Technoeconomic Analysis of a Lignocellulosic Biomass Indirect Gasification Process To Make Ethanol via Mixed Alcohols Synthesis

S. D. Phillips*

National Renewable Energy Laboratory, 1617 Cole Boulevard, Golden, Colorado 80401

A technoeconomic analysis of a 2000 tonne/day lignocellulosic biomass conversion process to make mixed alcohols via gasification and catalytic synthesis was completed. The process, modeled using ASPEN Plus process modeling software for mass and energy calculations, included all major process steps to convert biomass into liquid fuels, including gasification, gas cleanup and conditioning, synthesis conversion to mixed alcohols, and product separation. The gas cleanup area features a catalytic fluidized-bed steam reformer to convert tars and hydrocarbons into syngas. Conversions for both the reformer and the synthesis catalysts were based on research targets expected to be achieved by 2012 through ongoing research. The mass and energy calculations were used to estimate capital and operating costs that were used in a discounted cash flow rate of return analysis for the process to calculate a minimum ethanol selling price of $0.267/L ($1.01/gal) ethanol (U.S.$2005).

Introduction

The Advanced Energy Initiative (AEI) announced in 2006 has set aggressive goals for developing the breakthrough technologies needed to make lignocellulosic-based ethanol cost-competitive with corn-based ethanol by the year 2012.[1,2] One of the major technology routes to achieving the fuels production goals is via gasification of biomass followed by catalytic synthesis to liquid fuels. In previous reports by the National Renewable Energy Laboratory (NREL), the benchmark to be "cost competitive with corn-ethanol" has been defined as $0.283/L ($1.07 per gallon) ethanol production cost[3] in 2005 which translates to $0.346/L ($1.31/gal) in the year 2012 using 2007 U.S. dollars. This represents the low end of historical fuel ethanol prices[4] and is considered to be the cost of production (COP) selling price at which lignocellulosic-derived ethanol will be commercially viable.

Biomass can have a wide range of physical, fuel, and chemical properties that affect its conversion into useful forms of energy such as liquid fuels. One of the benefits of thermochemical conversion processes relative to fermentation technologies is the ability to convert a wide variety of biomass feed stocks regardless of its sugar and lignin contents. Recent studies[5] show there are significant quantities of biomass available in the United States that could be used to produce liquid fuels, including the forest resources chosen for this study. In the target year of 2012, the availability of forest resources will be sufficient to supply a 2000 tonne/day facility. This scale of biomass supply is comparable to that supplied to the pulp and paper industry for decades in the United States and is made up of residual wood after harvesting for lumber.

A process design and technoeconomic evaluation of a thermochemical ethanol process was done at NREL in late 2006.[6] This paper, based primarily on that report, addresses the conversion of woody biomass to ethanol via thermochemical (TC) pathways that are expected to be demonstrated at the pilot-scale level by 2012. It is not a "state-of-technology" design report since several key parameters used in the model are performance targets that have not been demonstrated yet.

An indirectly heated steam gasifier was selected for this study since it was the reactor type used in several earlier NREL reports that used wood as the feedstock.[7−9] A similar study using an oxygen-blown, high-pressure reactor is in progress. The general process areas include the following: feed preparation, gasification, gas cleanup and conditioning, alcohol synthesis, separations, heat, power, and cooling utilities.

The process was designed to be both energy self-sufficient and "electrical energy neutral". In lieu of burning fossil fuels or purchasing electricity from the external power grid to meet the overall process energy requirements, all energy requirements were met by burning combustible streams derived from the biomass fed to the plant. The two main sources of combustion fuels were the residual char from gasification and dirty syngas diverted immediately after the gasifier. A small fraction of the residual gases from the synthesis and separation sections were also burned to provide heat and power. This approach made the process energy self-sufficient with the penalty of lower product yields. Only enough syngas was diverted to meet the exact internal heat and power requirements, if possible. If other operating requirements prevented the elimination of "excess" electricity, it was sold as a coproduct.

Methodology

ASPEN Plus process simulation software (version 2004.1) was used to model and study several process designs to determine the best locations for returning recycle streams taken from various points in the process. Results from the process model simulations were used in an economic analysis as discussed in ref 10 to size process equipment and develop an estimate of capital and operating costs. A discounted cash flow rate of return (DCFROR) analysis was used to determine the minimum ethanol selling price (MESP) necessary to meet a small hurdle rate (IRR) of 10% for an *n*th-plant design. Model details are given in ref 6.

Capital costs were developed from a variety of sources. Unit operations that used well-known technology that can be purchased as modular packages (i.e., amine treatment, acid gas removal) used the overall cost for the package unit when available.[11−14] Costs for common equipment items (e.g., tanks, pumps) were estimated using the ASPEN Icarus Questimate

* To whom correspondence should be addressed. Tel.: (303) 384-6235. Fax: (303) 384 6363. E-mail: steven_phillips@nrel.gov.
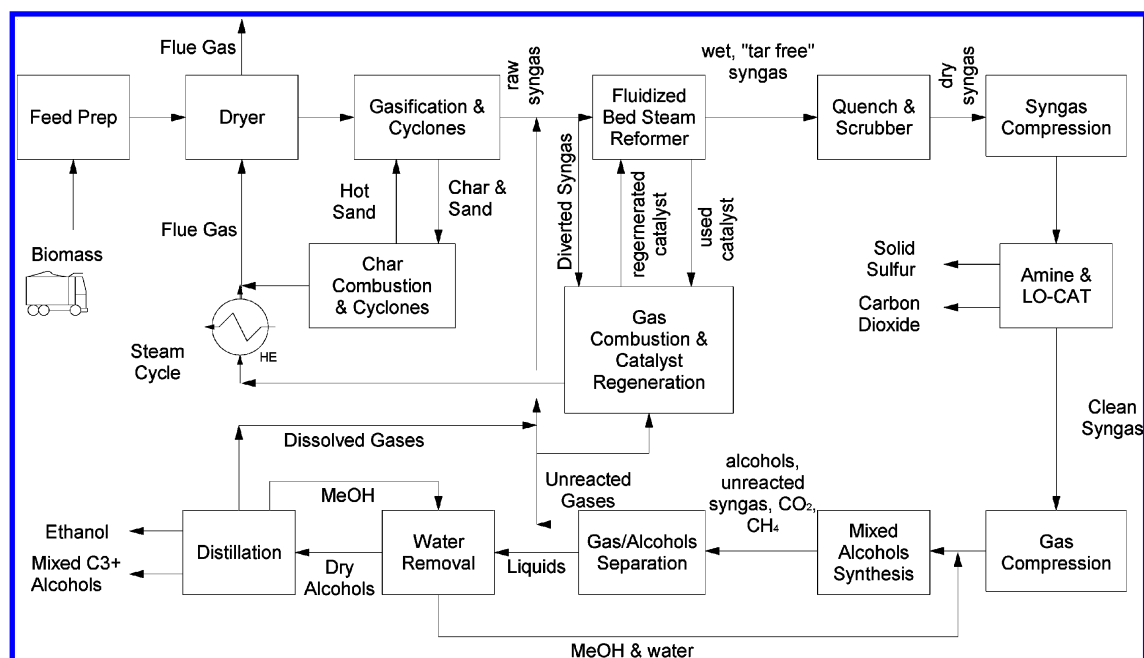
Figure 1. Block flow diagram of the biomass to ethanol process.

costing software. Other more specific unit operations (gasifier, molecular sieve, etc.) used cost estimates from other studies and/or from vendor quotes. The installed capital costs were developed using general plant-wide factors including items such as cost contributions for equipment installation, instrumentation and controls, piping, electrical systems, buildings, and yard improvements.

