



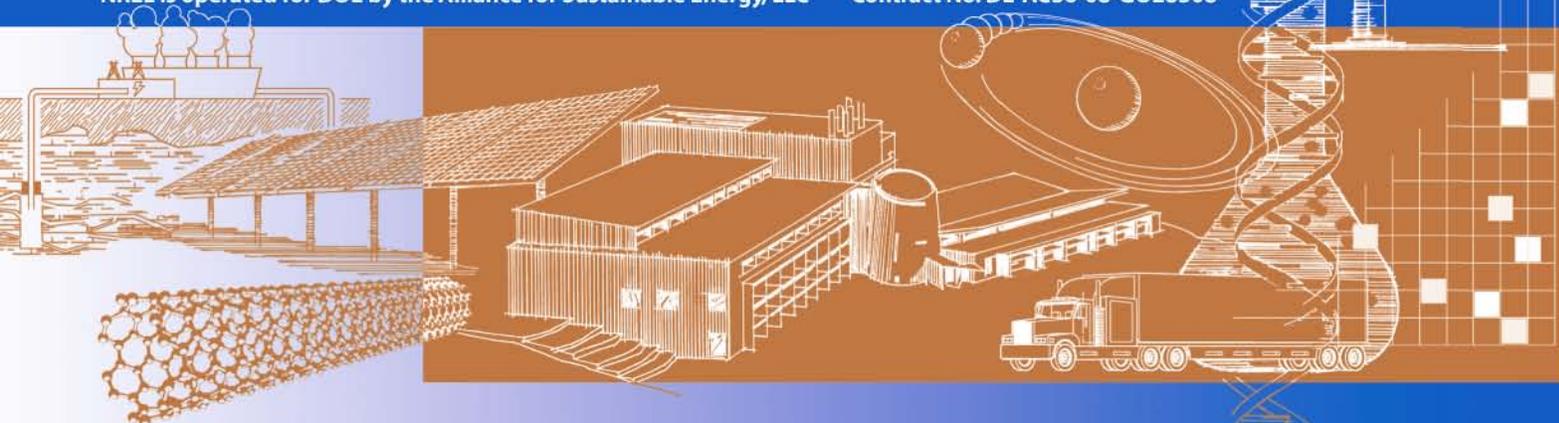
# Biochemical Production of Ethanol from Corn Stover: 2008 State of Technology Model

D. Humbird and A. Aden

*Technical Report*  
NREL/TP-510-46214  
August 2009

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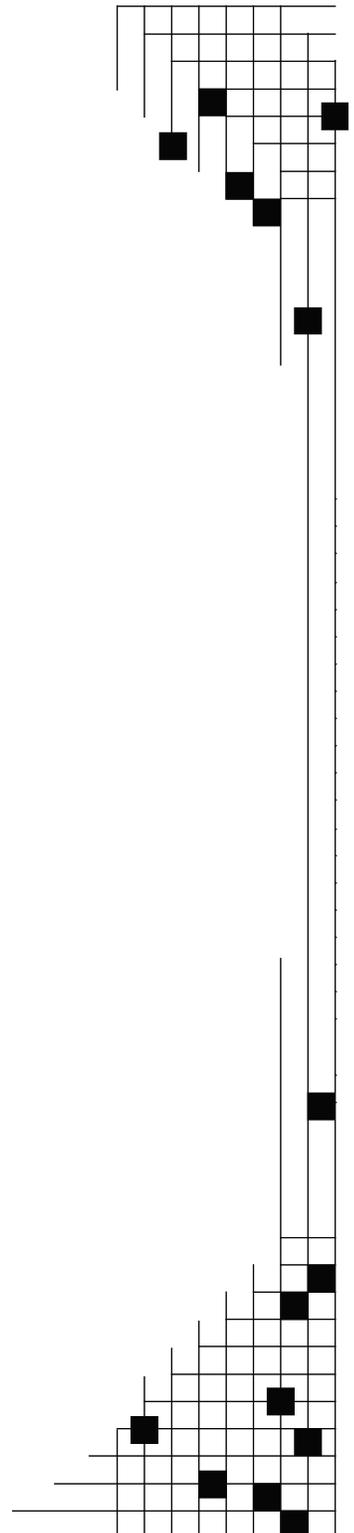
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# 1 Introduction

