



Hydrogen Production from Biomass via Indirect Gasification: The Impact of NREL Process Development Unit Gasifier Correlations

C.M. Kinchin, and R.L. Bain

Technical Report
NREL/TP-510-44868
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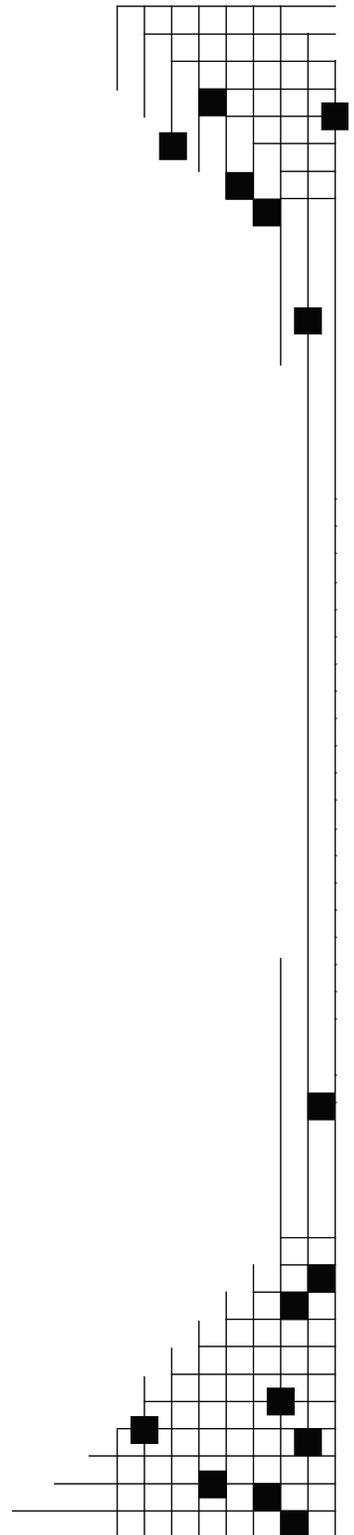
C.M. Kinchin, and R.L. Bain

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National Renewable Energy Laboratory
1617 Cole Boulevard, Golden, Colorado 80401-3393
303-275-3000 • www.nrel.gov

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Abbreviations and Acronyms

BCL	Battelle Columbus Laboratory
CO	carbon monoxide
CO ₂	carbon dioxide
DCFROR	discounted cash flow rate of return
H ₂	molecular hydrogen
H ₂ S	hydrogen sulfide
MHSP	minimum hydrogen selling price
NREL	National Renewable Energy Laboratory
PSA	pressure swing absorption
scf	standard cubic feet
TCPDU	Thermochemical Process Development Unit
TOC	total operating cost
TPI	total project investment

Executive Summary

NREL developed a new set of empirical gasifier correlations using data collected from the Thermochemical Process Development Unit. The new correlations replaced the old correlations in the Aspen Plus model presented in the Biomass to Hydrogen Design Report (Spath et al. 2005). The new correlations predict a slightly different dry gas composition, although the most significant difference is less char. The new correlations predict 0.10 lb of char per lb of dry feed, whereas the old correlations predict 0.22 lb of char; consequently, a portion of the raw syngas must be diverted to the char combustor to provide the heat necessary for gasification and drying the incoming biomass. The model predicts 2.6% more final hydrogen product when the new gasifier correlations are used. The new correlations also predict a minimum hydrogen selling price of \$1.24 per kg (2002 dollars), which equal to the \$1.24 per kg predicted by the old correlations. Using 2007 Biomass Program economic assumptions, the new correlations predict a minimum hydrogen selling price of \$2.14 per kg (2007 dollars).

Table of Contents

Abbreviations and Acronyms	iii
Executive Summary	iii
1 Introduction.....	1
2 NREL Gasification Tests	1
3 Results of Regression Analysis.....	3
3.1 Significance Testing.....	3
3.2 Regression Results	4
4 Comparison of New and Old Correlations.....	6
5 Results of Wood to Hydrogen Model using New Correlations	7
5.1 Overview of Model.....	7
5.2 Results of Goal Case Design with New Correlations	9
6 Updated Economic Analysis using New Correlations.....	11
6.1 Capital Costs	11
6.2 Operating Costs.....	11
6.3 Discounted Cash Flow Rate of Return Analysis (DCFROR).....	12
6.4 New Correlations versus Old Correlations in 2002 Dollars	12
7 Updated Economics	15
7.1 Economic Results Using 2005 Biomass Program Assumptions.....	15
7.2 Economic Results Using 2007 Biomass Program Assumptions.....	16
7.3 Economic Results using H2A Analysis	17
7.4 Summary of Economics.....	17
8 Conclusions.....	18
9 References.....	18
Appendix A. Correlation Values for Equations 1, 2 and 3	19

1 Introduction

In 2005, the National Renewable Energy Laboratory (NREL) developed an Aspen Plus model to examine the technical and economic feasibility of indirect steam gasification to produce hydrogen from wood (Spath et al. 2005). In the model, the gasifier products were predicted using empirical correlations developed by Bain in 1992 (Bain 1992). The empirical correlations were based on data collected from the Battelle Columbus Laboratory (BCL) 9 tonne/day facility. In 2007 and 2008, NREL conducted gasification tests using various biomass feedstocks for the purpose of developing empirical correlations. The new set of empirical correlations have been generated and assimilated into the Wood to Hydrogen model. This report presents the results of the model using the new correlations and is divided into the following sections:

1. A brief description of the NREL gasification tests
2. Results of the regression analysis used to develop the new correlations
3. Comparison of the new and old correlations
4. Results of the Wood to Hydrogen model using the new correlations
5. Updated economics analysis using new correlations
6. Updated economics using 2005 and 2007 Biomass Program assumptions
7. Updated economics using the H2A analysis.

2 NREL Gasification Tests

The NREL gasification tests examined the effects of several process variables on the gasifier product yields and compositions (char production rate, tar rates, dry gas composition, etc.). The following process variables were adjusted in the experiments:

1. Feedstock: wheat straw, Vermont wood, corn stover, oak wood, and pine wood
2. Gasification temperature: 600°C to 950°C
3. Steam to biomass mass ratio: 0.33 to 2.0.

Not all feedstock samples were tested over the entire range of gasification temperatures. Only oak and pine woods were tested at gasification temperatures above 875°C and steam to biomass ratios above 1.2. Also, residence time at the maximum gasification temperature was calculated and used as a process variable in the Aspen Plus model, although it was not controlled during the gasification experiments. The NREL gasifier is actually a two stage gasification process, consisting of a fluidized bed reactor with a maximum temperature of 750°C, followed by an electrically heated thermal cracker capable of temperatures up to 950°C.

The effects of the adjusted process variables (input variables) listed above were measured in the following product rates and compositions (output variables):

1. Total nitrogen-free dry gas yields, standard cubic feet (scf) per lb of moisture and ash free (MAF) feed

2. Hydrogen, mol%
3. Carbon dioxide, mol%
4. Carbon monoxide, mol%
5. Methane, mol%
6. Ethane, mol%
7. Ethylene, mol%
8. Acetylene, mol%
9. Propane, mol%
10. Propene, mol%
11. 1-Butene, mol%
12. 2-c-Butene. mol%
13. 2-t-Butene, mol%
14. Carbonyl sulfide, ppmv per lb of dry wood
15. Hydrogen sulfide, ppmv per lb of dry wood
16. Benzene, lb per lb of dry wood
17. Toluene, lb per lb of dry wood
18. Phenol, lb per lb of dry wood
19. Cresols, lb per lb of dry wood
20. Naphthalene, lb per lb of dry wood
21. Phenanthrene, lb per lb of dry wood
22. Total tars (sum of all tars except benzene), lb per lb of dry wood
23. Char, lb per lb of dry wood.