Overhead and contingency factors were used with the installed equipment costs to determine a total plant investment cost (TPIC). The TPIC and plant operating expenses were used in a discounted cash flow analysis to determine the cost of ethanol production with a specific discount rate, referred to here as the minimum ethanol selling price (MESP). The MESP was the primary value used to compare alternate designs during process synthesis.

Process Design Overview

The plant size for this study is 2000 dry tonne/day (2205 dry ton/day). At an expected 8406 operating hours per year, the annual feedstock requirement is 700 000 dry tonne/yr. This is a small portion of the 127 million dry tonne/yr of forest resources potentially available. The delivered feedstock cost was chosen to match recent analyses done at Idaho National Laboratory (INL) to target $38.58 per dry tonne ($35 per dry ton) by the year 2012.[15] A block flow diagram of the current design, depicted in Figure 1, has the following main process areas.

Feed Handling and Preparation. This area of the process accommodates the delivery of woody biomass, short-term on-site storage, and preparation. Hybrid poplar wood chips (field-processed through a hog mill) are delivered to the plant gate at 50 wt % moisture. Oversized wood chips, larger than 50 mm, are sent to a hammer mill for further size reduction and returned to the feed screening step.

The ultimate analysis for the feed is given in Table 1. The value for sulfur appears to be high relative to many published analyses for hybrid poplar which give values of 0.02 wt %.[16] It was kept at the previously indicated 0.09 wt % level for consistency with previous reports that used this value.[8,9]

Table 1. Ultimate Analysis of Hybrid Poplar Feed

component	wt %, dry basis
carbon	50.88
hydrogen	6.04
nitrogen	0.17
sulfur	0.09
oxygen	41.90
ash	0.92
heating value (kcal/kg):	4820 HHV

Front-end loaders transfer the wood chips from the feed sizing/reduction area to dryer feed bins where the biomass is dried from its as-received moisture level using the combined hot flue gases from the char and fuel combustors. The moisture content of the feed leaving the dryer was kept between 5 and 20 wt % corresponding to the practical limits for drying and feeding into the gasifier, respectively. The feed moisture level leaving the dryer in the process model was determined by evaluating which drying levels gave the best economic results for this process design.

The wood chips are dried in two rotary kilns, in parallel, by direct contact with hot flue gases from the char combustor and the fluidized bed steam reformer's (FBSR) fuel combustor. The humid exhaust gas exiting the dryer is sent through a cyclone and baghouse filter to remove particulates prior to being emitted to the atmosphere. The stack temperature of the flue gas is controlled to be 34 °C above its dew point temperature which varies, depending on processing conditions. The dry biomass is mechanically conveyed to the gasifier area.

Gasification. The reactor used in this analysis is a low-pressure, indirectly heated circulating fluidized bed (CFB) gasifier modeled with correlations based on run data from the Battelle Columbus Laboratory's (BCL) 9 tonne/day test facility.[17] Two parallel gasifier trains were assumed for this study. Heat for the endothermic gasification reactions is supplied by circulating hot synthetic olivine solids at a rate of 12.25 kg olivine/kg bone dry wood between the gasifier and char combustor vessels. Makeup olivine is added at a rate of 0.06% of the circulating rate to account for losses. Conveyors and hoppers are used to feed the biomass to the gasifier, and superheated steam extracted from the power cycle is injected into the gasifier at a ratio of 0.18 kg steam/kg bone dry biomass.

Table 2. Gasifier Operating Parameters, Exit Gas Composition, and Efficiency

gasifier variable	value	
temperature	889 °C	
pressure	1.6 bar	
H_2:CO molar ratio	0.60	
gasifier efficiency	76.6% HHV	76.1% LHV
gasifier outlet gas comp.	mol % (wet)	mol % (dry)
H_2	15.0	25.1
CO_2	7.4	12.4
CO	25.1	41.9
H_2O	40.2	
CH_4	9.0	15.1
C_2H_2	0.3	0.4
C_2H_4	2.5	4.1
C_2H_6	0.1	0.2
C_6H_6	0.1	0.1
tar ($C_{10}H_8$)	0.1	0.2
NH_3	0.2	0.3
H_2S	0.04	0.07

The steam is used to aid in stabilizing the entrained flow of biomass and olivine through the gasifier.

The gasifier and combustion reactors operate at a steady-state temperature that is dependent on the biomass feed type, its moisture content, feed rate, solids circulation rate, and pressure. The biomass feed mixes with the hot olivine and steam and thermally decomposes to a mixture of syngas components (CO, H_2, CO_2) in addition to methane (CH_4), tars (modeled as naphthalene), and solid "char" particles derived mainly from the less reactive "fixed" or nonvolatile carbon in the biomass. No additional oxidants are added to the gasifier which instead depends on the oxygen content of the biomass to produce the syngas. The gasifier temperature is too low for carbon and steam or water-gas-shift reactions to occur at significant rates.

Cyclones at the exit of the gasifier separate the char and circulating olivine from the syngas and direct these solids through a standpipe to the bottom of the char combustor where air at 1.2 times the stoichiometric oxygen requirement for total char combustion is blown into the bottom of the combustor. The air causes the char to burn and entrains the solids in the combustor, blowing it upward toward another pair of cyclone separators. The entrained olivine is heated to about 1000 °C from the char combustion and is separated from the hot flue gas and most of the ash at the series-connected cyclone separators. The first cyclone is designed to capture 99.99 wt % of the olivine, and the second cyclone is designed to capture 99.99 wt % of the ash and any smaller olivine particles escaping the first cyclone. Finally, an electrostatic precipitator is used to capture any residual dust particles escaping from the second cyclone. The hot olivine from the first cyclone flows by gravity back into the gasifier to complete the solids circulation loop. Ash and olivine particles captured in the second cyclone are cooled, moistened to 10 wt % water to minimize dust, and transported to a land fill facility.

A small amount of magnesium oxide is fed to the circulation loop to help prevent ash slag from forming at the high combustion temperatures. Potassium from the biomass has the potential of forming low-melting temperature slag with silica. The addition of MgO was initially added to the model when the bed material was silica sand which is more prone to forming the silicates of potassium.[8] Olivine, which is chemically different than silica sand, was selected as a sand replacement to avoid the potential agglomerations. However, the MgO addition was left in the model as a safety precaution and to avoid concerns about glass formation. The gasifier operating parameters and results are given in Table 2.