In 2002, the National Renewable Energy Laboratory (NREL) published a detailed report documenting a process design and economic analysis for the biochemical conversion of lignocellulosic biomass (corn stover) to ethanol [1]. The overarching process design chosen for the report was dilute acid pretreatment of corn stover followed by enzymatic saccharification and pentose/hexose cofermentation using recombinant *Zymomonas mobilis*. Aspen Plus chemical process simulation software [2] was used to compute material and energy balances for a biorefinery processing 2,000 dry metric tons of corn stover per day. Equipment and operating costs were computed from simulation results in a spreadsheet framework assuming “n<sup>th</sup>-plant” economics. Using a discounted cash flow rate of return analysis, the minimum ethanol selling price (MESP) required to meet a 10% internal rate of return was obtained.

The primary function of the design report was to determine the chemical and biochemical conversion targets required to attain the ethanol cost goals put forth by the U.S. Department of Energy (DOE) in terms of MESP. In the 2002 design report, the DOE goal for MESP in 2012 was \$1.07/gallon ethanol in 2000 dollars. When inflation was factored into capital, chemical, feedstock collection, and labor costs, the updated MESP target became \$1.49/gal ethanol in 2007 dollars.

The annual State of Technology (SOT) model is complementary to the design report. With minimal change to the unit operations in the Aspen Plus model, the projected conversions from the design report are replaced by present conversions that have been experimentally verified. All data in this report have been demonstrated experimentally, with a preference for data obtained in NREL’s ~1 ton/day pilot facility. Not all conversion parameters can be practically obtained in the pilot plant, so the SOT contains both laboratory-scale data and pilot-scale data. The SOT is arguably a misnomer since no commercial cellulosic ethanol plants exist today. The aim of the SOT analysis is not to calculate the current commercial cost of cellulosic ethanol. Instead, it is meant to reflect NREL’s best estimate of ethanol production costs in a hypothetical “n<sup>th</sup> plant” using our current best slate of demonstrated technical capabilities. By comparing the SOT year-over-year, research advances can be quantified in terms of economic improvements in the overall process. As such, progress toward the 2012 goal of \$1.49/gal can be tracked.

## 2 State of Technology for FY 2008

During Fiscal Year (FY) 2008, several key biochemical platform accomplishments were achieved and documented, spanning all major areas of research including feedstock, pretreatment, conditioning, enzymatic hydrolysis, and fermentation. Table 1 presents a sample of these accomplishments and their respective importance toward meeting programmatic goals. While significant analytical experience was gained on switchgrass, corn stover remained the targeted feedstock for this SOT update. Note that several of the highlighted accomplishments, particularly in the area of enzymatic saccharification, were new findings or new tools. For example, NREL has successfully developed a quantum mechanical model that describes how CBH1 molecules in the enzyme cocktail proceed down the length of a cellulose chain. Subsequent research plans for FY 2009 will use this

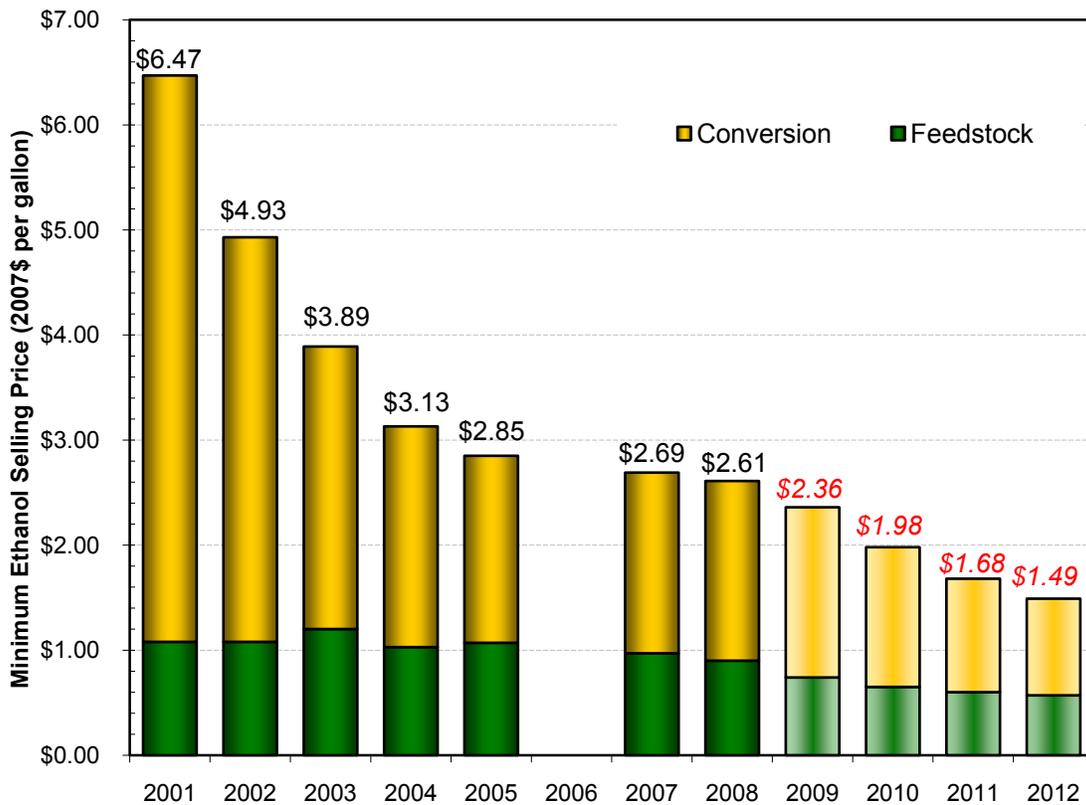
knowledge to improve the specific activity of enzyme systems, which in turn should reduce their overall cost to the process.

