.4537	2.3469
Ethanol unit production revenue (\$/gal)	2.4848	2.4855
ROI (%)	9.27	11.29
IRR (after tax) (%)	3.52	6.33
Input target ethanol sell price (\$/gal)	2.50	2.50

Information for utility costs was obtained from the website for the Nebraska Energy Office. The average industrial rate for electricity cost in the Dawson Public Power District (service provider for Ravenna, NE) in November 2016 was \$0.12 per kilowatt-hour (kW-h) [21].

This study assumed all labor workers in the plant were standard operators receiving the same wages of \$25 per hour. Information for operator salaries comes from the NREL studies [15,16] and from the website for the United States Bureau of Labor and Statistics information for chemical plant and systems operators [22].

It is important to note that the ethanol production cost accounts for all operations within the production process. Pretreatment, fermentation, distillation, and the utilities operations of the simulation all impact the ethanol production cost; it is not affected only by the enzymatic hydrolysis operation, even though this study aims to observe the impact of the enzymatic hydrolysis operation.

### 3. Results and Discussion

#### 3.1. Batch vs. Fed-Batch

Figure 2 shows the breakdown of annual operating costs for both SPD base cases, and Table 3 gives a side-by-side comparison of the values for the batch and fed-batch SPD base case models. The batch base case had higher values in every category. Equipment is the largest contributor to capital costs, and the batch process required one additional granular activated carbon (GAC) column and five additional stirred tank reactors, compared to the fed-batch process. The batch process also had a greater labor requirement, contributing to the higher total cost of labor for the batch process when compared to the fed-batch process. As seen in Figure 2, raw materials were by far the greatest contributor to annual operating costs for both batch and fed-batch processes.

**Table 3.** Side-by-side comparison of batch and fed-batch SuperPro Designer base case results.

	Batch	Fed-Batch
Capital cost (\$)	196,487,072	167,194,736
Ethanol produced (gallons/year)	47,900,000	47,600,000
Power required (kW-h/year)	50,386,181	50,151,126
Power cost (\$/year)	2,519,309	2,507,556
Total utilities cost (\$/year)	7,254,020	7,323,292
Labor requirement (hours/year)	189,541	182,941
Total labor cost (\$/year)	4,957,604	4,578,098
Facilities costs (\$/year)	34,888,000	29,634,000
Amount of enzyme used (metric tons per year)	2123	2091
Cost of enzyme (\$/year)	24,202,200	23,837,400
Ethanol unit production cost (\$/gallon)	2.4537	2.3469

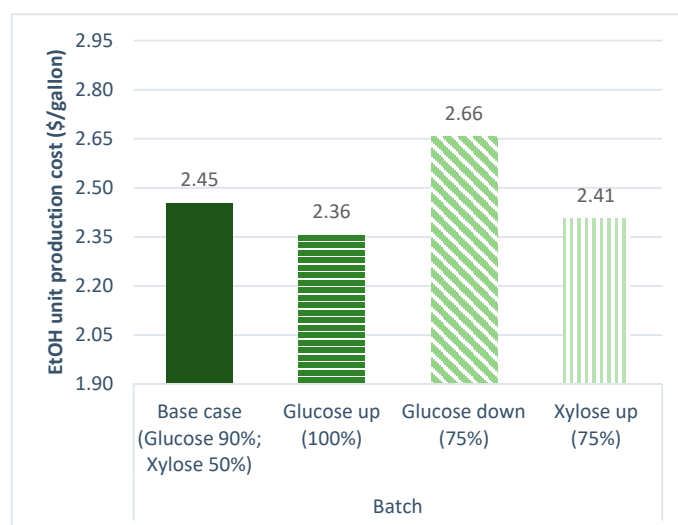


**Figure 2.** Annual operating cost breakdown for (a) batch SuperPro Designer base case and (b) fed-batch SuperPro Designer base case.

### 3.2. Sensitivity Analysis

#### 3.2.1. Sugar Conversion

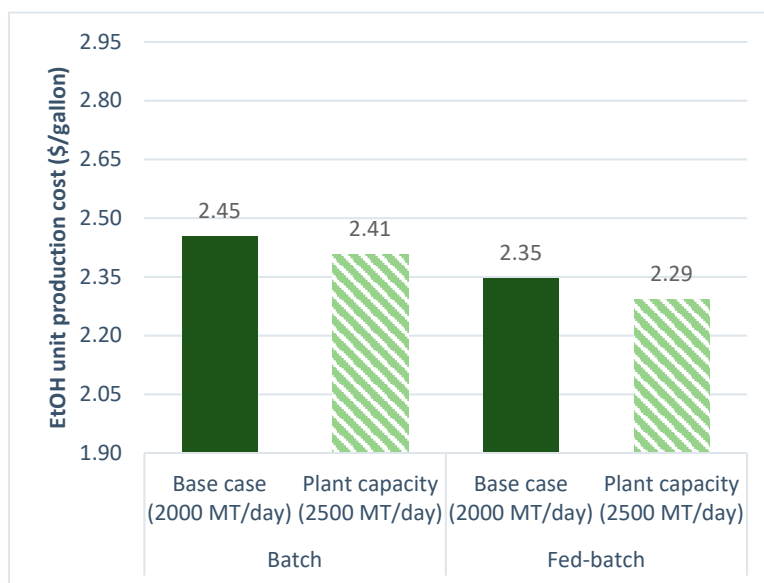
This study tested the effect of sugar conversion on the ethanol production cost. In the base case scenarios, it was assumed the glucan to glucose conversion reaction would reach 90% completion and the xylan to xylose conversion reaction would reach 50% completion. Unfortunately, the fed-batch operation in SPD is rigid, and the extent completion for the fed-batch process base case could not be altered. Therefore, only the batch process base case was used to test the sensitivity of the ethanol production cost to the sugar conversion. First, the glucan to glucose conversion extent was adjusted from 90% up to 100% and then down to 75%. Then, the xylan to xylose conversion extent was adjusted from 50% up to 75%. For both reactions, when extent of conversion increased, the ethanol production cost decreased relative to the base case. When the extent of conversion decreased for the glucan to glucose reaction, ethanol production cost increased relative to the base case. Figure 3 shows the change in ethanol production cost for each of the adjusted conversions.