Table 3. Current and Target Design Performance for Conversions to CO and H_2 in the FBSR

compound	experimental conversion (%)	target conversion (%)
methane	20	80
ethane	90	99
ethylene	50	90
tars (C_{10+})	95	99.9
benzene	70	99
ammonia	70	90

Table 4. Fluidized Bed Steam Reformer Operating Conditions and Exit Gas Composition at Target Hydrocarbon and Tar Conversion Extents

tar reformer variable	value	
tar reformer inlet temperature	890 °C	
tar reformer outlet temperature	890 °C	
tar reformer outlet gas composition	mol % (wet)	mol % (dry)
H_2	37.3	42.9
CO_2	9.9	11.4
CO	37.5	43.1
H_2O	13.0	
CH_4	1.2	1.4
C_2H_2	0.01	0.01
C_2H_4	0.11	0.13
C_2H_6	10.8 ppmv	12.4 ppmv
C_6H_6	2.7 ppmv	3.1 ppmv
tar ($C_{10}H_8$)	0.5 ppmv	0.6 ppmv
NH_3	0.01	0.01
H_2S	0.02	0.02
N_2	0.72	0.83
H_2:CO molar ratio	1.00	

Gas Cleanup and Conditioning. This area of the process is for cleaning and conditioning the syngas prior to the alcohol synthesis step. The extent of cleanup is dictated by the requirements of the synthesis catalyst and in the case of tar removal to protect process equipment. The basic operations in this section are as follows: (1) catalytic steam reforming of tars in the syngas, (2) removal of particulates and condensation of steam by wet scrubbing, (3) compression of the syngas to 30 bar, and (4) removal of acid gases (CO_2 and H_2S) using an amine scrubber followed by a LO-CAT operation to reduce the H_2S to sulfur.

Steam reforming of the tars and hydrocarbons is envisioned to occur in an isothermal fluidized bed reactor (FBSR) with continuous regeneration and replacement of deactivated reforming catalyst to keep the overall catalytic activity sufficiently high to reform the tars and methane.[18] The catalyst will be regenerated in a second vessel using a currently undetermined method. In the ASPEN simulation, the tar reformer operates isothermally at 890 °C. An implicit assumption is that the energy needed for the endothermic reforming reactions can be transferred into the catalyst bed from the combustion of gases in the reformer combustor. Heat from the combustion gases has to be exchanged with the reformer through some currently undefined mechanism such as catalyst circulation or using tubes submerged in the catalyst bed with hot flue gas passing through the tubes. The FBSR reactions were modeled in ASPEN by specifying the conversion of each hydrocarbon and tar compound of interest. The conversion targets used for the reformer are listed in Table 3. The composition of the gas exiting the tar reformer is given in Table 4.

The hot syngas is cooled in a syngas cooler integrated with the steam cycle to superheat steam. Additional cooling is carried out via water scrubbing which also removes impurities such as particulates, residual ammonia, and any residual tars. The scrubbing system consists of a venturi scrubber and quench chamber using a closed circulation loop of scrubber water. Heat

Table 5. Acid Gas Removal Design Parameters

acid gas removal parameter	value
amine used	monoethanolamine
amine concentration, wt %	35
amine circulation rate (L/s)	116
amine temperature in absorber (°C)	43
absorber pressure (bar)	31
stripper condenser temperature (°C)	100
stripper reboiler temperature (°C)	114
stripper pressure (bar)	4.5
stripper reboiler duty (MW)	36.4
stripper condenser duty (MW)	24.2
amine cooler duty (MW)	12.1
heat duty to remove CO_2 (kcal/kg)	1480

is rejected from the quench water by heat-exchange with cooling tower water. A 8.3 L/min blowdown is sent to an off-site wastewater treatment facility. Any solids collected in the quench system are also sent off-site for treatment as a sludge stream assumed to be 50 wt % solids. The syngas exits the quench at a specified temperature of 60 °C.

After cleanup and cooling, the syngas is compressed in a five-stage centrifugal compressor (78% polytropic efficiency) with forced-air interstage cooling. The syngas exits each interstage cooler at 60 °C. The compressor exit pressure is set to 30 bar to achieve improved amine scrubber efficiency in the next process step.

The extent of acid gas removal is dependent on the specific catalysts being used in the process downstream of the reformer. Allowable H_2S concentrations can be as low as 0.1 ppb or as high as 100 ppm depending on the catalyst used for synthesis. This design is based upon a sulfided molybdenum catalyst which requires between 50 and 100 ppm H_2S in the syngas to maintain catalyst activity; the ASPEN model used 50 ppm.

Acceptable carbon dioxide concentrations are also dependent on the specific synthesis catalyst.[20] Some synthesis catalysts require low levels of CO_2, while others such as the sulfided molybdenum catalysts can tolerate relatively high CO_2 levels. Since CO_2 is a major constituent of the syngas, significant amounts of CO_2 need to be removed upstream of the synthesis reactor. Since the catalyst selected for this study is tolerant of both sulfur and CO_2, an amine scrubber using monoethanolamine (MEA) was used for achieving both the CO_2 and H_2S specifications.[12]

The acid gas scrubber was simulated using a simplified model consisting of user-defined component splits specifying the amount of CO_2 and H_2S to be removed to meet the design specifications at the synthesis reactor inlet. The amine system heating and cooling duties were calculated using information taken from section 21 of the GPSA Data Handbook.[21] Power requirements for pumping and cooling fans were also calculated using GPSA recommended values. The acid gas scrubber operating values for the base case are given in Table 5.

The acid gases removed in the amine scrubber are further processed through a liquid-phase oxidation process to capture and reduce H_2S to elemental sulfur. The LO-CAT process was selected because of its progress in minimizing catalyst degradation and its environmentally benign catalyst. The elemental sulfur produced in the oxidizer is stockpiled for eventual disposal or sale. The CO_2 effluent from the LO-CAT process has a user-specified H_2S concentration of 10 ppmv.

Alcohol Synthesis. The alcohol synthesis reactor system is the heart of the entire process where the clean, conditioned syngas is converted to alcohols. At this point in the process the syngas has been reformed, quenched, compressed to an intermediate pressure, and treated to remove acid gases to appropriate

Table 6. Published and Target Process Conditions for Mixed Alcohols Synthesis

parameter	"state of technology"	target
temperature (°C)	~300	300
pressure (bar)	100–150	69
H_2/CO ratio	1.0–1.2	1.0
CO_2 concentration (mol %)	0–7	5.0
sulfur concentration (ppmv)	50–100	50

concentrations. The first steps in this area are to compress and heat the syngas to the synthesis reactor inlet conditions of 69 bar and 300 °C. A three-stage, steam-driven compressor with forced-air cooling between the first two stages is used in the model.

The ASPEN model did not require a specific reactor type to be defined. However, for costing purposes, a multitube fixed bed reactor using a modified Fischer–Tropsch catalyst (molybdenum-disulfide based) was chosen for this process design, as discussed in ref 6. Table 6 lists several process and syngas conditioning requirements for these synthesis reactions including both typical, experimentally verified conditions as well as targeted conditions from the U.S. Department of Energy's (DOE) Office of the Biomass Program (OBP) research plan that were used in the model. Although a specific catalyst type is discussed, any catalyst achieving the same results with the same syngas conditions would be acceptable.

In the ASPEN model, syngas was mixed with water and methanol recycled from the alcohols separation area before being vaporized and fed to the reactor. As discussed in the NREL report, methanol was fed to the reactor to improve the yield of higher alcohols. This mixture is catalytically converted to methanol, ethanol, and higher linear alcohols up to pentanol, water, methane, and minor amounts of other hydrocarbon byproducts. The water-gas-shift activity of the catalyst is significant and provides more H_2 as the reaction proceeds consuming CO from the syngas and water produced in the synthesis reactions.[21,22–28] The exothermic heat of reaction is removed by generating saturated steam for the steam cycle. The reactor effluent is cooled to 93 °C by exchange with cooler process streams followed by air-cooled exchangers to a temperature of 60 °C and finally cooled to 43 °C using cooling water.