**Table 1. Key FY08 Accomplishments in Biochemical Ethanol Research at NREL**

<b>Process Area</b>	<b>Accomplishment</b>	<b>Importance/Significance</b>
<b>Feedstock</b>	Developed and/or adapted dry corn stover analytical methods for switchgrass	Enables the addition of energy crops to the platform research activities
	Assessed processability of ensiled corn stover	Gives picture of how wet stored residues will convert
<b>Pretreatment</b>	Achieved target of 75% xylan conversion to xylose at continuous reactor scale at 30% solids loading	Critical step towards 2012 pilot scale run and cost competitive ethanol
<b>Conditioning</b>	Developed more detailed understanding of hydrolysate toxicity and its impact on fermentation	Overcoming this technical barrier is a key aspect of obtaining high yields and successful 2012 demonstration
	Confirmed ammonium hydroxide as preferred conditioning option vs. overliming	Reduces sugar losses and potential for fermentative enhancement, also begins determining process for 2012
<b>Saccharification</b>	Series of Joule milestones completed on accessory enzymes	First step towards understanding the potential of accessory and hemicellulase enzymes to improve overall yields
	First-of-a-kind quantum mechanical model of CBH1, from CBM linker to catalytic domain	Tells us exactly how this key enzyme functions, can be used to develop targeted enzyme improvements that can be used to enhance the enzyme performance
	New tools: protein crystallography	can now develop actual 3D structures of enzymes to better understand the structure/function relationship with activity
	Testing of current industrial enzymes	Important to understand how they perform under process relevant conditions (high solids)
<b>Fermentation</b>	Successful benchmarking of all ethanologen projects	Establishes baseline performance for current industry leading organisms, provides pathway for obtaining organism for use during 2012 demo

The key updates to the FY 2008 SOT model were in the pretreatment and fermentation areas and in feedstock collection. Advancements in NREL's Pretreatment and Enzymatic Hydrolysis Subtask [3] showed that certain process changes were necessary to meet target conversions of xylan to monomeric xylose. In the fermentation area, some positive results regarding fermentation of xylose were included in the model. In the feedstock collection area, new corn stover cost data provided by Idaho National Laboratory and Oak Ridge National Laboratory were incorporated into the overall economics. Figure 1

shows the MESP computed for the years 2001-2008 (with the exception of 2006, when this exercise was not commissioned) and the research targets to 2012. For consistency, values in Figure 1 before 2008 were updated to use 2007 dollars and the 2007 feedstock cost of \$69.60/ton. (The numbers may therefore appear different from previous versions of this figure in older publications.) The SOT MESP for FY 2008 was slightly down compared with FY 2007 at \$2.61/gal due to a reduction in the feedstock cost assumption. Conversion costs were flat compared with FY 2007 at \$1.71/gal; however, we note that some of the key conversion targets were demonstrated at pilot scale in a continuous fashion for the first time in FY 2008. In other words, the FY 2008 number is ‘harder’ than the FY 2007 number, which was based on laboratory-scale conversion data and therefore contained a higher level of uncertainty. It is believed that the technical targets needed to reach \$1.49/gal by 2012 are still achievable and that the learning from scale-up efforts in 2008 will translate directly into economic improvements for 2009 through 2012.