It should be noted that components 2 through 15 are reported on a dry, nitrogen-free, tar-free basis. They were measured after tars and water present in the syngas were condensed in scrubbers. Components 16 through 22 are considered tars and were measured prior to the scrubbers.

The purpose of this report is to provide the results of the Wood to Hydrogen Aspen Plus model with updated gasifier correlations; therefore, a detailed description of the experimental procedure used to collect the thermochemical process development unit (TCPDU) data will not be provided here. The detailed experimental procedure and results can be found in the Joule Milestone Report “Pilot-Scale Parametric Gasification of Wood, Switchgrass, and Wheat Straw to Develop Correlations for Input in Process Models” (Phillips 2007) and “Parametric Gasification of Oak and Pine Feedstocks using the TCPDU and Slipstream Water-Gas Shift Catalysts” (Hrdlicka 2008).

3 Results of Regression Analysis

3.1 Significance Testing

The data collected during the NREL gasification tests were subjected to a regression analysis using Unscrambler statistical software. The significance of several process variables was first tested. The process variables considered were:

1. Proximate analysis
2. Ultimate analysis
3. Ash composition (base ratio, dolomite ratio, Ca content)
4. Fluid bed temperature
5. Secondary entrained-flow thermal cracker temperature
6. Steam to biomass ratio
7. Residence time.

In addition to the process variables listed above, interactions between most of the process variables as well as squared effects were also tested for significance. All of the process variables were determined to be significant for at least a few of the output variables, and none were determined to significantly affect all output variables, although several process variables were found to be significant more often than others. The process variables that were most often significant, as well as generally the most significant in terms of the highest degree of significance were the following:

1. Ultimate analysis
2. Proximate analysis
3. Gasification temperature (secondary thermal cracker temperature)
4. Steam to biomass ratio
5. Residence time
6. Interactions between gasification temperature, steam to biomass ratio, and residence time
7. Squared effects of gasification temperature, steam to biomass ratio, and residence time.

The significance testing resulted in 18 significant process variables: five ultimate analysis terms (5), four proximate analysis terms (9), gasification temperature (10), steam to biomass ratio (11), residence time (12), three interactions terms (15), and three squared terms (18). A correlation equation requiring eighteen terms suggests over-fitting, but the correlation needs to be sufficiently robust to predict several gasification products (dry gases, tars, char) from a wide variety of feedstocks. If the correlation is asked to predict only dry gases from a single feedstock composition, less than eighteen terms would be required, but this model is expected to be more robust. Also, terms that show very small variation across feedstocks, such as mass percent hydrogen in the ultimate analysis, could arguably be eliminated from the correlations, but other ultimate analysis terms such as the nitrogen and sulfur mass percent make significant contributions to syngas cleanup operating costs.

3.2 Regression Results

After significant effects were determined, a separate regression analysis was run in Unscrambler for each of the output variables (product yields and compositions). The result of the regression analysis was an equation that predicts the production rates or gas composition as a function of the input process variables:

$$Y = B_{int} + X_M * B_M + X_{VM} * B_{VM} + X_{FC} * B_{FC} + X_A * B_A + X_C * B_C + X_H * B_H + X_O * B_O + X_N * B_N + X_S * B_S + X_{TC} * B_{TC} + X_{SB} * B_{SB} + X_{RT} * B_{RT} + S_{TC}^2 * B_{TC}^2 + S_{SB}^2 * B_{SB}^2 + S_{RT}^2 * B_{RT}^2 + I_{TC:SB} * B_{TC:SB} + I_{TC:RT} * B_{TC:RT} + I_{SB:RT} * B_{SB:RT} \quad (1)$$

where

Y = Predicted value of output variable (lb/lb of dry biomass for char and tars, mole percent for dry gases)

B_{int} = Intercept term

X_M = Moisture mass percent of the feed (proximate analysis)

B_M = Coefficient of moisture term

X_{VM} = Volatile matter mass percent of the feed (proximate analysis)

B_{VM} = Coefficient of volatile term

X_{FC} = Fixed carbon mass percent of the feed (proximate analysis)

B_{FC} = Coefficient of fixed carbon term

X_A = Ash mass percent of the feed (proximate analysis)

B_A = Coefficient of ash term

X_C = Carbon mass percent of the feed (ultimate analysis, wt% as received)

B_C = Coefficient of carbon term

X_H = Hydrogen mass percent of the feed (ultimate analysis, wt% as received)

B_H = Coefficient of hydrogen term

X_O = Oxygen mass percent of the feed (ultimate analysis, wt% as received)

B_O = Coefficient of oxygen term

X_N = Nitrogen mass percent of feed (ultimate analysis, wt% as received)

B_N = Coefficient of nitrogen term

X_S = Sulfur mass percent of feed (ultimate analysis, wt% as received)

B_S = Coefficient of sulfur term

X_{TC} = Gasification temperature, °C

B_{TC} = Coefficient of gasification temperature term

X_{SB} = Steam to biomass ratio

B_{SB} = Coefficient of steam to biomass ratio term

X_{RT} = Residence time, seconds

B_{RT} = Coefficient of residence time term

S_{TC}^2 = Gasification temperature squared term

B_{TC}^2 = Gasification temperature squared coefficient

S_{SB}^2 = Steam to biomass ratio squared value

B_{SB}^2 = Steam to biomass ratio squared coefficient

S_{RT}^2 = Residence time squared value

B_{RT}^2 = Residence time squared coefficient

$I_{TC:SB}$ = Gasification temperature – steam to biomass ratio interaction value

$B_{TC:SB}$ = Gasification temperature – steam to biomass ratio interaction coefficient

$I_{TC:RT}$ = Gasification temperature – residence time interaction value

$B_{TC:RT}$ = Gasification temperature – residence time interaction coefficient

$I_{SB:RT}$ = Steam to biomass ratio – residence time interaction value

$B_{SB:RT}$ = Steam to biomass ratio – residence time interaction coefficient

The linear terms for the ultimate analysis mass percents, proximate analysis mass percents, gasification temperature, steam to biomass ratio, and residence time are all calculated by simply multiplying the measured value for the input variable by its respective coefficient. The squared and interaction values, on the other hand, are calculated as follows:

If S_A^2 is the square value of input process variable A, and I_{AB} is the interaction value between input process variables A and B, then:

$$S_A^2 = (\text{Weight}_A * (X_A - \text{Center}_A))^2 \text{ and} \quad (2)$$

$$I_{AB} = \text{Weight}_A * (X_A - \text{Center}_A) * \text{Weight}_B * (X_B - \text{Center}_B) \quad (3)$$

where Weight_i and Center_i are values generated by Unscrambler used to calculate the squared and interaction terms. For every output variable a unique set of B intercept, B coefficient, weight, and center values are generated by Unscrambler. X_A and X_B are the measured values of the process variables (residence time, thermal cracker temperature, and steam to biomass ratio). Appendix A presents the correlation values for equations 1, 2, and 3, as well as the R^2 value of the correlation equation for each predicted variable.

A common method of measuring the accuracy of a correlation equation is to predict an output variable using measured conditions, and then plot the predicted values versus the measured values. An example of a predicted versus measured plot is provided in Figure 1. In Figure 1, the predicted hydrogen production rate is plotted versus the measured hydrogen production rate.

