Biomass Gasification for Chemicals Production Using Chemical Looping Techniques

FINAL TECHNICAL REPORT

AWARDED UNDER
DOE/EERE **DE-EE0007530**

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SUBMITTED TO

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Disclaimer

Executive Summary

The Ohio State University (OSU) is investigating the Biomass to Syngas (BTS) chemical looping technology to produce syngas for chemical production applications from biomass under US Department of Energy (DOE) Award #DE-EE0007530. The BTS process aligns with the programmatic area of interest of "Conversion, via biological, thermal, catalytic or chemical means, of acceptable feedstocks into advanced biofuels and/or biobased products including intermediate and end-use products". Compared to conventional biomass gasification processes, the BTS process eliminates the need for air separation units and tar reforming reactors, which leads to energy efficiency improvement and capital cost reduction. The overall objective is to ascertain the potential of biomass gasification based on the chemical looping technique through mitigation of the possible techno-economic challenges in the steps of scale up for commercialization.

The scope of work consists of 1) designing, constructing and operating a 10 kWth commercially scalable sub-pilot BTS system and; 2) completing a comprehensive techno-economic analysis (TEA) of the BTS process using methanol production as an example.

Over the course of the project, the project team completed the design, fabrication, and operation of a 10 kW_{th} sub-pilot scale test unit for the BTS process. Corn cob and wood pellets were successfully tested in the unit for high purity syngas generation in extended test campaigns that totals over 200 hours. Syngas purity (H_2 and CO) of >70% was achieved with a CH_4 concentration of <6%. The H_2/CO ratio was greater than 1.8.

A comprehensive techno-economic analysis was performed to compare the BTS process and a reference indirectly heated gasification process for methanol synthesis. The result, updated with experimental results for BTS process performance, shows a methanol required selling price (MSP) of \$1.15/gal, compared to \$1.28/gal for the reference case.

1. Background

Biomass is a domestically abundant renewable fuel source that can increase the U.S. energy independence. Biomass gasification is highly versatile thermochemical approach that can utilize biomass waste material to produce syngas as a building block for chemicals and liquid fuel production. However, the significant capital cost investment required for the gasification unit, tar reformer, and syngas condition systems prevents this process from being an economic alternative to fossil fuel based approaches. The Biomass-To-Syngas BTS process can effectively convert biomass into high quality syngas with high H_2:CO ratio, which is an important and widely accepted intermediate product for a variety of downstream chemical processes including methanol production and Fischer-Tropsch synthesis for fuel production. The BTS process eliminates the need for air separation unit and tar reforming resulting in a cost reduction for syngas production.

The proposed work aligns with the area of interest identified by BRDI as: “Conversion, via biological, thermal, catalytic or chemical means, of acceptable feedstocks into advanced biofuels and/or biobased products including intermediate and end-use products.” The BTS process is developed based on the innovative OSU chemical looping technology platform.

In the BTS process, as shown in Figure 1, the iron-titanium composite metal oxide (ITCMO) is used to convert biomass to syngas in a single co-current moving bed reactor, called the reducer. Biomass is pyrolyzed to release tars and volatiles, which are partially oxidized by the ITCMO oxygen carrier to form syngas. The devolatilized char or fixed carbon in the reducer is also gasified to syngas with steam and CO_2 where the ITCMO oxygen carrier serves as a catalyst for char gasification. Steam is introduced in the reducer to also tune the composition of syngas produced. The reduced ITCMO oxygen carrier exiting the reducer is regenerated in the oxidizer with air and is recycled to the reducer to complete the loop. The main reactions involved in the process are:

Reducer: $CH_xO_y + Fe_2O_3 + TiO_2 + H_2O \rightarrow CO + H_2 + FeTiO_3$

Oxidizer: $FeTiO_3 + O_2(Air) \rightarrow Fe_2O_3 + TiO_2$

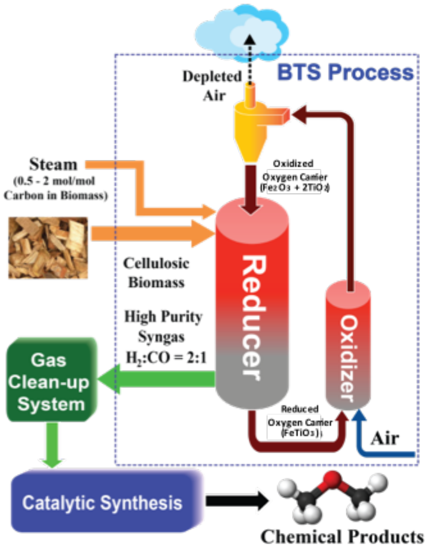


Figure 1 Concept of the BTS Process

The use of the oxygen carrier eliminates the need for an air separation unit (ASU) to produce molecular oxygen, significantly reducing the capital and operating costs for syngas generation from biomass. Also, the co-current moving bed reducer coupled with the highly reactive oxygen carrier produces high purity syngas, thus eliminating the need for a tar reformer and steam reformer as compared to the conventional biomass gasification processes. As a result, the BTS process reduces the cost for syngas generation from biomass.

The overall objective of this proposed project is to ascertain the potential of biomass gasification based on the chemical looping technique through mitigation of the possible techno-economic challenges in the steps of scale up for commercialization. The specific goals that the project seeks to achieve are to 1) design, construct and operate a 10 kW$_{th}$ commercially scalable sub-pilot BTS system; 2) complete a comprehensive techno-economic analysis (TEA) of the BTS process using methanol production as an example.

2. BTS Reactor Preliminary Design

One of the objectives of this project is to design and construct an integrated BTS system at sub-pilot scale with biomass as its feedstock.

As shown in Figure 2, the BTS sub-pilot reactor system comprises two reactors, i.e. a reducer and a combustor. The reactors are connected using non-mechanical gas sealing devices and a gas-solid separator. In addition, a receiver is located between the gas-solid separator and the reducer, serving as a particle reservoir. The reducer is a co-current moving bed reactor, in which solid and gas both travel downwards. The combustor is a fluidized bed reactor, connected with a pneumatic riser on top for solid transportation. Each reactor is heated externally with electric heaters to maintain isothermal operating temperatures.

During operation, biomass and steam are introduced from the top of the co-current reducer, and converted to syngas by oxygen carrier particles. Syngas is separated from the particles at the bottom of the reducer in the product gas separator. The oxygen carrier particles from the reducer are regenerated by air in the combustor and transported back to the receiver.

The detailed sizing of the reactors and interconnecting gas sealing device were finalized based on hydrodynamic calculations. To perform the calculation, a BTS reducer operation condition for full capacity was identified, as shown in Exhibit 3. The operation condition was chosen to meet the following criteria: (1) process a wood pellet biomass feedstock at 10 kW_{th} capacity; (2) produce a syngas with H_2:CO ratio; (3) maintain neutral heat of reaction. Note that due to the small scale of the sub-pilot unit and the large heat loss associated with it, this reactor will not be operated autothermally. Instead, the operation condition was chosen to give a net heat of reaction of zero, which was determined by heat balance calculation performed by an Aspen performance model of the BTS reactor system. The expected syngas composition was also determined by the Aspen performance model. Hydrodynamic calculation was performed on the given condition to ensure that the sub-pilot reactor system can be operated.

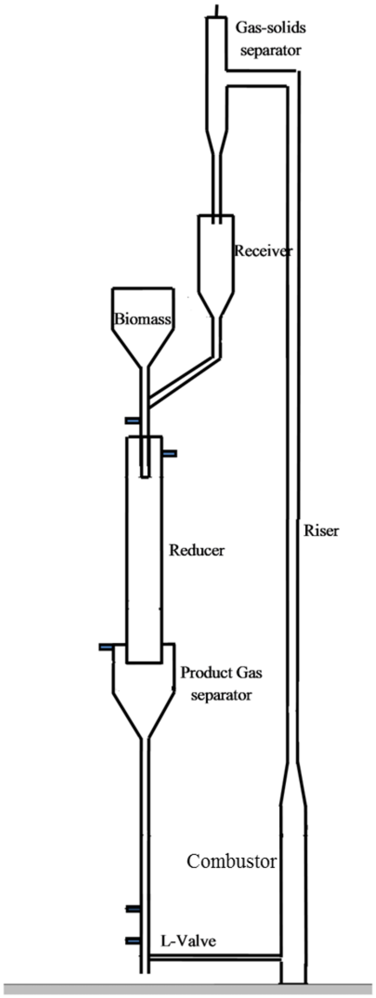


Figure 2. BTS sub-pilot reactor system

Table 1 Full capacity operating condition of BTS reducer

Parameter	Value
Operating Temperature	900ºC
Operating Pressure	1 atm
Feedstock	
Fuel	Wood Pellet
Fuel Higher Heating Value	8425 BTU/lb (19554 kJ/kg)
Fuel Flow Rate	2.76 kg/hr
Oxygen Carrier	Iron-titanium composite metal oxide
Oxygen Carrier Flow Rate	23.81 kg/hr
Steam Flow Rate	4.03 kg/hr
Product Syngas	
CO	52.64 mol/hr
	12.49%
H_2	108.66 mol/hr
	25.78%
CO_2	59.43 mol/hr
	14.1%
H_2O	200.80 mol/hr
	25.78%
Reduced Oxygen Carrier	
Conversion	33%
Oxidation state	Fe(II)
Oxygen Transferred	1.19 kg/hr

2.1 Sizing of reducer

The dimensions of the co-current reducer were restricted by the following considerations:

(1) The volume of the reactor should be large enough to provide sufficient residence time for both solid and gaseous species:

$$V_r \cdot (1 - \epsilon) \geq F_s \cdot \tau_s$$

$$V_r \cdot \epsilon \geq F_g \cdot \tau_g$$

where V_r is the volume of the reactor; ϵ is the voidage of the reactor; F_s is the volumetric flow rate of oxygen carrier particles; τ_s is the required residence time for oxygen carrier particles; F_g is the volumetric flow rate of gases; τ_g is the required residence time for gases.

(2) The cross-sectional area of the reactor should be large enough to avoid a high gas velocity:

$$\frac{F_g}{S_r} \leq 0.8 \cdot u_{mf}$$

where S_r is the cross-sectional area of the reducer; u_{mf} is the minimum fluidization velocity of the oxygen carrier particles. u_{mf} in the reducer was estimated to be 0.91 m/s at 900°C and 1 atm.

(3) Because the reactor is heated externally, the diameter of the reactor should be small enough to ensure even temperature distribution along the radial direction. The maximum diameter was chosen to be 6 inches.

(4) The height of the reactor should be minimized due to the height limit of the facility.

Based on the criteria listed above, it was determined that the reducer reactor should be a cylindrical column with diameter of 6 inches and height of 75 inches. Correspondingly, the solid and gas residence times in the reactor is 87 mins and 2.2 seconds, respectively. The pressure drop across the reducer is less than 73 inches of water during normal operation.

2.2 Sizing of product gas separator

Syngas produced in the reducer must be separated from the oxygen carrier particles. Thus, a gas-solid separator was used at the bottom of the reducer. The separator is a cylindrical column with a diameter larger than the reducer. The separator was sized such that the gas velocity in the annular region between the separator and the reducer is well below than the minimum fluidization velocity of the oxygen carrier particles. Thereby, syngas can be extracted from the gas outlet on top of the separator without entraining any oxygen carrier particles. The diameter of the separator was determined to be 10 inches.

2.3 Sizing of combustor and riser

The combustor reactor is a fluidized bed reactor where oxygen carrier particles are regenerated via the reaction with oxygen from air. Air introduced into the combustor serves three purposes: (a) provide oxygen for oxygen carrier regeneration; (b) fluidize the oxygen carrier particles in the combustor; (c) entrain the oxygen carrier particles in the riser back to the moving bed reactor. Therefore, the sizing of the combustor and riser was restricted by the following considerations:

(1) At the designed air flow rate, oxygen flow must be greater than the amount required by oxygen carrier regeneration (1.19 kg/hr);

(2) At the designed air flow rate, the gas velocity in the combustor should satisfy:

$$u_{mf} < \frac{F_{g,c}}{S_c} < u_t$$

where u_{mf} is the minimum fluidization velocity of oxygen carrier particles; $F_{g,c}$ is the volumetric flow rate of air introduced into the combustor; S_c is the cross-sectional area of the combustor; u_t is the terminal velocity of the oxygen carrier particles. Under 900 °C and 1 atm, u_{mf} and u_t in the combustor were estimated to be 0.57 m/s and 12.46 m/s, respectively.

(3) At the designed air flow rate, the gas velocity in the riser should be greater than the terminal velocity:

$$\frac{F_{g,c}}{S_{ri}} > u_t$$

where S_{ri} is the cross-sectional area of the riser. The terminal velocity in the riser is the same as that in the combustor, which is 12.46 m/s. Note that the mass flows of gas flowing through the combustor and the riser are identical. Thus, the difference in gas velocity is a result of the different cross section areas of the two parts.

(4) The volume of the combustor should be large enough to provide sufficient residence time for the oxygen carrier particles:

$$V_c \cdot (1 - \epsilon_c) \geq F_s \cdot \tau_c$$

where V_c is the volume of the combustor; ϵ_c is the voidage of the fluidized bed at the designed air flow rate; τ_c is the required residence time for the oxygen carrier particles.

Based on the criteria listed above, it was determined that the combustor reactor should be a cylindrical column with diameter of 6 inches and height of 20 inches. The pressure drop across the combustor is about 20 inches of water during normal operation. The riser diameter was determined to be 2 inches. The pressure drop across the riser is less than 5 inches of water.

2.4 Sizing of stand pipes

The stand pipes in the CLG system are used to interconnect different reactors. The stand pipes must allow solid to flow through while keeping the gas from the two ends segregated. To achieve desirable gas sealing, a small amount of nitrogen is introduced to the middle of every stand pipe. The nitrogen stream splits into two directions and flow towards both ends of the stand pipe, keeping the gas from either end from flowing into the stand pipe. The nitrogen flow in the stand pipe also creates a pressure drop, which is used to balance the pressure difference created by the reactors. The sizing of the stand pipes was restricted by the following considerations:

(1) The diameter should be large enough to avoid bridging of oxygen carrier particles, which stops normal solid flow in the stand pipes. The minimum stand pipe diameter is 1.5 inches.

(2) The gas velocity in the stand pipes should be lower than the minimum fluidization velocity of the oxygen carrier particles.

(3) The stand pipes should be long enough to provide sufficient pressure drop across the stand pipes in order to balance the system pressure.

(4) The diameter should be minimized to minimize the amount of nitrogen required.

Based on the considerations above, the diameter of the stand pipes were chosen to be 1.5 inches. The top zone seal was determined to be 24 inches long, while the bottom zone seal was sized to be 34 inches long.

3. Cold Flow Model Studies

A cold flow model (CFM) was fabricated and installed a supporting structure along with the associated instrumentation. Cylindrical acrylic columns of different sizes were procured to fabricate different parts of CFM such as reducer, combustor, riser, zone seals and gas-solid separators. The lengths and diameters of these columns were determined based on the CFM sizing calculations. After the parts were fabricated, the CFM unit was assembled. The completed and assembled unit is shown in Figure 3.

Figure 3 BTS Cold Flow Model Unit

Air is used as the only gas source into the SFM to simulate pressure balance conditions. A total of four mass flow controllers (MFC) procured from Alicat Scientific Inc. were installed to precisely control air flow rates into the CFM. The MFCs share a common air source and provide the air for the entrainment gas, aeration gas, top zone seal, and reducer gases. Each MFC is connected and controlled through the supervisory control and data acquisition (SCADA) software DAQFactory. The gas from the outlets of gas solid separators is sent to a baghouse via metallic hose.

To measure the pressure differentials and absolute pressures along various points of the CFM, a total of nine differential pressure transducers and two pressure transducers were procured from OMEGA Engineering. A diagram depicting the location of each differential and absolute pressure being measured is contained below in Figure 4. Each pressure transducer reads from designated ports in the CFM and sends the readings to DAQFactory for operator use and data logging. As can be seen, the pressure differentials cover the entire height of the unit on both sides granting complete knowledge on the pressure profile of the unit at all times.

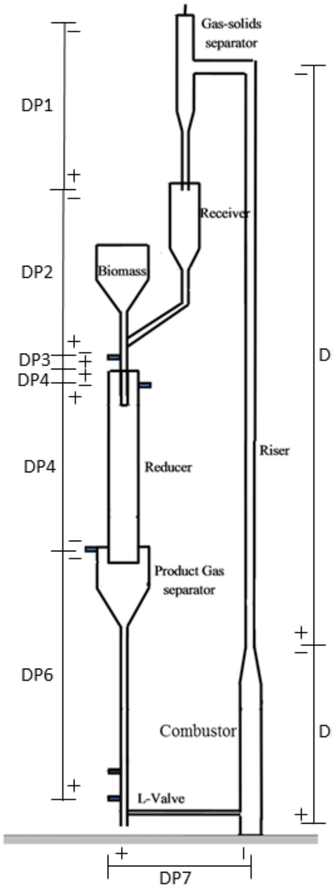


Figure 4 Pressure Drop Diagram of the CFM

Circulation tests were conducted to verify proper operation of the unit. Initial tests showed good circulation of solids within the unit, no gas leaks, and correct reading from each of the pressure transducers. Special attention was placed on the dynamics of the reducer. To properly represent the hot conditions of the sub-pilot BTS unit, a pressure drop of roughly 30 inches of water column across the entire reducer was desirable. Sample data from the test is located below in Figure 5.

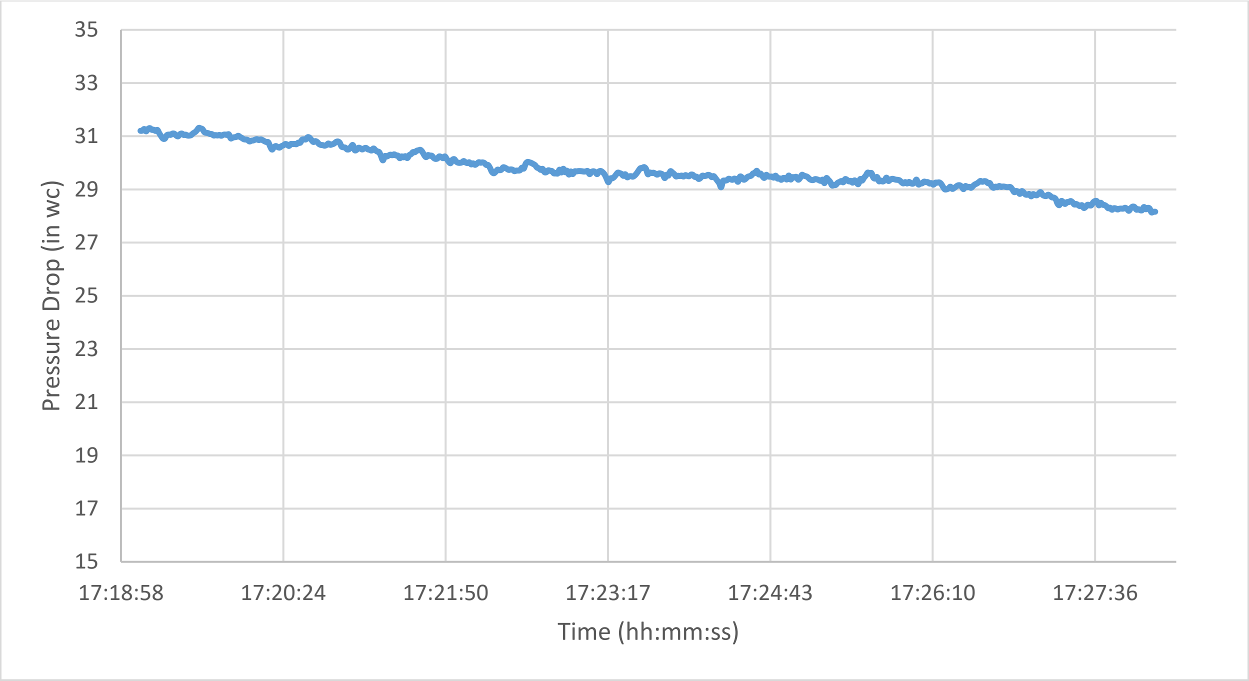


Figure 5 Reducer Pressure Drop

Proper pressure balance within the CFM unit can be verified by comparing the total pressure drop of the combustor side (L-valve, combustor, riser) to that of the reducer side (zone seals, reducer). This can be represented by the equation below.

$$DP7 + DP8 + DP9 = DP6 + DP3 + DP2 - DP5 - DP4$$

Shown in Figure 6 is the pressure balance taken from the experimental data. As can be seen, the pressure on the combustor side and reducer side agrees fairly well. The data confirms that the necessary pressure drop in the reducer is achievable and the desired pressure balance can be maintained in the CFM.

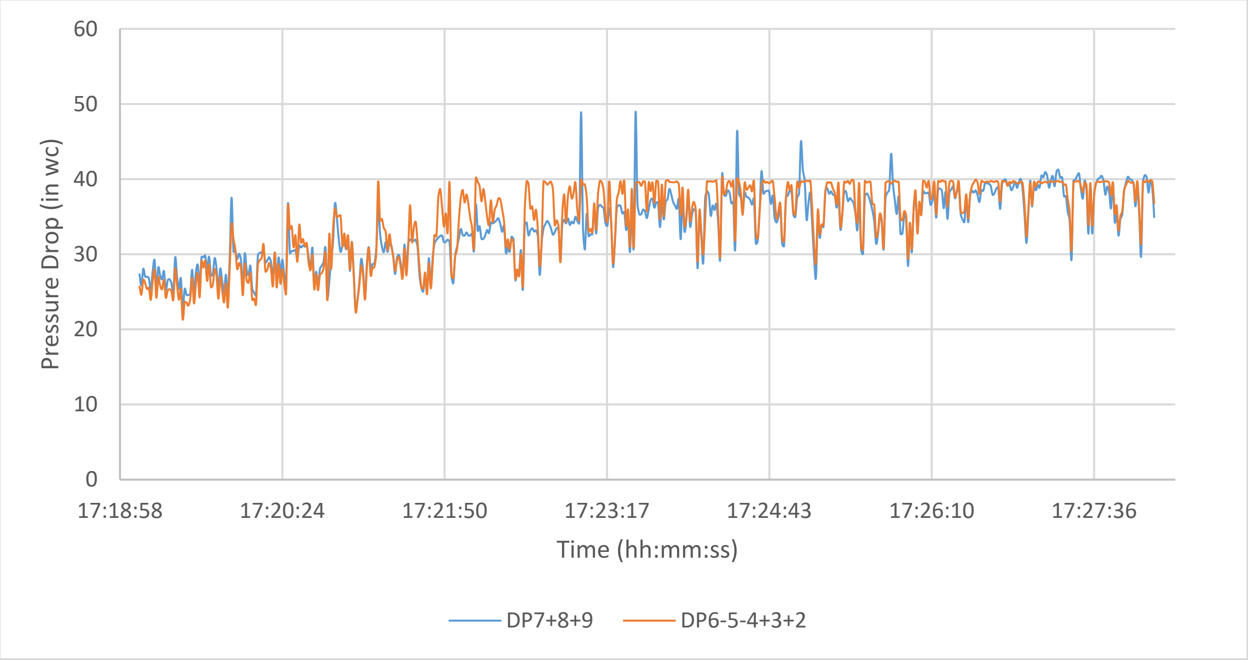


Figure 6 Pressure Balance in CFM

Biomass injection into the system was carried out in the CFM to study the distribution and flow of biomass pellets in the reducer.

A screw-feeder was used to inject biomass pellets into the reducer. Biomass was put into the screw-feeder and conveyed to the reducer through a helical screw. Figure 7 shows the setup for biomass injection.

Figure 8 shows how the biomass was made to enter the reducer. The red circle represents the point-of-entry of oxygen carriers into the reducer whereas green circle shows the point-of-entry of biomass. The rectangular box-type structure was designed to make sure that there is proper distribution of biomass in the reducer i.e. it spreads along the entire cross-section of the reducer.

Figure 7 Screw Feeder for Biomass Injection

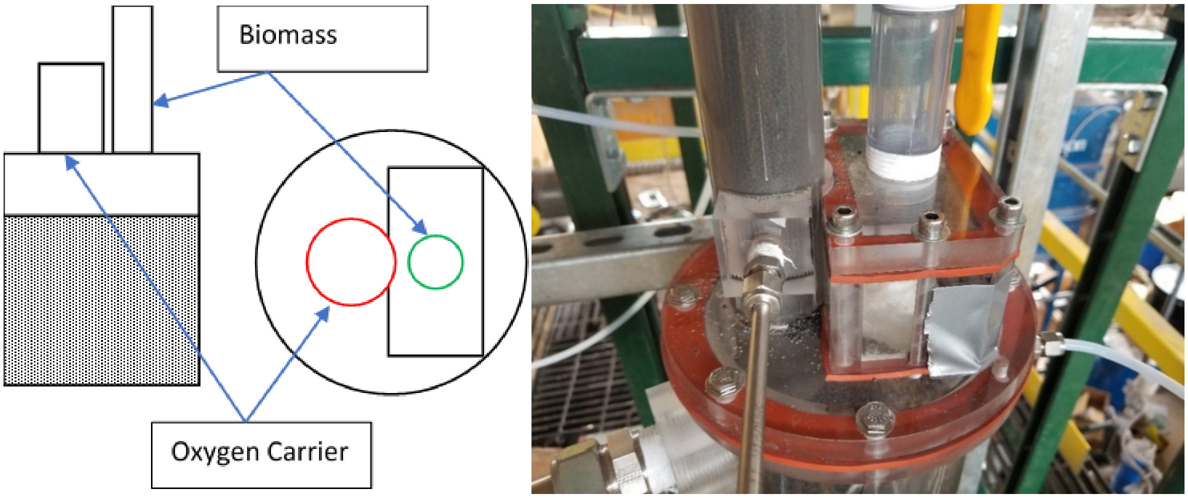


Figure 8 Biomass Injection Port at the Top of the Reducer

Multiple runs were made while injecting biomass into reducer under varying operating conditions in terms of oxygen carrier particles to biomass loading ratio. Air at was used as gas throughout the operation. To consider the effect of gas volume expansion at elevated temperatures, excessive gas was sent into the reducer to get the desired pressure drop.