**Figure 3.** Comparison of ethanol production cost for the batch SuperPro Designer base case and adjusted sugar conversion extents.

### 3.2.2. Plant Capacity

Plant capacity was increased to observe the effect on the ethanol production cost. Assuming the same land area could produce extra biomass, the transportation distance and transportation costs did not change. The assumed location of the plant near Ravenna, Nebraska means availability of biomass is not expected to be a bottleneck, as this area is high in corn production. Hence, we assume additional corn stover availability is not problematic and transportation distance and costs will not change significantly. After increasing the input parameter for production capacity of the plant, SPD automatically resized the equipment within the process design as necessary. By increasing the plant capacity to 2500 metric tons per day, the ethanol production cost decreased by approximately \$0.05/gal for the batch case and by \$0.06/gal for the fed-batch case. Figure 4 illustrates the change in ethanol production costs as a result of increasing plant capacity.

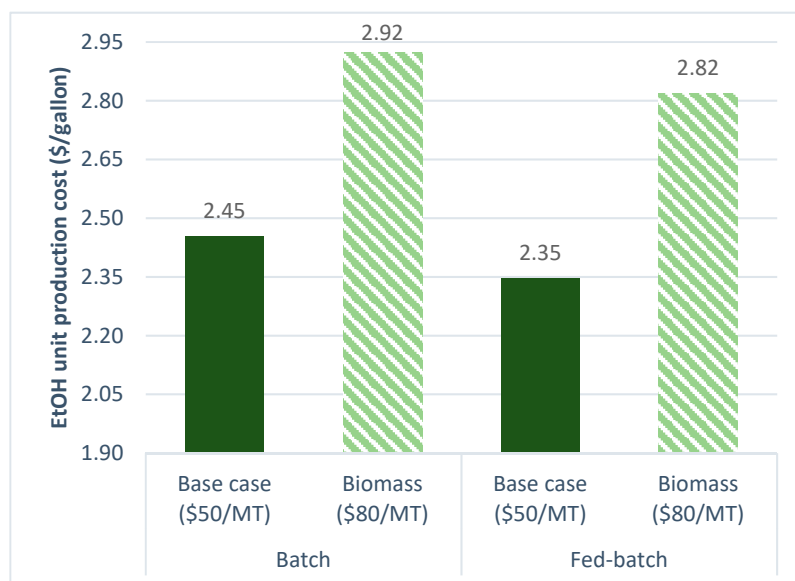


**Figure 4.** Comparison of ethanol production cost for the SuperPro Designer base case and adjusted plant capacity.

### 3.2.3. Biomass Cost

Biomass is a large contributor to raw material costs in cellulosic ethanol production. Other studies show raw materials comprise approximately 30% or more of total operating costs [12,15,17,19,20,23–25]. Results from this simulation indicate biomass comprised over 50% of annual operation costs. This study investigated how changing biomass costs affected the ethanol production cost. In the SPD base cases, the stover biomass was assumed to cost \$50 per metric ton. To test the sensitivity of the ethanol production cost to the cost of the biomass, the price was increased to \$80 per metric ton. For the batch process, the ethanol production cost increased by \$0.47/gal. For the fed-batch process, the ethanol production cost again increased by \$0.47/gal. Figure 5 illustrates the change in ethanol production cost as a result of changing the biomass cost.