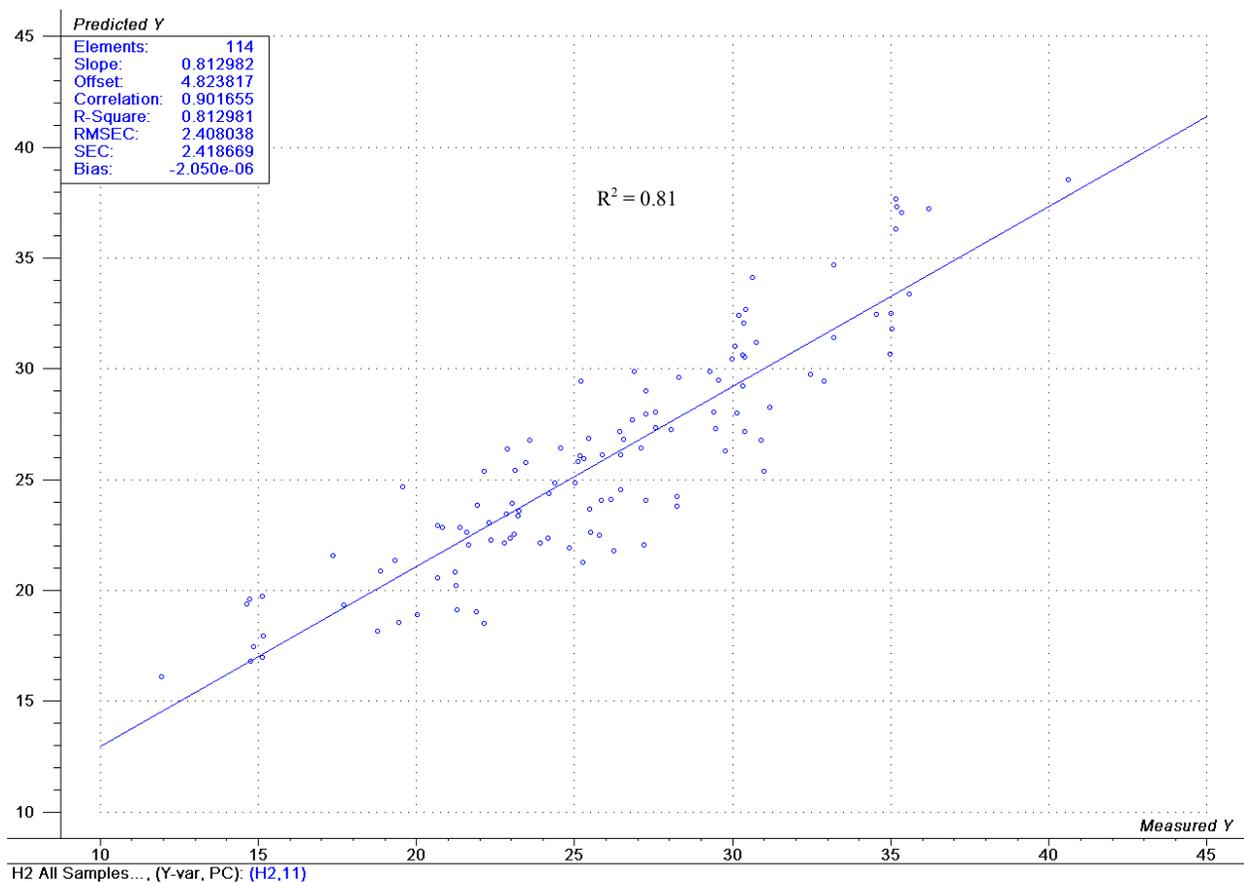


Figure 1. Sample Unscrambler regression plot, predicted versus measured H₂ concentration (mol %) in the dry, scrubbed gas

4 Comparison of New and Old Correlations

The new correlation equations predict most components better than the previous equations. In addition, several dry gas components and tar species not predicted with the old set of correlations can now be predicted. Table 1 below compares the performance of the new and old correlations using the R-square value, R^2 , of the regression line drawn through the predicted versus measured plot.

The adjusted R-square value, R^2_{adj} , is also reported in Table 1. The old correlation uses fewer terms and is based on a smaller data set than the new correlations. The adjusted R^2 calculation considers the size of the data set as well as the number of terms used in the correlation equation. Thus, the original R^2 values are “adjusted” to account for the size of the data set as well as the number of terms, which allows the two correlations to be compared more fairly. In this case, a comparison of the adjusted R^2 values for the old and new correlations is consistent with a comparison of original R^2 values; therefore the adjusted R^2 results do not provide any new information but are included for the sake of completeness.

Table 1. R^2 Values for New and Old Correlation Equations

Component	New Correlation R^2	Old Correlation R^2	New Correlation R^2_{adj}	Old Correlation R^2_{adj}
Hydrogen	0.81	0.92	0.79	0.91
Carbon Monoxide	0.64	0.40	0.60	0.36
Carbon Dioxide	0.77	0.42	0.74	0.38
Methane	0.81	0.70	0.79	0.68
Ethane	0.71	0.85	0.68	0.84
Ethylene	0.87	0.88	0.85	0.87
Acetylene	0.94	0.72	0.93	0.70
Propane	0.90		0.89	
Propene	0.92		0.91	
1-Butene	0.83		0.81	
2-c-Butene	0.73		0.70	
Hydrogen Sulfide	0.86		0.84	
Benzene	0.93		0.92	
Toluene	0.72		0.67	
Phenol	0.91		0.90	
Cresols	0.89		0.87	
Naphthalene	0.92		0.91	
Phenanthrene	0.85		0.83	
Heavy Tar, MW > 180	0.68		0.63	
Total Tar, MW > 78	0.82	0.89	0.79	0.88
Char	0.78	0.66	0.75	0.64
Nitrogen Free Dry Gas Yield	0.87	0.94	0.85	0.94

5 Results of Wood to Hydrogen Model using New Correlations

5.1 Overview of Model

In the model, wood is gasified using a low-pressure indirectly-heated circulating fluidized bed gasifier. Indirect heat is supplied by adding hot olivine to the gasifier. After gasification, olivine and char are separated from the syngas using cyclones. The char, mixed with olivine, is then combusted to add heat to the olivine, which is then re-circulated back to the gasifier. Steam is also added to the gasifier as a fluidizing agent. The syngas exiting the gasifier then passes through a catalytic tar reformer where hydrocarbons and tars are cracked and reformed with water to increase CO and H₂ yields. The syngas is then cleaned of CO₂ and H₂S. The cleaned syngas then passes through two shift reactors that increase the H₂ to CO ratio. Pressure swing absorption (PSA) is then used to separate hydrogen from the syngas.

The feedstock used for the model is hybrid poplar wood chips delivered at 50 wt% moisture. The capacity of the model is 2,000 bone dry tonnes/day. The model requires a small amount of natural gas, although more will probably be necessary for startup. The majority of steam and electricity necessary to operate the envisioned facility is produced by combusting byproducts.

The 2005 Biomass to Hydrogen Design Report (Spath et al, 2005) presents a techno-economic analysis for two design cases: a current design case and a goal design case. The current design case assumes tar conversion rates verified experimentally with no regeneration of the tar reforming catalyst. The current design case also includes a steam-methane reformer, prior to the shift reactors, to convert methane and hydrocarbons to CO and H₂. However, the goal design case assumes higher (goal) tar conversion, and includes a tar and methane reforming, catalyst regenerating reactor. As such, the goal design case does not include a steam-methane reformer because conversion of methane and hydrocarbons in the tar reformer is high enough that a steam-methane reformer is not justified.