The CFM testing showed that the distribution and mixing of the biomass pellets in the moving bed of oxygen carriers was dependent on how well the biomass pellets can cover of the top of the moving bed. In areas where biomass pellets were distributed on the top surface of the moving bed, the pellets were well mixed with the oxygen carriers along the height of the reducer. However, in areas without coverage of pellets on the top surface of the moving bed, such as the oxygen carrier feeding port, little biomass pellets were observed within the bed. The result indicated that the lateral mixing of solids in the moving bed reducer is limited. Thus, the design of the biomass injection port on the reducer should maximize the area on top of the reducer that can be covered by biomass feedstock. Internals in the moving bed reducer can also be used to enhance the distribution of biomass.

4. 10 kWth Sub-Pilot Test Unit Setup

4.1 Test Unit Detailed Design and Safety Review

Based on the preliminary design, sizing calculations and CFM studies, a list of necessary instrumentation, along with a set of piping and instrument diagrams (P&IDs), was generated. The sub-pilot scale test unit consists of a moving bed reducer, a fluidized bed combustor, a dilute phase riser, interconnecting non-mechanical gas seals, a biomass feeder, a set of product syngas gas coolers, and electric heaters for the reactors. The reducer, combustor, and riser are cylindrical vessels made from Incoloy 800H. The reducer has a diameter of 6” and a height of 80”. The combustor has a diameter of 6” and a height of 26”. The riser diameter is 2”.

The associated instrumentation includes sensors for temperature and pressure measurements, mass flow controllers for gas flow control, control valves for reducer and combustor gas outlets, and gas analyzers. Figure 9 and Figure 10 show the overview of the reducer and combustor P&IDs.

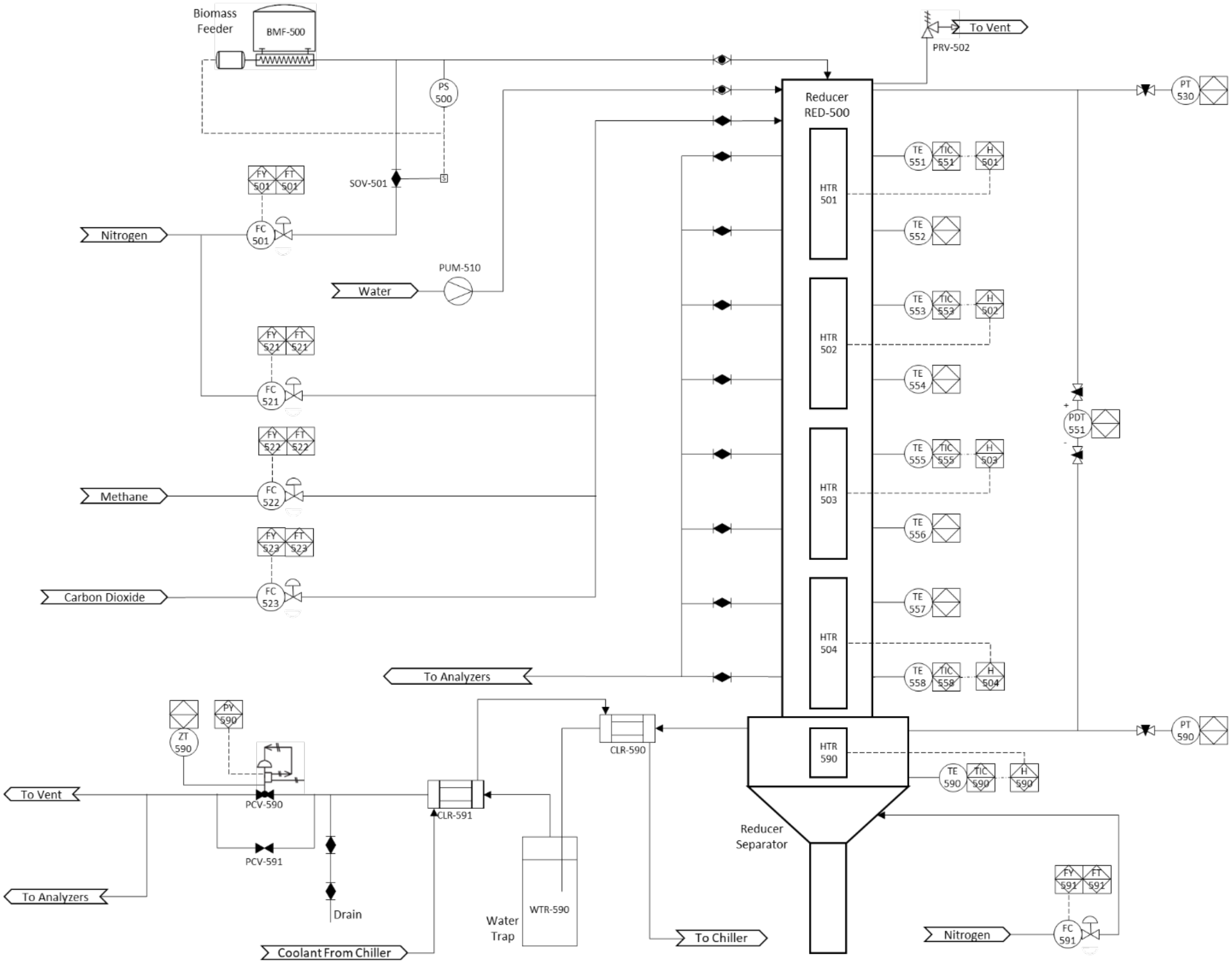


Figure 9 Overview of Reducer P&ID

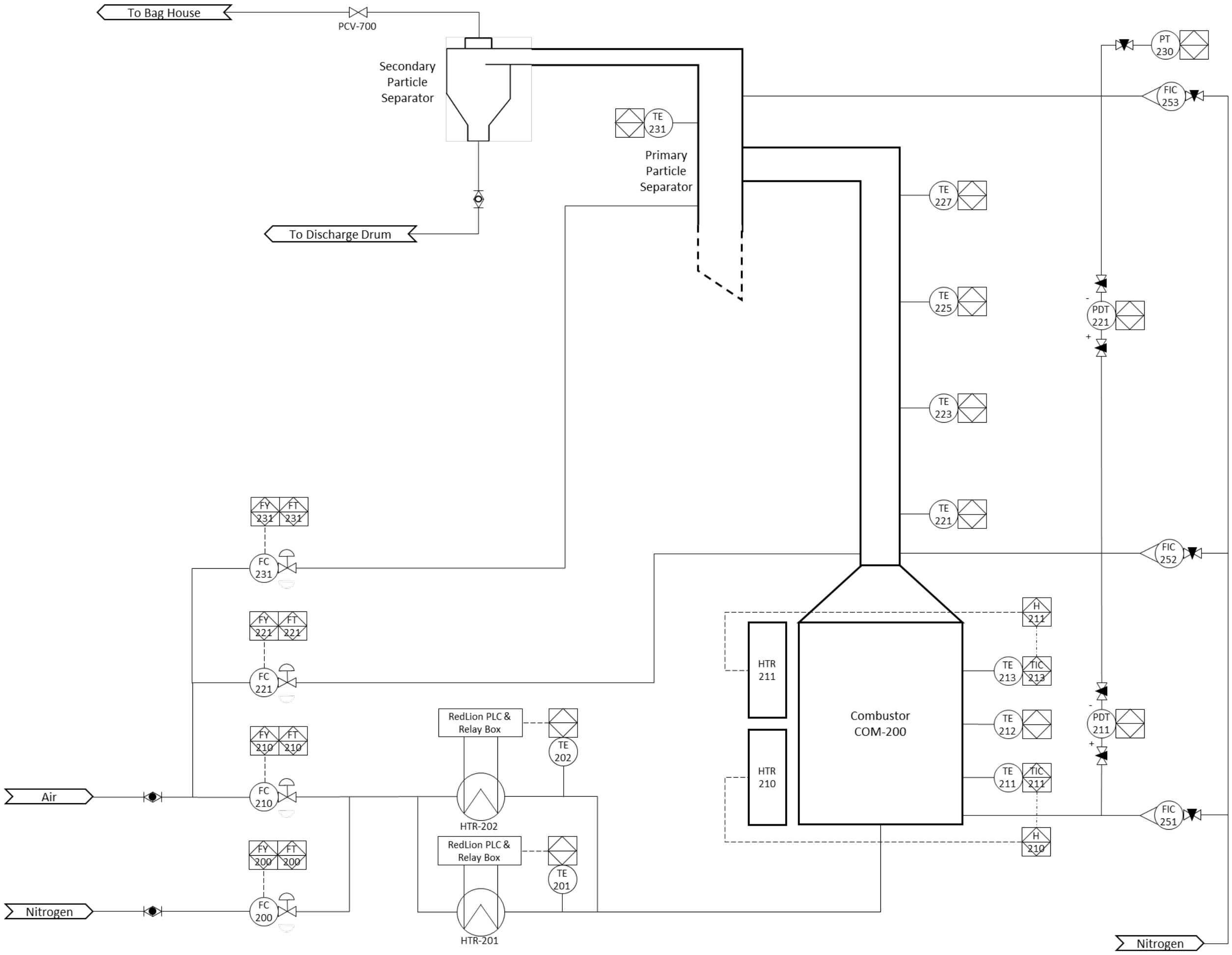


Figure 10 Overview of Combustor P&ID

A HAZOP review was performed to analyze the operation of the BTS sub-pilot reactor system. The HAZOP review was based on the current reactor design, P&ID drawings, and past operation experience of OSU's chemical looping units.

The BTS sub-pilot reactor system was divided into 3 sub-systems based on the P&ID drawings, i.e. the combustor, reducer, and zone seal/L-valve system. The study on each of the sub-system consist of a node of the HAZOP review. The review was performed by considering the deviation of the system from normal operation conditions. A series of guide words were used to define the deviations. Each of the guide words was applied to each components of the system, and the cause, frequency, consequences, severity, safeguards/protection factors, and mitigation strategies were considered. Figure 11 shows an exemplary node of the HAZOP review.

NODE: B Title: Reducer

Design Intent of Node: Convert biomass (20g/min) into syngas. Hazards: heat, flow, combustible gas

Guideword/ Deviation	Cause	Freq	Consquences	Severity	Safeguards/Protection Factors	Protec	Scenario		Action Priority	Rec #
						Factor	Freq	Risk		Comments
None of								---		
Biomass flow	BMF-500 Failure	0	No reaction	0		0	**0**	0	C	Order spare screws for biomass feeder. Monitor biomass feeder operation by weighing scale
	Depleted biomass supply	0	No reaction	0		0	**0**	0	C	
	Plugged line or injection port	1	No reaction; Biomass feeding damage; Pressure buildup in biomass feeder	1	Operator recognition	1	**0**	1	B	Install hard wired pressure switch to shut down biomass feeder when the pressure is high.
steam flow	PUM-501 Failure	-1	Low H2 concentration	0		0	**-1**	---		
biomass feeder N2 flow	FC-501 Failure	-1	Syngas leak into biomass feeder.	2		0	**-1**	1	B	In PLC: Show alarm if FC-501 flow is low
	N2 tank runout	0	same as above	2		0	**0**	2	A	Operation: check gas tank inventory every 2 hours; Install gas alarm near the biomass feeder; Leak test the biomass feeder before run
gas flow	FC-521 or FC-522 or FC-523 Failure; gas tank runout	-1	Incorrect reaction condition; inaccurate gas analysis	0		0	**-1**	---		
		-1	Low reactor pressure drop, causing air flow into reducer	3		0	**-1**	2	A	In PLC: trip system when zone seal flow direction reversed and switch gas flowing into combustor to N2 Operation: check gas tank inventory every 2 hours
solid flow	Plug in system / no solids circulation	2	Accumulation of biomass at the top of the reducer; sintering of oxygen carrier particles	0	Operator recognition	1	**1**	1	B	Add isokinetic solid circulation measurement device to the reducer
outlet gas flow	Gas outlet plug	0	Pressure buildup in system; Syngas leakage into combustor; potential skin burn	3	PPE and insulation/heater	1	**-1**	2	A	In PLC: trip system when zone seal flow direction reversed and switch gas flowing into combustor to N2 Add pressure releive valve at the top of the reducer
reactor temperature	Heater Failure	0	Loss of reaction; inability to operate	0		0	**0**	0		Stop biomass feed when reactor temperature is too low; Install heater protection TCs
pressure drop	PDIT-550 Failure	1	no indication of pressure drop accross reducer; difficulty in operation	0		0	**1**	1	B	Add purge gas to pressure ports
More of								---		
Biomass flow	Biomass Feeder Malfunction	-1	biomass feeding line plug; high reactor pressure; syngas leakage into combustor; potential skin burn	3	PPE and insulation/heater	1	**-2**	1	B	In PLC: trip system when zone seal flow direction reversed and switch off biomass feeder.
biomass feeder pressure	biomass feeding line plug	1	biomass feeder damage; pressure buildup in biomass	1	Operator recognition	1	**0**	1	B	Install hard wired pressure switch to shut down biomass feeder when the pressure is high.
steam flow	PUM-501 malfunction	-1	Pressure buildup in system; Syngas leakage into combustor; potential skin burn	3	PPE and insulation/heater	1	**-2**	1	B	In PLC: trip system when zone seal flow direction reversed and switch off water pump.
gas flow	MFCs malfunction	-1	Pressure buildup in system; Syngas leakage into combustor; potential skin burn	3	PPE and insulation/heater	1	**-2**	1	C	In PLC: trip system when zone seal flow direction reversed and switch off water pump.
reactor pressure	Outlet or system plug		Same as "None of: gas outlet flow"					---		
pressure drop	Excessive gas flow caused by excess gas or biomass		Same as "More or: biomass flow/steam flow/gas flow"					---		
As well as								---		
Liquid water flow into reducer	PUM-501 malfunction	-1	system pressure fluctuation	0	The reactor is heated. Liquid water will be evaporized when entering the reactor.	1	**-2**	---		Previous experience shows that liquid water is not going to cause significant fluctuation
moisture into analyzer	CLR-590 failure	-1	damage IR analyzers	0	Gas conditioning system	1	**-2**	---		
Other than								---		
air flow into reducer	Incorrect pressure balance	-1	biomass burning in reducer; local hot spot; reactor damage; potential skin burn	3	PPE and insulation/heater	1	**-2**	1	B	In PLC: trip system when zone seal flow direction reversed. switch gas flowing into combustor to N2
Part of								---		
Reducer integrity	High temperature or stress	-1	Syngas or high temperature solid leakage; potential skin burn or toxicity; potential fire	3	PPE and insulation/heater	1	**-2**	1	B	Install gas alarms around the reactor

Figure 11 HAZOP Analysis Node for the BTS Sub-pilot Unit Reducer

The findings of the findings/recommendations from the HAZOP review were incorporated into the BTS sub-pilot unit before the actual operation commence. The detailed reactor design drawings were generated based on the previous reactor sizing, results of the CFM testing, and the findings of the HAZOP review.

4.2 Test Unit Fabrication, Assembly, and Commissioning

Fabrication drawings were developed which serve as guideline of modification of current reactor system. A previous reactor vessel on a supporting structure was disassembled for modification and repair. The structure was cleaned up to prepare for the re-assembly of the reactor after the modification.

Raw Incoloy 800H alloy pipes/plates were purchased for completing the modification and assembly of reactor system according to the drawing. Figure 12 shows the details of modified parts at top reducer when assembled. The top zone seal standpipe was relocated off-center of the reducer top plate, allowing additional space for a ¾” biomass-feeding pipe to be welded nearby. One pressure port above were bended horizontally to provide space for the new added pipe. Re-location of the standpipe also created 3 degrees of rotation for the whole section of top zone seal and above, thus the riser pipe was cut and re-welded to adapt such difference.

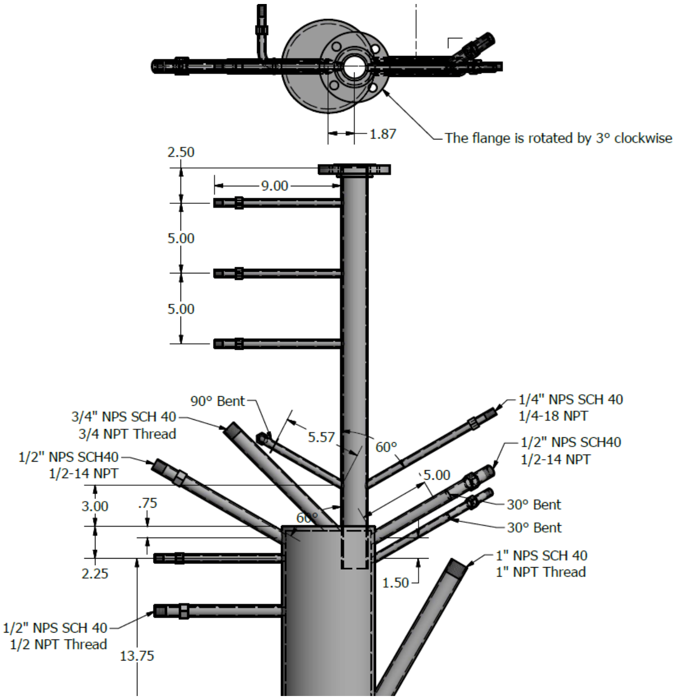


Figure 12 Detailed drawing of the top reducer section

All safeguards, including mechanical features and control system interlocks, were incorporated to the system, and tested for functionality.

After the completion of test unit assembly and control system programming, the solid circulation at ambient condition was tested to see if solids transport in an appropriate way to prepare for future operation at elevated temperature. Given 1700 standard liter per minute (SLPM) air flow through the combustor reactor, and 110 SLPM through L-valve aeration tap, the solids were circulated smoothly. Air was used as L-valve gas in this test. The test lasted for 1 hour, and no operation issue was noticed. The pressure at combustor side was balanced by the reducer side. The pressure profile in combustor, reducer, riser, top zone seal and bottom zone seal is shown in Figure 13.

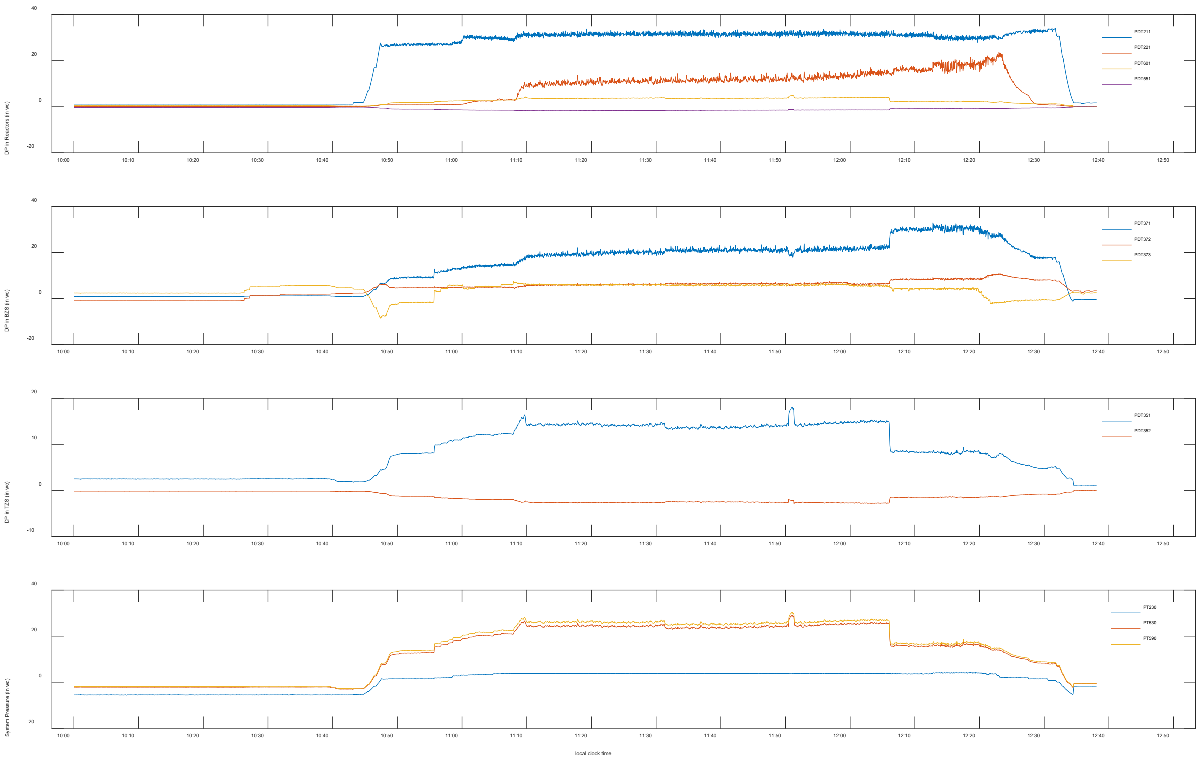


Figure 13 Pressure and differential pressure profile of system in cold solids circulation

A test operation at hot condition was performed to verify the functionality of the equipment and instrumentation. During the test operation, fluidization in the combustor reactor and solid circulation in the system were maintained by using proper amount of combustor air and L-valve aeration gas.

The reactor was gradually heated up to 980ºC in the reducer. The temperatures in the reducer and the combustor are shown in Figure 14. It was found that the combustor cannot be heated to sufficiently high temperature, due to the large amount of low temperature air being sent into the combustor. An additional air heater was later installed to supply additional heat to the combustor.

As shown in Figure 15 and Figure 16, the combustor air and L-valve aeration gas flow rate were gradually decreased as the reactor temperature decreases. This is due to the increase in gas velocity as system temperature increases.