The condensed alcohols are separated from the unconverted syngas in a knock-out vessel. Although the unconverted syngas could be recycled directly to the inlet of the alcohol synthesis reactor, this was not done in this design. Instead, it was recycled to the Gas Cleanup and Conditioning section primarily as feed to the FBSR. The reason for this approach was to avoid the accumulation of inert gas species (CO_2, CH_4) in the recycle stream fed to the reactor which would increase the reactor size for the same conversion extent. The hydrocarbons produced as a byproduct of the alcohol synthesis were assumed to be converted in the reformer at the same extent as the hydrocarbons derived from gasification. Residual alcohols in the recycled gas are assumed to be completely reformed to syngas. Carbon dioxide sent to the reformer inhibits the water-gas-shift reactions in the reformer which helps conserve CO in the syngas. Carbon dioxide accumulation is avoided by its removal in the amine scrubber. A portion of the recycled stream is purged to the fuel combustor to produce heat for the tar reformer. Since the recycled gas is at high pressure in the synthesis area, it is heated and expanded through a turboexpander to generate electricity before being fed to the reformer or combustor. The liquid alcohols are sent to the alcohol separation and purification area.

The reactor was modeled as a simple conversion-specified reactor using a series of normal alcohol and hydrocarbon

Table 7. Mixed Alcohol Synthesis Reaction Results Compared to Published "State of Technology" Performance Ranges

result	"state of technology"	target results
total CO conversion (per-pass)	10–40%	60%
total alcohol selectivity	70–80%	90%
gas hourly space velocity (h^{-1})	1600–12 000	4000
catalyst alcohol productivity (g/kg-catalyst/h)	150 – 350	600

Table 8. Mixed Alcohol Product Distributions from Literature and the NREL Model

alcohol	Dow[28] (wt %)	SRI[25] (wt %)	NREL model (wt %)
methanol	30–70	30.77	5.01
ethanol	34.5	46.12	70.66
propanol	7.7	13.3	10.07
butanol	1.4	4.14	1.25
pentanol +	1.5	2.04	0.17
acetates (C1 and C2)	2.5	3.63	
others			10.98
water	2.4		1.86
total	100	100	100

production reactions with target conversions and selectivity as listed in Table 7. The propane, butane, and pentane+ reactions are set to zero because the catalyst will likely not favor these reactions. The specific conversions of each of the other reactions were set to reach catalyst performance targets.

The reaction conversions were also set to achieve a certain product distribution of alcohols. The mixed alcohol products described in the literature are often high in methanol but contain a wide distribution of several different alcohols. The product distributions described by Dow and SRI[25] are shown in Table 8 along with the relative product concentrations calculated by the model. The most significant differences between the NREL model product distribution and those shown in literature are with regards to the methanol and ethanol distributions. This is primarily due to the almost complete recycle of methanol to the synthesis reactor to increase the production of ethanol and higher alcohols. While this concept has been proposed in the literature,[27] it has never been tested in an integrated setting with significant quantities of methanol fed to the reactor. This assumption needs to be experimentally demonstrated.

Alcohol Separation. The condensed alcohols stream from the synthesis section is depressurized to separate absorbed gases that are recycled mainly to the tar reformer in the Gas Cleanup and Conditioning section. The alcohol stream is vaporized, slightly superheated, and then dehydrated in a pressure-swing absorber (PSA) using molecular sieves. The dry mixed alcohols stream leaving the dehydrator enters into the first of two distillation columns where methanol and ethanol are separated from the incoming stream with 99% of the incoming ethanol being recovered in the overhead stream along with essentially all the methanol. The bottom stream consists of 99% of the incoming propanol, 1% of the incoming ethanol, and all of the butanol and pentanol. The mixed alcohol bottoms are considered a coproduct of the plant.

The second column separates the methanol from the binary methanol/ethanol mixture. The ethanol recovery is 99 mol % of the incoming ethanol and has a maximum methanol concentration of 0.5 mol % to meet ASTM ethanol fuel specifications. The ethanol, which exits from the bottom of the column, is cooled before being sent to product storage. The methanol and a small quantity of ethanol exiting the overhead are used to flush the molecular sieve column during its regeneration step and then recycled, along with the water removed during the back flush step, to the synthesis reactor.

Table 9. Plant Electricity and Shaft Power Requirements by Area

plant section	electric power (MW)	turbine shaft power (MW)
feed handling and drying	1.0	
gasification	4.5	
tar reforming, cleanup, and conditioning	2.4	34.0
mixed alcohol synthesis	0.20	7.0
alcohol separation and purification	0.20	
steam system and power generation	0.60 (used) 11.0 (gen)	
cooling water and other utilities	0.70	
miscellaneous	1.0	
total plant power requirement	11.0	41.0

Heat, Cooling, and Power. A conventional steam cycle produces steam for injection into the gasifier and reformer reactors, steam for all the reboilers, and electricity for internal power requirements. This process design includes a steam cycle that produces steam by recovering heat from the hot process streams throughout the plant. Steam demands for the process include the gasifier, amine system reboiler, alcohol purification reboilers, and LO-CAT preheater. Of these, only steam to the gasifier and fluidized bed steam reformer is directly injected into the process; the rest of the plant heat demands are provided by indirect heat exchange of process streams with the steam and have condensate return loops. Power for internal plant loads is produced from the steam cycle using an extraction steam turbine/generator. Power is also produced from the process expander in the synthesis area. The overall energy balance is managed to generate only the electricity required by the plant by burning a portion of the "dirty" unreformed syngas in the fuel combustor as needed.

Most of the process in this design is cooled using forced-air heat exchangers. For cooling lower temperature streams, a cooling water system is also included in the ASPEN Plus model to determine the requirements of each cooling water heat exchanger within the biomass conversion process as well as the requirements of the cooling tower. A mechanical draft cooling tower provides cooling water to several heat exchangers in the plant. The forced air and cooling water duties are shown in Table 10 along with their corresponding initial temperatures. Cooling tower specifications and results are shown in Table 11. The interstage coolers on the first syngas compressor account for 42% of the cooling duty provided by cooling fans. With an average gas temperature of 151 °C and nearly 49 MW of thermal energy potentially available, the interstage coolers are good candidates for alternate cooling methods or possibly power recovery using an organic rankine cycle (ORC) turbine. The largest percentage of cooling water duty (83%) is to condense the low-temperature vapors exiting the steam turbine at 43 °C—too cold for forced-air cooling or ORC.

Other ancillary equipment in the analysis includes an instrument air system, a firewater storage tank and pump, a diesel tank and pump, an olivine truck scale with dump and an olivine lock hopper, a MgO lock hopper, and a hydrazine storage tank and pump for cooling water chemical treatment.

A thermal pinch analysis was done to analyze the process energy network. The pinch concept offers a systematic approach to optimize the energy integration of the process and to detect any temperature crossing in heat exchangers. Details of the pinch analysis are discussed in the full design report.