More detailed descriptions of the current and goal case designs are presented in the 2005 Biomass to Hydrogen Design Report (Spath et al. 2005). A block flow diagram of the goal case design is presented in Figure 2.

5.2 Results of Goal Case Design with New Correlations

Table 2 compares the gasifier performance for the goal case design using the old and new correlations.

Table 2. Gasifier Operating Parameters, Yields, and Gas Compositions Using New and Old Correlation Equations

Gasifier Variable	Old Correlations		New Correlations	
Gasifier Temperature	1598°F (870°C)		1605°F (874°C)	
Pressure	23 psia (1.6 bar)		23 psia (1.6 bar)	
Steam to Biomass Ratio (biomass on a bone-dry basis)	0.4 lb/lb		0.4 lb/lb	
Gas Composition	mol % (wet)	mol % (dry)	mol % (wet)	mol % (dry)
H ₂	12.91	23.85	8.20	14.92
CO ₂	6.93	12.80	7.81	14.21
CO	22.84	42.18	21.82	39.67
H ₂ O	45.87	---	45.00	---
CH ₄	8.32	15.36	12.81	23.30
C ₂ H ₂	0.22	0.41	0.40	0.73
C ₂ H ₄	2.35	4.35	2.36	4.30
C ₂ H ₆	0.16	0.29	0.00	0.00
C ₃ H ₈	Not Predicted	Not Predicted	0.30	0.55
C ₃ H ₆	Not Predicted	Not Predicted	0.26	0.47
1-C ₄ H ₈	Not Predicted	Not Predicted	0.08	0.15
c-C ₄ H ₈	Not Predicted	Not Predicted	0.00	0.00
t-C ₄ H ₈	Not Predicted	Not Predicted	0.14	0.25
C ₆ H ₆	0.07	0.13	0.26	0.47
Tar (C ₁₀ H ₈)	0.13	0.23	0.35	0.64
NH ₃	0.18	0.32	0.15	0.28
H ₂ S	0.04	0.07	0.04	0.06
Dry Gas Yield, lbmol of dry gas/lb of dry feed	0.035		0.036	
H ₂ :CO molar ratio	0.57		0.38	
Gas Heating Value, Btu/lb	Wet: 4,759 HHV Dry: 8,019 HHV	4,401 LHV 7,416 LHV	Wet: 4,345 HHV Dry: 6,937 HHV	4028 LHV 6,430 LHV
Char Yield, lb/lb of dry feed	0.22		0.10	
Gasifier Efficiency	72.1% HHV 71.8% LHV		71.7% HHV 71.5% LHV	

After replacing the old correlations with the new correlations, the most significant change to the gasifier product composition is the char yield. Using the old correlations, about 0.22 lb of char is produced for every pound of moisture-free biomass fed to the gasifier. Using the new correlations, about 0.10 lb of char is produced. Ten percent char yield is more consistent with literature values than 22% for the conditions modeled (gasification temperature, pressure, steam-to-biomass ratio). Char is not a primary or secondary product of the biomass to hydrogen process, but because it is combusted to heat the olivine it does affect the heat balance of the indirect gasification system as well as the overall heat balance of the entire model. The amount of char produced according to the new correlations does not provide enough heat to maintain the gasifier above 1400°F, therefore 23% of the raw syngas must be combusted with the char to provide the indirect heat necessary to maintain a gasification temperature close to 1600°F. Also, flue gas from the char combustor is used to dry the incoming biomass from 50 wt% moisture to 12 wt% moisture. Less char requires less air for combustion, resulting in less flue gas available for drying. The amount of flue gas produced by the char combustor according to the new correlations does not provide enough heat to dry the incoming biomass to a moisture content of 12 wt%. Diverting 23% of the raw syngas to the char combustor results in enough flue gas to dry the incoming biomass to 12 wt%.

Less char is predicted using the new correlations because the new char correlation is based on char collected during the TCPDU experiments, whereas the old correlations are based on experiments that calculated the char yield using carbon balance equations. Char was not actually collected during the experiments that provided data for the old correlations.

The gasifier reaches an equilibrium temperature based on the amount of heat delivered indirectly from the char combustor. Applying the new correlations to the model resulted in an equilibrium gasification temperature of 1605°F (874°C), rather than 1598°F (870°C).

Compared to the old correlations, the raw syngas composition (including tars) predicted using the new correlations is significantly different for several components: more CH₄ and tars are produced, although less H₂ is produced. These changes may initially appear significant, but the differences in gas composition and dry gas yield exiting the gasifier are largely nullified by downstream operations. The additional CH₄ and tars predicted by the new correlation suggest more hydrogen will be bound to carbon and not available as H₂ product, however the reformer converts 80% of CH₄ and 99.9% of tars to CO and H₂. Also, the lower H₂ to CO ratio predicted by the new correlations suggest that the final H₂ product yield will decrease. However, the downstream shift reactors increase the ratio, making the ultimate effect on the final H₂ product yield negligible.

The envisioned facility is designed to be a stand-alone plant requiring very little external power or fuel supplies. Most electricity and steam are generated on-site by combusting byproducts such as char and PSA off gas. Therefore, improvements in dry gas yield or product yield almost inevitably result in less byproducts to power the facility. Consequently, a portion of the intermediate streams must be diverted to generate the required steam and power. In this case, the raw syngas exiting the gasifier was diverted, although other streams such as the syngas leaving the tar cracker or the incoming biomass feed stream are also candidates to supplement steam and power generation. For this reason, the overall efficiency and economics for this stand-alone plant do not change significantly when the new correlations are applied. This may not be the case for

all models. Table 3 compares the overall plant results for the old correlations versus new correlations.

Table 3. Overall Plant Performance Metrics for Old and New Gasifier Correlations

	Old Correlations	New Correlations	% Change
Hydrogen Production	15,322 lb/hr (71.5 MMSCFD)	15,728 lb/hr (73.4 MMSCFD)	+2.6%
Electricity Purchased from Grid	10,285 kW (13,792 HP)	4,477 kW (6,004 HP)	-56.8%
Natural Gas Use	3739 lb/hr	7198 lb/hr	+92.5%
Overall Plant Efficiency	53.3% HHV 47.7% LHV	53.9% HHV 48.5% LHV	+1.1% HHV +1.7% LHV

6 Updated Economic Analysis using New Correlations

The economic analysis consists of first estimating the capital and operating costs, then calculating a minimum hydrogen selling price using a discounted cash flow rate of return analysis. A brief explanation of the economic analysis is provided in this section. A more detailed explanation is provided in the 2005 Biomass to Hydrogen Technical Report (Spath et al. 2005). All capital and operating costs for this economic analysis are based on cost data from the same technical report.

6.1 Capital Costs

The purchased cost of most equipment came from literature sources and Questimate (AspenTech cost estimation software). The installed equipment costs were calculated by multiplying the purchased costs by installation cost factors in Peters and Timmerhaus (Peters and Timmerhaus 2003).

Indirect costs are non-process fixed capital investment costs, such as road and fence construction and legal fees. The indirect costs are calculated by multiplying the total purchased equipment costs by scaling factors in Peters and Timmerhaus (Peters and Timmerhaus 2003).

The sum of the total installed cost and the total indirect cost is the total project investment (TPI).

6.2 Operating Costs

For this economic analysis, both variable and fixed operating costs were considered. Variable operating costs are operating costs that can change when the process inputs or conditions change. Examples of variable operating costs are natural gas and boiler water chemicals. Variable operating costs are estimated on a per unit basis (per kg of feed, per scf of syngas, etc.), and then the total variable cost is calculated by multiplying per unit cost by the total number of units.