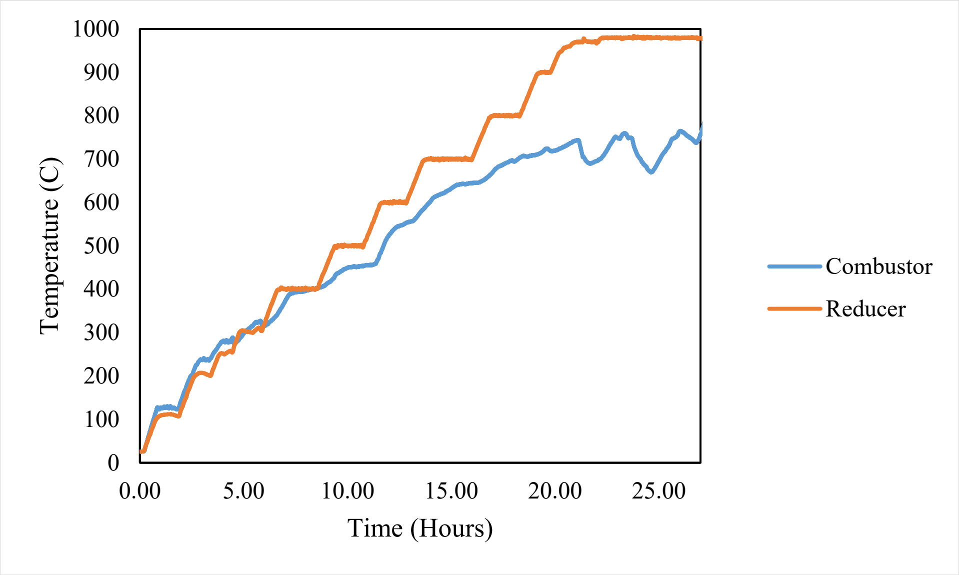


Figure 14 Reactor temperatures during test operation

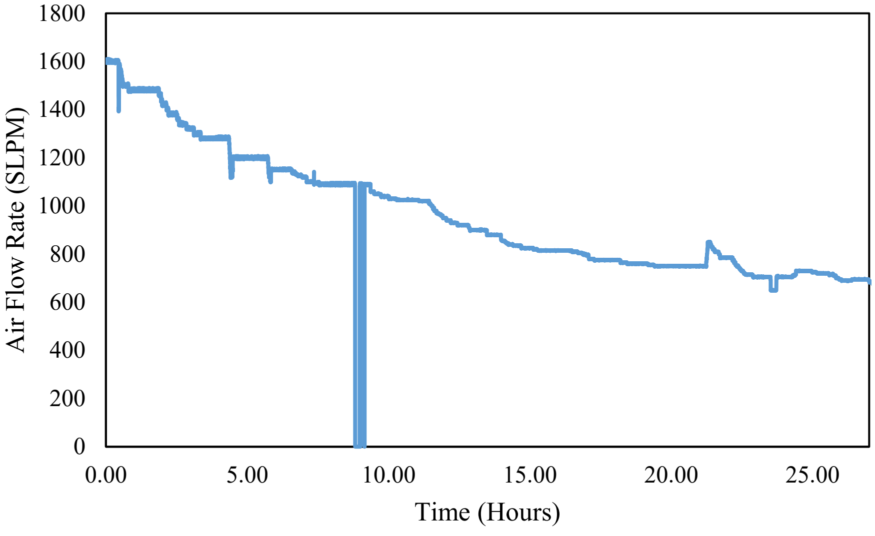


Figure 15 Combustor Air Flow Rate

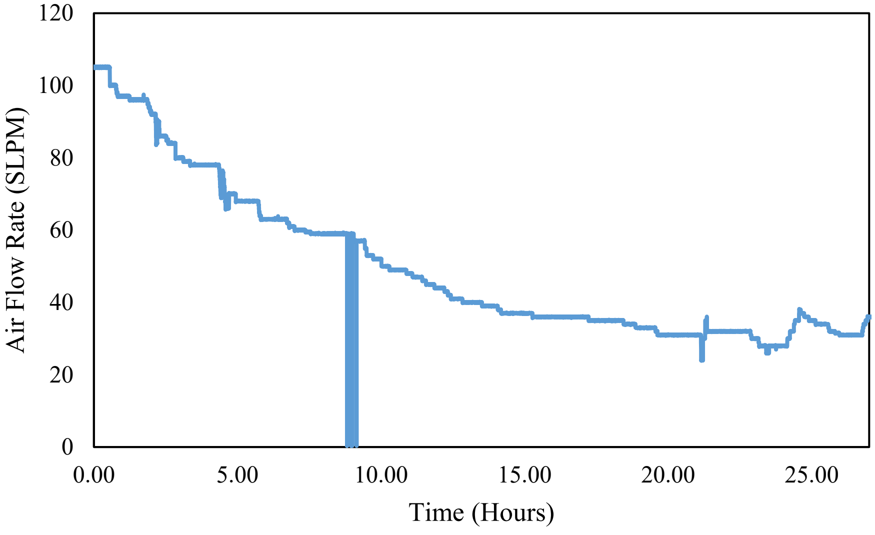


Figure 16 L-valve Gas Flow Rate

5. 10 kWth Sub-Pilot Test Unit Operation

5.1 Test Operation with Gaseous Fuel

The first hot operation under reactive condition was conducted with gaseous fuel, i.e. CH_4. After about 13 hours of heating, the reducer reactor was heated up to 1000ºC, while the combustor reached about 820ºC. The methane injection port installed in the combustor was used to further increase the combustor temperature to 900ºC. After fuel injection to the reducer, the combustor temperature further increased to 1000ºC. The temperature variation during the experiment is shown in Figure 17.

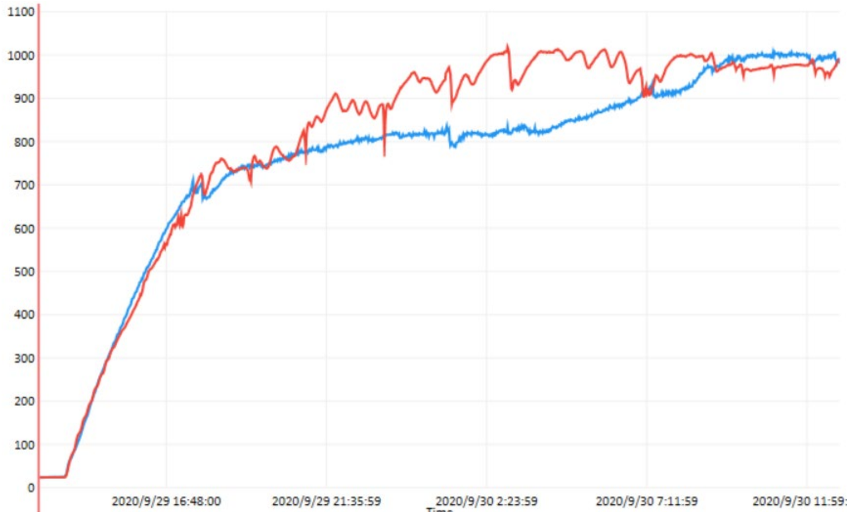


Figure 17 Reducer (red) and combustor (blue) temperature during operation

Methane was used as the test fuel to the reducer in order to verify the functionality of the fuel injection lines, syngas coolers, and gas analysis system. It was verified that the fuel injection system was functioning correctly, and the syngas coolers were able to remove the moistures in the syngas. The gas analysis system was able to respond to the variation in gas composition. Figure 18 shows the syngas composition produced after methane was introduced to the reducer. A full CH_4 conversion was achieved. A syngas purity of >80% was achieved from the conversion of CH_4

(dry base, N_2 free). Figure 19 shows the O_2 concentration in combustor effluent gas. The O_2 concentration decreased after fuel injection, indicating the regeneration of oxygen carrier materials in the combustor.

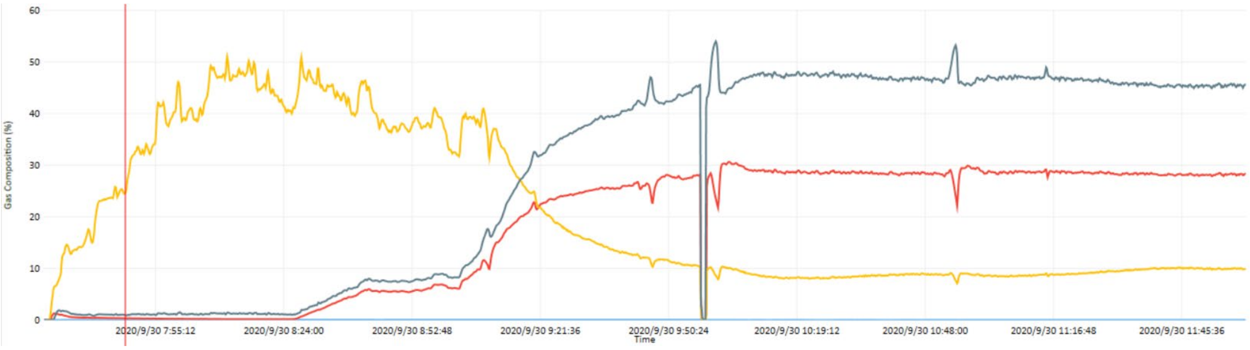


Figure 18 Composition of syngas produced from reducer. Blue: CH_4; Grey: H_2; Red: CO; Yellow: CO_2.

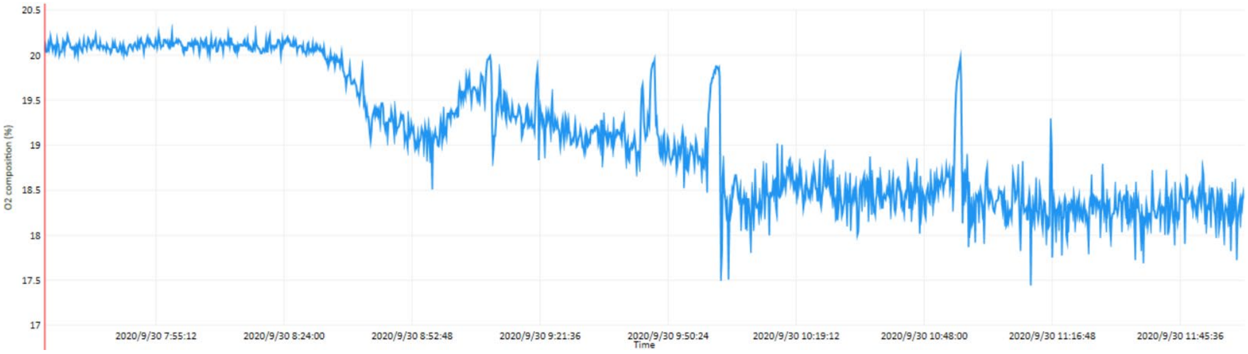


Figure 19 O_2 concentration in combustor effluent gas.

After a steady operation of about 2 hours, the solid circulation was interrupted, and the operators were unable to resume normal operation. It was found that the solid-state relay (SSR) that controls the electric heater around the L-valve (solid circulation device) section was malfunctioning. Despite the input signal being off, the SSR remained on (closed circuit). The temperature of the L-valve section was uncontrolled and heated to above 1140ºC. The excessive heating caused significant thermal stress at the section and caused a leakage in a flange, which leads to the interruption of operation. The operation was aborted afterwards. Biomass injection was not attempted. The SSR was replaced after the operation and the flange was re-sealed for next operation.

5.2 Syngas Generation with Biomass Fuel: Test Campaign 1

Following the initial operation using gaseous fuel, test campaigns on the BTS sub-pilot test unit using biomass fuel for syngas generation were performed. The biomass feedstock used in the first test was a pelletized corn cob (by Best Cob), with a pellet diameter of about 4mm. The composition of the corn cob is shown in Table 2. The molar ratio of the elements in the corn cob can be written as: $CH_{1.52}O_{0.63} \cdot 0.14H_2O$.

Table 2 Composition of corn cob biomass

Proximate Analysis		
	As Received	Dry Basis
Moisture (wt %)	9.36	
Ash (wt %)	2.67	2.95
Volatile (wt %)	70.67	77.97
Fixed Carbon (wt %)	17.29	19.08
BTU	7213	7958
MAFBTU	8200	8200
Sulfur (wt %)	0.14	0.15

Ultimate Analysis		
Moisture (wt %)	9.36	
Carbon (wt %)	44.38	48.96
Hydrogen (wt %)	5.63	6.21
Nitrogen (wt %)	0.64	0.71
Sulfur (wt %)	0.14	0.15
Ash (wt %)	2.67	2.95
Oxygen (Diff.) (wt %)	37.18	41.02

The sub-pilot unit was heated up to 950ºC before the biomass feeder was started.

The first few injections were performed under low fuel loading and high oxygen carrier circulation rate. Figure 20 shows the reducer product gas composition. At the early stage of biomass injection, due to the high oxygen carrier oxidation state in the reducer, the majority of the product was CO_2. As the oxygen carriers were reduced, the concentration of partial oxidation products (CO and H_2) increased. The periodic fluctuation in gas composition was probably due to the solid feeding mechanism.

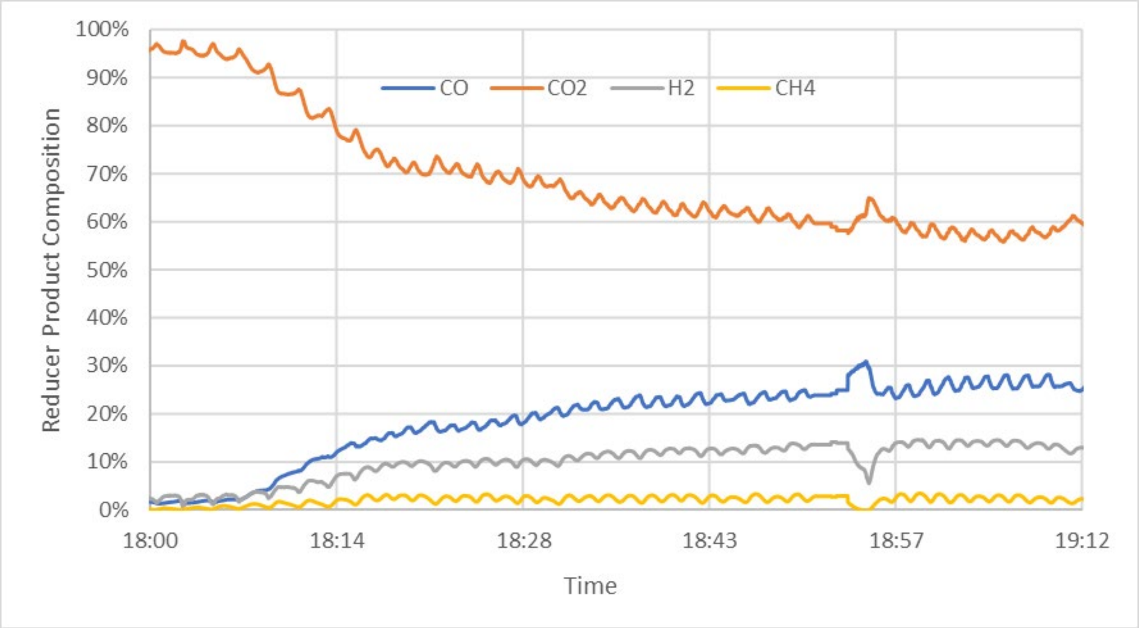


Figure 20 Reducer gas product in biomass injection, dry base, N_2 free

In a following biomass injection test, the biomass flow rate was increased. After the initial full oxidation stage, a high purity syngas was generated, with a composition of 42% CO, 37% H_2, 17% CO_2, and 4% CH_4. The syngas purity was close to 80%. The variation of reducer product gas composition is shown in Figure 21.

At 9:00 in Figure 21, steam was introduced to the reducer to increase the H_2:CO ratio of the syngas. The H_2:CO molar ratio was adjusted to slightly above 1. However, due to the limitation encountered in the unit, increasing the steam flow rate in the reducer was not possible.

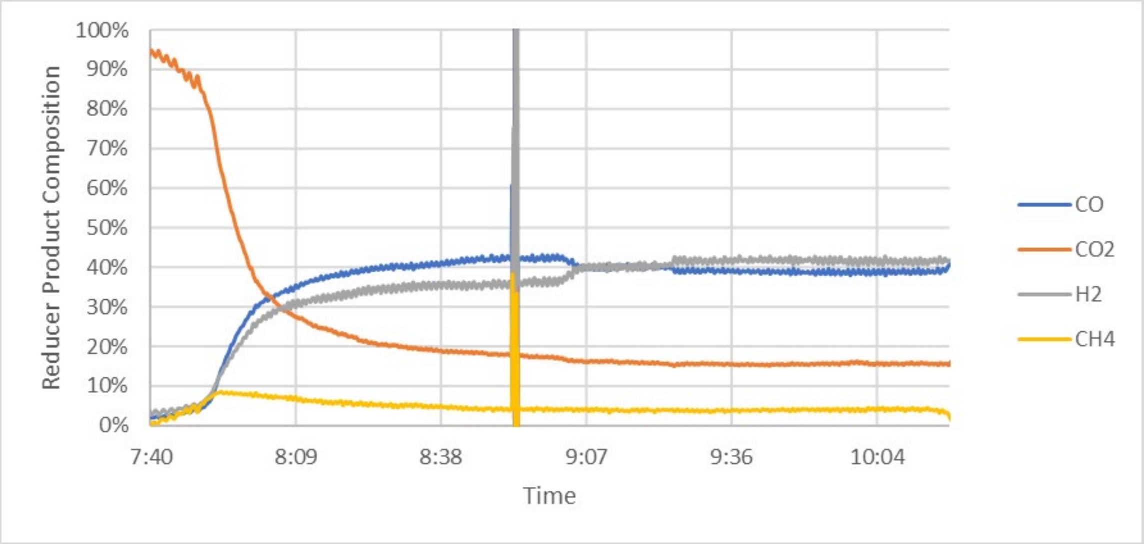


Figure 21 Reducer gas product in biomass injection, dry base, N_2 free

Despite the successful conversion of biomass and generation of high purity syngas, several issues were encountered during the operation.

1. The biomass feeder was connected to the reducer via a metal tube that was not flexible. This metal tube partially supported the weight of the biomass feeder. Thus, the weight scale underneath the biomass feeder was unable to correctly measure the weight variation during biomass feeding.
2. The reducer outlet pipes and valves were too small for the gas flow generated from the reducer. The high gas flow caused pressure build-up at the reducer

outlet. Thus, the project team was unable to further increase the biomass feeding capacity and steam flow rate in the system.

3. The biomass feeder was driven by a DC motor that was directly connected to the feeding screw. However, the normal rotation speed of the motor would cause an excess flow of biomass into the reducer. Thus, the project team has been operating the feeder motor in a "pulse width modulation (PWM)" manner, i.e. turn on the motor for 10%~20% of the 5-second cycle and turn it off for the rest of the of cycle. This non-continuous feeding manner may have caused significant pressure fluctuation in the system, as shown in Figure 22.

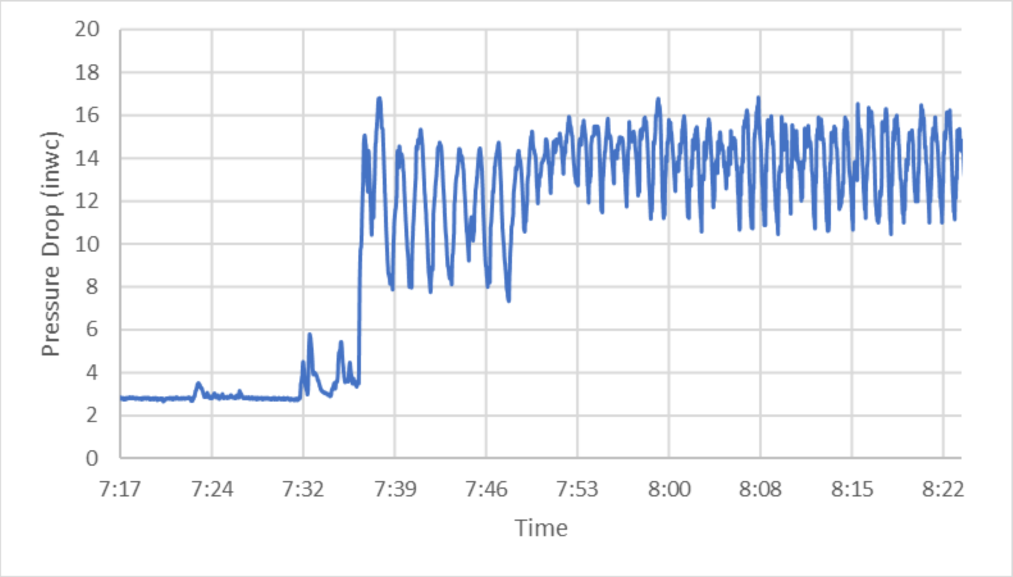


Figure 22 Pressure drop across the reducer during biomass feeding

To address the issues found during the test campaign, the following modifications were performed:

1. Insert a section of flexible tubing between the reducer and the biomass feeder to minimize the supporting effect of the reducer on the weight scale reading.

2. Install a reducing gear box between the DC motor and the feeding screw on the biomass feeder. The gear box was designed to allow the motor to rotate all the time at normal speed while the feeding screw will be rotating at a much slower speed. More steady feeding will minimize the pressure fluctuation in the reducer as well as other parts of the system.

3. Install a diaphragm pump at the reducer outlet to allow a higher biomass feeding rate and steam feeding rate.

5.3 Syngas Generation with Biomass Fuel: Test Campaign 2

Following the modification and improvement to the unit, an extended operation campaign was performed. Two types of biomass were tested during the test campaign, i.e. corn cob pellet (by Best Cob) and pine wood pellet (by Tractor Supply Co). Both types of biomass were pelletized to cylindrical shape with about 4mm diameter and 5-30 mm length. Figure 23 shows a picture of the wood pellet.

Figure 23 Wood pellet tested in sub-pilot reactor

The sub-pilot unit was heated up to 950°C before the biomass feeder was started. The test campaign lasted for 117 hours, during which the oxygen carrier particles in the sub-pilot reactor system were circulated continuously. Figure 2 shows the reducer temperature over the duration of the test campaign.

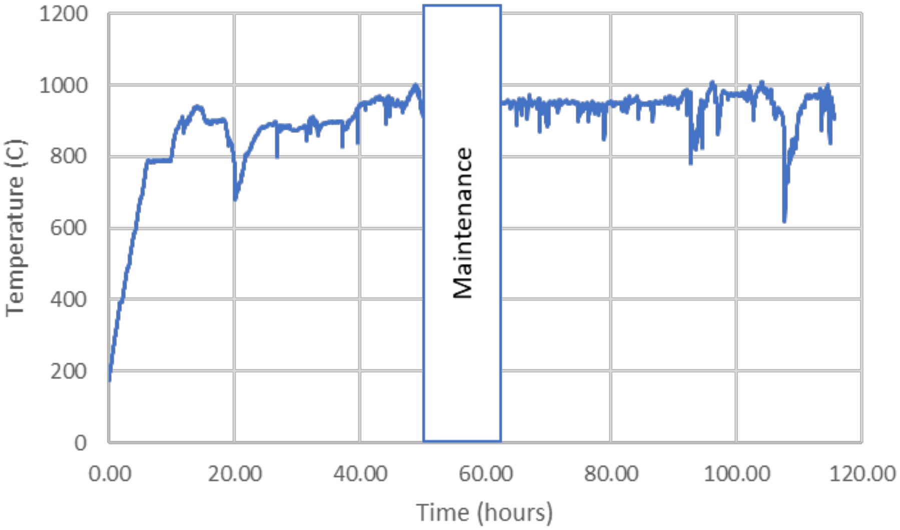


Figure 24 Reducer temperature during test campaign

Due to the small inventory of the biomass feeder, the project team was only able to perform intermittent biomass injection testing. Figure 25 and Figure 26 shows the gas concentration profile during the first biomass injection test with corn cob.

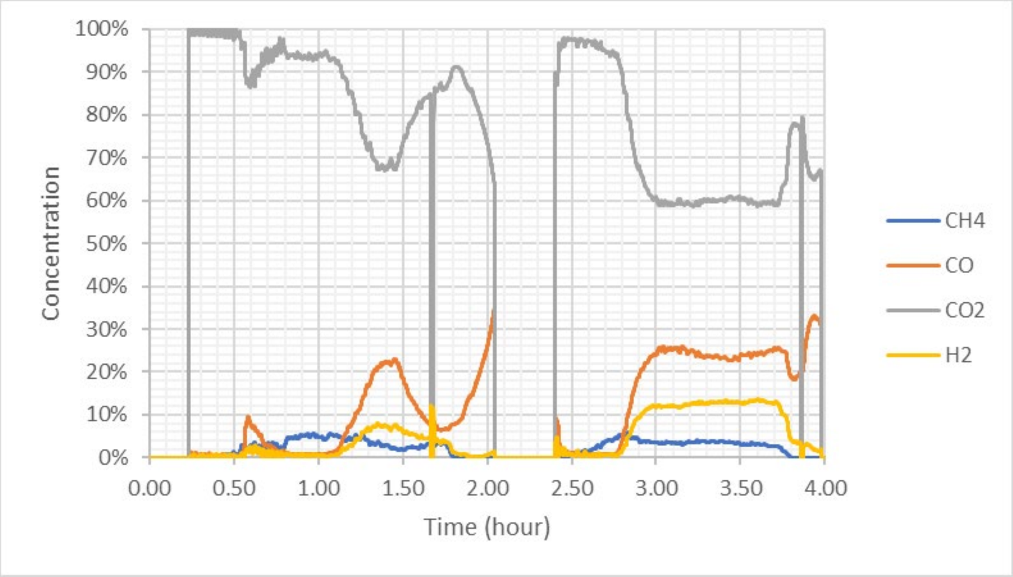


Figure 25 Reducer gas composition during biomass injection

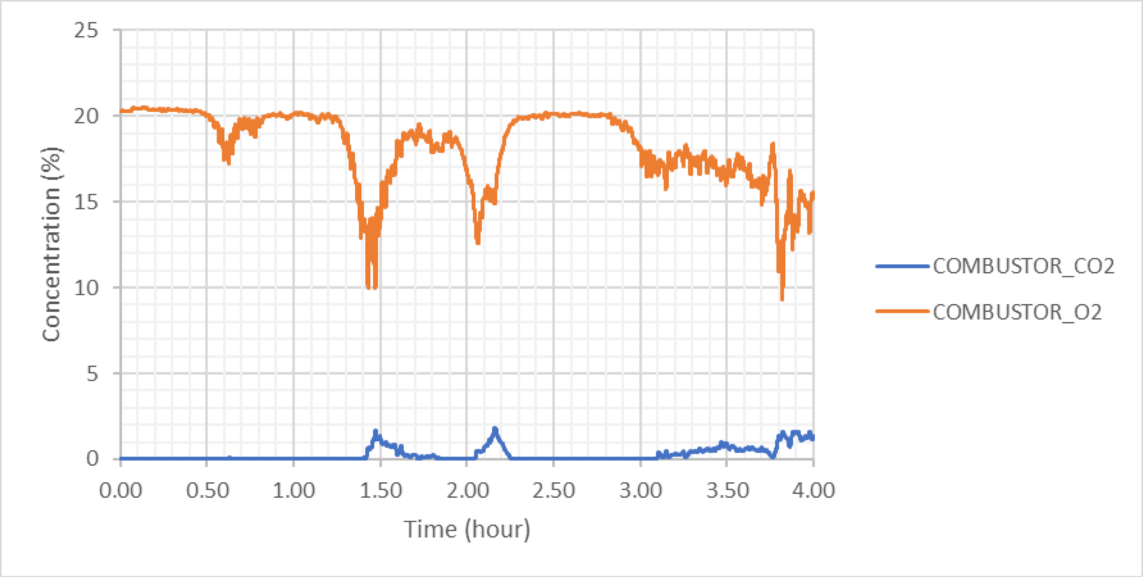


Figure 26 Combustor gas composition during biomass injection

The fluctuation of gas concentrations was potentially due to unsteady solid circulation rate. Due to the high oxygen carrier flow rate compared to the biomass feed rate, excessive oxygen was being supplied to the biomass in the reducer under this operating condition. As a result, the dominant product from the reducer is CO_2. In addition, as the oxygen carrier cannot be reduced sufficiently to low oxidation state, hydrocarbon (CH_4) conversion in the reducer was low. About 3% of CH_4 was detected from the reducer.