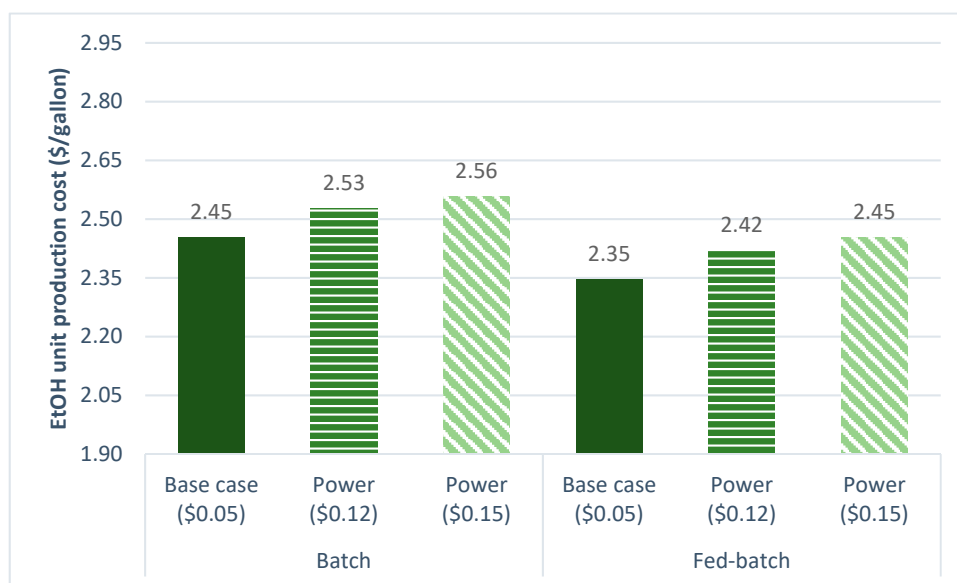




**Figure 5.** Comparison of ethanol production cost for the SuperPro Designer base case and adjusted biomass cost.

### 3.2.4. Power Cost

Power costs vary widely by location, and they are subject to change. After noticing the high energy requirements of cellulosic ethanol production, this study sought to observe the effect of power cost on ethanol production cost. The SPD base case assumed a power cost of \$0.05, which was close to the NREL base case assumption of \$0.06. According to the Nebraska Energy Office, Ravenna, NE is in the Dawson Public Power district (DPPD). The data on the Nebraska Energy Office website showed the average industrial energy cost for DPPD was \$0.12/kW-h. The power cost was increased to \$0.12 to match this data [21]. The increase in power cost caused the ethanol production cost in the batch process to increase by \$0.08/gal. The ethanol production cost in the fed-batch process increased \$0.07/gal. Figure 6 illustrates the change in ethanol production cost as a result of changing the power cost.

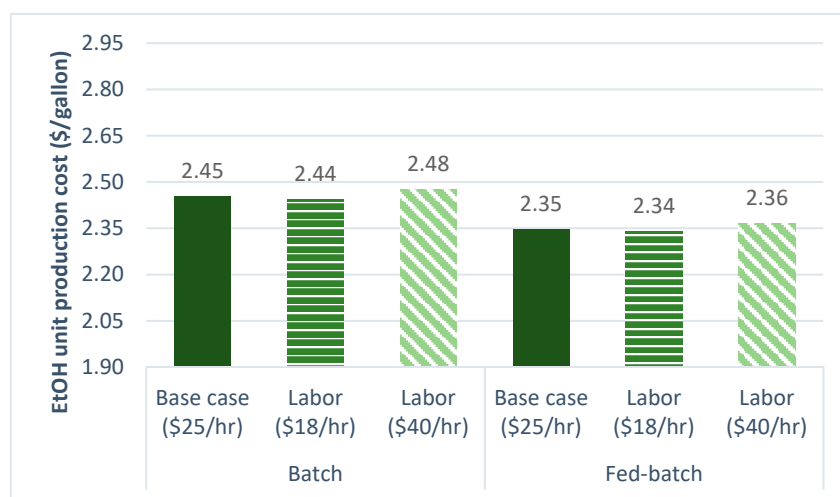


**Figure 6.** Comparison of ethanol production cost for the SuperPro Designer base case and adjusted power costs.



### 3.2.5. Labor Cost

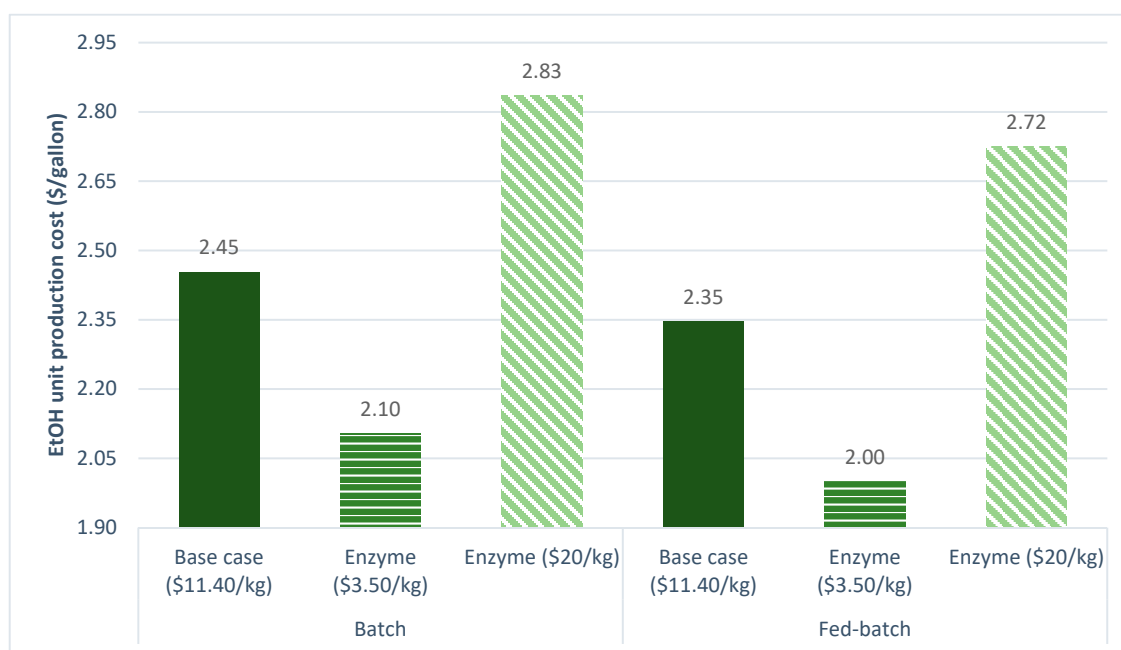
The sensitivity of ethanol production cost to labor cost was tested after noticing the high labor requirements for cellulosic ethanol production and how it changed greatly by using either batch or fed-batch enzymatic hydrolysis. Data from the US Bureau of Labor and statistics were used to realistically adjust salary costs for operators that would be working in the plant. The data showed that a \$40 hourly wage represented the 90th percentile of all chemical plant operators, so this was chosen as the high estimate, and an \$18 hourly wage was chosen as the low estimate for the sake of observing sensitivity. When the hourly wage for operators increased, the ethanol production cost for the batch process design increased \$0.08/gal. The ethanol production cost for the fed-batch process design increased only \$0.01/gal. When the hourly wage for operators decreased, the ethanol production cost for both the batch process design and the fed-batch process design decreased by \$0.01/gal. Note that this simulation maintains the assumption of the same hourly wage for all workers in the plant, even though this would not be the case. In reality, supervisors and managers would likely receive different wages than operators. Figure 7 illustrates the change in ethanol production cost as a result of changing labor costs.