Fixed operating costs are costs such as employee salaries, overhead, and maintenance. These costs were assumed equal to the 2005 Biomass to Hydrogen Technical Report (Spath et al. 2005).

The sum of the variable operating costs and the fixed operating costs is the total operating cost (TOC).

6.3 Discounted Cash Flow Rate of Return Analysis (DCFROR)

After the total capital cost and total operating cost have been estimated, a minimum hydrogen selling price (MHSP) can be calculated using a discounted cash flow rate of return analysis. Data from the Aspen Plus simulation are downloaded to a Microsoft Excel workbook that contains capital and operating cost data, as well as economic assumptions. The Excel goal seek function is then used to calculate a MHSP. Table 4 lists several economic assumptions used in the DCFROR analysis.

Table 4. Economic Assumptions Used in DCFROR Analysis

Assumption	Value
Internal rate of return (after-tax)	10%
Debt/equity	0%/100%
Plant life	20 years
General plant depreciation	200% Double Declining Balance
General plant recovery period	7 years
Steam plant depreciation	150% DDB
Steam plant recovery period	20 years
Construction period	2.5 years 8%
1 st 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 an expense in the 1 st construction year)

6.4 New Correlations versus Old Correlations in 2002 Dollars

The results of the model with both new and old correlations were used to calculate a MHSP using the DCFROR analysis spreadsheet. When the new correlations are used, the MHSP remains \$1.24 per kg in 2002 dollars. The new and old correlations were compared in 2002 dollars because the original economic analysis by Spath was done in 2002 dollars. While the final selling price remains unchanged, the operating costs, primarily natural gas and electricity purchased, are slightly different due to differences in intermediate stream flowrates and compositions. The results are presented in Tables 5 and 6.

Table 5. DCFROR Summary Sheet for Old Correlations in 2002 Dollars

Hydrogen Production Process Engineering Analysis

Design Report: Goal Case with Old Correlations in 2002 Dollars
 2000 Dry Metric Tonnes Biomass per Day
 BCL Gasifier, Tar Reformer, Sulfur Removal, HTS & LTS, PSA, Steam-Power Cycle
 All Values in 2002\$

Minimum Hydrogen Selling Price (\$/kg) \$1.24	\$8.66 (\$/GJ H2, HHV basis)
	\$10.34 (\$/GJ H2, LHV basis)
Hydrogen Production at operating capacity (MM kg / year)	58.4
Hydrogen Yield (kg / Dry US Ton Feedstock)	75.7
Delivered Feedstock Cost \$/Dry US Ton	\$30
Internal Rate of Return (After-Tax)	10%
Equity Percent of Total Investment	100%
	70.6 (Million SCF / day)
	2,116 (dry tons / day)
	at operating capacity

Capital Costs		Operating Costs (cents/kg hydrogen)	
Feed Handling & Drying	\$18,900,000	Feedstock	39.7
Gasification, Tar Reforming/Regeneration, & Quench	\$23,800,000	Natural Gas	5.9
Compression & Sulfur Removal	\$16,100,000	Catalysts	0.7
Shift, and PSA	\$16,500,000	Olivine	6.6
Hydrogen Compression	\$2,800,000	Other Raw Materials	0.9
Steam System and Power Generation	\$14,200,000	Waste Disposal	1.2
Cooling Water and Other Utilities	\$3,400,000	Electricity	7.1
Total Installed Equipment Cost	\$95,700,000	Fixed Costs	16.8
		Capital Depreciation	12.3
Indirect Costs	48,800,000	Average Income Tax	9.8
(% of TPI)	33.8%	Average Return on Investment	23.3
Total Project Investment (TPI)	\$144,400,000		
		Operating Costs (\$/yr)	
Loan Rate	N/A	Feedstock	\$23,200,000
Term (years)	N/A	Natural Gas	\$3,400,000
Capital Charge Factor	0.184	Catalysts	\$400,000
		Olivine	\$3,800,000
Maximum Yields (100% of Theoretical) based on composition		Other Raw Matl. Costs	\$500,000
Theoretical Hydrogen Production (MM kg/yr)	119.7	Waste Disposal	\$700,000
Theoretical Yield (kg/dry ton)	155.0	Electricity	\$4,100,000
Current Yield (Actual/Theoretical)	49%	Fixed Costs	\$9,800,000
		Capital Depreciation	\$7,200,000
Gasifier Efficiency - HHV	70.8%	Average Income Tax	\$5,700,000
Gasifier Efficiency - LHV	69.3%	Average Return on Investment	\$13,600,000
Overall Plant Efficiency - HHV	53.0%	Total Plant Electricity Usage (KW)	40259
Overall Plant Efficiency - LHV	47.1%	Electricity Produced Onsite (KW)	-29974
		Electricity Purchased from Grid (KW)	10284
		Plant Electricity Use (KWh/kg H2)	5.79
		Plant Steam Use (kg steam/kg H2)	19.5

Table 6. DCFROR Summary Sheet for New Correlations in 2002 Dollars

Hydrogen Production Process Engineering Analysis

Design Report: Goal Case with New Correlations in 2002 Dollars
 2000 Dry Metric Tonnes Biomass per Day
 BCL Gasifier, Tar Reformer, Sulfur Removal, HTS & LTS, PSA, Steam-Power Cycle
 All Values in 2002\$

Minimum Hydrogen Selling Price (\$/kg) \$1.24	\$8.63 (\$/GJ H2, HHV basis)
	\$10.30 (\$/GJ H2, LHV basis)
Hydrogen Production at operating capacity (MM kg / year)	60.0
Hydrogen Yield (kg / Dry US Ton Feedstock)	77.7
Delivered Feedstock Cost \$/Dry US Ton	\$30
Internal Rate of Return (After-Tax)	10%
Equity Percent of Total Investment	100%
	72.5 (Million SCF / day)
	2,116 (dry tons / day)
	at operating capacity

Capital Costs		Operating Costs (cents/kg hydrogen)	
Feed Handling & Drying	\$18,900,000	Feedstock	38.6
Gasification, Tar Reforming/Regeneration, & Quench	\$23,700,000	Natural Gas	11.0
Compression & Sulfur Removal	\$14,500,000	Catalysts	0.6
Shift, and PSA	\$16,700,000	Olivine	6.3
Hydrogen Compression	\$2,900,000	Other Raw Materials	0.9
Steam System and Power Generation	\$15,100,000	Waste Disposal	1.1
Cooling Water and Other Utilities	\$3,500,000	Electricity	4.5
Total Installed Equipment Cost	\$95,300,000	Fixed Costs	16.3
		Capital Depreciation	12.0
Indirect Costs	48,600,000	Average Income Tax	9.6
(% of TPI)	33.8%	Average Return on Investment	22.6
Total Project Investment (TPI)	\$143,800,000		
		Operating Costs (\$/yr)	
Loan Rate	N/A	Feedstock	\$23,200,000
Term (years)	N/A	Natural Gas	\$6,600,000
Capital Charge Factor	0.184	Catalysts	\$400,000
		Olivine	\$3,800,000
Maximum Yields (100% of Theoretical) based on composition		Other Raw Matl. Costs	\$500,000
Theoretical Hydrogen Production (MM kg/yr)	119.7	Waste Disposal	\$700,000
Theoretical Yield (kg/dry ton)	155.0	Electricity	\$2,700,000
Current Yield (Actual/Theoretical)	50%	Fixed Costs	\$9,800,000
		Capital Depreciation	\$7,200,000
Gasifier Efficiency - HHV	70.8%	Average Income Tax	\$5,700,000
Gasifier Efficiency - LHV	69.3%	Average Return on Investment	\$13,600,000
Overall Plant Efficiency - HHV	53.0%		
Overall Plant Efficiency - LHV	47.1%	Total Plant Electricity Usage (KW)	38952
		Electricity Produced Onsite (KW)	-32158
		Electricity Purchased from Grid (KW)	6794
		Plant Electricity Use (KWh/kg H2)	5.46
		Plant Steam Use (kg steam/kg H2)	19.8

7 Updated Economics

7.1 Economic Results Using 2005 Biomass Program Assumptions

In 2005, the biomass program updated several assumptions to make the DCFROR analysis more current, such as increasing the feedstock cost from \$30 to \$35 per dry U.S. ton and using 2005 dollars. The results of the updated DCFROR analysis with 2005 assumptions are presented in Table 7.

