

PROCESS SYNTHESIS AND ANALYSIS OF CATALYTIC BIOFUEL PRODUCTION FROM SEPARATE CONVERSION OF HEMICELLULOSE AND CELLULOSE USING LIGNIN-DERIVED ALKYLPHENOL SOLVENTS

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ABSTRACT

We develop a process in which the cellulose and hemicellulose fractions of lignocellulosic biomass are converted separately to jet fuel-range liquid hydrocarbon fuels (butene oligomers (BO)) while alkylphenol solvents (propyl guaiacol (PG) and propyl syringol (PS)) are produced from the lignin fraction through catalytic conversion subsystems. In this process dilute sulfuric acid (SA)-catalyzed pretreatment fractionates the biomass into insoluble cellulose and soluble hemicellulose-derived xylose, and they then are converted separately to levulinic acid (LA) using 2-sec-butylphenol (SBP) and lignin-derived alkylphenol solvents (LDS), respectively. The LA is converted catalytically to BO, passing through γ -valerolactone (GVL) and butene platforms. This process leads to a high biomass-to-fuels yield (34.8 mol%) at low concentrations of biomass derivatives using large volumes of solvents. Therefore, we design separation subsystems for recovering the alkylphenol solvents and biomass derivatives to be combined with the catalytic conversion subsystems of hemicellulose, cellulose and lignin. We then show a heat exchanger network (HEN) design to satisfy total energy requirements of the process from combustion of biomass degradation products. Finally, our economic analysis shows that the process design using corn stover feedstock leads to a minimum selling price of \$3.37 per gallon of gasoline equivalent for jet fuel-range liquid hydrocarbon fuels, which suggests that it is an economically competitive alternative to current biofuels production approaches as a result of iterative experimental and computational efforts.

INTRODUCTION

In general biofuel production processes from lignocellulosic biomass (biochemical[1-3], thermochemical[1, 4, 5], and catalytic[6, 7]) used to produce biofuels, cellulose and hemicellulose fractions of the lignocellulosic biomass can be converted to various biofuels (bioethanol, biobutanol, and biogasoline[8-11]), while the remaining fraction (lignin) used to process heat and power through combustion[7]. In catalytic biofuel production, Biofuel can be produced by catalytic upgrading of Biomass-derived platforms (Levulinic acid (LA) and γ -valerolactone (GVL)) at low concentrations using large

volumes of solvents[6, 9, 12, 13]; a huge energy is required for recovery of the solvents and extraction of the platforms. However, this approach has the advantages of milder processing and less expensive processing than various technologies (biochemical[1, 14] and thermochemical[15] processes).

Jet fuel-range liquid hydrocarbon fuels (BO) can be produced from LA by passage over a triple-bed catalyst system (RuSn(1:4)/C, SiO₂/Al₂O₃, and Amberlyst 70) via GVL and butene platforms [6, 16]. After fractionation of biomass using pretreatment[14, 17-21], LA can be produced separately either by dehydration of cellulose-derived glucose and dehydration of hemicellulose-derived xylose. Recently, according to reported research, an alkylphenol solvent (2-sec-butylphenol (SBP)) can effectively extract LA from the dilute sulfuric acid (SA)-catalyzed hydrolysate of cellulose[19]. In addition, according to Azadi et al.[22], LA can be obtained from hemicellulose-derived xylose using alkylphenol solvents (Lignin-derived alkylphenol solvent (LDS); propyl guaiacol (PG), and propyl syringol (PS)) obtained by depolymerization of lignin.

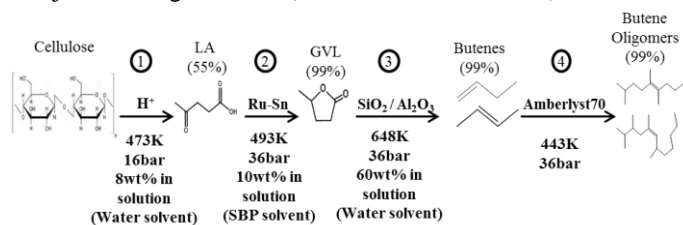
We present a strategy for separate catalytic conversion of hemicellulose (using LDS) and cellulose (using SBP alkylphenol solvent) fractions of lignocellulosic biomass to liquid hydrocarbon fuels. First, we develop an integrated catalytic conversion process for hemicellulose, cellulose, and lignin (including separation subsystems for recovering the alkylphenol solvents and biomass-derived platforms (LA and GVL)). And we design a heat exchanger network (HEN) to reduce the energy requirements of the integrated process. Finally, we prove the major technological bottlenecks and cost drivers of our developed process by technoeconomic analysis.

TECHNOLOGY OVERVIEW

Our strategy combines three methods to catalytically convert individual components of biomass to fuels/solvents: hemicellulose-to-fuels, cellulose-to-fuels, and lignin-to-solvents.