**Figure 7.** Comparison of ethanol production cost for SuperPro Designer base case and adjusted labor.

### 3.2.6. Enzyme Cost

Enzyme costs are difficult to estimate. Often the data needed to calculate enzyme cost and cost contributions are difficult to find and/or determine due to confidentiality surrounding commercial enzyme production. Cost of the enzymes are subject to change, so the effect of enzyme price on the ethanol production cost was tested (Figure 8). The SPD base case assumed an enzyme cost of \$11.40/kg protein, which translated to a cost of \$0.50 per gallon of ethanol produced. The NREL base case [15,16] suggested an enzyme cost of \$0.17 per gallon of ethanol produced would be more ideal. To lower the cost of enzyme per gallon to \$0.17, the cost of enzyme per kilogram of protein was estimated to decrease to about \$3.50, so a simulation was run with this lower cost. Because of the difficulties estimating enzyme cost, this study also tested the effect of nearly doubling the cost of enzyme by running a simulation with an enzyme cost of \$20/kg protein. At the lower enzyme cost, the ethanol production cost for both the batch and fed-batch process designs decreased by \$0.35/gal. At the higher enzyme cost, the ethanol production cost increased by \$0.38/gal and by \$0.37/gal for the batch and fed-batch process designs, respectively.



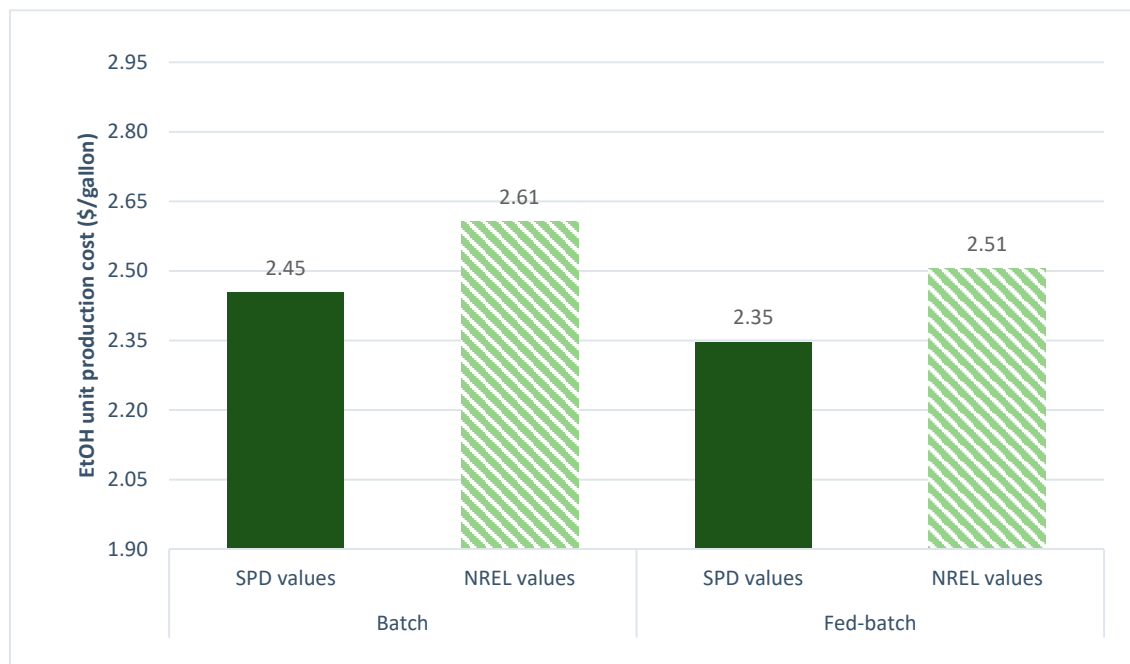
**Figure 8.** Comparison of ethanol production cost for the SuperPro Designer base case and adjusted enzyme costs.