Table 7. DCFROR Summary Sheet for New Correlations in 2005 Dollars

Hydrogen Production Process Engineering Analysis			
Design Report: Goal Case with New Correlations in 2005 Dollars 2000 Dry Metric Tonnes Biomass per Day BCL Gasifier, Tar Reformer, Sulfur Removal, HTS & LTS, PSA, Steam-Power Cycle All Values in 2005\$			
Minimum Hydrogen Selling Price (\$/kg) \$1.45		\$10.14	(\$/GJ H ₂ , HHV basis)
		\$12.11	(\$/GJ H ₂ , LHV basis)
Hydrogen Production at operating capacity (MM kg / year)		60.0	72.5 (Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)		77.7	2,116 (dry tons / day)
Delivered Feedstock Cost \$/Dry US Ton		\$35	at operating capacity
Internal Rate of Return (After-Tax)		10%	
Equity Percent of Total Investment		100%	
Capital Costs		Operating Costs (cents/kg hydrogen)	
Feed Handling & Drying	\$22,300,000	Feedstock	55.6
Gasification, Tar Reforming/Regeneration, & Quench	\$27,900,000	Natural Gas	12.7
Compression & Sulfur Removal	\$17,100,000	Catalysts	0.6
Shift, and PSA	\$19,700,000	Olivine	7.0
Hydrogen Compression	\$3,400,000	Other Raw Materials	1.1
Steam System and Power Generation	\$17,800,000	Waste Disposal	1.3
Cooling Water and Other Utilities	\$4,100,000	Electricity	5.6
Total Installed Equipment Cost	\$112,300,000	Fixed Costs	16.7
Indirect Costs	48,200,000	Capital Depreciation	11.8
(% of TPI)	33.8%	Average Income Tax	9.8
		Average Return on Investment	22.8
Total Project Investment (TPI)	\$142,800,000	Operating Costs (\$/yr)	
Loan Rate	N/A	Feedstock	\$33,300,000
Term (years)	N/A	Natural Gas	\$7,600,000
Capital Charge Factor	0.187	Catalysts	\$400,000
		Olivine	\$4,200,000
Maximum Yields (100% of Theoretical) based on composition		Other Raw Matl. Costs	\$700,000
Theoretical Hydrogen Production (MM kg/yr)	119.7	Waste Disposal	\$800,000
Theoretical Yield (kg/dry ton)	155.0	Electricity	\$3,400,000
Current Yield (Actual/Theoretical)	50%	Fixed Costs	\$10,000,000
		Capital Depreciation	\$7,100,000
Gasifier Efficiency - HHV	70.8%	Average Income Tax	\$5,900,000
Gasifier Efficiency - LHV	69.3%	Average Return on Investment	\$13,700,000
Overall Plant Efficiency - HHV	53.0%	Total Plant Electricity Usage (KW)	38952
Overall Plant Efficiency - LHV	47.1%	Electricity Produced Onsite (KW)	-32158
		Electricity Purchased from Grid (KW)	6794
		Plant Electricity Use (KWh/kg H ₂)	5.46
		Plant Steam Use (kg steam/kg H ₂)	19.8

When 2005 economic assumptions are used, the MHSP increases from \$1.24 to \$1.45 per kg. As expected, the new feedstock cost has a significant effect on the operating costs. According to sensitivity studies presented in the Biomass to Hydrogen Technical Report (Spath et al. 2005), the MHSP is more sensitive to feedstock cost than any other cost variable or process variable studied, therefore this is a reasonable and expected result.

Updating to 2005 dollars requires updating three cost indices, a plant cost index, an industrial inorganic chemical cost index, and a labor index. The plant cost index impacts capital costs, whereas the industrial inorganic chemical cost index and labor index impact operating costs.

Between 2002 and 2005, all three indices increased, causing an increase in both capital and operating costs.

7.2 Economic Results Using 2007 Biomass Program Assumptions

To make the DCFROR analysis consistent with 2007 Biomass Program assumptions, the feedstock price was increased to \$60 per dry U.S. ton, and the cost indices were updated to 2007 dollars. The results of the DCFROR analysis using the new correlations and 2007 Biomass Program assumptions are presented in Table 8.

Table 8. DCFROR Summary Sheet for New Correlations in 2007 Dollars

Hydrogen Production Process Engineering Analysis			
Design Report: Goal Case with New Correlations in 2007 Dollars			
2000 Dry Metric Tonnes Biomass per Day			
BCL Gasifier, Tar Reformer, Sulfur Removal, HTS & LTS, PSA, Steam-Power Cycle			
All Values in 2007\$			
Minimum Hydrogen Selling Price (\$/kg) \$2.14		\$14.91	(\$/GJ H ₂ , HHV basis)
		\$17.81	(\$/GJ H ₂ , LHV basis)
Hydrogen Production at operating capacity (MM kg / year)		60.0	72.5 (Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)		77.7	2,116 (dry tons / day)
Delivered Feedstock Cost \$/Dry US Ton		\$60	at operating capacity
Internal Rate of Return (After-Tax)		10%	
Equity Percent of Total Investment		100%	
Capital Costs		Operating Costs (cents/kg hydrogen)	
Feed Handling & Drying	\$25,100,000	Feedstock	103.4
Gasification, Tar Reforming/Regeneration, & Quench	\$31,400,000	Natural Gas	13.8
Compression & Sulfur Removal	\$19,200,000	Catalysts	0.6
Shift, and PSA	\$22,200,000	Olivine	7.6
Hydrogen Compression	\$3,800,000	Other Raw Materials	1.2
Steam System and Power Generation	\$20,100,000	Waste Disposal	1.5
Cooling Water and Other Utilities	\$4,700,000	Electricity	6.1
Total Installed Equipment Cost	\$126,500,000	Fixed Costs	20.1
		Capital Depreciation	16.0
Indirect Costs	64,500,000	Average Income Tax	12.8
(% of TPI)	33.8%	Average Return on Investment	30.5
Total Project Investment (TPI)	\$191,000,000	Operating Costs (\$/yr)	
		Feedstock	\$62,000,000
Loan Rate	N/A	Natural Gas	\$8,300,000
Term (years)	N/A	Catalysts	\$400,000
Capital Charge Factor	0.186	Olivine	\$4,600,000
		Other Raw Matl. Costs	\$700,000
Maximum Yields (100% of Theoretical) based on composition		Waste Disposal	\$900,000
Theoretical Hydrogen Production (MM kg/yr)	119.7	Electricity	\$3,600,000
Theoretical Yield (kg/dry ton)	155.0	Fixed Costs	\$12,100,000
Current Yield (Actual/Theoretical)	50%	Capital Depreciation	\$9,600,000
		Average Income Tax	\$7,700,000
Gasifier Efficiency - HHV	70.8%	Average Return on Investment	\$18,300,000
Gasifier Efficiency - LHV	69.3%	Total Plant Electricity Usage (KW)	38952
Overall Plant Efficiency - HHV	53.0%	Electricity Produced Onsite (KW)	-32158
Overall Plant Efficiency - LHV	47.1%	Electricity Purchased from Grid (KW)	6794
		Plant Electricity Use (KWh/kg H ₂)	5.46
		Plant Steam Use (kg steam/kg H ₂)	19.8

Between 2005 and 2007, the plant cost index and industrial inorganic chemical index both increased sharply, while the labor index increased less sharply. The final effect was an increase in both capital and operating costs, as expected, although the increase in feedstock cost from \$35 to \$60 per dry U.S. ton has the most significant impact. When 2005 assumptions are replaced with 2007 assumptions, the MHSP increases from \$1.45 to \$2.14 per kg. Most of the increase in MHSP can be attributed to increased feed cost.