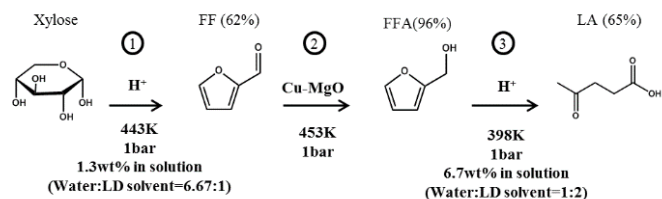
First, the cellulose-to-fuels conversion strategy (developed by Sen et al. [19]) consists of (1) LA production from cellulose hydrolysis and dehydration[23], (2) LA reduction to GVL[16], and (3) production of BO from GVL[6]. The cellulose fraction is converted with 55 mol% yield into LA in a batch reactor containing 8 wt% cellulose and 0.5M SA catalyst in a water

solution at 473K and 16 bar, while the remaining cellulose is degraded to humins[23] (Reaction 1 in Scheme 1). After extraction of LA from the aqueous SA solution using an alkylphenol solvent (SBP), 10 wt% LA is converted to GVL in the presence of SBP with a 99% molar yield over a RuSn(1:4)/C catalyst at 493K and 36 bar using internal H₂ released from formic acid(FA) decomposition [16] (Reaction 2 in Scheme 1). Finally, the **BO** can be produced from the GVL (with 40 wt% water) over two catalyzed bed reactors connected in series. The GVL in a water solution can be converted to butene with near quantitative yield over a SiO₂/Al₂O₃ catalyst at 648 K and 36bar (Reaction 3 in Scheme 1), and subsequently the most of butene in the absence of water can be converted to **BO** (C₈H₁₆, C₁₂H₂₄, C₁₆H₃₂, C₂₀H₄₀) over a Amberlyst-70 catalyst at 443 K and 36bar, resulting in a mixture of gasoline and jet fuel range alkenes (Reaction 4 in Scheme 1).



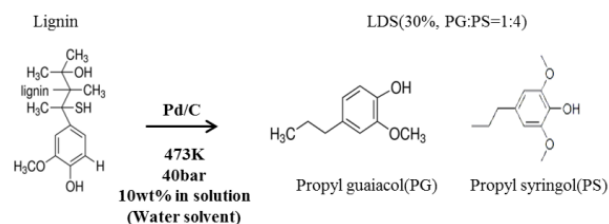
Scheme 1 Conversion of cellulose to **BO**[19]

The hemicellulose-to-fuels conversion strategy consists of five catalytic conversion subsystems for the (1) biphasic conversion of xylose to furfural (FF)[22], (2) hydrogenation of FF to furfuryl alcohol (FFA) [24], (3) biphasic hydrolysis of FFA to LA[22], (4) LA reduction to GVL[16], and (5) production of **BO** from GVL[6]. First, the hemicellulose-derived xylose can be converted to FF with a high yield (62 mol%) at 443K and 1 bar in a biphasic reactor using organic (LDS) and acidic aqueous (containing 0.1M hydrochloric acid(HCl) and sodium chloride (NaCl)) solvents (water: LDS=6.67:1, w/w) with 1.3 wt% xylose water, while 13% of the remaining xylose is degraded to humins [22](Reaction 1 in Scheme 2). In the biphasic reaction, the aqueous (water) phase is NaCl-saturated to increase the polarity difference between aqueous and organic phases, leading to high partitioning (84%) of FF into the LDS. After distillation of FF from the LDS, the distilled FF can be converted to FFA with a 96% molar yield in a fixed bed reactor with a Cu-MgO catalyst in an H₂-rich condition (H₂ : FF molar ratio = 2.5:1) at 453K and 1 bar[24] (Reaction 2 in Scheme 2). After this gaseous reaction, the FFA can be obtained with high yield (65 mol%) into LA at 398K and 1 bar in a biphasic reactor containing 6.7 wt% FFA and dilute SA catalyst (0.3 M) in an organic (LDS/aqueous (water)) solution (water: LDS=1:2, w/w), while the remaining FFA is degraded to humins[22] (Reaction 3 in Scheme 2). Using LDS as organic solvents in the biphasic reaction leads to a low concentration of FFA in the aqueous phase. This results in high yields of LA because of a decrease in side reactions as well as high partitioning (76%) of LA into LDS. Finally, **BO** can be produced from the LA via GVL and butene platforms based on the aforementioned reactions (Reactions 2-4 in Scheme 1).



Scheme 2 Conversion of xylose to LA

Azadi et al.[22] showed that LDS containing alkyl-substituted phenolics (PG and PS) can be produced from the lignin by catalytic depolymerization (Reaction 1 in Scheme 3). First, the lignin can be converted with 63% molar yield into phenolic monomers (PG, PS, guaiacyl propanol (GP) and syringyl propanol (SP)) in a one-pot reactor over a metal catalyst (Pd/C) using a H₂-rich condition containing 10 wt% lignin in a water solution at 473K and 40 bar, while the remaining lignin is degraded to humins. As a result, after separation of organic phase (PG and PS), we obtain the final LDS mixture (PS:PG=4:1w/w).