The gas composition in combustor confirmed that the oxygen carrier released oxygen in the reducer to convert the biomass fuel, and was regenerated in the combustor by air. The presence of CO_2 in the combustor was a result of unconverted char being carried over to the combustor by solid circulation. The spikes of CO_2 concentration corresponds to the carry over of unconverted char to the combustor during a surge of solid circulation rate.

The concentration of O_2 and CO_2 can be used to estimate the amount of oxygen transferred to char. The amount of char being consumed can be estimated by:

$$F_C = \frac{F_{N2}}{x_{N2}} x_{CO2}$$

where F_{N2} is the flow rate of N_2 in the combustor, x_i is the concentration of species i. The amount of oxygen being consumed can be estimated by:

$$F_{O2} = F_{N2}\left(\frac{21\%}{78\%} - \frac{x_{O2}}{x_{N2}}\right)$$

Thus, the fraction of oxygen consumed by char in the combustor is:

$$\frac{\frac{x_{CO2}}{x_{N2}}}{\frac{21\%}{78\%} - \frac{x_{O2}}{x_{N2}}}$$

This fraction over time is shown in Figure 27. In most of the time, less than 20% of the oxygen in the combustor was consumed by unconverted char, while greater than 80% of the oxygen was consumed by the reduced oxygen carrier. Since the reaction occurring in the reducer is partial oxidation, it is estimated that the carbon conversion in the reducer is greater than 80%.

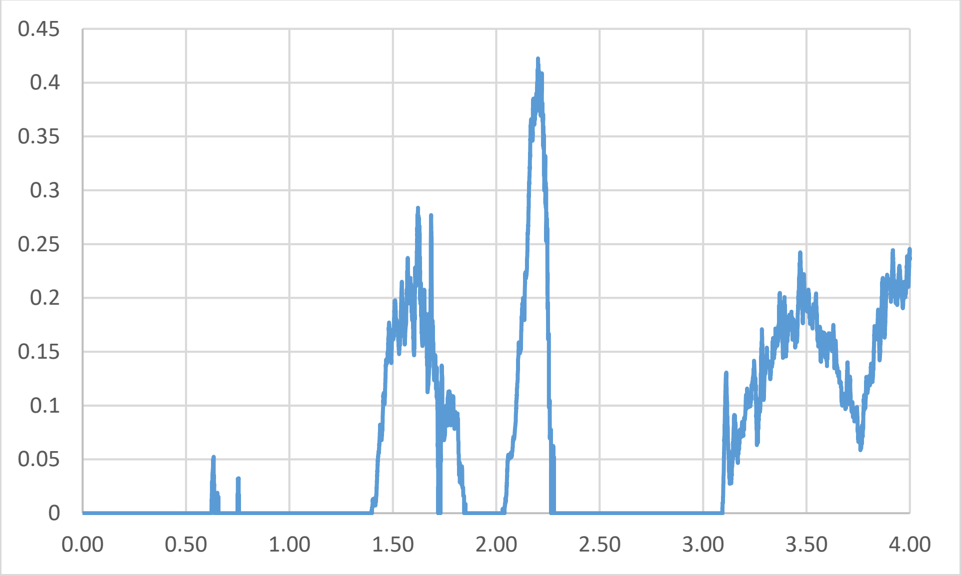


Figure 27 Fraction of oxygen consumed by char

In a later test condition, steam was used to increase the H_2/CO ratio to above 2, as required by downstream methanol synthesis. Figure 28 shows the reducer gas composition during the test.

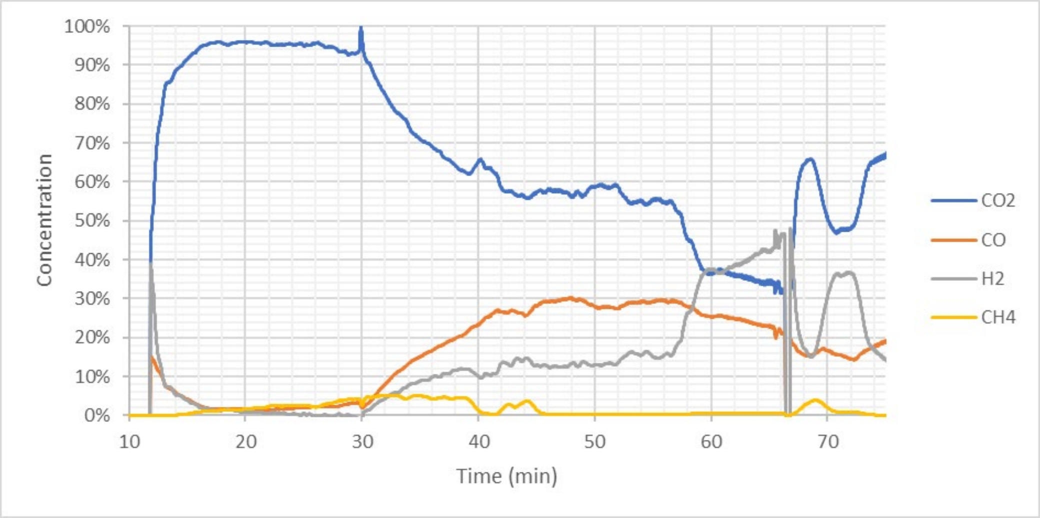


Figure 28 Reducer gas composition during biomass injection with steam

Steam injection started at about 55min of the plot. The injection of steam significantly increased the concentration of H_2. A H_2/CO ratio of greater than 2 was achieved soon after the start of steam. However, the condition was not maintained due to the failure of biomass feeder. The feeder screw was jammed by the biomass pellets and stopped rotating. This caused a sudden drop in gas flow in the system and caused a stop of solid circulation.

5.4 Syngas Generation with Biomass Fuel: Test Campaign 3

Based on experience from previous test campaigns, a number of repairs and modifications to the sub-pilot reactor system has been performed:

1. A 90V DC motor was installed in place of the previous 24V DC motor on the biomass feeder. The previous feeder could not provide sufficient torque and was occasionally jammed by large biomass pellets.
2. Gas bypass lines were installed at the reducer outlet to allow higher gas flow to exit the reducer. These lines allow the syngas to bypass the pressure control valve that may be limiting the gas flow.
3. Biomass pellets were crushed to smaller size by passing them through the screw feeder 2-5 times. During this process, the long pellets are crushed into smaller pellets. This ensures that the biomass pellets can be fed into the reactor smoothly.

Following the completion of repairs and modifications, a 4-day test campaign of the sub-pilot reactor system was performed. Two types of biomass, corn cob and wood pellets, were tested for syngas generation. The project team was able to achieve smooth operation during biomass injection with minimal system pressure fluctuations. This was potentially due to the smoother feeding of biomass fuel into the system from the high torque motor as well as smaller pellet size. A total of about 22 hours of biomass feeding was achieved.

Parametric testing with corn cob feeding was performed for a total of over 15 hours. A biomass feed rate of ~60g/min was maintained, while steam flow rate in the rage of 12 ~ 50 g/min was tested. Figure 29 shows the syngas composition from the reducer during one fuel feeding test with corn cob. Figure 30 shows the corresponding syngas quality.

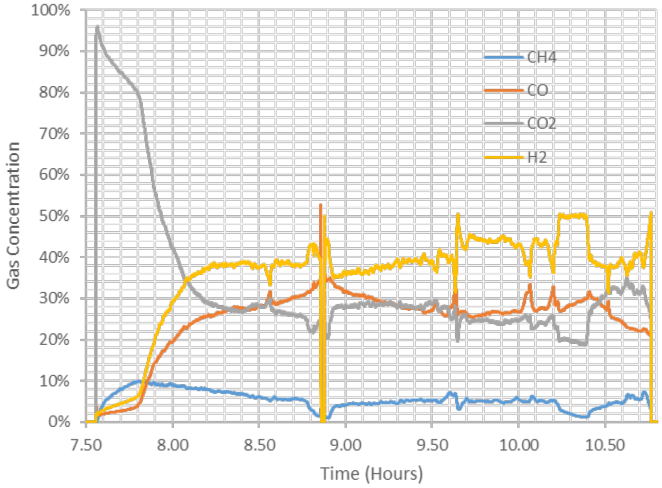


Figure 29 Syngas composition during corn cob feeding

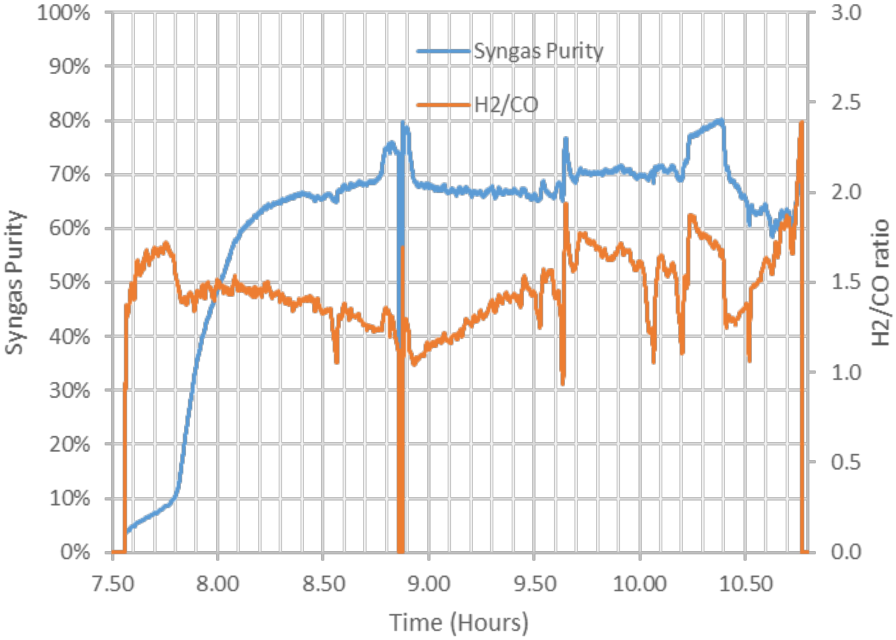


Figure 30 Syngas quality during corn cob feeding

Note that the steam flow rate was increased at about the 9-hour time mark in Figure 29 and Figure 30 to adjust the H_2/CO ratio of the syngas. The system was able to generate a syngas of 70% purity (dry base, N_2 free). At low steam flow rate, the H2/CO ratio was about 1.4, while at high steam flow rate, a H_2/CO ratio of 1.7 was achieved.

At the start of the biomass feeding when CO_2 was the predominant product, the CH_4 concentration was up to 10%, as the oxidation state of the oxygen carriers was not low enough to catalyze hydrocarbon conversion. However, as oxygen carrier oxidation state reduces and syngas was generated, CH_4 concentration dropped to below 6%. Overall, the syngas production performance is slightly lower than that predicted by thermodynamic analysis, which is potentially a result of the following factors:

1. Low reducer temperature due to steam injection. Reducer temperature dropped from 950°C to less than 900°C after steam injection started, due to the insufficient power of steam generator.
2. Insufficient gas residence time due to large gas flow rate.
3. Reduced oxygen carrier activity due to the existence of high concentration steam in the system

Wood pellets were also tested for syngas generation. A wood pellet feed rate of ~60g/min and steam feed rate of 42 g/min was used. The syngas generated was similar to that observed for corn cob, as shown in Figure 31 and Figure 32. A syngas purity of 70% was achieved with a CH_4 concentration of <6%. The H_2/CO ratio was greater than 1.8.

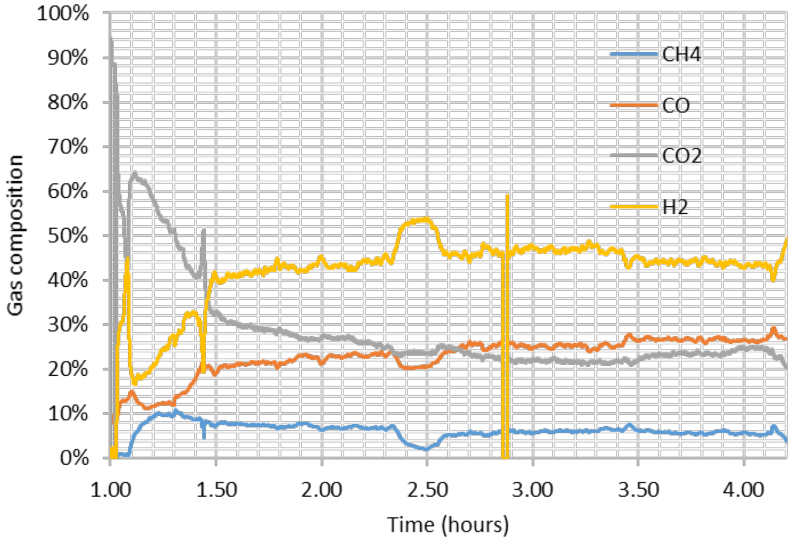


Figure 31 Syngas composition during wood pellet feeding

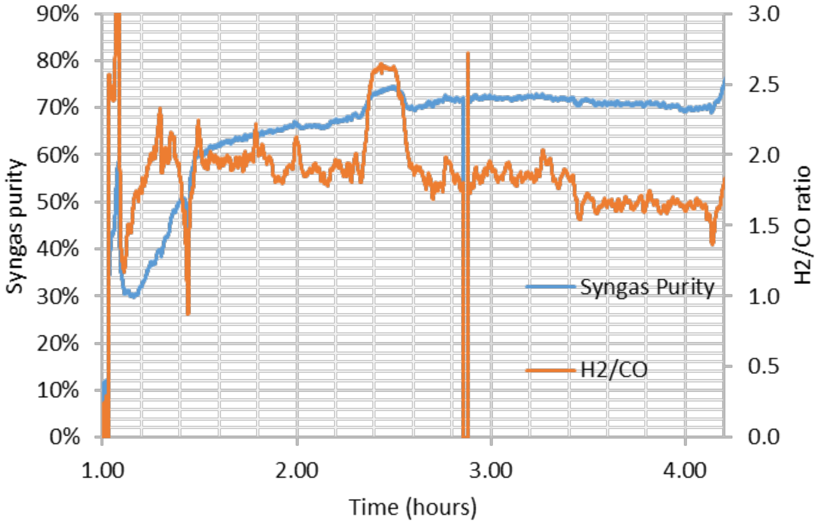


Figure 32 Syngas quality during wood pellet feeding

6. Techno-Economic Analysis on BTS Process

6.1 Preliminary Techno-Economic Analysis

Nexant performed a preliminary techno-economic analysis (TEA) was performed to evaluate and compare the OSU's BTS gasification process against a conventional biomass gasification process, for fuel-grade methanol production. The reference case design representing a conventional indirectly-heated biomass gasification process was selected from a recent DOE report entitled "Techno-economic Analysis for the Thermochemical Conversion of Biomass to Liquid Fuels, June 2011, US DOE Pacific Northwest National Laboratory (PNNL-19009)." This report, along with its companion 2015 study "Process Design and Economics for the Conversion of Lignocellulosic Biomass to Hydrocarbons via Indirect Liquefaction, March 2015, NETL/TP-5100-62402, PNNL-23822," contain a detailed set of plant heat and material balances and cost estimates for the various process units within an overall biomass to methanol plant. These were used to develop the performance and cost estimates to evaluate a BTS process-based biomass to methanol design.

For this preliminary TEA study, Nexant first modeled the PNNL indirectly-heated gasification biomass to methanol (IHGBTM) reference design using ASPEN, and benchmarked the heat and materials balance results and cost estimates using the published data from the PNNL reports. This was done to establish a working model, allowing the balance of plants to be identified, utility consumptions and the overall plant performance determined. The benchmarked model was then used as a tool to estimate the performance of the overall OSU BTS gasification-based biomass to methanol (OSUBTM) process, of which the BTS gasification system performance data and cost estimates were provided by the OSU. The BTS gasification system performance data used in this preliminary analysis is only preliminary and not yet fully optimized. The TEA study was later updated with actual data obtained from the experimental portion of the project.

The preliminary TEA report included as Appendix I to this report. In comparison with the IHGBTM Reference Design, the OSUBTM plant performance shows:

- Higher carbon efficiency – it consumes the same amount of biomass feed but produces approximately 20% more crude methanol, as a result of a higher amount of syngas that is generated by the BTS gasification process, but
- Lower overall plant thermal efficiency - the higher carbon efficiency for the OSUBTM plant negatively resulted in less energy available to support its overall in-plant steam utility and power consumption requirements. Main reasons for its higher power and fuel requirements are due to:
    - Increased syngas flow - resulting in a higher syngas compression power consumption downstream of the BTS process unit.
    - Increased CO_2 content (hence flow) in the syngas stream – resulting in a larger amine plant for CO_2 removal plant; and hence its associated energy requirement. The BTS syngas contains twice as much CO_2 as that in the IHGBTM Reference Design. This increases the amine plant utility requirements such as amine regeneration steam, condensing duty, and circulation load significantly. In the PNNL IHGBTM Reference Design, char and tar are formed in the gasifier. The char is carried by the circulating heat transfer medium to a combustor and is combusted to provide the heat for gasification and biomass drying. The CO_2 from char combustion is vented to the stack hence reducing the flow and CO_2 content of the syngas. Whereas in the OSUBTM design, all the biomass carbon is converted to carbon oxides in the gasifier and no char or tar is formed. The CO_2

from gasification is carried in the syngas and processed through heat recovery, syngas cleanup and compression before CO_2 removal.

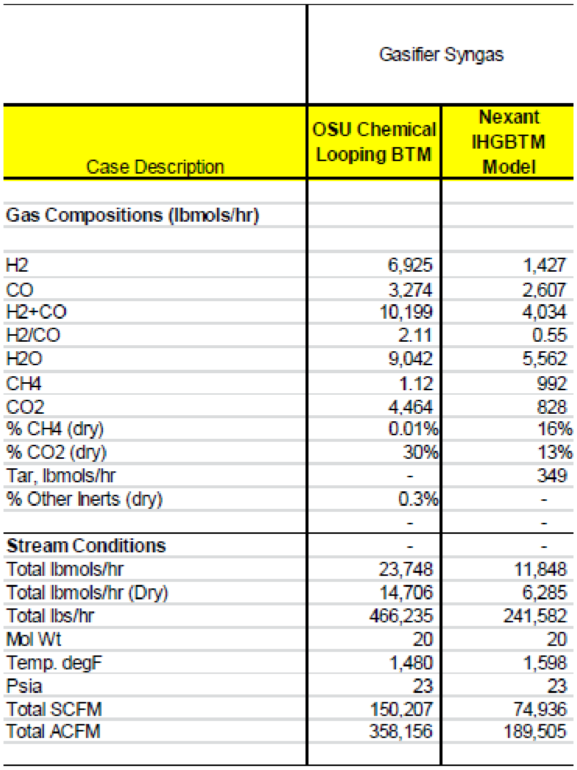
- Steam and air stream preheating requirement for the BTS gasification process, and
- A lower methane content in the syngas feed to methanol synthesis resulting in a lower heating value purge gas to be used as supplementary fuel for in-plant use. The net requirement is the need for importing natural gas as fuel, along with additional power.

Section 5 of the report provides a more detailed account of the OSUBTM model balance of plants results. The OSUBTM design has a higher estimated TPI cost at $347MM, of which the increase in the amine CO_2 removal plant cost due to a larger plant size requirement is the major culprit. The net result in the estimated RSP of methanol for the OSUBTM is at $1.38/gal, about 7.8% higher than that estimated for the IHGBTM design.

6.2 Update of Techno-Economic Analysis and Sensitivity Analysis

After reviewing the preliminary TEA report, the project team has identified a route for significant system optimization / improvement.

In previous process simulation, biomass was assumed to be fully converted to gaseous products in the reducer, which leads to a large amount of CO_2 generated in the reducer and a high steam consumption, as shown in Figure 33. As a result, the acid gas removal (AGR) system and steam generation were sized to be much larger than the reference case, which significantly increased the cost for the chemical looping system.



	Gasifier Syngas	
Case Description	OSU Chemical Looping BTM	Nexant IHGBTM Model
Gas Compositions (lbmols/hr)		
H2	6,925	1,427
CO	3,274	2,607
H2+CO	10,199	4,034
H2/CO	2.11	0.55
H2O	9,042	5,562
CH4	1.12	992
CO2	4,464	828
% CH4 (dry)	0.01%	16%
% CO2 (dry)	30%	13%
Tar, lbmols/hr	-	349
% Other Inerts (dry)	0.3%	-
	-	-
Stream Conditions	-	-
Total lbmols/hr	23,748	11,848
Total lbmols/hr (Dry)	14,706	6,285
Total lbs/hr	466,235	241,582
Mol Wt	20	20
Temp. degF	1,480	1,598
Psia	23	23
Total SCFM	150,207	74,936
Total ACFM	358,156	189,505

Figure 33 Simulated syngas composition from chemical looping system and reference case

The main strategy to reduce the amount of CO_2 in syngas is to entrain a portion of unconverted biomass char to the combustor, which is shown to be feasible in previous bench scale experiment (Xu et. al. Applied Energy 222 (2018): 119-131).

A process simulation in ASPEN Plus is performed to investigate the impact of transferring unconverted char, which is shown in Table 3. Leaving 20% of carbon in biomass unconverted could reduce oxygen carrier solid circulation rate by 34%, steam consumption by 40%, and CO_2 flow in syngas by 28%. Thus, the oxygen carrier attrition, reactor size, and downstream CO_2 removal demand are greatly reduced. On the other hand, the syngas (H_2+CO) yield is reduced by about 15% due to the reduction of carbon conversion in the reducer, which reduces the overall carbon efficiency.

Table 3 Effect of unconverted char on system performance

Unconv. Char	OC circulation rate	Steam Consumption (mol/mol C)	CO_2 in syngas (mol/mol C)	H_2+CO Yield (mol/mol C)
0%	100%	0.86	0.40	1.68
10%	83%	0.69	0.34	1.56
20%	66%	0.51	0.29	1.43

The TEA model was updated generate a syngas with composition and purity that is consistent with the experiment results. It was also assumed that the CH_4 produced from the reducer was inert in the following methanol synthesis process, which was subsequently burned to generate steam and/or power. 20% of char in biomass was assumed to be burned in the combustor to reduce steam consumption as well as provide additional heat. Thereby, the system was able to eliminate the need for natural gas import, as well as to reduce the amount of electricity imported. The cost was updated to remove the natural gas steam boiler, and to include the cost for additional steam and power generation device. The updated capital cost is shown in Table 4.

Table 4 Capital Cost for Various Components in Updated TEA Model

	Reference IHGBTM	OSU BTS Previous design	OSU BTS Char bypass
Biomass Prep	$ 28,159,000	$ 28,159,000	$ 28,159,000
Gasification	$ 15,229,000	$ 26,843,000	$ 19,330,138
Syngas cooling	$ 20,468,000	$ 6,677,000	$ 6,066,505
Feed compression	$ 9,636,000	$ 12,261,000	$ 10,942,565
Sulfur removal	$ 1,660,000	$ 1,660,000	$ 1,660,000
MP steam boiler		$ 6,907,000	
CO2 recovery	$ 59,205,000	$ 96,078,000	$ 76,340,799
Syngas compression	$ 12,189,000	$ 13,032,000	$ 11,630,659
Methanol synthesis	$ 24,918,000	$ 28,170,000	$ 24,918,000
Purge gas expander	$ 2,149,000	$ 2,031,000	$ 2,031,000
Steam and power	$ 18,601,000	$ 1,966,000	$ 18,601,000
Balance of plant	$ 5,454,000	$ 5,898,000	$ 5,898,000
Total	$ 197,751,000	$ 229,682,000	$ 205,577,665

Although the total plant cost is slightly higher than the reference case, the OSU BTS plant gives higher overall efficiency. In addition, the natural gas import is eliminated, while electricity import is significantly reduced. Modeling results showed that the operation cost for both the OSU BTS plant and the reference plant was similar. Thus, the overall cost for methanol was lower for the OSU BTS plant. The required selling price (RSP) of methanol was estimated to be $1.15/gal, as compared to $1.28/gal of the reference case.

A series of sensitivity analysis is also performed on various assumptions and cost factors. Figure 34 shows the result of these analysis.