### 3.2.7. SPD Parameters vs. NREL Parameters

The base case parameters within SPD are similar to the base case parameters in NREL studies [15,16]. To observe how the slight differences affected ethanol production cost, the costs of labor, power, enzyme, corn stover, and the annual operating hours were changed to better reflect the NREL studies. This sensitivity test has important implications for broader literature reviews of techno-economic studies. Different studies will use different input values to calculate their results, which poses challenges for making direct comparisons of the results across studies. The SPD base case values were compared with the NREL base case values in an effort to see how the different input parameters behave in exactly the same process flow sheet. This was done in hope of better understanding the implications of different input parameters when trying to compare separate studies. Table 4 shows the comparison between SPD base case values and NREL base case values. The NREL parameters resulted in a higher ethanol production cost for the batch and fed-batch processes, as seen in Figure 9. Of all the parameters changed, the feedstock cost had the greatest effect and the NREL base case values had a higher cost for the corn stover feedstock (see Table 4). This resulted in a higher cost for raw materials, and therefore, a higher production price for ethanol.

**Table 4.** Comparison of parameter values for SuperPro Designer base case and National Renewable Energy Laboratory (NREL) base case. Values for the batch enzymatic hydrolysis process are given first. Values for the fed-batch enzymatic hydrolysis process are given in parenthesis.

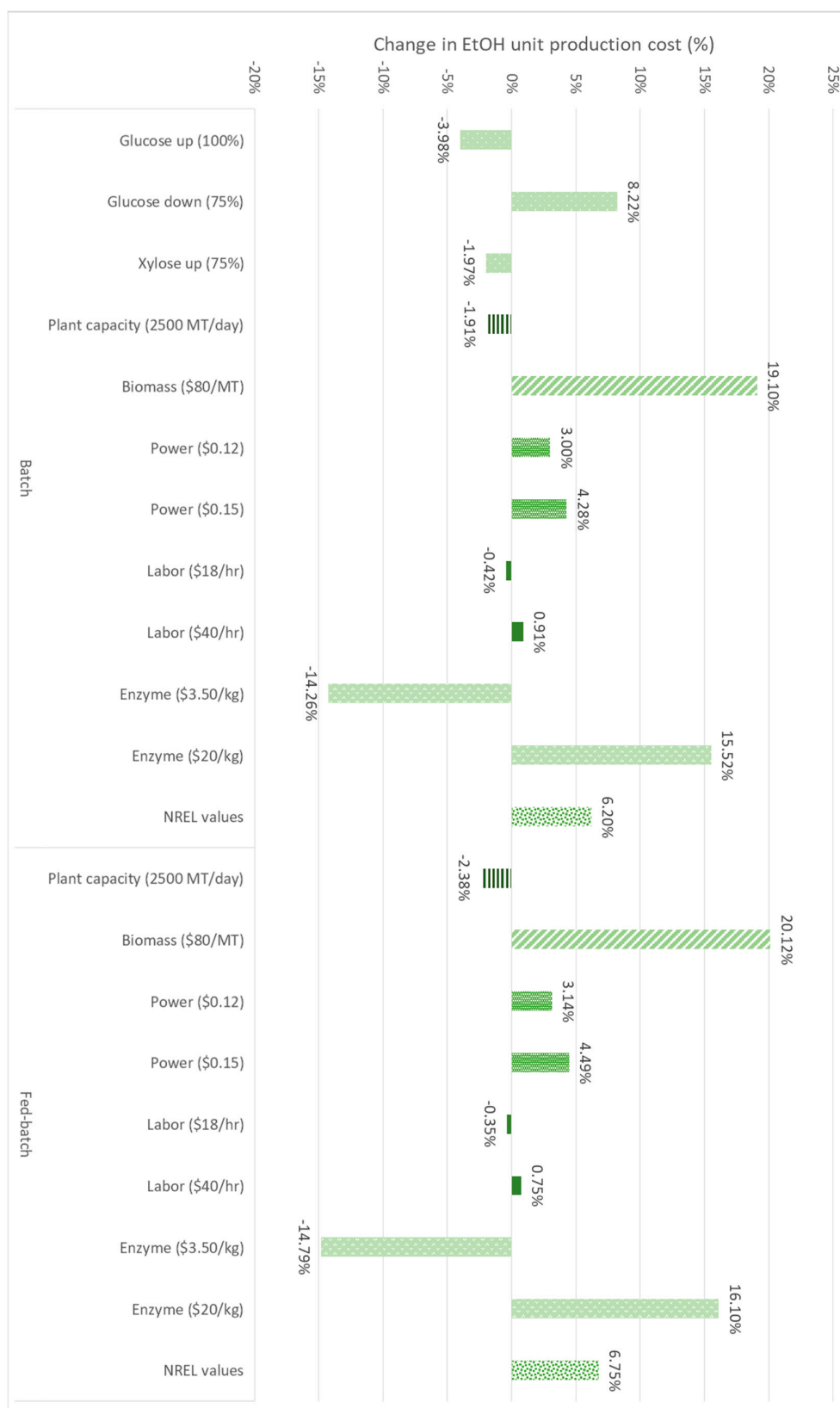
	SuperPro Designer Base Values	NREL Base Values
Labor (\$/h)	25	23
Standard power (\$/kW-h)	0.05	0.06
Enzyme cost (\$/kg)	11.40	4.00
Corn stover cost (\$/metric ton)	50	80
Annual operating hours (h/year)	7920	7880
Ethanol production cost (\$/gallon)	2.4537 (2.3469)	2.6059 (2.5053)



**Figure 9.** Comparison of ethanol production cost for SuperPro Designer and National Renewable Energy Laboratory base case values.