Scheme 3 Conversion of lignin to LDS

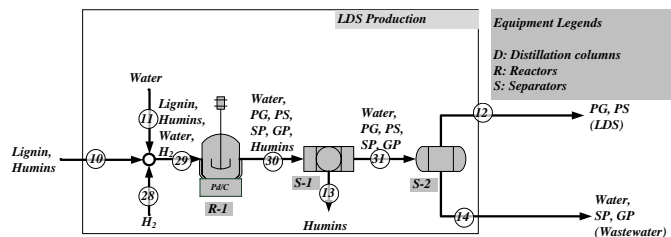


Figure 1 LDS production subprocess

As a result, we could obtain 46.3% of the overall “potential” cellulose/hemicellulose-to-**BO** experimental yield. The “potential” cellulose-to-**BO** and xylose-to-**BO** experimental yields were 53.4% and 36.4%, respectively.

PROCESS DEVELOPMENT

Process Synthesis

We developed an integrated catalytic process for monophasic conversion of cellulose to **BO** (using SBP solvent), biphasic conversion of hemicellulose to **BO** (using LDS), and monophasic conversion of lignin to LDS. The integrated process consists of nine major sections: biomass handling, pretreatment, wastewater treatment, boiler and turbogenerator, storage, cellulose conversion, hemicellulose conversion, butene production/butene oligomerization, and LDS production (Fig. 2).

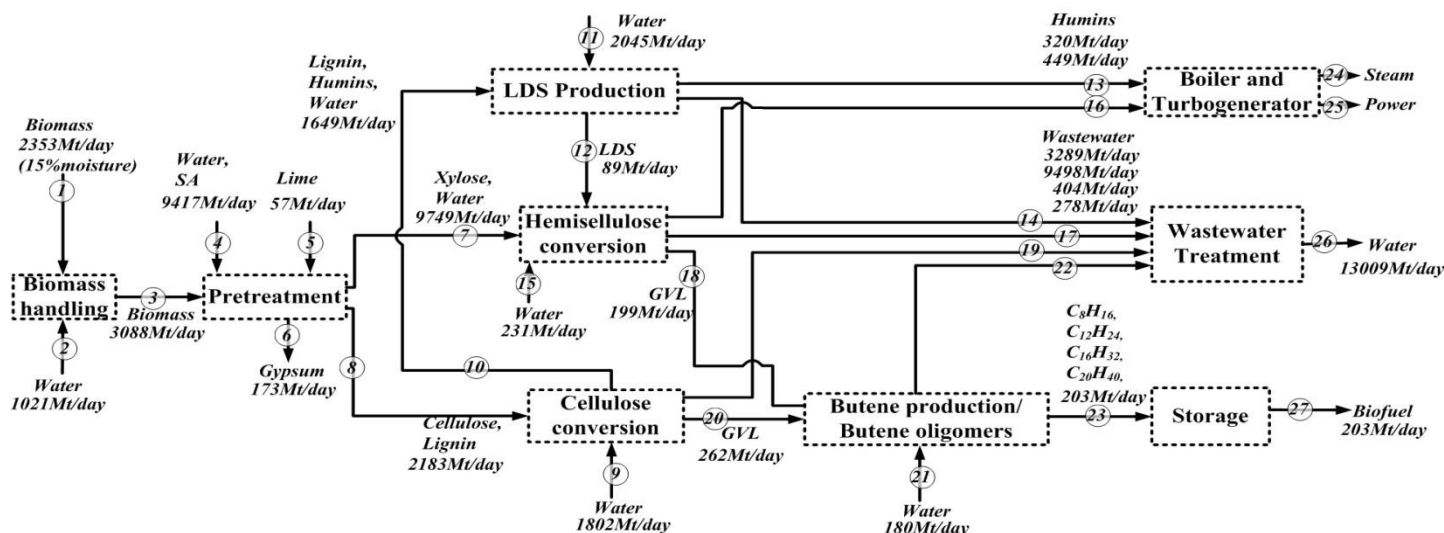


Figure 2 Integrated catalytic process

After size reduction of corn stover in the biomass handling section, the crushed corn stover is chemically pretreated using 0.5 M SA [1]. In the pretreatment section, most of the hemicellulose is converted to soluble xylose (92.5 mol%) and FF (5 mol%). Then, the liquid mixture (containing xylose, FF, SA, and water) is separated from insoluble materials (containing cellulose and lignin) through filtration. The insoluble materials are sent to cellulose (C_6) conversion subprocess, while the liquid mixture is sent to hemicellulose derived xylose (C_5) conversion subprocess after neutralization with $\text{Ca}(\text{OH})_2$ (lime) and removal of gypsum by precipitation.