**Figure 1. SOT MESP for 2001-2008, corrected for 2007 dollars and feedstock price. MESP beyond 2008 are forecast based on research targets.**

## 2.1 Feedstock

Feedstock, be it corn stover or other biomass, is a significant cost contributor to MESP. The cost basis used in previous SOT assessments for washed and milled corn stover delivered to the throat of the pretreatment reactor was \$60/dry ton through 2008 and

\$46/dry ton in the years 2009–2012. Idaho National Laboratory (INL) and Oak Ridge National Laboratory (ORNL) performed DOE research in 2008 regarding feedstock harvest and logistics and grower payments [3,4]. Their combined numbers were significantly higher than the cost assumptions used in previous SOTs, with feedstock cost estimated at \$69.60/dry ton in 2007 and decreasing modestly each year to reach \$50.90/dry ton in 2012. This cost increase over the previous assumption was incorporated into the 2008 SOT (and previous years, as noted above) and resulted in a higher MESP.

## 2.2 Pretreatment Area

In the pretreatment area, significant improvements in the conversion of xylan to xylose were demonstrated at the bench scale in FY 2007. For example, 75% conversion of xylan to xylose was demonstrated in high-solids experiments performed in bench-scale batch pretreatment reactors. The goal for FY 2008 was to demonstrate equivalent conversion to xylose (75%) and reduced loss to degradation products (8%) at the pilot scale in continuous operation. In other words, the FY 2008 targets for this area were intended to enhance process scale-up experience and not to provide a significant reduction in MESP.

The FY 2008 targets for the pretreatment area were:

- 75% conversion of xylan to monomeric xylose
- Loss of xylan to degradation products (mostly furfural) of less than 8%
- >30% solids loading
- Continuous operation at pilot-scale volume

Continuous pretreatment is more challenging than batch pretreatment because process conditions such as heat and mass transfer and residence time cannot be as tightly controlled. In their FY 2008 D Milestone [5], the Pretreatment and Enzymatic Hydrolysis Subtask reported that pretreatment runs carried out in the horizontal Sunds reactor [6] (modified for high solids operation) indicated a limit of conversion of xylan to monomeric xylose of ~60% in that reactor. Xylan conversion to the combined products of monomeric and oligomeric xylose was around 75%. Loss of xylan to furfural was 11%, and the balance (~14%) of xylan was unconverted. The simultaneous presence of high furfural and high unconverted xylan led to the hypothesis that poor residence time distribution control in the reactor was causing significant fractions of the feedstock to be either “undercooked” (unconverted) or “overcooked” (degraded to furfural) as it traveled through the reactor.

Since the targets for monomeric xylose yield and loss to furfural could not be met using the horizontal reactor in its current configuration, a subsequent secondary mild hydrolysis step was added. This “oligomer hold” step converted a significant fraction of the oligomeric xylose to monomeric xylose without generating additional degradation products. The oligomer hold step is essentially a hold of the first reactor effluent in an agitated tank at 130°C for 20 minutes. With this step, the overall monomeric xylose yield met the target of 75%. For the FY 2008 update, a new unit operation was added to the SOT model to account for this step.

In the Aspen Plus model, the new oligomer hold reactor was modeled as a simple stoichiometric reactor with a defined conversion. Three reactors were used in parallel, following the description of the pretreatment train from the 2002 design report. The 190°C effluent of the pretreatment reactor was flashed in an existing blowdown tank to about 2.5 atm to bring the temperature down to 130°C for the hold. The effluent was directed to the hold reactor, where the only reaction taking place was the conversion of xylose oligomers to monomeric xylose with a fractional extent of 70%. This brought the overall xylan-to-monomeric xylose yield to 75% between the two reactors. A new blowdown tank was added after the oligomer hold reactor to drop the pressure to 1 atm before feeding to the solid-liquid separator. The oligomer hold reactor was sized and costed using the Aspen Icarus Process Evaluator software [7]. Alloy 20-clad carbon steel was chosen for the reactor material due to the potential for corrosion from sulfuric acid. With further study on the corrosiveness of the hydrolyzate at the milder temperature of 130°C, it may be possible to substitute 316 stainless steel or Teflon-lined carbon steel for a modest savings. The cost of the new blowdown tank was assumed to be identical to the existing one. The total installed equipment cost to implement the oligomer hold step was \$2.1 million, or approximately 1.2 cents/gal of ethanol.