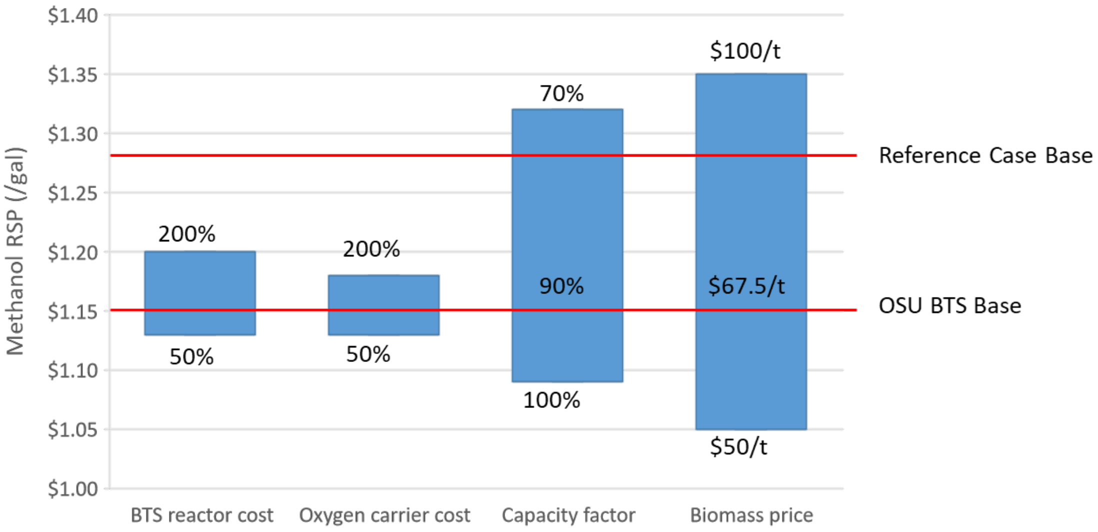


Figure 34 Sensitivity analysis for methanol production cost

It was shown that the methanol RSP is mostly driven by the cost of biomass feedstock. A 50% increase of biomass prices causes a 17% increase of methanol RSP Doubling the cost for the BTS reactor or the oxygen carrier material do not have a significant impact on the cost for methanol.

7. Conclusion

The project team has successfully completed the scope of the work and achieved the goals of the project. The project team has designed, fabricated, assembled, commissioned, and operated a 10 kW_{th} sub-pilot scale test unit for the BTS process. Multiple test campaigns, totaling to greater than 200 hours, were performed to test the conversion pelletized corn cob and wood biomass fuel in the BTS process. The project team was able to achieve the following in the 10 kW_{th} sub-pilot test unit:

1. Steady solid circulation, gas sealing between reactors, and controlled heating
2. Biomass fuel and steam injection
3. Syngas production from the reactor
4. Syngas (H_2+CO) purity in a dry, N_2 free base of >70%, and H_2:CO molar ratio of >1.7
5. Adjustable H_2:CO molar ratio
6. Unconverted hydrocarbon (CH_4) of <6%
7. Partial conversion of char in the combustor
8. Regeneration of oxygen carrier in the combustor

The project team also completed a comprehensive techno-economic analysis (TEA) for the production of methanol using the BTS process, and compared to the reference case of an indirectly heated gasification process. The BTS process model used in the TEA was updated with experimental data obtained from the 10 kW_{th} sub-pilot test unit, and optimized for energy integration. The result shows that the BTS process yields a higher carbon efficiency than the reference process. The required selling price of methanol for the BTS process is \$1.15/gal, as compared to \$1.28/gal for the reference case.

The project team concludes that the BTS process is a promising technology for clean fuel production from biomass with low cost and high biomass conversion efficiency. Future research and development in the following areas are recommended:

1. Oxygen carrier with enhanced reactivity for higher hydrocarbon conversion
2. Biomass feeding at large scale
3. Fate of sulfur, alkali, and ash
4. Extended operation in pilot scale
5. Process integration with Fischer-Tropsch process, due to lower H2:CO ratio requirement that reduces the need for steam and overall cost
6. Process configuration for H_2 production

8. Appendix: Techno-Economic Analysis Report

Nexant
The Power of Experience

Biomass Gasification for Chemicals Production Using Chemical Looping Techniques

Preliminary Techno-Economic Analysis Report

Submitted to

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By

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Acronyms and Abbreviations

ACFM	Actual Cubic Feet Per Minute
AGR	Acid Gas Removal
Ar	Argon
ASPEN	ASPEN Process Simulation Program
ASU	Air Separation Unit
BD	Blowdown
BENZE	Benzene
BFD	Block Flow Diagram
BFW	Boiler Feed Water
BM	Biomass
BOP	Balance of Plant
BTS	Biomass to Syngas
BTM	Biomass-to-Methanol
Btu	British Thermal Unit
C	Carbon
CAPEX	Capital Cost
CCF	Capital Charge Factor
CEPCI	Chemical Engineering Plant Cost Index
CH_3OH	Methanol
CH_4	Methane
C_2H_2	Acetylene
C_2H_4	Ethylene
C_2H_5OH	Ethanol
C_2H_6	Ethane
C_3H_8	Propane
C_4H_{10}	Butane
CH_3OCH_3	Dimethyl Ether
CF	Capacity Factor
Cl_2	Chlorine
CO_2	Carbon Dioxide
CO	Carbon Monoxide
COS	Carbonyl Sulfide
CS_2	Carbon Disulfide
CT	Cooling Tower
CuO	Copper Oxide
CW	Cooling Water
°C	Degree Celsius
DIMET	Dimethyl Ether

DOE	Department of Energy
DOE/NETL	Department of Energy/ National Energy Technology Laboratory
°F	Degree Fahrenheit
Fe	Iron
Fe_2O_3, Fe_3O_4	Iron Oxides
$FeTiO_3$	Iron Titanium Oxide
gal	gallon
gpm	Gallons per Minute
HC	Hydrocarbons
HCl	Hydrogen Chloride
HCN	Hydrogen Cyanide
HHV	Higher Heating Value
HMB	Heat and Material Balance
HP	High Pressure
HPSH	High Pressure Superheated
H_2	Hydrogen
H_2O	Water
H_2S	Hydrogen Sulfide
IHGBTM	Indirectly-heated gasification biomass-to-methanol
ISBL	Inside Battery Limit
ITCMO	Iron-Titanium Composite Metal Oxide
kW, kWe	Kilo-Watt
kWh	Kilo-Watt Hour
lb	Pound Mass
Lb/hr	Pounds per Hour
Lbmols/hr	Pound Moles per Hour
LHV	Lower Heating Value
Lo-Cat	Lo-Cat sulfur removal process
LP	Low Pressure
MDEA	Methyl Diethanolamine
MeOH	Methanol
MFSP	Minimum Fuel Selling Price
MM	Million
MMBtu/hr	Million Btu per Hour
MMGal/Yr	Million Gallons per Year
Mol%	Mole Percent
MP	Medium Pressure
MU	Makeup
MWe	Megawatt electric
MWh/yr, MWH	Megawatt Hour per Year
N_2	Nitrogen

NAPHT	Naphthalene
NETL	National Energy Technology Laboratory
NFOPEX	Non Fuel Variable Operating Cost
NG	Natural Gas
NH_3	Ammonia
NO, NO_2	Nitrogen Oxides
NREL	National Renewable Energy Laboratory
NR	Not Required
O&M	Operation and Maintenance
OC_{fix}	Fixed Operating Cost
OC_{var}	Variable Operating Cost
OPEX	Operating Cost
OSBL	Outside Battery Limit
OSU	The Ohio State University
OSUBTM	Ohio State University Biomass to Methanol Process
OSUBTS	Ohio State University Biomass to Syngas Process
PNNL	Pacific Northwest National Laboratory
ppm, ppmV	Parts per Million, Parts per Million by Volume
Psia, psig	Pounds per Square Inch Absolute, Pounds per Square Inch
RSP	Required Selling Price
SCF	Standard Cubic Feet
SMR	Steam Methane Reformer
SO_2	Sulfur Dioxide
STG	Steam Turbine Generator
TEA	Techno-Economic Analysis
TIC	Total Installed Cost
TiO_2	Titanium Oxide
TOC	Total Overnight Cost
Tpd, Tons/D, T/D	Short tons per day
TPEC	Total Purchased Equipment Cost
TPI	Total Project Investment
USDA	U.S. Department of Agriculture
wt%	Weight percent
WGS	Water gas shift
ZnO	Zinc Oxide
ZnS	Zinc Sulfide

Section 1 Introduction

1.1 BACKGROUND

With U.S. Department of Agriculture (USDA) and U.S. Department of Energy (DOE) Office of Energy Efficiency and Renewable Energy (EERE) funding from Cooperative Agreement Award DE-EE0007530, the Ohio State University (OSU) is developing a biomass-to-syngas (BTS) chemical looping process for efficient production of value-added chemicals and liquid fuels from biomass. The chemical looping BTS process is a unique gasification process developed at OSU that can provide a highly cost-effective and energy-efficient alternative to the conventional biomass gasification processes. In the BTS process, gasification of biomass is accomplished through the unique combination of a co-current moving bed reactor and iron-titanium composite metal oxide (ITCMO) oxygen carrier developed at OSU. It allows for conversion of biomass to a high quality syngas with a H_2:CO ratio of 2:1 in a single step without the use of molecular oxygen and capital intensive units like tar reformer, water gas shift (WGS) reactor, and air separation unit (ASU). Such high quality syngas generated permits the downstream processing to be readily conducted to produce chemicals and liquid fuels without requiring costly syngas conditioning systems for tar cracking and/or hydrogen upgrading.

To support this development, Nexant is tasked to conduct a preliminary techno-economic analysis of the BTS chemical looping process for biomass-to-methanol (BTM) production applications. The analysis is based on OSU's preliminary modeling assessment of the BTS process at its current stage of development. It is to be updated with actual process performance data obtained from the laboratory pilot plant of which experimental testing is still in progress.

1.2 STUDY OBJECTIVES

The overall objective of this study is to (1) assess how best to integrate the OSU's BTS chemical looping technology for BTM production applications, and (2) perform a preliminary techno-economic analysis (TEA) to evaluate the benefits of such an integrated BTS process in comparison with a conventional biomass-to-methanol process using a conventional low-pressure indirectly-heated biomass gasification process.

Section 2 Biomass to Methanol Plant Design Basis

2.1 DESIGN REFERENCES

The conventional biomass gasification process to be used for this TEA study as a reference design of which to compare against the OSU's BTS technology is selected from a recent DOE report entitled "*Techno-economic Analysis for the Thermochemical Conversion of Biomass to Liquid Fuels*, June 2011, US DOE Pacific Northwest National Laboratory (PNNL-19009)." The PNNL-19009 contains a series of biomass-to-liquid fuels designs based on both indirect and direct heated biomass gasification processes. The reference case used for comparison against the OSU BTS chemical looping system is the biomass-to-methanol case in PNNL-19009, based on a conventional indirect heated biomass gasification technology.

Besides PNNL-19009, the following reports were also be used to supplement and provide more design guidelines for this study:

- "*Baseline Analysis of Crude Methanol Production from Coal and Natural Gas*, Revised July 29, 2014, DOE/NETL. 341/020514" (Crude Methanol Study)
- "*Process Design and Economics for the Conversion of Lignocellulosic Biomass to Hydrocarbons via Indirect Liquefaction*, March 2015, NREL/TP-5100-62402, PNNL-23822",
- "Quality Guidelines for Energy System Studies: *Cost Estimation Methodology for NETL Assessments of Power Plant Performance*, April 2011, DOE/NETL. 2011/1455", and
- "*Capital Cost Scaling Methodology*, January 2013, DOE/NETL. 341/013113."

These reports, together, contain a set of detail design bases and assumptions, as well as reference plant costs and economic evaluation guidelines; and they were used to supplement the design bases and cost data from the PNNL-19009 report.

2.2 CASE CONFIGURATIONS

To identify and determine the advantages of integrating the OSU BTS chemical looping technology, two biomass-to-methanol (BTM) plant design cases were developed, of which a process matrix is shown in Table 2-1. One of these cases is the Reference Case BTM design using the indirectly-heated biomass gasification process that was selected from PNNL-19009; it was designated as Case 1. Nexant modeled this Case 1 Reference Design and benchmarked the heat and material balance results using the published data from the PNNL-19009 study report. This was done in order to fill in the process design details and establish an overall utility balance for the BTM plant. The benchmarked model was then used as a detailed tool to analyze and compare against the OSU's BTS chemical looping technology. The indirectly-heated gasification process was chosen over the directly-heated gasification process as the reference design because the PNNL-19009 study concluded that it is economically more viable. If the biomass-to-methanol plant utilizing OSU's BTS chemical looping technology can perform better than the indirectly-heated biomass gasification process, it will establish OSU's BTS process as the best available biomass gasification technology.

The case of interest for this preliminary TEA study is Case 2, which is a biomass-to-methanol (BTM) production process based on OSU's BTS chemical looping technology. The specific process plant or component technology associated with both of these two BTM designs are identified and compared in a matrix as shown in Table 2-1. In comparison, Case 2 represents a significant amount of process simplification as an overall BTM process.

Table 2-1
Case Study Matrix for Biomass-to-Methanol (BTM)

Major Process Plant	Case 1[1]	Case 2
Feed Handling and Preparation		
Steam-Heated Rotary Dryer	✓	✓
Biomass Gasification Technology		
Indirectly-Heated Gasifier	✓	
OSU BTS Chemical Looping		✓
Tar Reforming		
Bubbling Fluidized-Bed Reactor	✓	NR[2]
Syngas Cleanup		
LO-CAT & Zinc Oxide Sulfur Removal	✓	✓
Steam Reforming	✓	NR[2]
Amine Unit for CO_2 Removal	✓	✓
Methanol Production	✓	✓
Power Generation via Steam Turbine	✓	✓

[1] Reference Case based on Nexant's benchmark simulation of the Biomass-to-Methanol case in PNNL-19009

[2] NR = Not Required.

2.2.1 Case 1: Reference Indirectly-Heated Gasifier Biomass-to-Methanol Plant

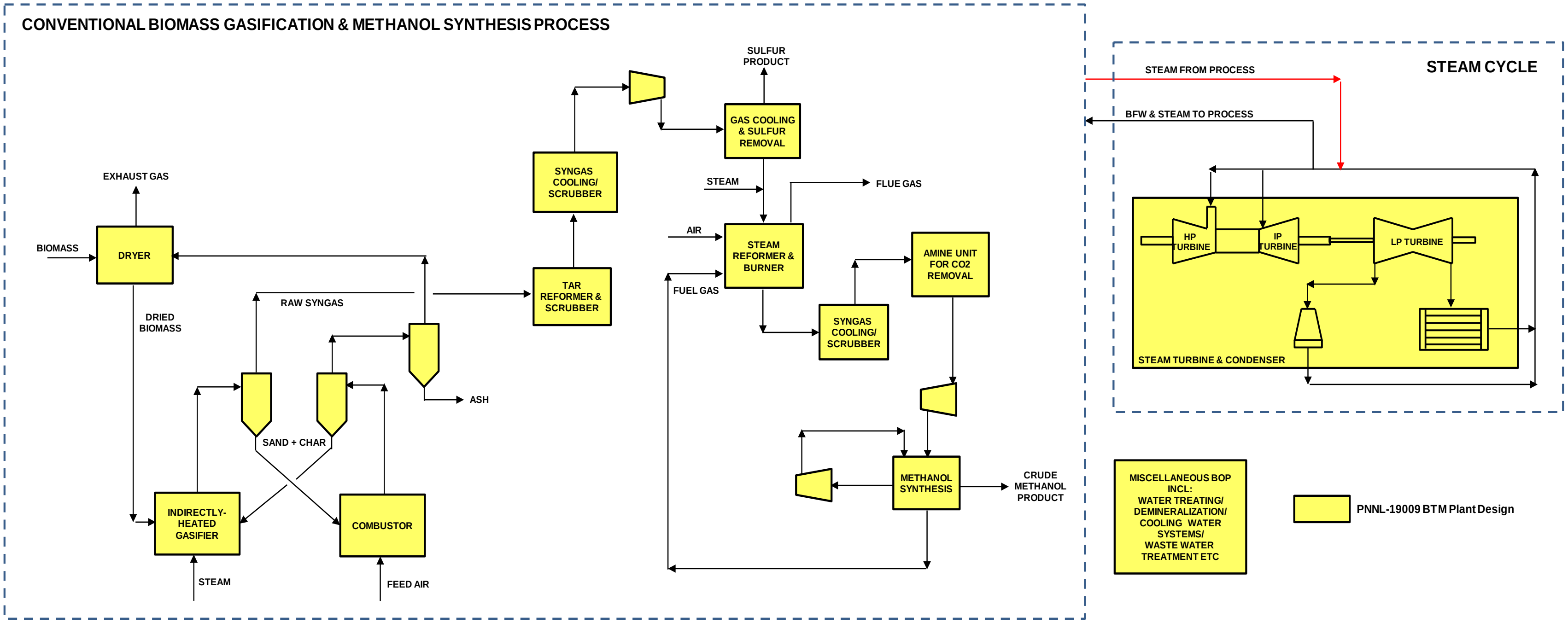
The indirectly-heated gasification biomass-to-methanol (IHGBTM) plant Reference Design uses hybrid poplar wood chips as feed. It is designed for processing 2,200 dry tons per day (tpd) of biomass.

In this process, biomass is assumed to be received in the form of wood chips with a moisture content of 50 wt%. The wood chips are dried in a directly heated rotary drier to a moisture content of 12 wt% before feeding to the gasifier. For the indirectly heated gasifier, the heat from the char combustor hot flue gas is used to produce steam and dry the biomass. Wood chips are converted to raw gas in a gasifier. The raw syngas is sent onto a tar reformer and scrubber. A large portion of the tar and light hydrocarbons in the raw syngas react with steam in the tar reformer and is converted to CO and H_2. The syngas is then sent to a wet scrubbing unit to remove the particulates in the syngas. The scrubbed syngas is compressed and sulfur in the syngas is removed. The clean syngas is sent to a steam reformer where methane and other light hydrocarbon gases are reformed to CO and H_2 in the desired ratio for methanol synthesis. The reformed syngas is cooled and then goes through an amine-based absorption system to remove CO_2. The purified syngas is then compressed to the desired pressure for methanol synthesis. Both the gasification and methanol synthesis processes generate a large amount of heat that is recovered, in the form of steam for process requirements and power generation. In the PNNL-

19009 design, a methanol purification process is included in order to produce a high purity chemical grade methanol as the product. For the current TEA study, the interest is in producing a crude methanol product for fuel applications.

A block flow diagram (BFD) of the Case 1 (IHGBTM) plant design is shown in Figure 2-1. This Reference Design, along with the Case 2 BTM plant using OSU's BTS chemical looping technology, are assumed to operate with an annual on-stream factor of 90 percent or 7,884 hrs/year at full capacity.

Figure 2-1
Case 1: Indirectly-Heated Gasification BTM Plant Design - Simplified BFD



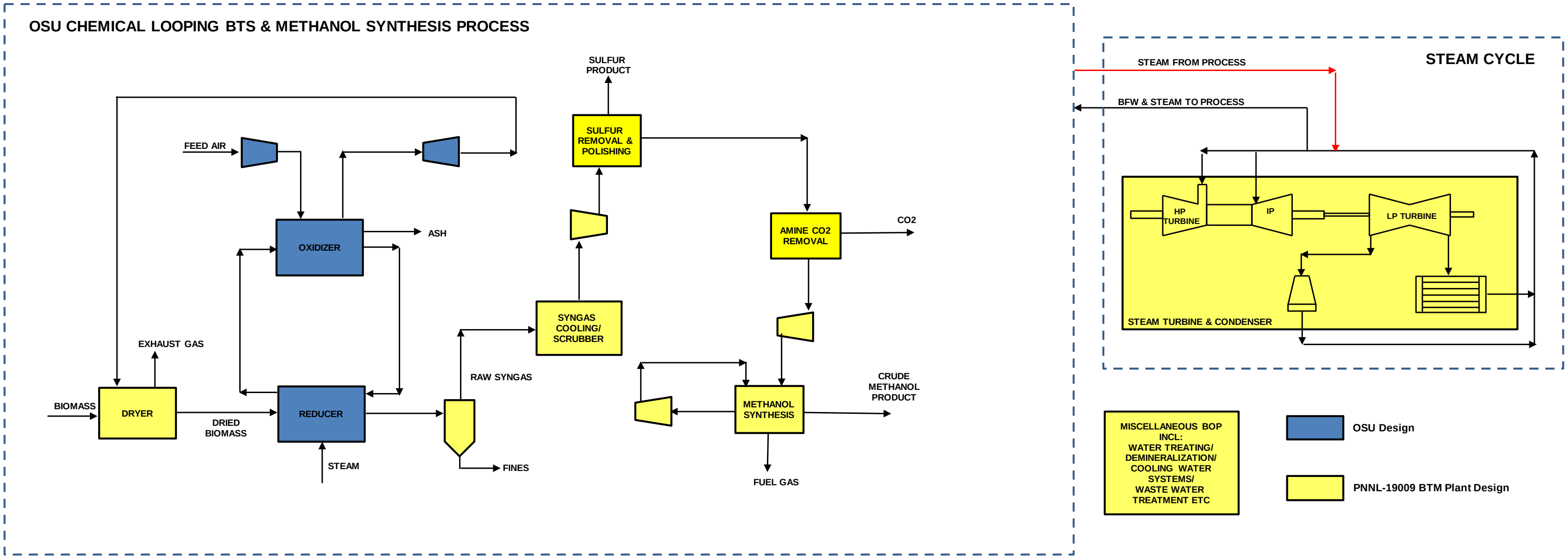
2.2.2 Case 2: OSU BTS Chemical Looping System Biomass-to-Methanol Plant

Case 2 represents an integrated OSU’s BTS chemical looping BTM plant design. Similar to Case 1, the Case 2 BTM plant feed rate is fixed at 2,200 dry tpd of biomass.

In Case 2, gasification of biomass is accomplished through a combination of co-current moving bed reactor and iron-titanium composite metal oxide (ITCMO) oxygen carrier developed at OSU. The multifunctional nature of the ITCMO oxygen carrier developed at OSU allows for both biomass gasification and tar cracking to syngas in a single reactor system, allowing for conversion of biomass to syngas with a H_2:CO ratio of 2:1 in a single step without the need of tar and steam reforming that are associated with the conventional biomass gasification technology of IHGBTM.

The Case 2 BTM plant BFD is shown in Figure 2-2. This figure serves to demarcate the battery limit and highlights the interfaces between OSU’s proprietary systems (colored in blue) and the rest of the BTM processes (in yellow) that are derived from BTM case from the PNNL-19009 report. The blue blocks represent OSU’s chemical looping reactors, which replaces the indirectly-heated gasifier, tar reformer and steam reformer in the reference Case 1 BTM plant.

Figure 2-2
Case 2: OSU BTS Chemical Looping BTM Plant - Simplified BFD



2.3 DESIGN CRITERIA

2.3.1 Heat and Material Balance

Per agreed upon design basis and study approach, Nexant carried out a simulation of the Case 1 and Case 2 BTM design to estimate their overall process heat and material balances (HMB). The detailed set of HMB are needed to determine the overall plant utility balance, including process waste heat recovery, generation from the power cycle, as well as cooling water load breakdown, etc., all of which lead to a better estimation of the overall BTM plant performance and costs.

The Case 1 Reference IHGBTM Design was simulated largely based on process data and information provided in the PNNL-19009 study report. Engineering design adjustments were made in some situations to ensure the reasonableness of the data provided. The resulting stream flows, heat and material balances, methanol production and power generation from the steam cycle were benchmarked and cross-checked against the PNNL-19009 BTM data to ensure that the results are within reasonable agreement. The Case 1 Reference Design benchmarked model provides a more coherent breakdown of the overall BTM plant process utility requirements and balance of plants than what had been provided in the PNNL-19009 study report. These are needed to assess how best to establish a similar BTM plant design, using OSU's BTS technology. The results of the Case 1 IHGBTM Reference Design are presented in more detail in Section 4.

Based on the Case 1 Reference Design results, Nexant provided OSU with the stream conditions and flows to the BTS chemical looping system. Using these as inputs, OSU developed a set of HMB around its proprietary BTS system, determined its overall utilities consumption and equipment sizes. These outputs were then transmitted to Nexant to complete the BTM plant simulation modeling of Case 2.

2.3.2 OSU BTS Chemical Looping System

To complete the Case 2 BTS BTM modeling and carry out its preliminary techno-economic analysis, OSU provided Nexant with the major stream flows and conditions into, within, and out of their proprietary BTS chemical looping process in order to enable Nexant to integrate these processes into its model. OSU also provided Nexant with a cost of their BTS chemical looping process. These were used as inputs to establish the overall BTM plant performance, total project investment and variable operating costs for Case 2.

2.3.3 Other Systems

The costs for the remaining BTM plant systems not directly related to the OSU BTS chemical looping process were developed by Nexant via capacity factor from the corresponding system costs derived from the PNNL study references cited in Section 2.1.

2.4 BIOMASS CHARACTERISTICS

As cited previously, design biomass feed to the BTM power plant is based on hybrid poplar wood chips with 50 wt% moisture content. The composition of the wood chips is listed in Table 2-2. The wood chips are dried to 12 wt% moisture in a rotary dryer before feeding to the biomass gasifier.