C_6 conversion subprocess comprises six sections, as shown in Fig. 3: C_6 -derived LA production, SA recovery for C_6 conversion, SBP recovery for C_6 conversion, C_6 -derived GVL production, butene production, and butene oligomerization. After LA production from cellulose (R-2 in Fig. 3; Reaction 1 in Scheme 1), the liquid mixture (containing LA, FA, SA, and

water) is separated from insoluble materials (lignin and humins) by a liquid–solid separator (S-3 in Fig. 3) with addition of water (feed : water mass ratio = 1:1.16). In this step, most of the lignin (with humins) is sent for LDS production (R-1 in Fig. 1; Reaction 1 in Scheme 3). Separation of LA is separated from the acidic liquid mixture in six stages of extraction using SBP (feed:SBP mass ratio = 1:4.4) at 453K and 16 bar, while most LA is recovered with other components (29% FA and 97% recycled GVL) in the organic (SBP) phase, and the SA-aqueous phase containing the remaining materials is recycled back to C_6 -derived LA production (R-2 in Fig. 3). Following GVL production from LA (R-3 in Fig. 3; Reaction 2 in Scheme 1), 80% of CO_2 is removed by a flash tank at 380K (S-5 in Fig. 3). After SBP/GVL distillation, 98% of GVL is recovered from the top of a 18-stage distillation column (D-1 in Fig. 3) at 392K and 1 bar, and 99.8% of SBP is recovered from the bottom at 512K. In two catalyst bed reactors with an platform butene– H_2O

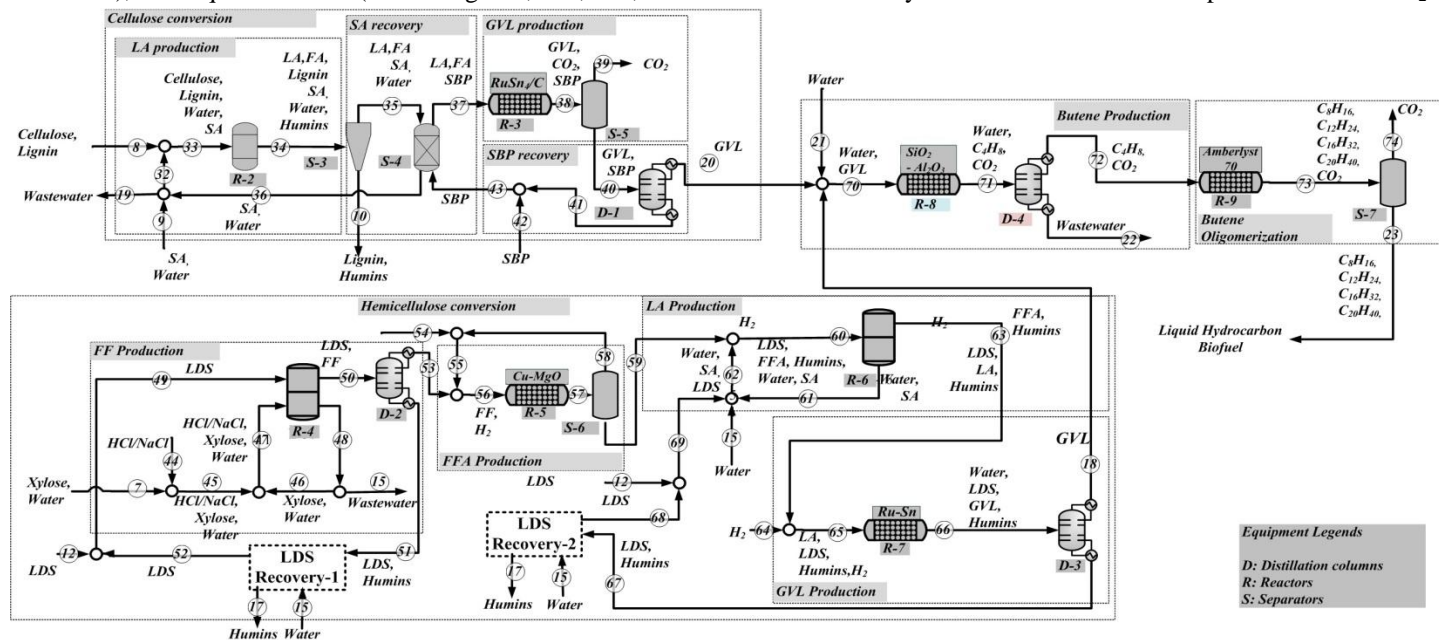


Figure 3 C_5 and C_6 conversion subprocesses

separator (D-4 in Fig. 3), the distilled GVL is converted to a mixture of BO and CO₂ (R-8, R-9 in Fig 3; Reactions in 3-4 Scheme 1) and is then condensed at 295K to separate liquid fuels (C₈H₁₆, C₁₂H₂₄, C₁₆H₃₂, C₂₀H₄₀) from the vapor stream (99.3%) of CO₂ (S-7 in Fig. 3).