### 3.2.8. Sensitivity Analysis Results

Throughout the sensitivity analysis, the response to the change in variables had about the same magnitude for both batch and fed-batch operations. The percent change in ethanol production cost from the base case to the adjusted parameter scenarios was roughly equivalent between the batch and the fed-batch processes (Figure 10). Since the changes were so uniform, the researchers conclude that SPD evaluates the batch and fed-batch processes in a similar manner. Figure 10 shows that the cost of biomass and cost of enzymes have the greatest impact on the unit production cost of ethanol.



**Figure 10.** Comprehensive figure showing the sensitivity of ethanol unit production cost to each variable tested in the study. The figure shows the percent change in unit production cost for each scenario compared to the SuperPro Designer base case value.

#### 4. Conclusions

This study examined the techno-economic differences between cellulosic ethanol production using batch enzymatic hydrolysis and production using fed-batch enzymatic hydrolysis. The sensitivity analysis found that cost of biomass had the greatest effect on ethanol production cost, which caused a 20% increase in ethanol production costs. Enzyme cost had the second greatest effect, decreasing ethanol production costs by 15% when cost of enzyme decreased by 70%. Ethanol production cost increased 16% when cost of enzyme increased by 75%.

The results support the proposition that fed-batch enzymatic hydrolysis does improve the techno-economics of cellulosic ethanol production, even if not in the ways expected. In the future, the researchers hope to implement a custom feeding profile for the simulation of fed-batch enzymatic hydrolysis operation, based on a feeding profile previously developed by our lab group. The feeding profile for a fed-batch operation in SPD is not customizable, so it was not possible to implement the feeding profile in this study. An optimized feeding profile could further improve enzymatic hydrolysis and the techno-economics of cellulosic ethanol production overall.

**Author Contributions:** Methodology, E.A. and D.R.K.; Formal Analysis, E.A. and D.R.K.; Writing-Original Draft Preparation, E.A.; Writing-Review & Editing, D.R.K.

**Funding:** This project is based on research that was partially supported by the Nebraska Agricultural Experiment Station with funding from the Hatch Multistate Research capacity funding program (Accession Number 1011053) from the USDA National Institute of Food and Agriculture.

**Acknowledgments:** The authors acknowledge Demetri Petrides (Intelligen, Inc.) for his valuable insight into SuperPro Designer.

**Conflicts of Interest:** The authors declare no conflicts of interest.

#### Appendix A

The process flow sheet in Figure A1 is captured from the SuperPro Designer software. It shows the design of the ethanol plant used for this simulation study. This particular process flowsheet came from the fed-batch plant, though it looks identical to the batch process flowsheet. Hydrolysis reactions occur in the reactor labeled P-30/R-102 Vessel Procedure. To differentiate the batch from the fed-batch reaction, the operation data are changed. In the case of SuperPro Designer, this is as simple as checking the box marked, “Consider fed-batch supply of reactants” and modifying the input reaction parameters appropriately.