Energy Balance

Energy integration is extremely important to the overall economics and efficiency of this process. Detailed energy balances around the major process areas were done using data

Table 10. Plant Cooling Duties for Forced-Air-Cooled and Water-Cooled Process Equipment

plant use	forced air cooling duty and initial temperature		cooling water duty and initial temperature	
	duty (MW)	temp (°C)	duty (MW)	temp (°C)
amine stripper condenser	27.4	100		
amine liquid cooler	13.7	81		
first syngas compressor interstage coolers	48.7	151 (av)		
second syngas compressor interstage coolers	3.64	88 (av)		
distillation column condensers	20.0	95 (av)		
molecular sieve condenser	0.84	107		
sand/ash cooler			0.255	995
quench water recirculation cooler			0.444	44
first syngas compressor aftercooler			1.56	60
LO-CAT absorbent solution cooler			0.008	49
synthesis raw product cooler	3.16	93	1.31	60
end product finishing coolers			0.130	49
blowdown water-cooled cooler			0.898	275
steam turbine condenser			22.2	43
total cooling duty	117		26.9	

Table 11. Process Water Demands for Ethanol Production

fresh water demands	flow rate (kg/h)
cooling tower makeup	54 480
boiler feed makeup	1700
sand/ash wetting	100
total	56 280
overall water demand (L water/L ethanol)	1.9

from the ASPEN Plus simulation. The major process energy inputs and outlets are illustrated in Figure 2, along with their energy flow values.

Consistent with previous discussion, the largest energy flows are associated with the forced-air interstage cooling of the compressors and from several other air-cooled heat exchangers including distillation column condensers. Together, these two categories represent 30% of the energy lost from the process. The "other" category not only consists primarily of other losses from the cooling tower system (drift and blowdown) but also accounts for energy losses due to ambient heating effects and mechanical work such as pump and compressor efficiency losses. Overall, the process is approximately 43% efficient on an LHV basis for moisture-free biomass. Figure 2 shows that approximately 55% of the energy in the wet raw biomass is recovered in the two alcohol products. Improvements in these energy efficiencies could potentially result in additional cost savings to the process and increased yields.

Water Issues

Because of water concerns surrounding other biomass to ethanol processes,[29] minimizing fresh water use was a primary design consideration for this process. When feasible, air-cooling was used in place of cooling water (e.g., distillation condensers, compressor interstage cooling, etc.) Water is required primarily as a fluidizing agent and reactant in the gasifier and fluidized bed steam reformer; however, process cooling is the major factor for water loss from the process. Over 75% of the fresh water use in the process is because of losses from the cooling tower via evaporation and drift. Since the steam turbine exit requires cooling water to condense the vapors, little can be done to reduce the cooling tower water losses without altering the turbine operating conditions. When water is a major concern for a specific plant location, additional water savings are available by using a higher exit pressure from the steam turbine and forced-air cooling to condense the steam. This will result in lower yields but may be necessary in some cases. The overall thermochemical ethanol process water demand of 1.9 L of water

per L of ethanol is considerably less than current fermentation-based ethanol plants.

Process Economics

The total project investment, based on total equipment cost (TEC) as well as variable and fixed operating costs, is developed from mass and energy balance information taken from the ASPEN simulation results. With these costs, a discounted cash flow analysis was used to determine the production cost of ethanol when the net present value of the project is zero. This section describes the cost areas and the assumptions made to complete the discounted cash flow analysis.

Capital Costs

The capital cost estimates are based as much as possible on the previous design work by Spath et al.[8] and Aden et al.[7] For these cost estimates, the purchased cost of the equipment was calculated, and cost factors were applied to determine the installed equipment cost.[30] This method of cost estimation has an expected accuracy of roughly +30% to −10%. The factors used in determining the total installed cost (TIC) of each piece of equipment and the resulting installed equipment costs by area are shown in Tables 12 and 13, respectively. Indirect costs (nonmanufacturing fixed-capital investment) were also estimated as per Spath et al. using cost factors, shown in Table 14, that are given as percentages of total purchased equipment cost, total installed cost (TIC), and total project investment (TPI) which is the sum of the TIC and the total indirect costs.

Operating Costs

The fixed operating costs for a biochemical ethanol facility given in ref 3 were used as a starting point to develop fixed costs for this thermochemical design with details given in the full report. The total labor costs[31] for the thermochemical process were estimated to be $2.270MM ($2005.) Other fixed costs were added to cover benefits, general plant maintenance, plant security, janitorial services, and communications. To make the estimate more closely match the PEP yearbook average,[32] an overhead factor of 95% was used resulting in an average loaded labor rate of $36/h. The fixed operating costs totaled $12.32MM in $2005 as given in Table 15. Variable operating costs, other than the feedstock costs, are given in Table 16.

Value of Higher Alcohol Coproducts and Minimum Ethanol Selling Price

The alcohol synthesis process will create higher molecular weight alcohols with a range of possible values depending upon
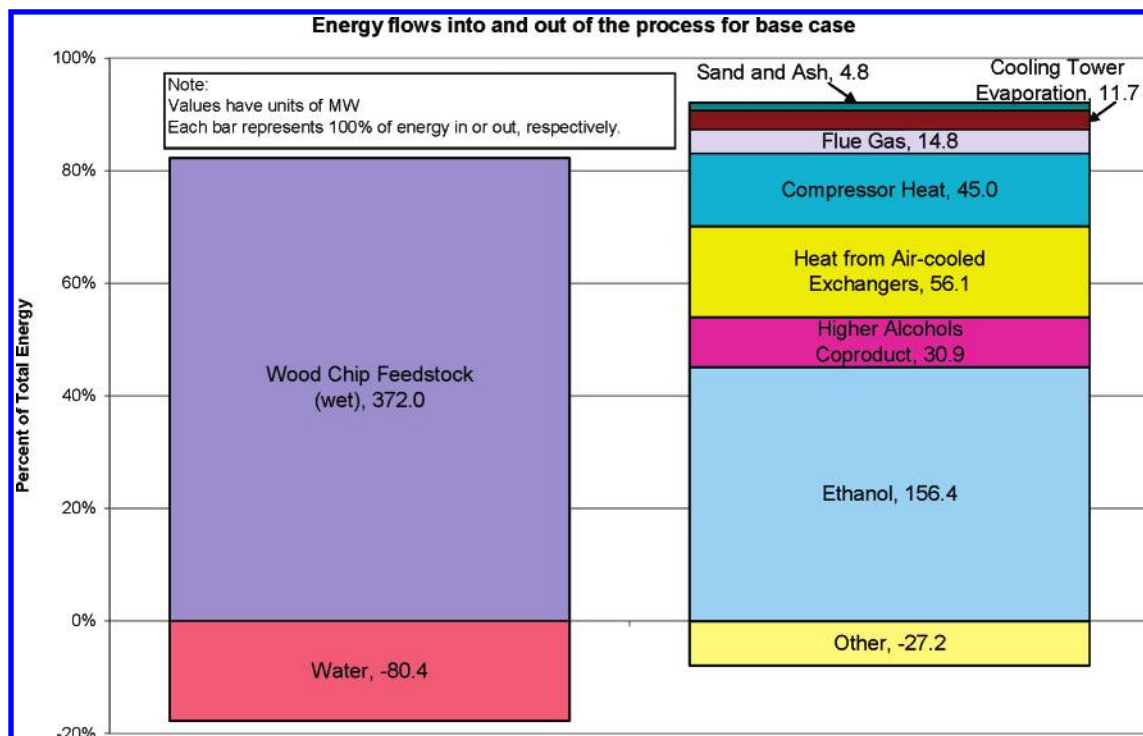
Figure 2. Chart of relative energy flows into and of the overall ethanol process. Small energy flows (<1%) have been omitted for clarity.