C₅ conversion subprocess comprises six sections, as shown in Fig. 4: C₅-derived FF production, C₅-derived FFA production, LDS recovery 1 for C₅ conversion, C₅-derived LA production, C₅-derived GVL production, LDS recovery 2 for C₅ conversion. Pretreated xylose is converted to FF in a biphasic (LDS/water) reactor (R-4 in Fig. 3; Reaction 1 in Scheme 2), and then the organic (LDS) stream (containing FF and humins) is separated from the aqueous stream (containing unconverted xylose) via a biphasic reactor with a biphasic decanter equipped. 99.9% of FF is recovered from the top of a 20-stage distillation column (D-2 in Fig. 3) at 440K and 1 bar, and 99.9% of LDS (with 13.3 wt% humins) is recovered from the bottom at 542K. After passing FF/LDS distillation, most of the humins are removed from bottom stream using a separation subsystem for LDS recovery 1 (including distillation, pressure filtration, and decanting) and used to generate heat and electricity in the boiler and turbogenerator, and the recovered LDS (99.5% of the LDS) are recycled back to the biphasic reactors (R-4 in Fig. 3). The distilled FF is converted into FFA in a fixed bed reactor (R-5 in Fig 3; Reaction 2 in Scheme 2), which operates in an H₂-rich condition. Most H₂ is removed passing through a flash tank at 300K (S-6 in Fig. 3) and is then recycled back to R-5. The FFA is converted to LA in a biphasic (LDS/water) reactor (R-6 in Fig 3; Reaction 3 in Scheme 2), and 76% of LA is then extracted by LDS. The LA is converted into GVL in a fixed bed reactor (R-7 in Fig. 3) that operates in an H₂-rich condition. To reduce the LDS requirement, LDS have to be reused in the biphasic reaction, thus we designed a separation subsystem for LDS recovery (including distillation, pressure filtration, and decanting) similar to LDS recovery subsystem 1. After GVL/LDS distillation, 99.9 wt% GVL is separated from the top of the 20-stage distillation column (D-3 in Fig. 3) at 393K and 1 bar and the distilled GVL is subsequently converted to BO (R-4, R-5, D-4, S-7 in Fig. 3) as described in the previous sections.

As a result, the overall cellulose/hemicellulose-to-BO numerical yield is 41.9%; cellulose-to-BO and hemicellulose-to-BO numerical yields are 43.1% and 39.9%, respectively. The numerical yield of cellulose-to-BO is lower than the experimental yield due to some loss of biomass derivatives (LA, GVL, and BO) during their recovery via separation systems. Especially 1.8% loss of LA occurs in SA recovery section. On the other hand, the numerical yield of hemicellulose-to-BO is higher than the experimental yield because the unconverted xylose is reused in FF production section. The comparison of experimental and numerical yields are shown in Table 1.

Table 1 Comparison of experimental yields and numerical yields

Unit: %	Cellulose-to-BO yields	Hemicellulose-to-BO yields	Overall yields
Experimental	53.4	36.4	46.3
Numerical	43.1	39.9	41.9
Difference	-10.3	+3.5	-4.4

Heat Integration

When 2000 tons of corn stover per day are processed (energy content of 358 MW)[7], our integrated process requires 273 MW of heat, 281 MW of cooling, and 2 MW of electricity. The energy content of liquid hydrocarbon fuel is 114 MW, which is 31.9% of the energy content of biomass feedstock (358.4 MW). The total heating and electricity requirements of our integrated process cannot be satisfied by burning residue, because the energy content of biomass residue is determined to be 186 MW. Thus, we carried out heat integration between hot and cold process streams for reduce the energy requirements. After heat integration, we design a HEN with 49 heat exchangers, which resulted in significant energy recovery (197 MW), thereby reducing the heating requirements of the process by more than 72% to 76 MW, as shown in Table 2. Also, reduced total heating requirements can be satisfied by combustion of biomass residues (132MW).

As a result, the energy flow diagram in Fig 4 describes the energy efficiency (37.9%), which is the ratio of the energy output to energy input.

Table 2 Energy requirements of our strategy before and after heat integration.