## Corn Stover to Ethanol

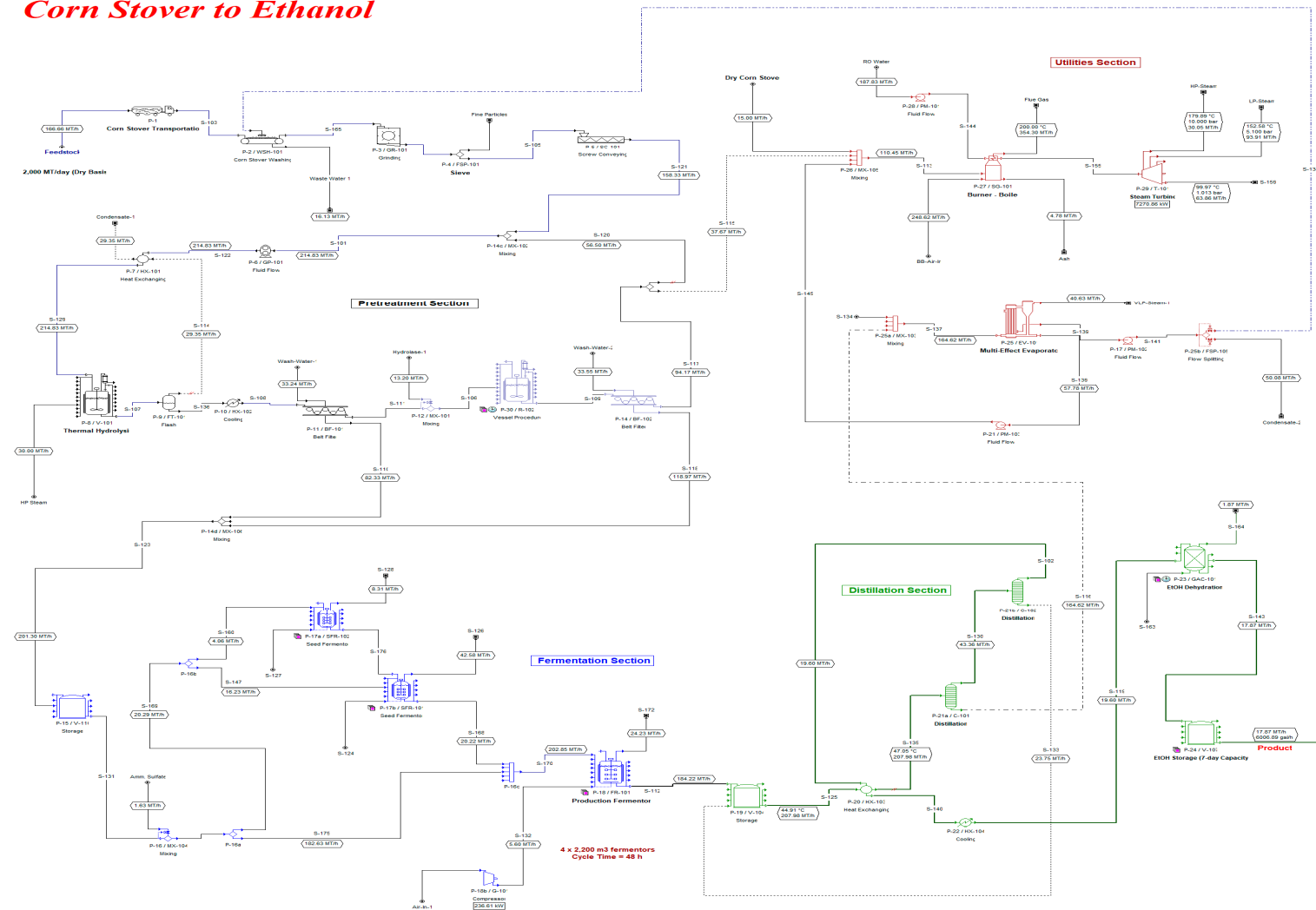


Figure A1. A sample process flowsheet from SuperPro Designer as was used in this study.