Table 2 shows the effect of the oligomer hold step on MESP. When comparing the xylan conversion assumptions in FY 2007 with the ones made here (total conversion of xylan to xylose of 75% and a loss to degradation products of 11%), the MESP is essentially flat when the same feedstock cost is used. The additional expense of the oligomer hold equipment is offset by an increase in xylose-to-ethanol conversion demonstrated in the fermentation area (discussed in Section 2.4). This is compared with an alternative case using the typical performance of the vertical Sunds reactor, in which the initial monomeric xylose yield is higher at 70% but no oligomer hold step is used to convert the additional 6% oligomers to monomeric xylose. In this case, there would have been an increase in MESP of 4 cents/gal over FY 2007 due to the decreased xylose concentration entering fermentation, even with the additional ethanol conversion.

**Table 2. Comparison of MESP (2007\$) for Different Pretreatment Scenarios**

	<b>FY07 SOT Case (FY08 feedstock \$)</b>	<b>FY08 SOT Oligomer Hold</b>	<b>Vertical Sunds No Oligomer Hold</b>
<b>First Stage Pretreatment</b>			
Xylan to oligomer	6.0%	21.0%	6.0%
Xylan to xylose	75.0%	60.0%	70.0%
Xylan to furfural	13.3%	11.0%	11.0%
<b>Second Stage (Oligomer Hold)</b>			
Oligomer to xylose	N/A	70%	N/A
<b>Overall Xylan to Xylose</b>	<b>75%</b>	<b>75%</b>	<b>70%</b>
<b>MESP (\$/gal)</b>	<b>\$ 2.63</b>	<b>\$ 2.61</b>	<b>\$ 2.67</b>
<i>Xylose to ethanol in fermentation</i>	76%	80%	80%

At the level of xylan conversion discussed here, the oligomer hold step provides minimal savings but is presently considered a necessary addition in order to meet the targeted monomeric xylose yield at the pilot scale. Planned modifications to the pilot plant reactor in FY 2009 will tighten the residence time distribution and should improve the

conversion of xylan to monomeric xylose. However, even with improved residence time distribution, meeting the much more aggressive monomeric xylose (ultimately 90%) and xylan degradation (ultimately 5%) targets in the upcoming years seems unlikely without a strategy of targeting maximum oligomer yields (e.g., less severe pretreatment conditions) and an additional step to convert the oligomers to monomeric xylose.

It should be noted that the oligomer hold step is not the committed solution for reaching higher xylan-to-xylose conversion. Other options for converting the oligomers are being explored, e.g., targeted enzymatic conversion. These solutions will all likely contribute some additional cost to the process and will be evaluated against each other as more information about their cost and efficacy becomes known. In the meantime, the Aspen Plus model will be used to investigate different possible modes of operation between the pretreatment (first-stage) and oligomer hold (second-stage) reactors. For example, if the conditions of the first stage of pretreatment were shifted to favor convertible oligomers instead of monomers, then the first-stage reactor—currently about \$20 million or 12 cents/gal—could be designed for a lower temperature and pressure, resulting in a significant capital savings.

### **2.3 Enzymatic Saccharification Area**

For the FY 2008 SOT, the assumed cost of cellulase enzymes was updated from \$0.32/gal to \$0.35/gal of ethanol to reflect inflation, since the older value was derived in 2005 using the cost-metric calculations developed for a series of DOE enzyme research subcontracts. DOE has negotiated awards with several enzyme manufacturers and developers [8] who, in the coming years, will further improve the cellulase enzymes needed for cellulosic biofuels production.

DOE and NREL have both received verbal feedback from the industry that the equivalent purchase price of cellulase enzymes exceeds \$0.35/gal. However, more data are required before this cost can be confidently updated in the techno-economic model. Under the enzyme award projects, DOE and NREL will be auditing technical and economic progress toward meeting program goals. As the performance of these projects is validated, those data can be used to better define a more accurate state of technology cost for enzymes in the biochemical process.