Table 12. General Cost Factors Used To Determine Total Installed Equipment Costs

	% of TPEC
total purchased equipment cost (TPEC)	100
purchased equipment installation	39
instrumentation and controls	26
piping	31
electrical systems	10
buildings (including services)	29
yard improvements	12
total installed cost (TIC)	247

Table 13. Installed Capital Equipment Costs by Process Area

process area	installed capital equipment cost ($2005)	% of total
feed handling and drying	23 200 000	17
gasification	12 900 000	9
FBSR and quench	38 400 000	28
acid gas and sulfur removal	14 500 000	11
alcohol synthesis - compression	16 000 000	12
alcohol synthesis - other	4 600 000	3
alcohol separation	7 200 000	5
steam system and power generation	16 800 000	12
cooling water and other utilities	3 600 000	3
total installed equipment cost	137 200 000	

Table 14. Cost Factors Used To Determine Indirect Costs

indirect costs	% of TPEC	% of TIC	% of TPI
engineering	32	13	9
construction	34	14	10
legal and contractors fees	23	9	7
project contingency	7.4	3	2
total indirect costs	96.4	39	28

their end market. At the high-value end, the chemical-market value could be $0.977/L to $1.11/L ($3.70/gal to $4.20/gal). At the low-value end, the coproduct should command at least a value similar to a residual fuel oil or about $0.225/L ($0.85/gal).[33] An intermediate value was chosen for the base case in this report. It is anticipated that the higher alcohols would make an excellent gasoline additive or gasoline replacement in its own

Table 15. Other Fixed Costs

cost item	factor	cost ($)
salaries		2 270 000
general overhead	95% of total salaries	2 155 000
maintenance	2% of total project investment	3 948 000
insurance and taxes	2% of total project investment	3 948 000

right upon EPA fuel certification. For the fuel market it should command a price similar to gasoline. Using the target $0.267/L ($1.07/gal) MESP as a scaled reference gasoline price and adjusting to n-propanol's heating value (a major constituent of the higher alcohol stream), its value would be $0.330/L ($1.25/gal). However, since the mixed alcohol product has not been refined to meet likely fuel specifications, its value was discounted by $0.026/L ($0.10/gal) to a price of $0.304/L ($1.15/gal).

Once the capital and operating costs were determined, a MESP for ethanol was determined using a discounted cash flow rate of return analysis. The MESP is the selling price of ethanol at the plant gate that makes the net present value of the process equal to zero using a 10% discounted cash flow rate of return over a 20-year plant life. The base case economic parameters used in this analysis are given in Table 17.

Process Economics, Sensitivity Analyses, and Alternate Scenarios

The cost of ethanol was derived using technology that has been developed and demonstrated or is currently being developed as part of the OBP research program at NREL. These targets represent the parameters in the technoeconomic model that give an estimated MESP of $0.267/L ($1.01/gal) ethanol— slightly less than the a priori $0.283/L ($1.07/gal) target. A summary of the breakdown of costs by process area is depicted in Figure 3. The cost year for the analysis was 2005.

Process costs are determined from various assumptions based on 2012-targeted technology, markets, and various financial assumptions. When a research target is not met, or a market or

Table 16. Variable Operating Costs

variable	information and operating cost
FBSR catalyst	To determine the amount of catalyst inventory, the FBSR was sized for a gas hourly space velocity (GHSV) of 2476/h based on the operation of the tar reformer at NREL's TCPDU where GHSV is given at standard temperature and pressure. initial fill then replacement of 1% per day of the total catalyst volume; price: $10.30/kg
alcohol synthesis catalyst	initial fill then replaced every 5 years based on typical catalyst lifetime; catalyst inventory based on GHSV of 6000/h; price: $11.57/kg
gasifier bed material	synthetic olivine and MgO; delivered to site by truck equipped with self-contained pneumatic unloading equipment; disposal by landfill; olivine price: $190.59/tonne; MgO price: $402.34/tonne
solids disposal cost	price: $19.81/tonne
diesel fuel	usage: 37.9 L/h plant wide use; price: $0.26/L
chemicals	boiler chemicals − price: $3.09/kg; cooling tower chemicals − price: $2.20/kg; LO-CAT chemicals − price: $150/tonne of sulfur produced
wastewater	The wastewater is sent off-site for treatment. price: $0.96/m^3

Table 17. Economic Parameters Used To Calculate MESP for Ethanol

assumption	value
internal rate of return (after-tax)	10%
debt/equity	0%/100%
plant life	20 years
general plant depreciation	200% DDB
general plant recovery period	7 years
steam plant depreciation	150% DDB
steam plant recovery period	20 years
construction period	2.5 years
first 6 months expenditures	8%
next 12 months expenditures	60%
last 12 months expenditures	32%
start-up time	6 months
revenues	50%
variable costs	75%
fixed costs	100%
working capital	5% of total capital investment
land	6% of total purchased equipment cost (cost taken as anexpense in the first construction year)

financial assumption does not hold, then the MESP is affected to varying degrees. Additionally, uncertainty about equipment design and installation and construction costs will impact the economics. Understanding how the variation in key parameters impacts the MESP helps determine which parameters can have the greatest impact on results. From this understanding of the key parameter's impact, research can be focused on the areas providing the biggest potential returns on research funding or that lower the risk of entering the market.

Feedstock cost is the largest contributor to the plant operating cost and is 43% of the MESP. Gas cleanup accounts for 31% of the MESP primarily because of its high capital equipment cost. Accordingly, two major areas of research in thermochemical ethanol production are feedstock cost reduction and gas cleanup and conditioning technology improvements to lower the capital equipment and catalyst costs. The large negative cost contribution in the synthesis area is because of the electricity generated by a turboexpander used to recover energy from the unused syngas being recycled to the FBSR and fuel combustor.

The results for the sensitivity analysis are illustrated in Figure 4 arranged from largest overall sensitivities at the top to smallest at the bottom. Nearly all of these ranges represent variations of a single variable at a time with two exceptions: (1) the feedstock comparison of corn stover to lignin required varying the ultimate elemental analysis, ash content, and moisture content simultaneously, and (2) the Combined Reformer Conversions incorporated all of the individual sensitivity ranges for the methane, benzene, and tar simultaneously. Note that every item in the chart has an associated value; if a bar is not apparent, then its effect over the specified range was insignificantly small. The

bars are color-coded to differentiate between technical research and development (R&D) and financial variables.

Financial Scenarios

Financial parameters have a significant effect on the MESP, yet R&D has the smallest direct effect on them. In particular, the required ROI for the project could more than double the calculated MESP. Successful R&D and demonstration projects can reduce the ROI requirements of corporations and lending institutions, decrease capital costs by implementing lessons learned in future designs, and reduce the required MESP toward the baseline case in this report. Since this study is based on an *n*th-plant design scenario, ROI requirements are likely to be closer to the base case value than the extreme case in the sensitivity analysis.

A conceptual design like this is normally thought to give accuracy in the capital requirements of −10% to +30%. Using this range for the total project investment (TPI) gives an MESP range of −7% to +20%.