Energy required [MW]	Before heat integration		After heat integration	
	Heating	Cooling	Heating	Cooling
Pretreatment	47.9	5.5	0.0	5.5
C ₆ conversion				
LA production	61.1	60.9	6.3	29.4
SA recovery	0.0	11.5	0.0	0.0
SBP recovery	5.2	3.8	0.0	0.0
GVL production	4.2	0.0	0.2	0.0
C ₅ conversion				
FF production	87.2	102.8	18.2	28.6
FFA production	0.4	2.5	0.4	2.5
LDS recovery 1	16.8	16.8	16.8	0.0
LA production	0.0	14.7	0.0	0.0
GVL production	12.3	4.9	8.4	0.0
LDS recovery 2	16.8	13.2	16.8	0.0
Butene production	14.7	13.6	8.4	0.0
Butene oligomerization	0.4	3.5	0.4	0.6
LDS production	5.7	27.3	0.0	7.0
Total	272.6	281.0	75.9	84.3

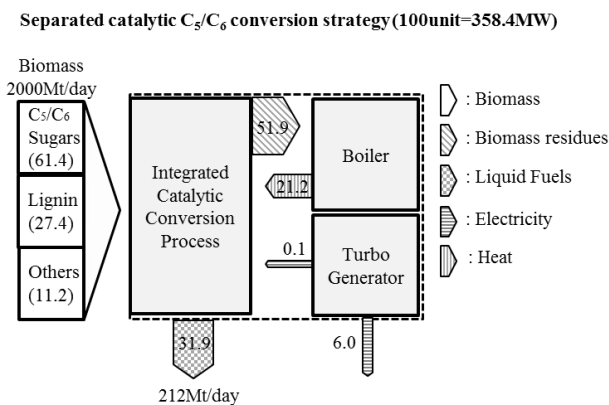


Figure 4 Energy flow diagram

TECHNOECONOMIC EVALUATION

Design basis and assumptions

To compare our strategy reported here with previous strategies for catalytic **BO** production (C₆ conversion strategy[19] and co-catalytic C₅/C₆ conversion strategy[7]), we follow a similar approach to that used in the benchmark lignocellulosic ethanol production strategy proposed by the National Renewable Energy Laboratory (NREL) [14]: we use the same feedstock processing rate (2000 dry metric tonne (Mt) of corn stover per day) and the same economic assumptions .

We developed a simulation model for our strategy using the ASPEN Plus Process Simulator[25]. Based on the experimental data and simulation results, equipment costs of cellulose conversion, hemicellulose conversion, and LDS production subprocesses were estimated using the Aspen Process Economic Analyzer[26], while the equipment costs of the remaining sections (biomass handling, pretreatment, boiler and turbogenerator, wastewater treatment, and storage) were estimated using an exponential scaling expression based on the equipment size and cost data in NREL [1, 14]. Equipment and utility costs of the HEN were also estimated using the Aspen Energy Analyzer[27].

Capital and Operating Costs

Table 3 shows the annualized capital and operating costs of Separated catalytic C₅/C₆ conversion (our strategy), C₆ conversion [19] and co-catalytic C₅/C₆ conversion strategy[7]. The total annualized capital costs for all strategies were determined to be \$32-39 million per yr. The largest portion of total annualized capital cost for all strategies was the boiler and turbogenerator (25-37%). The total annualized capital cost of our strategy was determined to be \$39 million per yr, which is 19-22% higher than the costs of the other strategies. Because our strategy requires hemicellulose conversion and LDS production/recovery subprocesses (\$2.4 million per yr difference) when compared with the C₆ conversion, and a pretreatment step (\$8.4 million per yr of difference) when compared with the co-catalytic C₅/C₆ conversion.

The largest portion of total operating cost for all strategies was feedstock cost (51-62%). The total operating cost of our strategy was estimated to be \$88.8 million per yr, which is 7.5-22.1% higher than that of the other strategies (co-catalytic

C₅/C₆ conversion: \$82.6 million per yr and C₆ conversion: \$72.1 million per yr). Because our strategy involves a larger conversion process, including C₅ conversion/LDS production subprocesses (newly added HCl, NaCl, and more H₂ use), and requires a pretreatment step (more SA and lime use). This means that our strategy has (1) higher material costs (H₂, HCl, NaCl, SA, lime, \$17.7 million per yr) and (2) higher fixed operating costs (\$9.7 million per yr), calculated based on the capital cost. As a result, the total production cost for our strategy, which is the sum of annualized capital and operating costs, is about \$127.6 million per yr, which is 11.9-21.3% higher than that of the other strategies.

Table 3 comparison of annualized capital and operating costs for all strategies

	Co-catalytic C ₅ /C ₆ conversion		C ₆ conversion		Separated catalytic C ₅ /C ₆ conversion	
	C	O	C	O	C	O
Pretreatment	0.0	0.0	8.4	5.8	8.4	4.9
LDS production	0.0	0.0	0.0	0.0	0.3	2.9
LA production	8.6	1.2	4.6	2.5	6.3	8.9
GVL production	2.6	24.7	1.6	7.3	2.7	12.3
SA recovery	0.0	0.0	1.2	0.0	1.2	0.0
GVL recovery	4.0	0.0	0.0	0.0	0.0	0.0
SBP recovery	0.0	0.0	1.3	0.0	1.3	0.0
LDS recovery	0.0	0.0	0.0	0.0	2.1	0.0
BO production	4.4	0.0	1.9	0.0	3.0	0.0
Boiler and turbogenerators	10.2	1.0	12.2	1.3	9.7	2.6
Other processing sections	1.7	1.6	1.3	1.7	3.9	2.3
Fixed operating cost	0.0	8.9	0.0	8.9	0.0	9.7
Feedstock cost	0.0	45.2	0.0	45.2	0.0	45.2
Total capital/operating cost	31.5	82.6	32.5	72.7	38.8	88.8
Total cost	114.0		105.2		127.6	