Table 2-2
Biomass Feedstock Composition

Feedstock	Hybrid Poplar Wood Chips
Main composition, dry basis, wt%	
Carbon	50.60
Hydrogen	6.08
Oxygen	40.75
Nitrogen	0.61
Sulfur	0.02
Chlorine	0.01
Ash	1.93
Total	100.00
Moisture, wt% as received	50
HHV, Btu/lb, dry basis	8,590

It should be noted that the biomass HHV shown in Table 2-2, at 8,590 Btu/lb, is about 5% higher than the HHV of 8,178 Btu/lb, as stated in the PNNL-19009 report. In Nexant's review of the PNNL-19009 BTM process, it was determined that in order to arrive at the gasifier syngas outlet temperature of 1,598 °F as shown in the PNNL-19009 report, it was necessary to increase the biomass HHV to 8,590 Btu/lb. This value is in-line with the Boie correlation[1] that was used as the default method in calculating the HHV of solid fuel. The estimated HHV using the Boie correlation is at 8,574 Btu/lb.

2.5 POWER GENERATION & AUXILIARY LOADS

The PNNL-19009 report provided a limited breakdown of the BTM reference case (Indirectly-Heated Gasifier BTM Plant) auxiliary loads, as well as steam turbine power generation. For their estimates, Nexant determined the power generation based on its steam cycle model, which takes into account the process steam generation from waste heat recovery in the BTM island, as well as steam consumptions in both the BTM and power islands. Auxiliary loads were estimated, wherever applicable, by pro-rating from the PNNL-19009 BTM reference case using relevant scaling parameters obtained from the model's heat and material balance.

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[1] Biomass HHV calculated by Boie correlation = $100*(151.2*x_{carbon} + 499.77*x_{hydrogen} + 45*x_{sulfur} - 47.7*x_{oxygen} + 27*x_{nitrogen}) - 189 = 8573.9$ Btu/lb

Table 2-3 shows the power production and auxiliary load breakdown of the PNNL-19009 study BTM reference plant, of which Case 1 model were benchmarked against. There are other process units that consume power but not shown in the PNNL; these were estimated in more detail based on the model H&M balances.

Table 2-3
PNNL-19009 BTM Reference Case Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	DOE Report PNNL 19009
Gas Turbine Power	0
Gas Expander Power	0
Steam Turbine Power	23,300
TOTAL POWER, kWe	**23,300**
AUXILIARY LOAD SUMMARY, kWe	
Feed Handling & Preparation	
Biomass Handling & Preparation	0
Gasification	
Char Combustion Air Blower	4,700
Char Combustion Flue Gas Blower	0
Syngas Cleanup & Compression	
ReformerCombustor Air Compressor	1,200
Reformer Flue Gas Blower	2,400
Scrubber Water Makeup Pump	0
Scrubber Condensate Transfer & Recirculation Pump	8
Scrubber Syngas Compressor	15,900
Syngas Compressor Condensate Pump	0
Miscellaneous Syngas Cleanup	0
Acid Gas Removal & Methanol	0
Amine AGR / Lo-Cat /ZnO	0
Lo-Cat Regeneration Air Compressor	0
Methanol Feed Compressor	3,900
Methanol Purge Gas Recycle Compressor	2,200
Power Generation	
Steam Turbine Auxiliaries	400
Miscellaneous Balance of Plant	0
TOTAL AUXILIARIES, kWe	**30,708**
NET POWER, kWe	**-7,408**
Thermal Efficiency, % LHV (Crude Methanol)	52%
(Refined Methanol)	**48%**

2.6 COOLING WATER AND AIR COOLING

With both simulation cases modeling, cooling water is minimized through the use of air fin coolers where applicable, as was done in the PNNL-19009 study. The minimum stream cooling temperature by air is assumed to be 150 °F (66 °C) and the minimum cooling temperature by cooling water is assumed to be 110 °F (43 °C). The steam turbine condenser is assumed to be an air-cooled condenser to minimize water demand.

2.7 OTHER SITE SPECIFIC REQUIREMENTS

Although the following design parameters are considered site-specific, and were not quantified for this preliminary TEA study, allowances for normal conditions and construction shall be included in the cost estimates as appropriate, for future analysis.

- Flood plain considerations
- Existing soil/site conditions
- Water discharges and reuse
- Rainfall/snowfall criteria
- Seismic design
- Buildings/enclosures
- Fire protection
- Local code height requirements
- Noise regulations – Impact on site and surrounding area.

Per agreed upon study basis and approach, the following assumptions and methodology were used to estimate the Total Project Investment (TPI) and Operation and Maintenance (O&M) costs for the BTM plant. The estimates are based on 2011 costs, similar to the PNNL-19009 study.

3.1 BTM PLANT CAPITAL COST ESTIMATION METHODOLOGY

3.1.1 Case 1 Installed Equipment Cost Estimate Criteria

For the BTM plant, the PNNL-19009 report provided cost results, in 2008 dollars that were broken down into only six (6) major systems (Table 3-1). There are no further cost breakdown details beyond these six major system categories in the report. Their cost for methanol synthesis plant included a process purification step to produce a chemical grade methanol, which had to be separated it out, as only crude methanol production is of interest in the current TEA study.

Table 3-1
PNNL-19009 BTM Cost Estimate Basis

	PNNL	PNNL
COST YEAR	2008	Escalated to 2011 (by CEPCI index ratio)
CAPITAL COSTS	$MM	$MM
ISBL		
Feed Prep & Drying	$11.20	$11.40
Gasification with Tar Reforming / Heat Recovery / Scrubbing	$15.20	$15.47
Syngas Cleanup & Compression	$28.70	$29.21
Methanol Synthesis & Purification	$12.60	$12.83
Steam System & Power Generation	$9.40	$9.57
OSBL		
Balance of Plant & Offsites	$2.20	$2.24
Total Purchase Equipment Cost (TPEC)	$79.30	$80.72
Other Direct Costs		
Total Installed Cost (TIC)	$195.87	$199.38
Total Indirect Cost	$99.92	$101.71
Fixed Capital Investment		
Land Purchase Cost		
Working Capital		
Total Project Investment	$295.79	$301.08

A separate report, “*Process Design and Economics for the Conversion of Lignocellulosic Biomass to Hydrocarbons via Indirect Liquefaction*, March 2015, NREL/TP-5100-62402, PNNL-23822”, provides a more detail individual equipment costs for the equipment associated with converting biomass via indirectly-heated gasification to high-octane hydrocarbons, of which methanol is a precursor to the production of these hydrocarbons. These data were used as a supplementary data basis for the current TEA study.

3.1.2 Case 2 Installed Equipment Cost Estimate Criteria

For Nexant modeling of the PNNL reference BTM plant, capital cost for systems where additional cost details are required, scaling ratios were drawn from cost details described in the PNNL-23822 report to the extent possible. In general, this cost estimation methodology involved using the scaling parameters and size ratio (capacity factor), as well as the reference cost and baseline capacity from the PNNL-23822 report. Once these have been established, the capital cost can be estimated based on the revised capacity from the HMB developed by Nexant’s model of the Case 1 BTM plant.

For the OSU’s BTM plant, costs for equipment in the OSU BTS chemical looping system that are proprietary were provided by OSU. Nexant used and reported these costs on an as-provided basis.

3.1.3 Total Capital Investment Estimation Methodology

The total capital investment were factored from installed equipment costs per the same methodology as that used in the PNNL-19009 report and shown here in Table 3-2 below.

Table 3-2
PNNL-19009 BTM Reference Case Project Investment Factors

Project Investment Factors	% 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%
Indirect Costs	
Engineering	32%
Construction	34%
Legal and Contractors Fees	23%
Project Contingency	37%
Total Indirect	126%
Total Project Investment	373%

3.2 OPERATION & MAINTENANCE COSTS

The operation and maintenance (O&M) costs are pertaining to those charges associated with operating and maintaining the plant over its expected life. These costs include:

- Operating labor
- Maintenance – material and labor
- Administrative and support labor
- Consumables
- Fuel
- Waste disposal

There are two O&M cost components; fixed O&M, which is independent of methanol production, and variable O&M that is proportional to methanol production. Variable O&M costs were estimated based on an on-stream factor of 90%.

3.2.1 Fixed Costs

Fixed operating costs consider employee salaries and benefits, overhead, plant maintenance costs, insurance and taxes. The fixed operating costs for the current TEA study were estimated based on the O&M cost methodology in the DOE/NETL report 341/020514, "Baseline Analysis of Crude Methanol Production from Coal and Natural Gas", July 29, 2014. The number of operators and other fixed costs basis are as shown in Table 3-3 below.

Table 3-3
Nexant IHGBTM Reference Design Fixed Operating Cost Basis

Fixed Operating Cost Components		Remarks
Operating Labor Requirements per Shift	Positions	
Skilled Operator	2	
Operator	6	
Foreman	1	
Laboratory technicians	2	
Total Operating Positions	11	
2011 Base Hourly Rate	$39.7 / hr	
Labor Burden	30%	
Other Fixed Operating Costs		
Administrative / Support Labor	25%	of O&M labor
Maintenance Materials + Labor	2.4%	of TPI
Maintenance Labor	40%	of maintenance materials + labor
Insurance and taxes	2%	of TPI

3.2.2 Variable Costs

The cost of consumables, including fuel, were determined based on the individual rates of consumption, the unit cost of each specific consumable commodity, and the plant annual operating hours. Waste quantities and disposal costs were evaluated similarly to the consumables.

The unit costs for major consumables and waste disposal were selected from the PNNL-19009 and PNNL-23822 report. For this study, it is assumed that the 2011 delivered cost of the hybrid poplar chips is at $67.51/dry ton, as escalated from the 2008 unit costs for wood residues listed in the PNNL-19009 report.

3.3 BTM PLANT FINANCIAL MODELING BASIS

3.3.1 Required Selling Price (RSP)

The figure-of-merit used to evaluate the BTM plant's overall financial performance is the methanol RSP, expressed in $/gal of crude methanol. All costs are expressed in "first-year-of-construction" year dollars, and the resulting RSP is also expressed in "first-year-of-construction" year (2011) dollars. The RSPs for each case are calculated assuming a financial structure that is representative of a commercial fuels project, with a financial structure with assumptions as shown in Table 3-4.

Table 3-4
Financial Assumptions for Methanol RSP Calculation

Parameter	Value
TAXES	
Income Tax Rate	38% (Effective, 34% Federal , 6% State)
Capital Depreciation	20 years, 150% declining balance
Investment Tax Credit	None
Tax Holiday	None
FINANCING TERMS	
Repayment Term of Debt	15 years
Grace Period on Debt Repayment	None
Debt Reserve Fund	None
TREATMENT OF CAPITAL COSTS	
Capital Cost Escalation During Construction (nominal annual rate)	3.6%
Distribution of Total Overnight Capital over the Capital Expenditure Period (before escalation)	3 Year Period: 10%, 60%, 30%
Working Capital	Zero for all parameters
% of Total Overnight Capital that is Depreciated	100%
INFLATION	
RSP, O&M, Fuel Escalation (nominal annual rate)	3.0% RSP, O&M, Fuel
FINANCIAL STRUCTURE (COMMERCIAL FUELS PROJECTS)	
Type of Security	**Percent of Total**
Debt	45
Equity	55

A simplified capital charge factor (CCF)-based financial modeling methodology, as reported in the 2014 DOE NETL 341/020514 study of "Baseline Analysis of Crude Methanol Production from Coal and Natural Gas" and also described in the 2014 DOE/NETL 2011/1455 "Quality Guideline for Energy System Studies: Cost Estimation Methodology for NETL Assessments of Power Plant Performance", is used to calculate the methanol product RSP. The CCF used is 0.111 and it was pre-calculated by DOE NETL based on a detailed discounted cash flow financial model using the financial structure of a commercial project with assumptions as presented in Table 3-4 above.

The simplified equation used to calculate methanol product RSP is shown in the equation below:

$$RSP = \frac{(CCF)(TOC) + OC_{fix} + (CF)(OC_{var})}{(CF)(gal/yr\ MeOH)}$$

where:

CCF = Capital Charge Factor
TOC = Total Overnight Cost
OC_{fix} = Fixed Operating Cost
CF = Capacity Factor
OC_{var} = Variable Operating Cost

The total overnight cost (TOC), as defined, comprises of the total project investment (TPI) cost plus all owner's costs. The owner's costs are defined to consist of the followings:

- Preproduction costs, including startup, consisting of –
    - 6 months of all labor cost
    - 1 month of maintenance materials
    - 1 month of non-fuel consumables including power imports
    - 1 month of waste disposal
    - 25% of 1 month fuel cost at 100% capacity factor
    - 2% of TPI cost to cover other expenses
- Inventory capital –
    - 60 day supply of fuel and consumables
    - 0.5% of TPI cost
- Initial cost for catalyst and chemicals
- Land cost
- Other owner's costs at 15% of the TPI cost to cover project development costs such as feasibility study, front-end engineering design, economic development, legal and permitting, owner's engineering cost, and etc.
- Financing Costs at 2% of the TPI cost.

Section 4 Nexant Benchmark of PNNL Reference IHGBTM Design

The main objective of this preliminary techno-economic analysis (TEA) is to evaluate the techno-economic benefits of OSU's BTS process for fuel-grade methanol production in comparison with a conventional biomass-to-methanol process based on a conventional low-pressure indirectly-heated biomass gasification technology. As agreed upon, the conventional indirectly-heated gasification biomass to methanol (IHGBTM) process to be used as the Reference Design of which to compare against the OSU's chemical looping BTS technology was selected from the recent DOE study entitled "Techno-economic Analysis for the Thermochemical Conversion of Biomass to Liquid Fuels, June 2011, US DOE Pacific Northwest National Laboratory (PNNL-19009)." A simplified block flow diagram (BFD) of the PNNL IHGBTM Reference Design is shown in Figure 2-1, in Section 2 of this report.

Nexant modeled the PNNL IHGBTM Reference Design (Nexant Reference Design) and benchmarked the heat and material balance (HMB) results using the published data from the PNNL study report. This was done to allow the overall utility balance for the BTM plant to be developed and its overall plant performance determined. Calculated HMB stream flows were served as the basis for capital cost estimation of the various process plants within the IHGBTM design, by capital factoring from the PNNL published cost data. A brief process description of the PNNL IHGBTM design and its design assumption are presented in the section below. More details can be found in the PNNL-19009 report.

4.1 PNNL REFERENCE IHGBTM DESIGN PROCESS DESCRIPTION

The PNNL IHGBTM Reference Design used hybrid poplar wood chips as feed. The main processing plants include feed handling and preparation, indirectly-heated gasification, tar reforming and wet scrubbing, syngas compression and cleanup, steam reforming and heat recovery, CO_2 removal, methanol synthesis, steam cycle power generation and balance of plants, as depicted in Figure 2-1.

Feed Handling and Preparation - Within the feed handling and preparation step, biomass is assumed to be received with a moisture content of 50 wt%. The BTM design is based on a feed rate of 2,200 dry tons per day (2,000 dry metric tons per day) of wood chips. Pre-drying the wood chips as received feedstock is required before feeding it to the gasifier. This was done, in part, to minimize the heat load required for gasification. For the PNNL IHGBTM design, the wood chips were dried to a moisture content of 12 wt% before feeding it to the gasifier. The heat from the char combustor hot flue gas was used to dry the biomass and produce steam.

Indirectly Heated Gasification – As cited in the PNNL-19009 report, the technology was based on that developed by Battelle Columbus Laboratory, and later demonstrated by Future Energy Resources Company at Burlington, Vermont. In their TEA analysis, PNNL modeled the process using stoichiometric reactor models based on correlations developed by DOE NREL in 2005. Nexant took the same approach in our benchmarking modeling exercise.

The indirectly-heated gasification system consists of a gasifier and a separate combustor. Dried wood is fed into the gasifier that is operating at low pressure. Steam from the utility plant is used to fluidize the bed and supply a portion of the heat required for gasification. The gasifier is operated at 1598 ^{o}F (870 ^{o}C) and 23 psia. Heat is supplied by circulating hot olivine particles between the gasifier and the combustor.

A series of cyclone separators are used to remove particulates and char from the raw syngas leaving the gasifier. The olivine, char and ash are separated in the gasifier cyclone and sent onto the combustor. The raw syngas from the gasifier cyclone is sent to the tar reformer. The char is burned in the combustor to reheat the olivine. The flue gas from the combustor is sent to the primary combustor cyclone separator where 99.9% of the olivine are separated and returned to the gasifier. The hot flue gas containing residual fines and ash is sent to a secondary combustor cycle separator for final recovery and disposal. The hot flue gas is sent onto Feed Handling and Preparation plant for biomass drying.

Tar Reforming and Wet Scrubbing – As cited in PNNL-1009 study, during conventional biomass gasification, a small fraction of the biomass will be converted into tars that consist mainly of functionalized aromatic and poly-aromatic hydrocarbons. These tars not only will foul the downstream process equipment if allowed to condense, but they also contain significant quantities of carbon and hydrogen, which are important sources for CO and H_2. Thus, cracking these tars into small hydrocarbons is preferred over condensing them in the downstream water scrubber. The process takes place in a tar reformer of which the technology has been investigated by DOE NREL. Tar reforming is a catalytic process in which the tar, along with methane and other light hydrocarbons in the raw syngas react with steam and converted into CO and H_2, while NH_3 is decomposed to N_2 and H_2.

In their study, PNNL modeled the tar reformer process using a stoichiometric reactor model, based on a reactor yield correlation developed by DOE NREL. The gas enters the tar reformer at the gasifier outlet temperature and exits the tar reformer at about 1,383 ^{o}F (750 ^{o}C). But, Nexant found that the PNNL tar reformer outlet gas composition does not correspond to the equilibrium composition at 1,383 ^{o}F (750 ^{o}C). A lower reformer outlet temperature of 1,223 ^{o}F (662 ^{o}C) was calculated. Therefore, additional heat would be required to bring the tar reformer temperature to 1,383 ^{o}F (750 ^{o}C). Nexant modeled the tar reformer as a Gibbs/equilibrium reactor model and adjusted the approach to equilibrium to yield a reformer outlet temperature of 1,383 ^{o}F (751 ^{o}C). This approach yields a tar reformer outlet gas composition which is different than the PNNL gas composition. The composition, however, can be adjusted in the steam methane reformer downstream to yield a similar methanol synthesis feed composition. After the tar and other organic impurities are reduced in the tar reformer, the raw syngas is cooled to 390 ^{o}F (199 ^{o}C) via heat exchange. The raw syngas is then scrubbed to remove impurities such as particulates, and residual NH_3 and tars.

Syngas Compression and Cleanup – The syngas leaving the scrubber is compressed to about 450 psia using a centrifugal compressor with interstage cooling. The compressed gas contains a sufficient amount of methane and other light hydrocarbons, which represents a significant fraction of the total carbon and hydrogen in the biomass feed. Thus, steam reforming is used to convert these compounds to CO and H_2. Before reforming, however, the syngas has to be cleaned to remove the H_2S and other organic sulfur (COS, CS_2, mercaptans) that will poison the reforming catalysis. This is accomplished via a Lo-Cat process, follow by a ZnO sulfur polishing bed.

Lo-Cat is a liquid-phase oxidation bulk sulfur removal process, commercially available and licensed by Merichem. Lo-Cat was selected for H_2S removal because it can handle the low H_2S concentration in a biomass generated syngas. In this process, H_2S is absorbed from a gas stream in a liquid solution and directly oxidized to sulfur in the solution. The syngas is heated to about 120 ^oF (49 ^oC) and then sent to the Lo-Cat process where the H_2S is absorbed and converted to elemental sulfur in an iron chelate-based solution. In the PNNL study, the Lo-Cat process is assumed to remove the sulfur in the syngas to a concentration of 10 ppm H_2S.

Following the Lo-Cat process, a fixed-bed ZnO desulfurization unit is used to polish the syngas by reducing the sulfur level to less than 1 ppm. ZnO sulfur polishing is a common commercial practice. H_2S is captured by the ZnO sorbent as shown:

$$ZnO + H_2S \rightarrow ZnS + H_2O$$

In this study, both the Lo-Cat and ZnO sulfur polishing process were modeled as a component separation block.

Steam Methane Reforming (SMR) and Heat Recovery – Syngas leaving the ZnO sulfur polishing plant is sent onto a steam methane reformer to convert the remaining methane and light hydrocarbon to CO and H_2. Steam methane reforming is a well-established commercial catalytic process of which the two main reactions are represented as follows:

$$C_nH_m + nH_2O \rightarrow (n+m/2)H_2 + nCO \quad \text{(Reforming), and}$$

$$CO + H_2O \rightarrow CO_2 + H_2 \quad \text{(Water gas shift)}$$

High temperature steam from the offsite utility plant is used as feed to the reformer and to adjust and obtain a proper H_2:CO ratio required by the downstream methanol synthesis step. Reforming is endothermic and the heat required for the process was supplied by combustion of the off-gas from the methanol synthesis plant, supplemented by imported natural gas, as needed. Waste heat from the reforming plant is recovered to generate steam for in-plant use and power generation. The reforming process was modeled using a Gibbs reactor.

CO_2 Removal – The syngas leaving the reforming plant is cooled and then sent onto an amine unit to remove the CO_2 before compressing it to a higher pressure required for methanol synthesis. While not specific mentioned in the PNNL report, it is assumed that the design is based on a methyl diethanolamine (MDEA) CO_2 capture process of which the technology is commercially available and in wide practice. 99% removal of the CO_2 from the reformed syngas is assumed. After CO_2 removal, the syngas is compressed and sent onto methanol synthesis. The MDEA CO_2 removal process is modeled as a component separation block.

Methanol Synthesis – The compressed syngas, combined with the internal recycled stream from the methanol reactor, is sent onto the methanol synthesis plant of which the principle reactions are:

$$CO + 2H_2 \rightarrow CH_3OH$$

$$CO_2 + 3H_2 \rightarrow CH_3OH + H_2O$$

The first reaction is the main methanol synthesis reactor. There are minor side reactions leading to the formation of higher alcohols and dimethyl ether as byproducts. These reactions include:

$$nCO + 2nH_2 \rightarrow C_nH_{2n+1}OH + (n-1)H_2O$$

$$2CH_3OH \rightarrow H_2O + CH_3OCH_3$$

Syngas based methanol production is a well-developed commercial process. In the PNNL study, methanol synthesis is assumed to take place in the gas phase at 500 ^oF (260 ^oC) and at 850 psia (59 bar), using a fixed-bed reactor, packed with ZnO/CuO catalyst. Waste heat is recovered via generating a medium pressure steam. The vapor phase methanol product is cooled by heat exchange with the compressed syngas feed and then further cooled by air and cooling water. When cooled, the product stream is sent to a HP and a MP flash tank where the liquid raw methanol product is separated from the non-condensable gases. Methanol synthesis has a relatively low per pass conversion, recycling of the unconverted syngas is required in order to achieve a reasonable yield for the process. It was stated in the PNNL study that ninety-five percent of the vapor phase from the product stream is compressed and recycled back to the methanol reactor.

The methanol synthesis unit was modeled as a two stage synthesis reactor (Gibbs) with interstage cooling. The reactor effluent from the second stage was cooled to 130 ^oF (54 ^oC) and flashed at high pressure of 795 psia (55 bar) followed by medium pressure flash at 415 psia (29 bar). The flashed vapor is compressed and recycled as reactor feed. About 5% of the flashed gas is purged to remove the inert. The high pressure purge gas contained appreciable fuel value. It was sent through a power recovery turbine before used as fuel for the SMR combustor. The liquid from the medium pressure flash was collected as crude methanol product.

Further methanol refining by distillation would be required if chemical grade methanol product is desired. The PNNL study included a methanol product purification step to produce a high purity chemical grade methanol. This was deemed not necessary for the current TEA study, as the objective is to produce a raw methanol as a fuel blending stock

Power Generation and Balance of Plants – Different pressure levels of steam were generated and consumed throughout the BTM process. These were balanced and controlled via a utility header within a steam cycle power generation system. Major steam flows include high pressure superheated (HPSH) steam generated from syngas and combustor flue gas cooling in the gasification, SMR and tar reforming units. The HPSH steam is used to generate electric power in a steam turbine generator (STG). IP steam is extracted from the STG for injection to the SMR. LP steam is extracted for process reboiling and deaerator stripping uses. MP steam is also generated in methanol product gas cooling for injection to the gasifier and the SMR unit. The power generation/steam system was modeled as a three stage (HP/IP/LP) steam turbine generator (STG) with boiler feed water (BFW) deaeration and BFW pumps. A power generation/steam system flow diagram is shown in Figure 4-2 in section 4.4.4.

4.2 NEXANT'S BENCHMARK OF PNNL IHGBTM DESIGN (NEXANT REFERENCE DESIGN)

The PNNL IHGBTM design was simulated to establish a Reference Design with which it can be used to compare and evaluate the benefits of OSU's BTS technology. Nexant's simulation model was developed following the design details and data provided in the PNNL-19009 study report. While the PNNL-19009 report provides a reasonable set of HMB and cost estimates for the overall IHGBTM process, design details within each process systems are not always complete or available. Hence, some Nexant process design inputs are necessary to complete the modeling. The IHGBTM Reference Design model results are benchmarked (compared) against the data and performance results report in the PNNL-19009 study report.