7.3 Economic Results using H2A Analysis

The Department of Energy (DOE) Hydrogen Program uses a separate economic evaluation spreadsheet, called H2A, available on the DOE H2A Analysis website. The Hydrogen Program H2A Cash Flow Analysis uses slightly different economic assumptions, such as plant capacity factor and life of project, than the Biomass Program DCFROR Analysis, therefore the results of the two analyses will vary slightly.

The H2A analysis resulted in a hydrogen cost of \$1.56 per kg in 2005 dollars using the new correlations versus \$1.52 using the old correlations. The major process and cost data used in the H2A analysis is summarized in Table 9.

Table 9. Summary of H2A Results

	Old Correlations	New Correlations
Biomass Usage, kg dry biomass per kg H ₂	11.99	11.68
Electricity Usage, kWh per kg H ₂	1.48	0.95
Natural Gas Usage, Nm ³ per kg H ₂	0.38	0.63
Cooling Water Usage, gal per kg H ₂	2.37	2.46
Process Water Usage, gal per kg H ₂	1.60	2.23
Total Capital Costs, 2005 Dollars	\$155.3 million	\$156.0 million
Annual Fixed Operating Costs, 2005 Dollars	\$10.1 million	\$10.2 million
Annual Variable Operating Costs, 2005 Dollars	\$43.7 million	\$47.5 million
Selling Price, 2005 Dollars per kg H ₂	\$1.52	\$1.56

7.4 Summary of Economics

The results of the DCFROR analyses are presented in Table 10.

Table 10. Summary of DCFROR Analyses

	Hydrogen Program H2A Analysis		Biomass Program DCFROR Analysis	
	Old Correlations	New Correlations	Old Correlations	New Correlations
2002 Dollars			\$1.24	\$1.24
2005 Dollars	\$1.52	\$1.56	\$1.47	\$1.45
2007 Dollars				\$2.14

8 Conclusions

The new correlations produce significantly different results for the gasifier, but only slightly different results for the overall process. Table 11 summarizes the differences.

Table 11. Summary of Conclusions

Gasifier or Process Variable	Old Correlations	New Correlations
Gasifier Dry Gas Yield, lbmol of dry gas/lb of dry feed	0.035	0.036
Gasifier H ₂ :CO molar ratio	0.57	0.38
Gasifier Gas Heating Value Btu/lb	Wet: 4,759 HHV 4,401 LHV Dry: 8,019 HHV 7,416 LHV	Wet: 4,345 HHV 4028 LHV Dry: 6,937 HHV 6,430 LHV
Gasifier Char Yield, lb/lb of dry feed	0.22	0.10
Gasifier Efficiency	72.1% HHV 71.8% LHV	71.7% HHV 71.5% LHV
Overall Process Hydrogen Production	15,322 lb/hr (71.5 MMSCFD)	15,728 lb/hr (73.4 MMSCFD)
Electricity Purchased from Grid	10,285 kW (13,792 HP)	4,477 kW (6,004 HP)
Natural Gas Usage	3739 lb/hr	7198 lb/hr
Overall Plant Efficiency	53.3% HHV 47.7% LHV	53.9% HHV 48.5% LHV
MHSP, 2002 Dollars and Assumptions	\$1.24	\$1.24
MHSP, 2007 Dollars and Assumptions	N/A	\$2.14

9 References

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Appendix A. Correlation Values for Equations 1, 2 and 3

Component Statistic		C	H	N	O	S	TC	SB	RT	TC:SB	TC:RT	SB:RT	TC ²	SB ²	RT ²	R ²	
1-Butene	B0	2.194															
C4H8	B	0.004	0.016	-0.036	-0.015	0.956	-0.002	-0.232	0.147	0.022	-0.013	0.022	-0.048	-0.024	0.005	0.880	
	Center						750.000	0.630	1.488								
	Weight						0.010	3.614	4.610								
2-c-Butene	B0	0.250															
C4H8	B	0.000	0.013	0.065	-0.005	-0.740	0.000	0.000	-0.004	0.003	-0.002	-0.002	0.015	0.001	0.002	0.712	
	Center						745.944	0.632	1.494								
	Weight						0.010	3.577	4.391								
2-t-Butene	B0	0.354															
C4H8	B	-0.002	-0.036	-0.075	0.007	1.296	0.000	-0.033	-0.014	0.005	0.000	0.002	0.022	-0.006	0.006	0.713	
	Center						748.861	0.618	1.489								
	Weight						0.009	3.658	4.308								
Carbon Dioxide	B0	38.897															
CO2	B	0.077	1.525	5.268	-0.416	-73.339	-0.024	-2.212	5.363	0.529	-0.563	0.023	0.961	0.198	-0.218	0.812	
	Center						745.989	0.636	1.498								
	Weight						0.010	3.562	4.362								
Carbon Monoxide	B0	18.479															
CO	B	-0.091	-2.750	-11.893	0.724	160.964	0.002	-4.578	-0.022	0.063	0.417	0.297	-0.258	0.592	-0.052	0.689	
	Center						745.989	0.636	1.498								
	Weight						0.010	3.562	4.362								
Ethane	B0	0.293															
C2H6	B	-0.001	0.002	0.102	-0.003	-0.991	0.000	-0.027	0.076	0.011	-0.015	-0.003	0.006	0.000	-0.002	0.716	
	Center						748.607	0.628	1.488								
	Weight						0.010	3.592	4.514								
Ethene	B0	-2.721															
C2H4, Ethylene	B	-0.016	-0.303	-0.585	0.027	12.660	0.010	-0.411	-0.043	-0.217	0.180	0.106	-0.179	0.050	-0.074	0.958	
	Center						745.989	0.636	1.498								
	Weight						0.010	3.562	4.362								
Ethyne	B0	7.390															
C2H2, Acetylene	B	-0.003	-0.037	0.057	-0.005	0.572	-0.007	-0.587	0.426	0.099	-0.046	0.032	-0.240	-0.002	0.018	0.964	
	Center						745.989	0.636	1.498								
	Weight						0.010	3.562	4.362								
Helium	B0	3.892															
He	B	0.001	-0.059	0.025	-0.041	3.704	-0.002	0.452	0.348	0.050	-0.073	-0.068	0.084	-0.157	0.038	0.846	
	Center						745.989	0.636	1.498								
	Weight						0.010	3.562	4.362								