During FY 2009, NREL will be adding more technical detail to the basis of these enzyme costs. Instead of a fixed \$/gal ethanol equivalent basis, the techno-economic model will begin to incorporate specifics on the price and composition of the actual enzyme product and its required feed rate. In this way, the technical progress in this area can be more accurately tracked and translated into cost reductions.

### **2.4 Fermentation Area**

The FY 2008 targets for the fermentation area were:

- Increase in xylose-to-ethanol conversion from 76% to 80%
- Increase minor sugar conversion (arabinose, galactose, mannose) to ethanol from 0% to 40%

Research in the fermentation area is currently being performed outside of NREL through an ongoing ethanologen strain solicitation project. These data are protected under a Cooperative Research and Development Agreement (CRADA) and cannot be shared here, but a preliminary review of the results for the SOT report indicated that 80% xylose-to-ethanol conversion has been adequately demonstrated. This conversion was committed to the model and was used to generate the FY 2008 SOT MESP of \$2.61/gal.

In general, the ethanologen awardee companies had not made significant progress toward demonstrating improved conversion of any of the other minor sugars by the time of the SOT assessment. Some progress was shown on fermentation of arabinose under idealized conditions, but we were reluctant to include these results in the technoeconomic model until they can be demonstrated under more process-relevant conditions. Minor sugar conversion accounted for a significant part of the planned reduction in MESP for FY 2008. If conversion of all the minor sugars at the target level of 40% were realized, the MESP would have been \$2.51/gal, including the oligomer hold equipment from the previous section. This target has been pushed to FY 2009. However, achieving the more aggressive fermentation targets in future years (e.g., 85% conversion of all sugars to ethanol on our specific pretreated material) will require a strategy that allows NREL to better evaluate the performance of the improved strains on process-relevant hydrolyzates throughout the strain development process.

### **3 Progression to \$1.49/gal MESP in 2012**

Table 3 shows the MESP for the FY 2007 and FY 2008 SOT cases along with all planned development targets leading up to 2012. The targeted case and the final SOT case are both shown for FY 2008. The MESPs were computed in individual Aspen models using the conversion targets and feedstock and enzyme costs shown for the interim years. The 2012 case shown here is a direct progression from the FY 2008 SOT process and is different from the 2012 market target process (i.e., the design report process [1]) in a few respects, but it still meets the \$1.49/gal target. The process additions for oligomer conversion discussed here are of course included. Additionally, hydrolyzate conditioning with ammonium hydroxide instead of lime has been chosen as the research path going forward. Treatment of the whole hydrolyzate slurry with ammonia is ultimately planned, which eliminates a solid-liquid separation step. These changes will be reflected in an update to the design report scheduled for FY 2009.

**Table 3. Development Targets and Associated MESP** (cases from 2007 to 2012 in 2007\$, cases after 2008 based on FY08 model)

	2007 SOT	2008 Targets	2008 SOT	2009	2010	2011	2012
<b>Minimum Ethanol Selling Price</b>	<b>\$2.69</b>	<b>\$2.50</b>	<b>\$2.61</b>	<b>\$2.36</b>	<b>\$1.98</b>	<b>\$1.68</b>	<b>\$1.49</b>
Feedstock Contribution (\$/gal)	\$0.97	\$0.86	\$0.90	\$0.74	\$0.65	\$0.60	\$0.57
Conversion Contribution (\$/gal)	\$1.72	\$1.64	\$1.71	\$1.62	\$1.33	\$1.08	\$0.92
Yield (Gallon/dry ton)	72	76	73	78	83	87	90
<b>Technical Targets</b>							
<b>Feedstock</b>							
Feedstock Cost (\$/dry ton)	\$69.60	\$65.30	\$65.30	\$57.50	\$53.70	\$52.00	\$50.90
<b>Pretreatment</b>							
Solids Loading (wt%)	30%	30%	30%	30%	30%	30%	30%
Xylan to Xylose	75% <sup>a</sup>	75% <sup>b</sup>	75% <sup>b</sup>	80%	85%	88%	90%
Xylan to Degradation Products	13% <sup>a</sup>	8% <sup>b</sup>	11% <sup>b</sup>	8%	6%	5%	5%
<b>Conditioning</b>							
Ammonia Loading (mL of 30wt% per L hydrolyzate)	50	50	50	50	50	35	25
Hydrolyzate solid-liquid separation	yes	yes	yes	yes	yes	yes	no
Xylose Sugar Loss	2%	2%	2%	2%	2%	1%	1%
Glucose Sugar Loss	1%	1%	1%	1%	1%	1%	0%
<b>Enzymes</b>							
Enzyme Contribution (\$/gal EtOH)	\$0.35	\$0.35	\$0.35	\$0.35	\$0.17	\$0.12	\$0.12
<b>Saccharification &amp; Fermentation</b>							
Total Solids Loading (wt%)	20%	20%	20%	20%	20%	20%	20%
Sugar loss to contamination	7%	7%	7%	7%	7%	5%	3%
Combined Saccharification & Fermentation Time (d)	7	7	7	7	5	3	3
Corn Steep Liquor Loading (wt%)	1%	1%	1%	1%	1%	0.6%	0.25%
Overall Cellulose to Ethanol	85%	85%	85%	85%	85%	85%	85%
Xylose to Ethanol	76%	80%	80%	80%	80%	85%	85%
Minor Sugars to Ethanol	0%	40%	0%	40%	80%	85%	85%
<b>Reference Aspen model</b>	DW0810R	DW0810Q	DW0810Z	DW0810Y	DW0810T	DW0810U	DW0810V