Benchmarked results include process material balance, power balance (performance), overall carbon balance, overall sulfur balance, overall cooling water duty, overall water flow diagram, Total Project Investment (TPI), operating (O&M) cost, and methanol required selling price (RSP) are presented below.

4.3 NEXANT IHGBTM REFERENCE DESIGN MATERIAL BALANCE

The block flow diagram shown in Figure 4-1 depicts the flow scheme for the Nexant IHGBTM Reference Design. The corresponding ASPEN simulation process material balance major stream flows are shown in Table 4-1. The description for each process and power blocks was provided in section 4.1.

Figure 4-1
Block Flow Diagram – Nexant IHGBTM Reference Design Model

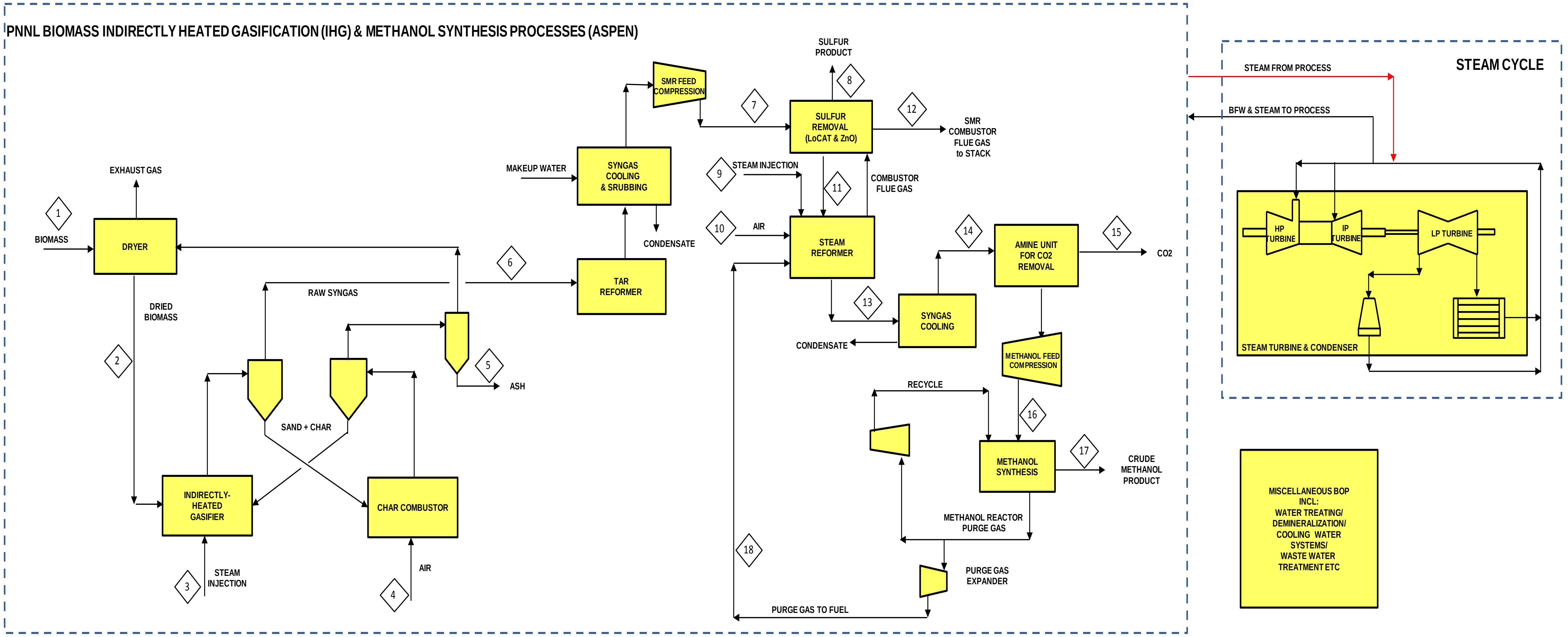


Table 4-1
Nexant IHGBTM Reference Design Model Stream Flows

Stream No.		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18
Description		As Received Biomass	Dried Biomass	Steam Injection to Gasifier	Air to Char Combustor	Ash from Gasifier	Raw Syngas from Gasifier	Syngas to Sulfur Removal	Sulfur Product	Steam Injection to SMR	Air to SMR Furnace	Feed to SMR	SMR Combustor Flue Gas to Stack	Syngas from SMR	Cooled SG to CO2 Removal	CO2 Vent	Methanol Synthesis Feed Gas	Crude Methanol Product	Methanol Purge Gas to Fuel
Vapor/Liquid Flows, lbmoles/hr	Mol Wt																		
AR	39.95	-	-	-	155.34	-	-	-	-	-	84.14	-	84.14	-	-	-	-	-	-
CH4	16.04	-	-	-	-	-	992.12	964.27	-	-	-	964.27	0.00	351.65	351.65	-	351.65	14.19	337.46
CO	28.01	-	-	-	-	-	2,606.75	2,171.22	-	-	-	2,171.22	0.00	2,914.85	2,914.85	-	2,914.85	2.74	445.72
CO2	44.01	-	-	-	5.51	-	828.42	2,142.45	-	-	2.98	2,142.45	881.72	2,011.36	2,010.86	1,809.77	201.09	24.07	66.45
COS	60.07	-	-	-	-	-	-	0.02	-	-	-	-	-	0.00	0.00	-	0.00	0.00	0.00
H2	2.02	-	-	-	-	-	1,427.25	4,524.12	-	-	-	4,524.12	0.00	6,227.58	6,227.57	-	6,227.57	1.49	961.54
H2O	18.02	-	-	4,058.79	165.77	-	5,562.25	25.12	-	4,977.94	89.79	25.12	1,784.94	4,521.59	32.11	32.11	-	112.30	0.40
H2S	34.08	-	-	-	-	-	1.15	1.12	-	-	-	-	0.00	0.00	0.00	-	0.00	0.00	0.00
N2	28.01	-	-	-	12,988.92	-	-	39.86	-	-	7,035.67	39.86	7,073.05	38.80	38.80	-	38.80	0.74	38.06
NH3	17.03	-	-	-	-	-	80.01	0.06	-	-	-	0.06	0.00	2.17	0.35	-	0.35	0.27	0.08
O2	32.00	-	-	-	3,484.46	-	-	-	-	-	1,887.42	-	464.41	0.00	-	-	-	-	-
SO2	64.06	-	-	-	-	-	-	0.00	-	-	-	0.00	0.00	0.00	-	-	-	-	-
S	32.06	-	-	-	-	-	-	-	1.12	-	-	-	-	0.00	-	-	-	-	-
HCN	27.03	-	-	-	-	-	-	0.00	-	-	-	0.00	-	0.01	0.01	-	0.01	0.01	0.00
C	12.01	-	-	-	-	70.11	-	-	-	-	-	-	-	-	-	-	-	-	-
CL2	70.91	-	-	-	-	-	-	-	-	-	-	-	0.00	0.00	-	-	-	-	-
NO	30.01	-	-	-	-	-	-	-	-	-	-	-	1.43	0.00	0.00	-	0.00	-	0.00
NO2	46.01	-	-	-	-	-	-	-	-	-	-	-	0.01	0.00	-	-	-	-	-
HCL	36.46	-	-	-	-	-	0.52	0.52	-	-	-	0.52	0.31	0.52	0.52	-	0.52	0.20	0.31
CH3OH	32.04	-	-	-	-	-	-	0.00	-	-	-	0.00	-	0.00	0.00	-	0.00	2,543.73	29.02
C2H2	26.04	-	-	-	-	-	26.34	0.00	-	-	-	0.00	-	0.00	0.00	-	0.00	0.00	0.00
C2H4	28.05	-	-	-	-	-	280.82	0.00	-	-	-	0.00	-	0.01	0.01	-	0.01	0.00	0.01
C2H5OH	46.07	-	-	-	-	-	-	0.00	-	-	-	0.00	-	0.00	0.00	-	0.00	2.05	0.02
C3H8O-01	60.10	-	-	-	-	-	-	-	-	-	-	-	-	0.00	0.00	-	0.00	0.03	0.00
C2H6	30.07	-	-	-	-	-	18.85	0.01	-	-	-	0.01	-	0.03	0.03	-	0.03	0.00	0.02
C3H8	44.10	-	-	-	-	-	-	0.00	-	-	-	0.00	-	0.00	0.00	-	0.00	0.00	0.00
C4H10-1	58.12	-	-	-	-	-	-	0.00	-	-	-	0.00	-	0.00	0.00	-	0.00	0.00	0.00
DIMET-01	46.07	-	-	-	-	-	-	-	-	-	-	-	-	0.00	0.00	-	0.00	0.00	0.00
BENZE-01	78.11	-	-	-	-	-	8.18	-	-	-	-	-	-	0.00	0.00	-	0.00	0.00	0.00
NAPHT-01	128.17	-	-	-	-	-	14.96	-	-	-	-	-	-	0.00	-	-	-	-	-
FE2O3	159.70	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
FE3O4	231.55	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
FE0.947O	52.89	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
FE	55.85	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
FETIO3	151.75	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
TIO2	79.90	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Solid Flows, lbs/hr																			
Biomass		367,437	208,771	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ash		-	-	-	-	3,546	-	-	-	-	-	-	-	-	-	-	-	-	-
Total lbmoles/hr		-	-	4,058.79	16,800.00	70.11	11,847.61	9,868.76	1.12	4,977.94	9,100.00	9,867.62	10,290.01	16,068.57	11,576.74	1,841.88	9,734.86	2,701.83	1,879.08
Total lbs/hr		367,437	208,771	73,120	484,798	4,388	241,582	181,324	36	89,679	262,599	181,284	287,378	270,963	190,030	80,226	109,804	85,026	24,779
Mole Wt		-	-	18.02	28.86	62.59	20.39	18.37	32.07	18.02	28.86	18.37	27.93	16.86	16.41	43.56	11.28	31.47	13.19
Temperature, degF		60	241	457	59	1,598	1,598	110	300	634	59	707	100	1,466	110	110	300	131	66
Pressure, psia		25.00	23.00	450.00	14.70	23.00	23.00	465.00	439.00	450.00	14.70	450.00	15.00	430.00	406.50	14.70	905.00	415.00	25.00
Enthalpy, MMBtu/hr		(1,673.71)	(573.61)	(412.01)	(20.34)	0.65	(696.41)	(497.42)	0.09	(496.76)	(11.01)	(448.51)	(314.62)	(773.53)	(491.00)	(310.36)	(168.19)	(279.70)	(45.96)
H2/CO		-	-	-	-	-	0.55	2.08	-	-	-	2.08	1.29	2.14	2.14	-	2.14	0.54	2.16
H2+CO+CO2, lbmols/hr		-	-	-	5.51	-	4,862.42	8,837.78	-	-	2.98	8,837.78	881.72	11,153.80	11,153.28	1,809.77	9,343.50	28.30	1,473.70
CH4, mol%		-	-	-	-	-	8.4	9.8	-	-	-	9.8	0.0	2.2	3.0	-	3.6	0.5	18.0
C4 and Heavier HC, mol%		-	-	-	-	-	2.9	0.0	-	-	-	0.0	-	0.0	0.0	-	0.0	0.0	0.0
CO2, mol%		-	-	-	0.0	-	7.0	21.7	-	-	0.0	21.7	8.6	12.5	17.4	98.3	2.1	0.9	3.5
H2S, ppmV		-	-	-	-	-	97	114	-	-	-	-	0	0	0	-	0	0	0
HCl, mol%		-	-	-	-	-	0.004	0.005	-	-	-	0.005	0.003	0.003	0.004	-	0.005	0.008	0.017
					Note 1				Note 2		Note 1					Note 1		Note 3	

Notes: 1) P & T at atmospheric condition
2) Molten sulfur
3) P& T from MP flash.

4.4 NEXANT IHGBTM REFERENCE DESIGN PERFORMANCE

The overall performance results of the Nexant IHGBTM reference design are shown in Table 4-2 below. For comparison, the DOE/PNNL IHGBTM performance is also shown. In general, the performance of the Nexant Reference Design, as modeled, is in alignment with that of the PNNL report, except at a slightly lower overall plant efficiency due to a combined effect of a lower power generation from the steam turbine power plant and higher estimated plant auxiliary loads.

The Nexant IHGBTM reference design thermal efficiency is 51% vs. 52% for the DOE/PNNL IHGBTM on a LHV basis. The 1% efficiency difference can be attributed to the lower steam turbine power output (about 21%) and a higher total auxiliary load (about 18%) as modeled by the Nexant Reference Design. Both designs have to import power to satisfy in-plant requirement, with an estimated 17.98 MWe for the Nexant Reference Design.

The total power produced for the Nexant Reference Design is 18.3 MWe vs 23.3 MWe for the DOE/PNNL 19009 design, as reported. The difference is mainly due to a lower steam turbine power output of 7.667 MWe for the Nexant Reference Design. A methanol purge gas expander was incorporated in the Nexant Reference Design to recover additional 2.679 MWe of power before sending the purge gas to the steam reformer plant as fuels, a feature that was not specifically mentioned or accounted for in the PNNL report.

The difference in steam turbine power output is probably due to the differences in accounting for the syngas and combustor flue gas heat recovery, as part of the overall heat integration balance of plants of which the DOE/PNNL 19009 report did not provide sufficient details to follow and understand what had been done in their modeling

The Nexant Reference Design has a higher overall estimated auxiliary load than what was reported for the DOE/PNNL design, by about 5.9 MWe. Approximately 2.6 MW of which is due to selected loads accounted for in the Nexant Reference Design, as it was modeled, but were not shown or mentioned in the DOE/PNNL 19009 report. These included:

Biomass Handling & Preparation – 742 kWe vs 0 kWe
Misc. Syngas cleanup (Air Fans & Solution Pumps) – 204 kWe vs 0 kWe
AGR (MDEA Circ. Pumps) – 778 kWe vs 0 kWe
Miscellaneous Balance of Plant (CW Cir. Pumps) – 874 kWe vs 0 kWe

The remaining load differences are in the compression power and pump requirements. These are possibly due to the differences in heat integration exchanger configuration, compressor efficiency, and pressure drop assumptions used for the calculations. The differences are shown as follows:

Char Combustion Air / Flue Gas blowers – 5,237 kWe vs 4,700 kWe
Reformer Combustor Air / Flue Gas blowers – 1,977 kWe vs 3,600 kWe
Scrubber SG Compressor – 19,731 kWe vs 15,900 kWe
Methanol Purge Gas Recycle Compression – 2,510 kWe vs 2,200 kWe.

Table 4-2
Nexant IHGBTM Reference Design Model Plant Performance Summary

Performance Summary BTM Plant	Nexant IHGBTM Model	DOE Report PNNL 19009
Feed		
Dry Wood Chips, Tons/D	2,205	2,200
AR Wood Chips (50% Moisture), Tons/D	4,409	4,409
Natural Gas, SCFH	0	0
Products		
Crude Methanol, lbs/hr	85,026	86,629
Tons/D	1,020	1,040
MMGal/Year	113	115
Methanol (100%), lbs/hr	81,507	82,958
Power Consumption, kW		
Gas Turbine Power	0	0
Gas Expander Power	2,679	0
Steam Turbine Power	15,633	23,300
Power Generation, kWe	**18,313**	**23,300**
AUXILIARY LOAD SUMMARY, kWe		
Feed Handling & Preparation		
Biomass Handling & Preparation	742	0
Gasification		
Char Combustion Air Blower	4,324	4,700
Char Combustion Flue Gas Blower	914	0
Syngas Cleanup & Compression		
ReformerCombustor Air Compressor	1,315	1,200
Reformer Flue Gas Blower	662	2,400
Scrubber Water Makeup Pump	63	0
Scrubber Condensate Transfer & Recirculation Pump	13	8
Scrubber Syngas Compressor	19,731	15,900
Miscellaneous Syngas Cleanup	204	0
Acid Gas Removal & Methanol		
Amine AGR / Lo-Cat /ZnO	778	0
Lo-Cat Regeneration Air Compressor	28	0
Methanol Feed Compressor	4,096	3,900
Methanol Purge Gas Recycle Compressor	2,510	2,200
Power Generation		
Steam Turbine Auxiliaries	370	400
Miscellaneous Balance of Plant	874	0
Total Auxiliary Loads, kWe	**36,621**	**30,708**
Net Power, kWe	**-18,309**	**-7,408**
Thermal Efficiency, % LHV (Crude Methanol)	**50.6%**	**51.7%**
Carbon Efficiency, % (Crude Methanol)	**33%**	**34%**
Water Demand/Discharge		
Water Demand/Discharge, gpm		
Cooling Tower Makeup	705	530
Boiler Feed Water Makeup	351	424
Scrubber Water Makeup	198	0
Total Water Demand	1,254	954
Waste Water, gpm	176	363

4.4.1 Nexant IGHBTM Reference Design Model Overall Carbon Balance

Table 4-3 shows the carbon balance for the Nexant IHGBTM Reference Design model. The carbon input to the plant consists of carbon in the air in addition to carbon in the biomass. Carbon leaves the plant as unburned carbon in the slag, methanol product, and as CO_2 in the stack gas and vent gas.

Table 4-3
Nexant IHGBTM Reference Design Model Carbon Balance

Overall Carbon Balance, lb/hr	**In**	**Out**
Biomass Feed	**92,962**	
Combustion Air	102	
Makeup Water	0	
MeOH Product		31,096
CO2 Removed		21,737
Combustion Flue Gas		39,382
Waste Water		7
Slag & Sulfur		842
Deaerator Vent		0
LoCAT Regenerator Vent		0
Convergence Tolerance		-1
Total	**93,064**	**93,064**

4.4.2 Nexant IHGBTM Reference Design Model Overall Sulfur Balance

Table 4-4 shows the sulfur balance for the Nexant IHGBTM reference design. Sulfur input comes solely from the sulfur in the biomass. Sulfur output includes the sulfur recovered in the LoCAT and the ZnO units, and sulfur emitted in the stack gas. Sulfur in the slag is assumed to be negligible.

Table 4-4
Nexant IHGBTM Reference Design Model Overall Sulfur Balance

Overall Sulfur Balance, lb/hr	In	Out
Biomass Feed	**37**	
Combustion Air	0	
Makeup Water	0	
MeOH Product		0
CO2 Removed		0
Combustion Flue Gas		0
Waste Water		0
Slag & Sulfur		37
Deaerator Vent		0
LoCAT Regenerator Vent		0
Convergence Tolerance		0
Total	**37**	**37**

4.4.3 Nexant IHGBTM Reference Design Model Cooling Water and Cooling Tower Requirements

Table 4-5 shows the comparison of the cooling water duty breakdowns for the Nexant IHGBTM reference design vs. the DOE/PNNL 19009 design. As shown, major water cooling duty is mainly from the tar reformer, syngas cooling and compression, steam reforming, amine overhead condenser and the methanol synthesis plants. As with the PNNL design, air cooling is used for steam turbine condensate cooling. As shown in Table 4-5, the total IHGBTM reference design cooling water (CW) duty, as modeled by Nexant, is significantly higher than that reported in the DOE/PNNL 19009 study. The main difference is in the scrubber overhead syngas compression cooling, the amine CO_2 stripper overhead condenser and the methanol synthesis unit syngas cooling requirements. The Scrubber overhead syngas compression cooling and the amine overhead condenser duties were not shown in the DOE/PNNL 19009 report which accounted for 215 MMBtu/hr of CW duty. In Nexant's reference design, the methanol synthesis gas is trimmed cooled from 207 ^{o}F (97 ^{o}C) to 130 ^{o}F (54 ^{o}C) using cooling water only. This cooling water duty could possibly be reduced by splitting the duty between air and cooling water, with an increase in capital cost.

Table 4-5
Nexant IHGBTM Reference Design Model Cooling Water Loads

Process Cooling Water Duties (MMBtu/hr)	**Nexant IHGBTM Model**	**DOE Report PNNL 19009**
Tar Reformer SG Cooling	77	87
Scrubber Overhead SG Compression	79	NA
LoCAT	0.1	3
Steam Reformer	9	9
Methanol Production	62	24
CO2 Stripper Ovhd Condenser	136	0
Total Process Cooling Water Duty	364	123
STG Surface Condenser Duty *	0	0
Total CW & CT Loads, MMBtu/hr	364	123
* Air Cooled		

4.4.4 Nexant IHGBTM Reference Design Model Overall Water/Steam Diagrams

Figure 4-2 shows the overall water flow diagram for the Nexant IHGBTM Reference Design Model.

The water demand comes mainly from the scrubber, deaerator and the cooling tower water makeup requirements. The total water demand is 1,253 gpm. Internal condensate and BFW blowdown are recycled to make up 487 gpm of the water demand. The balance of 766 gpm is supplied by water withdrawal from water sources. Process water discharge to waste treatment facility consists of 176 gpm of cooling tower blowdown.

The total water demand of 1,253 gpm (Table 4-2) compares to 954 gpm for the DOE/PNNL 19009 report. The PNNL 19009 water demand consisted mainly of cooling tower and boiler feed water makeup requirements. No details were shown for internal recycle. However, the waste water discharge of 363 gpm (Table 4-2) is higher compared to the 176 gpm from the Nexant Reference Design which may indicate a lower internal water recycle rate.

Figure 4-3 shows the overall steam flow diagram for the Nexant IHGBTM Reference Design Model. High pressure superheated steam (840 psig, 900 ^{o}F) is generated from heat recovery from the hot char combustor flue gas, tar reforming and the SMR units and is fed to the STG for power generation. The STG is a 3-stage turbine generator (HP, MP and LP). Total power generated is 15.6 MWe.

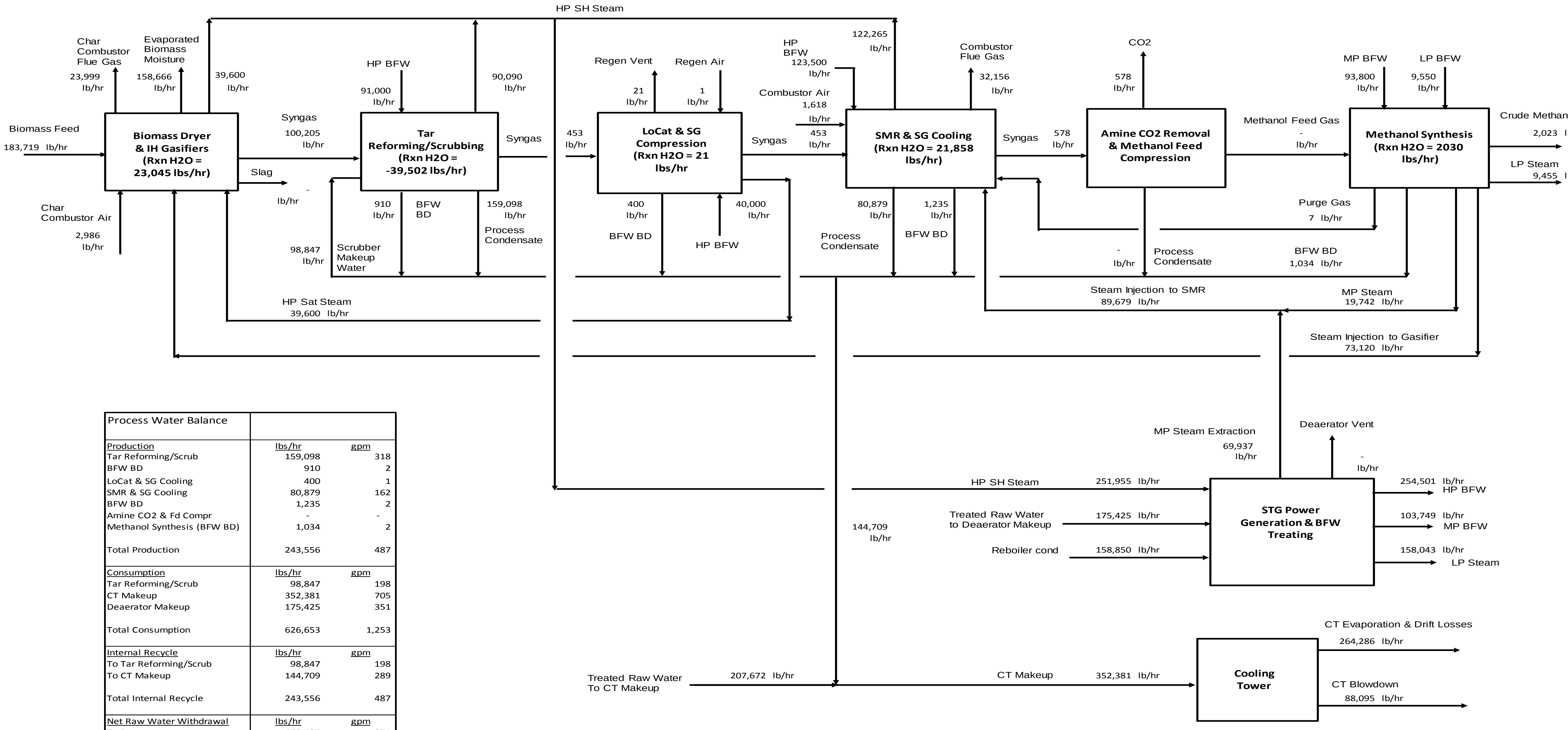
Medium pressure steam (435 psig, 684 ^{o}F) is extracted at the HP turbine exhaust for the SMR steam requirement. Low pressure steam is extracted at the MP turbine exhaust for amine reboilers and other LP steam users.