Minimum Selling Price (MSP)

We determined a minimum selling price (MSP) of liquid hydrocarbon fuels (**BO**) for a net present value equal to zero by using a discounted cash flow analysis[14] based on the capital and operating costs. Figure 5 shows the MSPs expressed in terms of gallon of gasoline equivalent (GGE) for all strategies. The MSP of liquid hydrocarbon fuels for our strategy (\$4.54 per GGE) is \$0.72 per GGE lower than the C₆ conversion strategy (\$5.26 per GGE) because of higher fuel production (66.5%; 25.8 × 10⁶ GGE per yr) despite its higher total annualized cost (21.3%; \$127.6 million per yr) compared to the C₆ conversion strategy (15.5 × 10⁶ GGE per yr; \$105.2 million per yr). However, the MSP for our strategy is \$0.25 per GGE higher than the co-catalytic C₅/C₆ conversion strategy (\$4.29 per GGE) because of the lower total annualized cost (11.8%; \$114.1 million yr⁻¹) despite similar fuel production (0.8%; 26.1 × 10⁶ GGE per yr) of the co-catalytic C₅/C₆ conversion strategy.

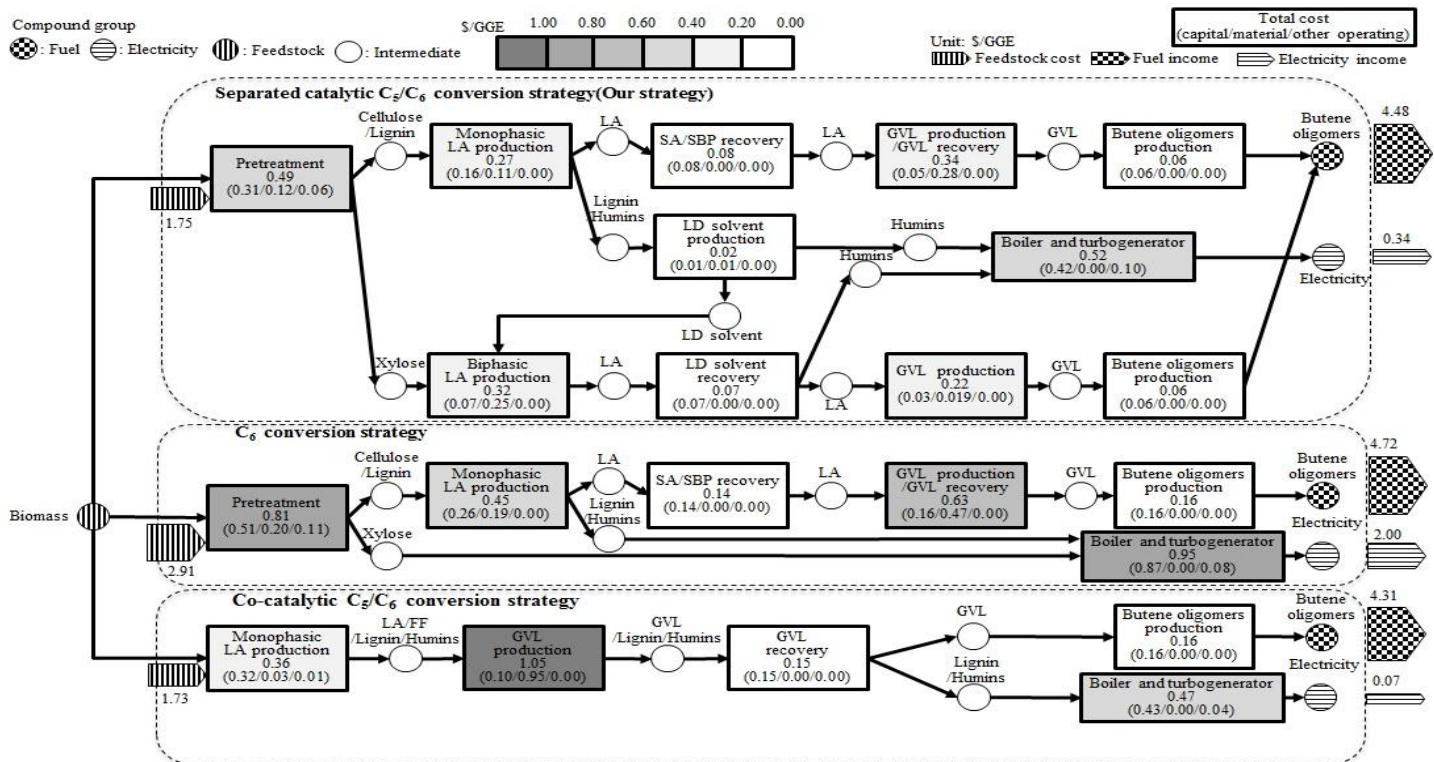


Figure 5 Capital and operating cost contributions of each processing section to the MSP for catalytic biofuel production strategies.