Component Statistic		C	H	N	O	S	TC	SB	RT	B _{TC:SB}	B _{TC:RT}	B _{SB:RT}	B _{TC} ²	B _{SB} ²	B _{RT} ²	R ²	
Hydrogen	B0	47.139															
H2	B	-0.030	1.559	10.667	-0.352	-139.271	0.005	20.304	-19.415	-1.294	0.719	1.057	0.206	-0.238	-1.311	0.780	
	Center						745.989	0.636	1.498								
	Weight						0.010	3.562	4.362								
Methane	B0	0.087															
CH4	B	0.043	0.030	-2.272	0.011	24.372	0.012	-4.429	3.565	-0.374	0.440	0.654	-0.093	0.252	-0.346	0.846	
	Center						745.989	0.636	1.498								
	Weight						0.010	3.562	4.362								
Propane	B0	-1.671															
C3H8	B	0.001	-0.018	-0.126	0.002	1.784	0.002	0.018	-0.005	-0.029	0.013	0.008	0.116	0.012	-0.016	0.901	
	Center						745.989	0.636	1.498								
	Weight						0.010	3.562	4.362								
Propene	B0	4.045															
C3H6	B	-0.002	-0.047	-0.116	0.002	2.400	-0.004	-0.364	0.247	0.054	-0.014	0.035	-0.150	-0.002	0.008	0.954	
	Center						745.989	0.636	1.498								
	Weight						0.010	3.562	4.362								
Carbonyl Sulfide	B0	2.784															
COS	B	0.000	-0.026	-0.005	-0.013	1.522	-0.001	-0.158	-0.863	-0.003	-0.019	0.042	-0.151	0.120	0.118	0.497	
	Center						750.424	0.588	1.468								
	Weight						0.010	3.688	4.712								
Hydrogen Sulfide	B0	6.159															
H2S	B	-0.036	-0.892	0.702	-0.294	34.016	0.019	4.599	-4.983	0.091	0.529	-0.254	0.424	-0.040	0.443	0.850	
	Center						742.692	0.618	1.472								
	Weight						0.010	3.633	4.372								
benzene	B0																
C6H6	B	-6.09E-05	-3.60E-04	1.08E-03	1.03E-04	-4.52E-03	6.22E-05	4.34E-03	-1.62E-03	-3.77E-04	-4.56E-05	-6.14E-04	2.33E-03	3.03E-04	-4.59E-04	0.971	
	Center						748.214	0.636	1.487								
	Weight						0.010	3.545	4.257								
toluene	B0	1.60E-03															
C7H8	B	-3.13E-05	-2.51E-04	5.60E-04	2.22E-05	8.76E-04	7.08E-06	1.38E-03	-4.72E-04	-1.15E-04	5.59E-06	-2.41E-04	-3.84E-04	7.32E-05	-7.27E-05	0.833	
	Center						755.682	0.642	1.497								
	Weight						0.010	3.551	4.569								
phenol	B0	1.96E-02															
C6H6O	B	-1.12E-05	-1.59E-04	2.08E-04	-2.55E-05	3.97E-03	-1.65E-05	-6.91E-05	8.21E-04	-5.77E-06	-1.17E-04	4.36E-06	-1.33E-03	-8.73E-05	5.64E-05	0.932	
	Center						748.214	0.636	1.487								
	Weight						0.010	3.545	4.257								

Component Statistic		C	H	N	O	S	TC	SB	RT	B _{TC:SB}	B _{TC:RT}	B _{SB:RT}	B _{TC} ²	B _{SB} ²	B _{RT} ²	R ²	
cresols	B0	1.84E-02															
C7H8O	B	-6.76E-06	-1.32E-05	1.17E-04	2.41E-05	-1.88E-03	-1.93E-05	-1.00E-03	2.68E-04	-7.04E-06	-1.02E-04	1.49E-04	-1.02E-03	-1.97E-04	1.70E-04	0.942	
	Center						748.214	0.636	1.487								
	Weight						0.010	3.545	4.257								
naphthalene	B0	-1.33E-02															
C10H8	B	-2.72E-05	-2.04E-04	4.85E-04	2.40E-05	2.91E-04	2.24E-05	9.26E-04	-4.87E-04	-1.85E-04	3.32E-05	-7.65E-05	1.18E-03	1.20E-04	-1.29E-04	0.978	
	Center						746.377	0.631	1.482								
	Weight						0.010	3.563	4.300								
"other tar"	B0	3.81E-02															
Sum of toluene through phenanthrene	B	-6.12E-05	-4.77E-04	1.10E-03	4.73E-05	1.55E-03	-3.26E-05	-3.14E-04	1.53E-03	3.23E-04	-2.66E-04	-2.83E-04	8.53E-04	-2.72E-04	9.14E-05	0.939	
	Center						750.368	0.629	1.482								
	Weight						0.010	3.541	4.321								
phenanthrene	B0	-4.60E-03															
C14H10	B	-7.58E-06	-5.88E-05	1.36E-04	5.98E-06	1.55E-04	7.46E-06	1.87E-04	-1.54E-04	-6.12E-05	1.41E-05	-6.67E-06	4.93E-04	3.02E-05	-4.06E-05	0.976	
	Center						744.485	0.625	1.484								
	Weight						0.010	3.583	4.281								
"heavy tar"	B0	6.10E-03															
Sum of everything	B	3.34E-05	2.51E-04	-5.97E-04	-2.64E-05	-4.95E-04	2.27E-06	-1.20E-03	1.66E-03	-4.64E-04	3.82E-04	-1.02E-04	1.30E-03	-2.39E-04	-1.47E-04	0.549	
With MW > 180	Center						745.833	0.618	1.470								
	Weight						0.010	3.584	4.460								
Total (>78)	B0	6.53E-02															
Sum of everything except benzene	B	-1.07E-04	-8.92E-04	1.92E-03	5.50E-05	5.46E-03	-2.87E-05	3.06E-04	2.58E-03	-2.01E-04	-2.72E-04	-5.94E-04	1.01E-03	-4.72E-04	-9.48E-05	0.767	
	Center						746.591	0.618	1.479								
	Weight						0.010	3.584	4.350								
Char	B0	4.47E-01															
	B	7.01E-04	-8.88E-03	-1.11E-02	-7.38E-03	7.23E-01	-5.78E-05	-1.22E-02	2.18E-02	-2.84E-03	-3.74E-03	-2.84E-03	-5.23E-04	-1.12E-03	-3.63E-03	0.740	
	Center						748.214	0.636	1.487								
	Weight						0.010	3.545	4.257								
Nitrogen Free	B0	52.870															
Outlet Dry Gas	B	0.030	0.483	-0.542	0.104	-12.498	0.006	5.017	-30.121	0.012	-0.163	-1.432	0.232	-2.177	2.235	0.981	
Flow Rate	Center						744.080	0.619	1.486								
	Weight						0.010	3.627	4.398								

REPORT DOCUMENTATION PAGE

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14. ABSTRACT (Maximum 200 Words) NREL developed a new set of empirical gasifier correlations using data collected from the Thermochemical Process Development Unit. The new correlations replaced the old correlations in the Aspen Plus model presented in the Biomass to Hydrogen Design Report (Spath et al. 2005). The new correlations predict a slightly different dry gas composition, although the most significant difference is less char. The new correlations predict 0.10 lb of char per lb of dry feed, whereas the old correlations predict 0.22 lb of char; consequently, a portion of the raw syngas must be diverted to the char combustor to provide the heat necessary for gasification and drying the incoming biomass. The model predicts 2.6% more final hydrogen product when the new gasifier correlations are used. The new correlations also predict a minimum hydrogen selling price of \$1.24 per kg (2002 dollars), which equal to the \$1.24 per kg predicted by the old correlations. Using 2007 Biomass Program economic assumptions, the new correlations predict a minimum hydrogen selling price of \$2.14 per kg (2007 dollars).						
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