Feedstocks

Because this process has been designed for utilization of forest resources there may be little control over the feedstock quality coming to the plant. The two most important feedstock quality parameters that can impact the process economics are moisture and ash content.

The high range of the ash content examined here are more indicative of agricultural residues or lignin-rich biochemical process residues; forest resources should have ash contents near that of this baseline case (about 1%). Keeping the feedstock cost constant on a moisture and ash free (maf) basis decreased the MESP; however, since the biomass feedstock rate to the plant was arbitrarily capped at 2000 tonne/day regardless of feedstock characteristics, increasing the ash percentage decreased the convertible material entering the plant resulting in a loss of alcohol yield as shown in Figure 5. The yield loss had a more significant impact on the MESP than did the price adjustment for ash content. This one variable-at-a-time sensitivity to ash misses another characteristic of agricultural residues, low moisture content, that would also help negate the yield losses since less or possibly no drying would be required. To negate the loss of product resulting from higher ash contents, the feed to the plant would need to be kept at a constant energy rate that would inherently account for variations in feedstock characteristics.

The biomass feed moisture higher than the baseline 50% is a problem. Drier feedstocks will give a lower MESP because of decreased drying requirements and lower raw syngas diver-
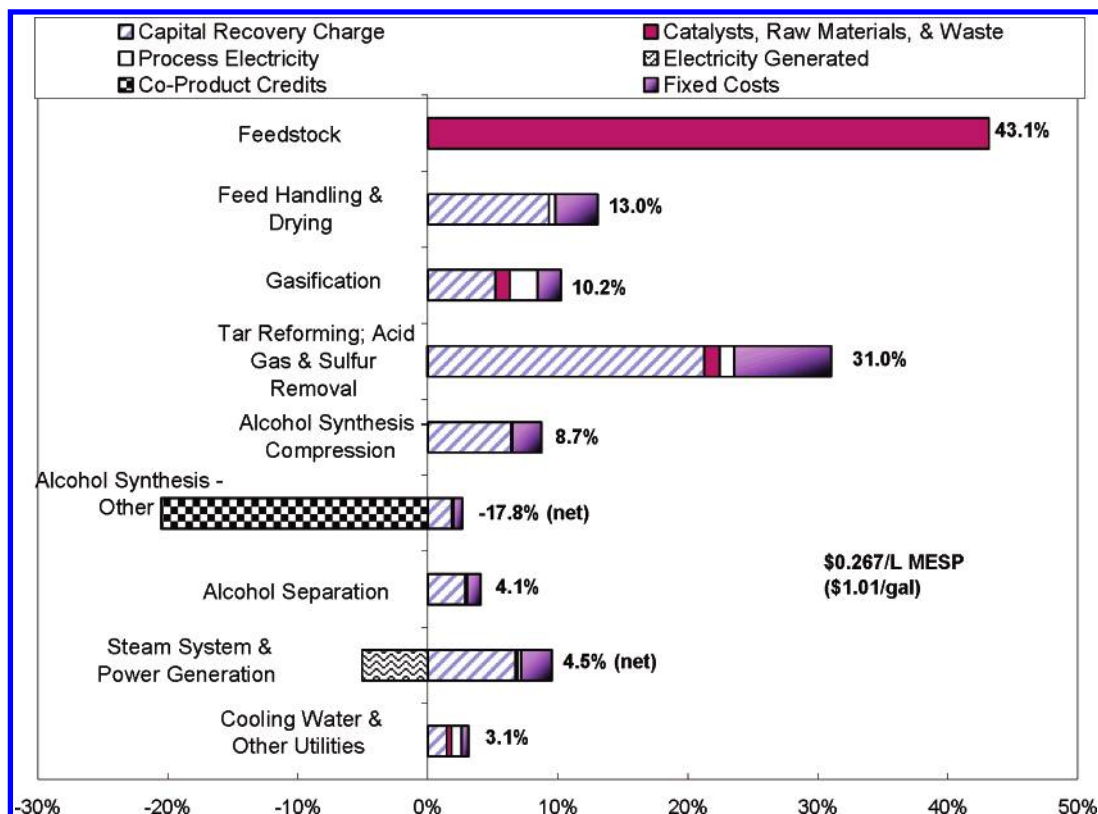
Figure 3. Cost contributions to the ethanol minimum ethanol selling price (MESP) for each process area.
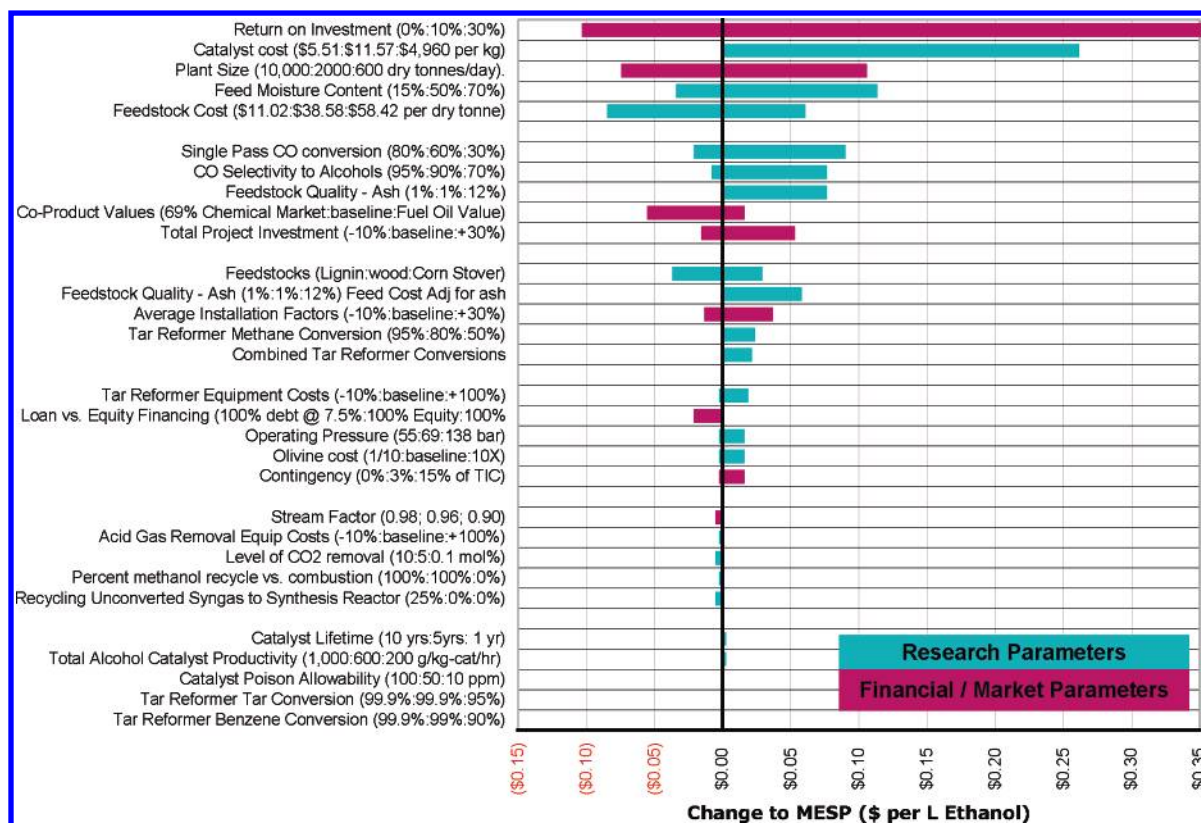

Figure 4. Results from sensitivity analyses of key technical and economic model parameters for the biomass to ethanol process.

sion required for heat and power resulting in higher product yields. As moisture content increases, the alcohol yield decreases because more raw syngas must be diverted for heat. Note in Figure 6 that decreasing moisture contents do not give corresponding increases in alcohol yields because flue gas is used for the drying and other process heating requirements dictate

the amount of raw syngas diverted for heat and power, not the feedstock drying.