a = bench scale

b = pilot scale/continuous operation

Targeted Improvements from 2007 to 2008

Targeted Future Improvements Year-Over-Year

At the end of FY 2008, the projected MESP for 2012 was \$1.33/gal, as reported in an NREL internal milestone report [9], and this figure may be seen in Biomass Program goal documents that were produced around that time. The shift from \$1.33/gal to \$1.49/gal seen here was a result of using (1) the actual cost indices for 2007 instead of the estimated ones, (2) updated feedstock cost assumptions (\$50.90/ton, up from \$46/ton), and (3) an inflation correction to the enzyme cost contribution (\$0.12/gal, up from \$0.10/gal). The costs for the 2012 case are summarized in Table 4. The technical targets and conversions were the same in each case.

**Table 4. Comparison of 2012 MESP Adjusting for Year-dollars and Feedstock Cost Assumption**

Year-dollars	Feedstock \$/dry ton	Enzymes \$/gal EtOH	MESP \$/gal
2007 estimated	\$46/ton	\$0.10/gal	\$1.33/gal
2007 actual	\$46/ton	\$0.12/gal	\$1.44/gal
2007 actual	\$50.90/ton	\$0.12/gal	\$1.49/gal

## 4 Concluding Remarks

In FY 2008, the MESP was slightly down compared with FY 2007 at \$2.61/gal due to a modest reduction in feedstock cost. Conversion costs were flat from FY 2007 to FY 2008 at \$1.71/gal. Pilot-scale research in the pretreatment area ran into a limit of xylan-to-xylose conversion for the continuous horizontal Sunds reactor used. An oligomer hold

step had to be added to convert xylose oligomers into xylose monomers to reach the target conversion of 75% xylan to xylose. This was a relatively inexpensive addition to the process and was offset by an increase in conversion of xylose to ethanol (from 76% to 80%), shown by ethanologen research performed for DOE outside of NREL. The largest planned reduction in MESP for FY 2008 was 40% conversion of minor sugars to ethanol, but this target was not included in the FY 2008 SOT due to the limited results shown so far.

Development targets are in place to reach \$1.49/gal by 2012. In the pretreatment area, learning in FY 2008 has led to some pilot plant modifications designed to improve the residence time distribution in the horizontal Sunds reactor. We also have a better understanding of the importance of obtaining high oligomeric xylose yields with a subsequent step to convert oligomers to monomers. In the saccharification and fermentation areas, the minor sugar-to-ethanol conversion target of 40% has been pushed to FY 2009, and a strategy involving NREL and DOE is being pursued to allow for better evaluation of strain performance on process-relevant material going forward. Concurrently, baseline validations of the enzyme solicitation awardee companies in FY 2009 will help to verify the accuracy of the enzyme cost estimates and performance capabilities used in the techno-economic model.

## 5 References

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[4] Perlack, B. "Supply Forecast and Analysis." Presented at the 2009 DOE Biomass Program Feedstocks Platform Review. [http://www.bcsmain.com/mlists/files/biomass/obpreview2009/feedstocks/documents//Feedstock\\_Analysis\\_Perlack\\_1.6.1.3.pdf](http://www.bcsmain.com/mlists/files/biomass/obpreview2009/feedstocks/documents//Feedstock_Analysis_Perlack_1.6.1.3.pdf). April 8, 2009.