## References

- Hill, S.; Hanson, S. Rising Corn Prices and Oversupply Push Ethanol Operating Margins to Multiyear Lows—Today in Energy—U.S. Energy Information Administration (EIA). Available online: <https://www.eia.gov/todayinenergy/detail.php?id=40813> (accessed on 3 October 2019).
- Hanson, S.; Hill, S. EPA Finalizes Renewable Fuel Standard for 2019, Reflecting Cellulosic Biofuel Shortfalls—Today in Energy—U.S. Energy Information Administration (EIA). Available online: <https://www.eia.gov/todayinenergy/detail.php?id=37712> (accessed on 3 October 2019).
- US EPA. Cellulosic Waiver Credits under the Renewable Fuel Standard Program. Available online: <https://www.epa.gov/renewable-fuel-standard-program/cellulosic-waiver-credits-under-renewable-fuel-standard-program> (accessed on 4 October 2019).
- Ballesteros, M.; Oliva, J.M.; Manzanares, P.; Negro, M.J.; Ballesteros, I. Ethanol Production from Paper Material Using a Simultaneous Saccharification and Fermentation System in a Fed-Batch Basis. *World J. Microbiol. Biotechnol.* **2002**, *18*, 559–561. [[CrossRef](#)]
- Hodge, D.B.; Karim, M.N.; Schell, D.J.; McMillan, J.D. Model-Based Fed-Batch for High-Solids Enzymatic Cellulose Hydrolysis. *Appl. Biochem. Biotechnol.* **2008**, *152*, 88. [[CrossRef](#)] [[PubMed](#)]
- Rudolf, A.; Alkasrawi, M.; Zacchi, G.; Lidén, G. A Comparison between Batch and Fed-Batch Simultaneous Saccharification and Fermentation of Steam Pretreated Spruce. *Enzyme Microb. Technol.* **2005**, *37*, 195–204. [[CrossRef](#)]
- Tai, C.; Arellano, M.G.; Keshwani, D.R. Epidemic Based Modeling of Enzymatic Hydrolysis of Lignocellulosic Biomass. *Biotechnol. Prog.* **2014**, *30*, 1021–1028. [[CrossRef](#)] [[PubMed](#)]
- Bansal, P.; Hall, M.; Realff, M.J.; Lee, J.H.; Bommarius, A.S. Modeling Cellulase Kinetics on Lignocellulosic Substrates. *Biotechnol. Adv.* **2009**, *27*, 833–848. [[CrossRef](#)] [[PubMed](#)]
- Gan, Q.; Allen, S.J.; Taylor, G. Kinetic Dynamics in Heterogeneous Enzymatic Hydrolysis of Cellulose: An Overview, an Experimental Study and Mathematical Modelling. *Process Biochem.* **2003**, *38*, 1003–1018. [[CrossRef](#)]
- Laureano-Perez, L.; Teymouri, F.; Alizadeh, H.; Dale, B.E. Understanding Factors That Limit Enzymatic Hydrolysis of Biomass. *Appl. Biochem. Biotechnol.* **2005**, *124*, 1081–1099. [[CrossRef](#)]
- Zhang, Y.-H.P.; Lynd, L.R. Toward an Aggregated Understanding of Enzymatic Hydrolysis of Cellulose: Noncomplexed Cellulase Systems. *Biotechnol. Bioeng.* **2004**, *88*, 797–824. [[CrossRef](#)] [[PubMed](#)]
- Kazi, F.K.; Fortman, J.A.; Anex, R.P.; Hsu, D.D.; Aden, A.; Dutta, A.; Kothandaraman, G. Techno-Economic Comparison of Process Technologies for Biochemical Ethanol Production from Corn Stover. *Fuel* **2010**, *89*, 20–28. [[CrossRef](#)]
- Intelligen, Inc. *SuperPro Designer*; Intelligen Suite; Intelligen, Inc.: Scotch Plains, NJ, USA, 2016.
- Intelligen, Inc. *SuperPro Designer User Guide*; Intelligen Suite, Inc.: Scotch Plains, NJ, USA, 1991.
- Aden, A.; Ruth, M.; Ibsen, K.; Jechura, J.; Neeves, K.; Sheehan, J.; Wallace, B.; Montague, L.; Slayton, A.; Lukas, J. *Process Design Report for Stover Feedstock: Lignocellulosic Biomass to Ethanol Process Design and Economics Utilizing Co-Current Dilute Acid Prehydrolysis and Enzymatic Hydrolysis for Corn Stover*; NREL/TP-510-32438; National Renewable Energy Lab. (NREL): Golden, CO, USA, 2002. [[CrossRef](#)]
- Davis, R.; Tao, L.; Scarlata, C.; Tan, E.C.D.; Ross, J.; Lukas, J.; Sexton, D. *Process Design and Economics for the Conversion of Lignocellulosic Biomass to Hydrocarbons: Dilute-Acid and Enzymatic Deconstruction of Biomass to Sugars and Catalytic Conversion of Sugars to Hydrocarbons*; NREL/TP-5100-62498; National Renewable Energy Lab. (NREL): Golden, CO, USA, 2015. [[CrossRef](#)]
- Aden, A.; Foust, T. Technoeconomic Analysis of the Dilute Sulfuric Acid and Enzymatic Hydrolysis Process for the Conversion of Corn Stover to Ethanol. *Cellulose* **2009**, *16*, 535–545. [[CrossRef](#)]
- Anex, R.P.; Aden, A.; Kazi, F.K.; Fortman, J.; Swanson, R.M.; Wright, M.M.; Satrio, J.A.; Brown, R.C.; Daugaard, D.E.; Platon, A.; et al. Techno-Economic Comparison of Biomass-to-Transportation Fuels via Pyrolysis, Gasification, and Biochemical Pathways. *Fuel* **2010**, *89*, 29–35. [[CrossRef](#)]
- Eggeman, T.; Elander, R.T. Process and Economic Analysis of Pretreatment Technologies. *Bioresour. Technol.* **2005**, *96*, 2019–2025. [[CrossRef](#)] [[PubMed](#)]
- Klein-Marcuschamer, D.; Oleskowicz-Popiel, P.; Simmons, B.A.; Blanch, H.W. Technoeconomic Analysis of Biofuels: A Wiki-Based Platform for Lignocellulosic Biorefineries. *Biomass Bioenergy* **2010**, *34*, 1914–1921. [[CrossRef](#)]



21. 2015 Utility Bundled Retail Sales-Industrial. Available online: [http://www.neo.ne.gov/statshtml/114\\_Industrial\\_Sector.pdf](http://www.neo.ne.gov/statshtml/114_Industrial_Sector.pdf) (accessed on 9 September 2017).
22. Chemical Equipment Operators and Tenders. Available online: <https://www.bls.gov/oes/current/oes519011.htm> (accessed on 23 September 2017).
23. Juneja, A.; Kumar, D.; Murthy, G.S. Economic Feasibility and Environmental Life Cycle Assessment of Ethanol Production from Lignocellulosic Feedstock in Pacific Northwest U.S. *J. Renew. Sustain. Energy* **2013**, *5*, 023142. [[CrossRef](#)]
24. Kumar, D.; Murthy, G.S. Impact of Pretreatment and Downstream Processing Technologies on Economics and Energy in Cellulosic Ethanol Production. *Biotechnol. Biofuels* **2011**, *4*, 27. [[CrossRef](#)] [[PubMed](#)]
25. Tao, L.; Aden, A.; Elander, R.T.; Pallapolu, V.R.; Lee, Y.Y.; Garlock, R.J.; Balan, V.; Dale, B.E.; Kim, Y.; Mosier, N.S.; et al. Process and Technoeconomic Analysis of Leading Pretreatment Technologies for Lignocellulosic Ethanol Production Using Switchgrass. *Bioresour. Technol.* **2011**, *102*, 11105–11114. [[CrossRef](#)] [[PubMed](#)]



© 2019 by the authors. Licensee MDPI, Basel, Switzerland. This article is an open access article distributed under the terms and conditions of the Creative Commons Attribution (CC BY) license (<http://creativecommons.org/licenses/by/4.0/>).