Thermal Conversion

Two gasification scenarios were examined. First, increasing the olivine cost by an order of magnitude (due to increased
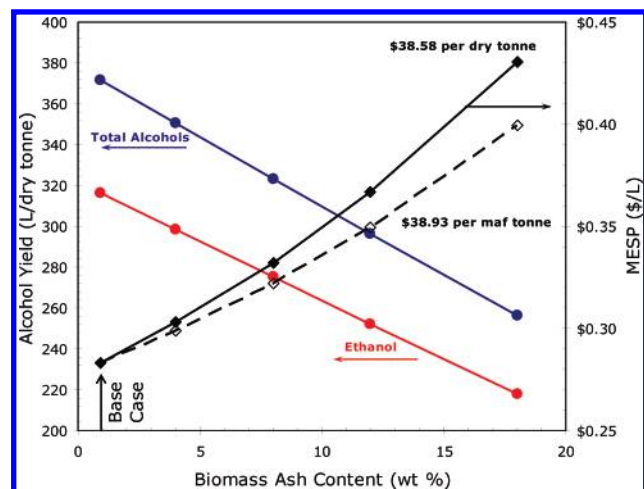
Figure 5. Impact of variation in biomass feedstock ash content on MESP and product yields at a constant feed rate independent of ash content.
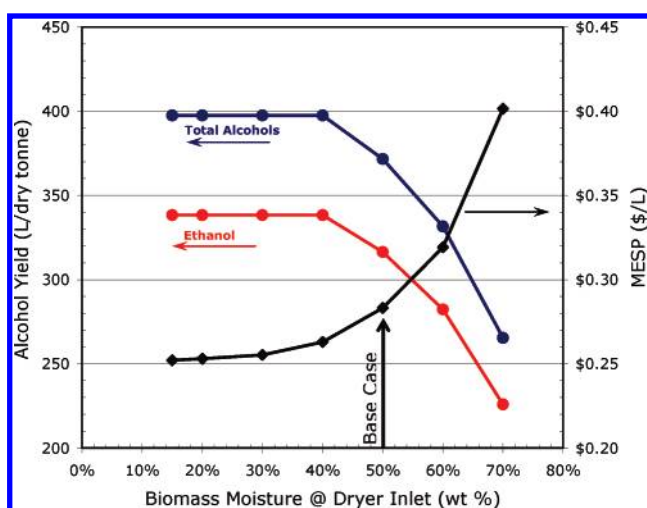

Figure 6. Impact of variation in biomass feedstock moisture content on MESP and product yields at a constant feed rate independent of moisture content.

olivine makeup and/or increased cost because of catalytic modification) could increase the MESP by 7%. The effect of reducing methane production in the gasifier increased the MESP by 0.1%.

Cleanup and Conditioning

These scenarios appeared to have an imperceptible effect on the MESP. However, these results are misleading. The scenarios show cost effects due to the material and energy balances. Since the amount of tar is small compared to the amount of CO and H_2, these effects are small. In reality, Cleanup and Conditioning is absolutely required for acceptable performance of gas compressors, wastewater treatment, and alcohol synthesis catalysts. Excessive tars in the syngas would significantly impact compressors and wastewater treatment with severe consequences to equipment and increased operating costs that are not rigorously modeled here. So, not meeting these targets would give poor performance, leading to greater cost effects than reflected by the sensitivity analysis for this area. The FBSR also provides for the retention of carbon from residual syngas streams with hydrocarbons, alcohols, and carbon dioxide in the process to improve product yields. Without the FBSR, residual gas streams could only be used for combustion resulting in increased electricity production and decreased liquid products and higher MESP.

Fuels Synthesis

These scenarios show the importance of the R&D for the synthesis catalysts. Poor performance could increase MESP by 25% or more. Whether this is due to actual nontarget catalyst formulations or due to poor performance in Cleanup and Conditioning that leads to poor alcohol synthesis catalyst performance, the cost effects are major. The catalyst cost sensitivity range was extremely large, from $5.51/kg to over $4960/kg. This was done to bracket a variety of potential catalyst systems, not just cobalt moly sulfide. Exotic metals such as rhodium (Rh) or ruthenium (Ru) can add considerable cost to a catalyst system even at relatively low concentrations. At low catalyst costs, total CO conversion and alcohol selectivity (CO_2-free basis) have the largest impact on the overall MESP. The catalyst productivity (g/kg/h) did not show much impact over the sensitivity range chosen. In reality, all of these catalyst performance indicators are tightly linked. It is unlikely that research could change one without affecting the others.

Another sensitivity evaluated was to burn the methanol/water mixture from the molecular sieve flush step instead of recycling to the synthesis reactor. This option increased the MESP by only $0.008/L ($0.03/gal), because the heat from the alcohol combustion offset the need to divert syngas for process heat.

Markets

Crediting the coproduct higher alcohols with the lower fuel oil value increases the MESP by about 6%; this is still reasonable to meet the qualitative "cost competitive" target. Of even more significance is that selling these higher alcohols for even 69% of their chemical market value will lead to a significant reduction of MESP (about 40%). This shows that the first thermochemical conversion plants could get a significant economic advantage in their early life from higher alcohol coproduct sales.

Conclusions

This technoeconomic study shows that biomass-derived ethanol from a thermochemical conversion process has the potential to be produced in a manner that is "cost competitive with corn-ethanol" by 2012 using commercially available unit operations (such as acid gas removal.) However, gas cleanup and catalytic synthesis are two critical process areas needing significant R&D advances to reach the $0.267/L ($1.01/gal) of ethanol MESP presented in this paper. The potential for significant improvements in process design and plant energy efficiency are evident from the current study and should lead to lower MESP and increased ethanol yields independent of meeting other performance targets. Areas for further development are suggested in the next section.

Future Work

The design discussed in this paper is only one of many possible thermochemical process configurations for converting biomass into liquid fuels. Future R&D work to develop and demonstrate reforming and synthesis catalysts is an obvious path to pursue. Better process models for gasification, alcohol synthesis, and steam reforming of tars are vital for a better understanding of the process. Different gasifier types, such as oxygen blown, are also of interest for converting biomass to fuels. Other possible research areas include the following: (1) examine the trade-offs of water cooling vs air cooling vs organic rankine cycle for cooling and power production, (2) understand

the trade-offs between an energy neutral, alcohol production facility to one that could also supply heat and electricity to a colocated biochemical conversion facility, (3) examine the potential of increased overall energy efficiency, and (4) examine the emissions profile from the plant and explore alternate emissions control equipment and their impact on alcohol yield.

Acknowledgment

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