The overall cost of liquid hydrocarbon fuels, which is the sum of the MSPs of liquid hydrocarbon fuels and electricity credit, is \$4.94 per GGE for our strategy, and 8.1% of the overall cost is electricity credit (\$0.40 per GGE), also Total cost (=fuel income+electricity income=\$4.94 per GGE) includes utilities (capital = \$0.02 per GGE, operating = \$0.04 per GGE), storage (capital = \$0.02 per GGE, operating = \$0.00 per GGE), wastewater treatment (capital = \$0.13 per GGE, operating = \$0.05 per GGE), and fixed operating costs (\$0.37 per GGE).

Sensitivity Analysis

We studied how key economic parameters affect the MSP for our strategy and co-catalytic C_5/C_6 conversion strategy [7] (Fig. 6). Feedstock cost, discount rate, H_2 cost, tax rate, and electricity price were selected as key economic parameters; variations in the selected parameters reflect values reported recently in the literature. First, we observed that feedstock cost had a similar impact on the MSP for both strategies: MSP decreased by 11.0~11.7% (\$0.50 per GGE) when the feedstock price decreased from \$65/ton (base case) to \$46/ton [7]. We also studied the impact of discount rate on the MSP for both strategies. A decrease in discount rate from 10% to 6.74% [28] resulted in a higher improvement (7.0%) of the MSP for our strategy than the other strategy, because annualized capital cost is strongly related to the discount rate and the annualized capital cost of our strategy (\$38.8million yr^{-1}) is higher than that of the other strategy (\$31.5million yr^{-1}). In addition, the MSP for our strategy decreased by 4.4% (\$0.26 per GGE), which is higher than that of the other strategy (1.6%; \$0.07 per GGE), when the H_2 cost (\$2.0 per kg) decreased to \$1.1 per

kg [29], because the amount of H_2 required by our strategy (16.6 Mt/day) is about 2.9-fold higher than that required by the other strategy (5.7 Mt/day). However, a decrease in tax rate from 35% to 13% [46] resulted in a similar improvement (1.9~2.2%; \$0.08~0.10 per GGE) in the MSP for both strategies due to similar fuel sale revenues. Finally, an increase in electricity price from \$0.0572 per kWh to \$0.0639 per kWh [30] did not lead to substantial improvements in the MSP for either strategy due to low net surplus electricity. When all parameters were adjusted simultaneously, the MSP for our strategy decreased by 25.8% to \$3.37 per GGE, which is \$0.01 per GGE lower than that of the co-catalytic C_5/C_6 conversion strategy and is close to that of conventional fuel.

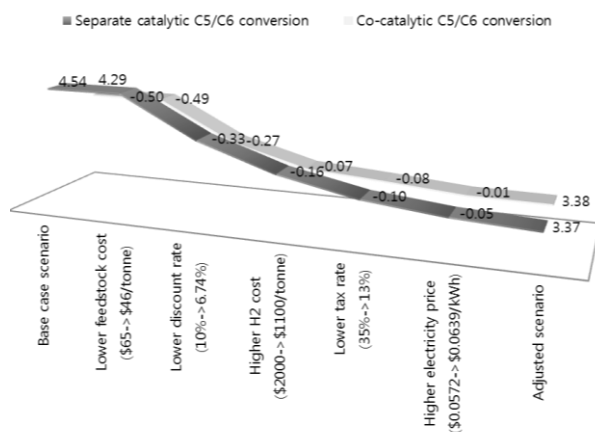


Figure 6 Impacts of adjusting key economic parameters (\$ per GGE)

CONCLUSION

We developed a strategy for separate catalytic conversion of hemicellulose and cellulose fractions of lignocellulosic biomass to jet fuel-range liquid hydrocarbon fuels (BO) using LD and SBP alkylphenol solvents. The proposed strategy has a high biomass-to-fuels yield (34.8 mol%) at low concentrations (1.3–30 wt% solids) using large volumes of solvents; a large amount of energy is therefore required for extraction of biomass-derived platforms (LA and GVL) and recovery of the solvents. To reduce the energy requirements, we designed effective separation subsystems and HEN. We were able to recover most of the solvents (99%) and reduce the total heating requirements by 72%, thereby satisfying all the energy requirements of the proposed strategy by combusting biomass residues (mostly degradation products). Our techno-economic analysis showed that the proposed strategy results in a minimum selling price of \$3.37 per GGE for jet fuel-range liquid hydrocarbon fuels if using 2,000 Mt per day of corn stover feedstock processing and updated economic parameters.

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