[5] Nagle, N., et al. *Validate >75% xylan yield and <8% degradation from whole corn stover using a high solids continuous pretreatment reactor*. NREL FY08 D Milestone Report, Sept. 30, 2008, Doc. #10341. <http://devafdc.nrel.gov/bcfdoc/10341.pdf>.

[6] We note that Sunds technology is currently owned by Metso Corporation. <http://www.metso.com/>.

[7] AspenTech. Aspen Icarus Process Evaluator.

<http://www.aspentech.com/products/aspens-icarus-process-evaluator.cfm>.

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<http://www.energy.gov/news/6015.htm>.

[9] Humbird, D.; Aden A. *FY08 State of Technology Update*. NREL FY 2008 Milestone Completion Report, Oct. 31, 2008, Doc. #10367.

<http://devafdc.nrel.gov/bcfcdoc/10367.pdf>.

# Appendix

## Ethanol Production Process Engineering Analysis

Corn Stover Design Report Case: FY08 State of Technology  
 Dilute Acid Prehydrolysis with Saccharification and Co-Fermentation  
 All Values in 2007\$

Minimum Ethanol Selling Price **\$2.61**

Ethanol Production (MM Gal. / Year) 56.0 Ethanol at 68°F  
 Ethanol Yield (Gal / Dry US Ton Feedstock) 72.6  
 Feedstock Cost \$/Dry US Ton \$65.3  
 Internal Rate of Return (After-Tax) 10%  
 Equity Percent of Total Investment 100%

Capital Costs		Operating Costs (cents/gal ethanol)	
Feed Handling	\$0	Feedstock	90.0
Pretreatment	\$25,300,000	Biomass to Boiler	0.0
Neutralization/Conditioning	\$11,400,000	CSL	15.6
Saccharification & Fermentation	\$22,200,000	Cellulase	35.0
On-site Enzyme Production	\$0	Sulfuric Acid	4.1
Distillation and Solids Recovery	\$27,100,000	Ammonia	20.7
Wastewater Treatment	\$5,700,000	Other Raw Materials	2.8
Storage	\$4,200,000	Waste Disposal	2.1
Boiler/Turbogenerator	\$55,000,000	Electricity	-13.2
Utilities	\$6,500,000	Fixed Costs	17.2
Total Installed Equipment Cost	\$157,300,000	Capital Depreciation	24.3
Added Costs	\$115,200,000	Average Income Tax	18.3
(% of TPI)	42%	Average Return on Investment	44.4
Total Project Investment	\$272,500,000		
Installed Equipment Cost/Annual Gallon	\$2.81	Operating Costs (\$/yr)	
Total Project Investment/Annual Gallon	\$4.86	Feedstock	\$50,400,000
Loan Rate	N/A	Biomass to Boiler	\$0
Term (years)	N/A	CSL	\$8,700,000
Capital Charge Factor	0.179	Cellulase	\$19,600,000
Denatured Fuel Prod. (MMgal / yr)	58.7	Other Raw Matl. Costs	\$15,500,000
Denatured Fuel Min. Sales Price	\$2.53	Waste Disposal	\$1,200,000
Denaturant Cost (\$/gal denaturant)	\$0.739	Electricity	-\$7,400,000
Maximum Yields (100% of Theoretical)		Fixed Costs	\$9,600,000
Ethanol Production (MM Gal/yr)	87.0	Capital Depreciation	\$13,600,000
Theoretical Yield (Gal/ton)	112.7	Average Income Tax	\$10,200,000
Current Yield (Actual/Theoretical)	64.4%	Average Return on Investment	\$24,900,000
		Excess Electricity (KWH/gal)	3.30
		Plant Electricity Use (KWH/gal)	2.09
		Plant Steam Use (kg steam/gal)	20.2
		Boiler Feed -- LHV (Btu/lb)	2,091
		Boiler Feed -- Water Fraction	0.501
		Specific Operating Conditions	
		Enzyme Loading (mg/g cellulose)	20.0
		Saccharification Time (days)	5.0
		Conversion Cellulose --> Glucose	0.9000
		Fermentation Time (days)	2.0

# REPORT DOCUMENTATION PAGE

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