MP steam is also generated by syngas heat recovery to provide steam for the gasifier and the SMR requirements.

The STG LP condensate, process condensates and makeup water are sent to the deaerator where the mixture is deaerated by stripping with a small amount of steam and treated for boiler feed water (BFW) uses. Makeup water from storage is heated by recovering process heat before mixing with the condensates. The deaerator operates at 7 psig and 228 ^{o}F.

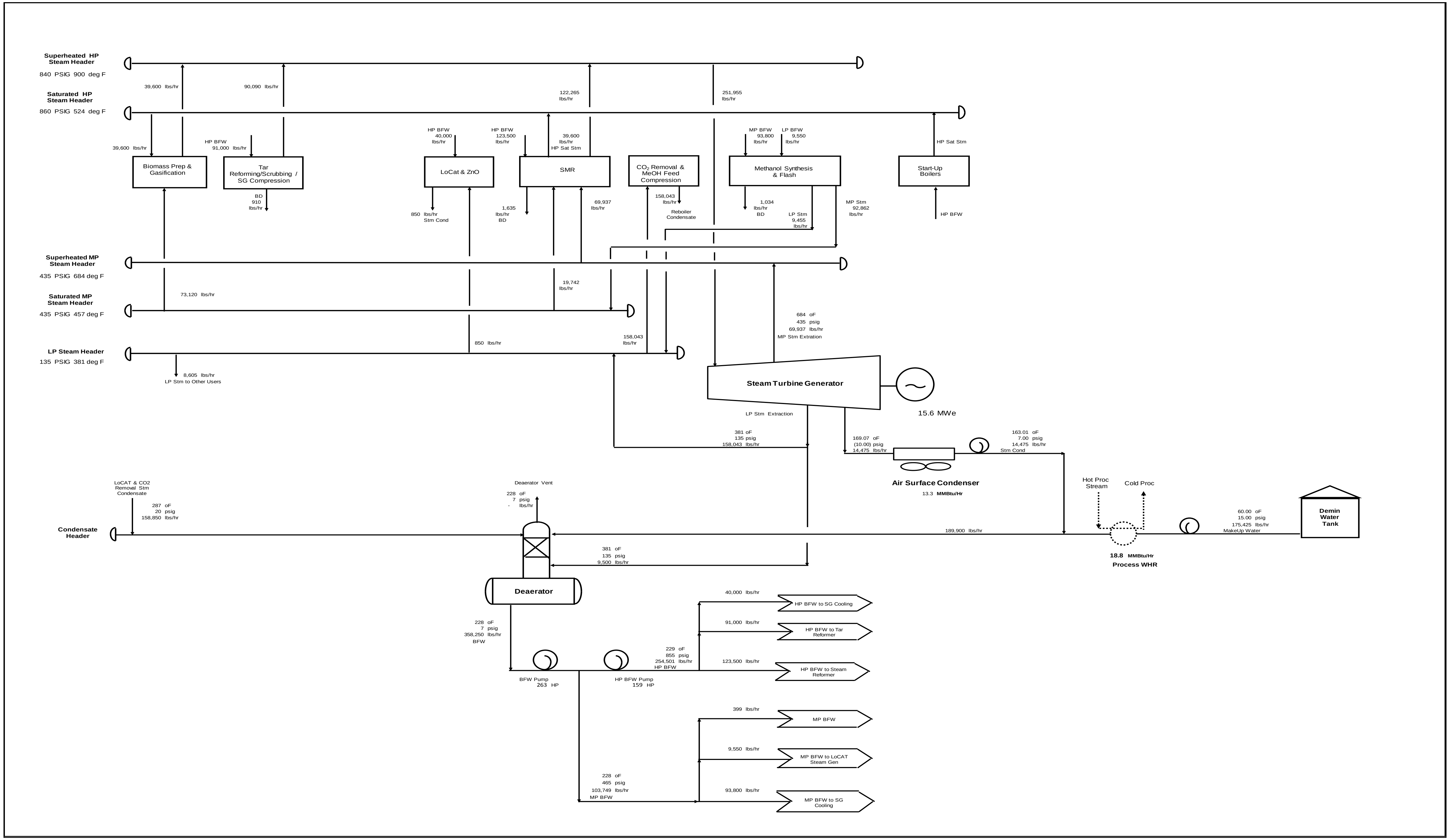
Figure 4-2

Nexant IHGBTM Reference Design Model Overall Water Flow Diagram



Process Water Balance		
Production	lbs/hr	gpm
Tar Reforming/Scrub	159,098	318
BFW BD	910	2
LoCat & SG Cooling	400	1
SMR & SG Cooling	80,879	162
BFW BD	1,235	2
Amine CO2 & Fd Compr	-	-
Methanol Synthesis (BFW BD)	1,034	2
Total Production	243,556	487
Consumption	lbs/hr	gpm
Tar Reforming/Scrub	98,847	198
CT Makeup	352,381	705
Deaerator Makeup	175,425	351
Total Consumption	626,653	1,253
Internal Recycle	lbs/hr	gpm
To Tar Reforming/Scrub	98,847	198
To CT Makeup	144,709	289
Total Internal Recycle	243,556	487
Net Raw Water Withdrawal	lbs/hr	gpm
To Deaerator	175,425	351
To CT Makeup	207,672	415
Total Raw Water Withdrawal	383,097	766
Net Discharge to Waste Water	lbs/hr	gpm
From CT BD	88,095	176

Figure 4-3
Nexant IHGBTM Reference Design Model BFW/Steam/Steam Turbine Generator Flow Diagram



4.5 NEXANT IHGBTM REFERENCE DESIGN CAPITAL COST

Estimated capital costs are summarized in tables 4-6, 4-7 and 4-8.

Table 4-6 shows the Total Project Investment (TPI) cost summary for the Nexant Reference Design which was developed from the total purchased equipment costs (TPEC) from the DOE/PNNL 19009 report and was supplemented by cost details from the DOE/PNNL 23822 report, per methodology as described in section 3.1. The cost basis year for the PNNL 19009 costs was 2008. These were escalated to year 2011 using the chemical plant cost index. Costs are broken down and presented in a format that is similar to what DOE had done in their 2014 “Baseline Analysis of Crude Methanol Production from Coal and Natural Gas” study report. In this format, the relative cost contribution to the overall methanol production can be separated and compared. The estimated Total Project Investment (TPI) for the Nexant Reference Design is $299 MM.

As stated in section 3.1, the DOE/PNNL 19009 report costs were broken down into only six major systems and needed to be supplemented by cost details from DOE/PNNL 23822. For example, in the PNNL 19009 report, the TPEC for the gasification system was lumped together with the tar reforming, heat recovery, scrubbing, syngas compression and sulfur removal costs in a single sum of $15.2MM. In order to better identify and distribute the cost, equipment cost details from PNNL 23822 were used to distribute the $15.2MM TPEC into the following process units:

- Gasification
- Tar Reforming & Scrubbing
- SMR Feed Compression
- Sulfur Removal (LoCAT & ZnO)

Additionally, it was not certain that the cost of the steam methane reformer (SMR) was included in the PNNL 19009 report. Therefore, for the Nexant Reference Design, the SMR cost is added to the total TPEC cost using Nexant’s SMR cost data.

Table 4-7 shows a comparison of the TPI cost as determined for the Nexant IHGBTM Reference Design with that was reported in the DOE/PNNL 19009 study. The TPI cost for the reported PNNL value of $292 MM has been adjusted for crude methanol production, excluding the methanol purification cost.

The following key TPI differences are noted:

- Gasification with Tar Reforming/Heat Recovery/Scrubbing - ($71 MM vs $58 MM) – $13 MM higher TPI due to the inclusion of SMR cost in the Nexant Reference Design, and
- Steam System & Power Generation ($28 MM vs $36 MM) – $8 MM lower TPI due to the smaller steam and STG system for the Nexant Reference Design.

Table 4-8 shows the addition of owner's costs to determine the total overnight cost (TOC) which is used to calculate the required sales price (RSP) for methanol, per methodology as described in Section 3.3. The TOC includes preproduction costs, inventory capital, initial costs for catalyst and chemicals, land and financing costs. The estimated TOC is $383 MM.

Table 4-6
Nexant IHGBTM Reference Design Model Total Project Investment Estimate Summary

Cost Basis – Year 2011

Item/Description	Totral Purchased Equipment Cost (TPEC)	Bulk*	Purchased Equipment Installation	Total Installed Cost (TIC)	Eng'g CM H.O., Fees	Contingencies	TOTAL PROJECT INVESTMENT	
	$1,000	$1,000	$1,000	$1,000	$1,000	$1,000	$1,000	$/TPD MeOH
BIOMASS PREP & DRYING	$11,400	$12,313	$4,446	$28,159	$10,146	$4,218	$42,524	$41,677
GASIFICATION WITH TAR REFORMING/HEAT RECOVERY/SCRUBBING								
Gasification	$6,166	$6,659	$2,405	$15,229	$5,487	$2,281	$22,998	$22,540
Tar Reforming & Scrubbing	$4,733	$5,112	$1,846	$11,691	$4,212	$1,751	$17,654	$17,303
SMR Feed Compression (to 460 psia)	$3,901	$4,213	$1,521	$9,636	$3,472	$1,443	$14,552	$14,262
Sulfur Removal (LoCAT & ZnO)	$672	$726	$262	$1,660	$598	$249	$2,507	$2,457
Steam Methane Reforming (SMR)	$3,553	$3,838	$1,386	$8,777	$3,163	$1,315	$13,254	$12,990
SUBTOTAL 2.	$19,026	$20,548	$7,420	$46,993	$16,933	$7,039	$70,965	$69,553
SYNGAS CLEANUP & COMPRESSION								
Aminne CO2 Recovery	$23,970	$25,887	$9,348	$59,205	$21,333	$8,869	$89,407	$87,627
Methanol Feed Compression	$4,935	$5,330	$1,925	$12,189	$4,392	$1,826	$18,407	$18,040
SUBTOTAL 3.	$28,904	$31,217	$11,273	$71,394	$25,725	$10,695	$107,814	$105,667
METHANOL SYNTHESIS & PURIFICATION								
Methanol Synthesis (Crude Methanol)	$10,088	$10,895	$3,934	$24,918	$8,979	$3,733	$37,629	$36,880
Crude Methanol Distillation	$0	$0	$0	$0	$0	$0	$0	$0
Purge Gas Expander	$903	$976	$352	$2,231	$804	$334	$3,369	$3,302
SUBTOTAL 4.	$10,992	$11,871	$4,287	$27,149	$9,782	$4,067	$40,998	$40,182
STEAM SYSTEM & POWER GENERATION	$7,531	$8,133	$2,937	$18,601	$6,703	$2,786	$28,091	$27,531
BALANCE OF PLANT & OFFSITES (OSBL)	$2,208	$2,385	$861	$5,454	$1,965	$817	$8,236	$8,072
CALCULATED TOTAL COST	$80,061	$86,466	$31,224	$197,751	$71,254	$29,623	$298,627	$292,683

Table 4-7
Nexant IHGBTM Reference Design Model Total Project Investment Comparison

Item/Description	Nexant IHGBTM Model	DOE Report PNNL 19009
Total Project Investment (Yr 2011)	$MM	$MM
BIOMASS PREP & DRYING	$42.52	$42.52
GASIFICATION WITH TAR REFORMING/HEAT RECOVERY/SCRUBBING		
Gasification	$23.00	$23.00
Tar Reforming & Scrubbing	$17.65	$17.65
SMR Feed Compression (to 460 psia)	$14.55	$14.55
Sulfur Removal (LoCAT & ZnO)	$2.51	$2.51
Steam Methane Reforming (SMR)	$13.25	$0.00
SUBTOTAL 2.	$70.97	$57.71
SYNGAS CLEANUP & COMPRESSION		
Amine CO2 Recovery	$89.41	$90.36
Methanol Feed Compression	$18.41	$18.61
SUBTOTAL 3.	$107.81	$108.97
METHANOL SYNTHESIS & PURIFICATION		
Methanol Synthesis (Crude Methanol)	$37.63	$41.10
Crude Methanol Distillation	$0.00	$0.00
Purge Gas Expander	$3.37	$0.00
SUBTOTAL 4.	$41.00	$41.10
STEAM SYSTEM & POWER GENERATION	$28.09	$35.69
BALANCE OF PLANT & OFFSITES (OSBL)	$8.24	$6.36
CALCULATED TOTAL PROJECT INVESTMENT	$299	$292

Table 4-8
Nexant IHGBTM Reference Design Model Total Overnight Cost Summary

Owner's Costs	$1,000
Preproduction Costs	
6 months All Labor (notes 1 & 2)	$4,900
1 Month Maintenance Materials (Note 2)	$398
1 Month Non-Fuel Consumables	$1,626
1 Month Waste Disposal	$76
25% of 1 Months Fuel Cost at 100% CF	$0
2% of TPI	$5,973
Subtotal	$12,973
Inventory Capital	
60 day supply of fuel at 100% CF	$8,930
60 day supply of non-fuel consumables at 100% CF	$674
0.5% of TPI (spare parts)	$1,493
Subtotal	$11,097
Initial Cost for Catalyst and Chemicals	$6,458
Land	$900
Other Owner's Cost (note 3)	$44,794
Financing Costs	$8,063
Total Owner's Cost	**$84,285**
Total Project Investment (TPI)	**$298,627**
Total Overnight Costs (TOC)	**$382,912**

Notes:
1) Includes operating and administrative labor plus maintenance labor which is 40% of the maintenance material + labor
2) Maintenance material + labor is 2.4% of TPI
3) Other owner's cost is15% of TPI including preliminary studies, front end engineering, owner's engineers, owner's management reserve and legal & permitting costs

4.6 NEXANT IHGBTM REFERENCE DESIGN O&M COST ESTIMATES

Table 4-9 shows the estimated annual O&M cost breakdown for the Nexant Reference Design. Table 4-10 compares the operating costs between the Nexant IHGBTM Reference Design and the DOE/PNNL 19009 case. The Reference Design O&M cost estimates are essentially based on that from the DOE/PNNL 19009 report, with minor adjustments to account for different capacity in selected plants as appropriate.

The total fixed annual operating and maintenance cost which includes the annual operating, maintenance and administration labor costs, property taxes and insurance is estimated to be $15.8 MM compared to $21.1 MM which was reported in the DOE/PNNL case (Table 4-10).

The annual variable operating cost including feed and fuel is estimated at $71.6 MM, on a 90% capacity factor basis. The biomass feed cost represents almost 70% of the total cost. The non-fuel portion is $22.7 MM which includes maintenance material cost, water, catalyst and chemicals, waste disposal and power import costs. The biomass feed cost is $48.9 MM annually based on $67.5/ton of dry biomass feed on year 2011 basis. The total variable cost is $71.6 MM vs $64.3 MM for the DOE/PNNL 19009 case (Table 4-10).

Table 4-10 compares the operating costs for the two cases. The major difference is in the non-fuel variable operating cost (NFOPEX). Nexant Reference Design has a NFOPEX of $22.7 MM vs DOE/PNNL 19009 case cost of $15.8 MM. The higher electricity import accounts for $6 MM of the differences. The chemicals and catalyst costs account for the balance of the differences. The higher catalyst cost is assumed due to the inclusion of the SMR catalyst cost in the Nexant Reference Design. No detail catalyst and chemicals breakdown was available from the DOE/PNNL 19009 report.

Table 4-9
Nexant IHGBTM Reference Design Model Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES

Case:	Nexant IHGBTM Model Reference Design			
Plant Size, Crude Methanol	85,026	lbs/hr	**Cost Basis (Year)**	2011
	1,020	Tons/D	**Capacity Factor (%)**	90
	113	MMgal/year		
Primary/Secondary Fuel:	Biomass/Natural Gas			

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (base):	$39.70	$/hr
Operating Labor Burden:	30.00	% of base
Labor Overhead Charge	25.00	% of labor

Operating Labor Requirements per Shift	units/mod	Total Plant
Skilled Operator	2.0	2.0
Operator	6.0	6.0
Foreman	1.0	1.0
Lab Tech's etc	2.0	2.0
TOTAL Operating Jobs	11.0	11.0

	Annual Cost $
Annual Operating Labor Cost	$4,973,140
Maintenance Labor Cost	$2,866,822
Administration & Support Labor	$1,959,990
Property Taxes and Insurance	$5,972,546
TOTAL FIXED OPERATING COSTS	**$15,772,498**

VARIABLE OPERATING COSTS

Maintenance Material Cost	**$4,300,233**

Consumables	Consumption Initial	Consumption /Day	Unit Cost	Initial Fill Cost	
Water(/1000 gallons)	0	1,805	3.50	**$0**	**$2,073,906**
Chemicals					
MU & WT Chem (lbs)	0	97	1.22	$0	$38,972
Olivine Makeup (ton)	0	7	274.83	$0	$601,655
Amine Solvent (lbs)	0	97	1.38	$0	$44,204
Tar Reforming Catalyst (lbs)	127,113	172	8.61	$1,094,626	$486,861
Steam Reforming Catalyst (lbs)	127,113	172	27.08	$3,441,732	$1,530,794
ZnO Catalyst (lbs)	61,246	282	5.88	$360,098	$544,642
Methanol Synthesis Catalyst(lbs)	150,398	198	10.38	$1,561,518	$673,916
LoCAT Chemicals (ton)	0	1	457.88	$0	$219,671
Subtotal Chemicals				**$6,457,973**	**$4,140,716**
Other					
Supplemental Electricity (MWh consumed)	0	439	78.58	$0	$11,342,967
Gases, N2 etc.(/100scf)	0	0	0.00	$0	$0
LP Steam (/1000 lbs)	0	0	0.00	$0	$0
Subtotal Other				**$0**	**$11,342,967**
Waste Disposal:					
Flyash (ton)	0	0	0.00	$0	$0
Slag (ton)	0	54	46.62	$0	$824,702
Subtotal Waste Disposal				**$0**	**$824,702**
By-products & Emissions					
Sulfur (ton)	0	0	0.00	$0	$0
Supplemental Electricity (MWh generated)	0	0	-78.58	$0	$0
Subtotal By-Products				**$0**	**$0**
TOTAL VARIABLE OPERATING COSTS				**$6,457,973**	**$22,682,524**
Biomass (T/D, Dry)	**0**	2205	67.51	**$0**	**$48,892,032**
Natural Gas (1000 SCF)	**0**	0	9.78	**$0**	**$0**

Table 4-10
Nexant IHGBTM Reference Design Model vs. PNNL Operating Cost

OPERATING COSTS, 2011 $MM/YR	**Nexant IHGBTM Model**	**DOE Report PNNL 19009**
FIXED OPERATING COSTS		
Annual Operating Labor Cost	$5.0	NA
Maintainence Labor Cost	$2.9	NA
Administration & Support Labor	$2.0	NA
Property Taxes and Insurance	$6.0	NA
TOTAL FIXED OPERATING COSTS	$15.8	$21.1
VARIABLE OPERATING COSTS		
NON-FUEL VARIABLE OPERATING COSTS		
Maintenance Material Cost	$4.3	$5.3
Water	$2.1	Included below
Chemicals		
MU & WT Chemicals	$0.04	Included below
Chemicals & Catalysts	$4.1	$3.2
Supplemental Electricity	$11.3	$5.3
Solid Waste Disposal	$0.8	$2.1
TOTAL NON_FUEL VARIABLE OPERATING COSTS	$22.7	$15.8
FUEL		
Biomass	$48.9	$48.5
Natural Gas	$0.0	$0.0
TOTAL VARIABLE OPERATING COSTS	$71.6	$64.3
TOTAL OPERATING COSTS	**$87.3**	**$85.4**

4.7 NEXANT IHGBTM REFERENCE DESIGN METHANOL REQUIRED SELLING PRICE

Table 4-11 shows a summary comparison of the capital expenditure (CAPEX), operating expenditure (OPEX), power output, and methanol required sales price (RSP) for the Nexant Reference Design and the DOE/PNNL 19009 case. The Nexant Reference Design RSP is $1.28/gal of crude methanol compared to $1.26/gal for the DOE/PNNL 19009 case. A breakdown of the cost components for the RSP is shown in Table 4-12. The lower fixed OPEX offsets the higher cost of electricity for the Nexant IHGBTM case as shown in Table 4-12.

Table 4-11
Nexant IHGBTM Reference Design Model Plant Performance and Economic Summary

Cost & Performance	**Nexant IHGBTM Model**	**DOE Report PNNL 19009**
CAPEX, $MM		
Total Direct Cost (TDC)	$198	$195
Total Project Investment (TPI)	$299	$292
Total Overnight Cost (TOC)	$383	$375
OPEX, $MM/yr		
Fixed Operating Cost (OC_{Fix})	$16	$21
Variable Operating Cost Less Fuel (OC_{VAR})	$23	$16
Fuel Cost (OC_{Fuel})	$49	$49
Power Production, MWe		
Expander	2.7	0.0
Steam Turbine	15.6	23.3
Total Power Output	18.3	23.3
Auxiliary Power Consumption	36.6	30.7
Net Power Output	-18.3	-7.4
Power Generated, MWh/yr (MWH)	-160,384	-64,896
Thermal Efficiency,% LHV	50.6%	51.7%
Crude Methanol Production		
Tons/D	1,020	1,040
MMgal/Yr	113	115
RSP, $/gal MeOH	**$1.28**	**$1.26**

Table 4-12
RSP Cost Components

RSP Cost Components	**Nexant IHGBTM Model**	**DOE Report PNNL 19009**
CAPEX	$0.42	$0.43
Fixed OPEX	$0.16	$0.20
Variable OPEX	$0.11	$0.10
Biomass Cost	$0.48	$0.47
Power Import Cost	$0.11	$0.05
RSP Total ($/gal)	**$1.28**	**$1.26**

Section 5 OSU BTS Chemical Looping BTM Design

5.1 OSU CHEMICAL LOOPING BTM DESIGN ANALYSIS

The Nexant benchmarked ASPEN model of the PNNL BTM Reference Design, described in Section 4, was used to analyze the overall BTM plant performance using OSU's chemical looping BTS gasification technology, instead of the indirectly heated biomass gasification that was used in the PNNL Reference Design. The overall process scheme for the OSU chemical looping integrated BTM plant (OSUBTM), as originally envisioned, was presented in Figure 2-2. The syngas generation step of the indirectly-heated biomass gasification process of the PNNL Reference Design is replaced with OSU's proprietary BTS gasification system, of which its performance was modeled separately by the OSU team. The BTS gasification system performance data (heat and material flows and process conditions) provided by OSU was then used in the overall benchmarked BTM ASPEN model (replacing the indirectly-heated gasification system which includes the tar and steam methane reformers) to analyze the overall OSUBTM process performance. Major stream flow heat and material balances for the overall OSUBTM plant were developed, along with balance-of-plant utility consumptions. For most of the processing plants that are common within both the OSUBTM and the PNNL Reference Design, utilities (e.g., plant auxiliary load) and cost estimates are estimated, by pro-rating from the PNNL Reference Design data using relevant scaling parameters and the heat and material balances stream flows developed from the OSUBTM plant ASPEN model.

5.2 OSU CHEMICAL LOOPING INTEGRATED BTM PLANT (OSUBTM)

The OSUBTM design used the same hybrid poplar wood chips as feed. The main processing plants include feed handling and preparation, the Ohio State University (OSU) iron-titanium composite metal oxide (ITCMO) chemical looping gasification process, wet scrubbing, syngas compression and cleanup, heat recovery, CO_2 removal, methanol synthesis, steam generation and balance of plants. A block flow diagram (BFD) of the OSUBTM integrated plant is shown in Figure 5-2, and the corresponding ASPEN process material balance major stream flows are shown in Table 5-1. Figure 5-2 BFD serves to demarcate the battery limits and highlights the interfaces between the OSU's proprietary BTS gasification systems (colored in blue) and the rest of the BTM processes (in yellow) that are derived from the benchmark model of the PNNL Reference Design.

The BTS gasification process was modeled separately by the OSU team, and the resulting major process stream flows (Streams 3, 4, 5 & 6 of Table 5-1) were provided as input for integrating into the overall OSUBTM plant model. It is understood that the BTS gasification modeling effort is an ongoing activity and that the conditions and results used in this preliminary TEA analysis are not fully optimized. The BTS plant processes the same amount of dried biomass feed at 12wt% moisture as in the PNNL Reference Design. The provided core BTS process modeling stream flow results from OSU were integrated into the overall OSUBTM plant model to develop its overall processing and utility support requirements. The core BTS ASPEN model was adjusted slightly by Nexant to facilitate integration to the overall OSUBTM model by:

- Allowing a reasonable pressure drop across the various processing units within the

BTS plant

- Utilizing the hot flue gas leaving the oxidizer to help dry the raw biomass feed down to 12wt% moisture, and
- Preheating the required air feeding to the oxidizer.

It is noted that the above process and heat integration steps did not significantly alter the raw syngas flow and composition produced from the BTS gasification plant, as originally predicted by the preliminary OSU ASPEN BTS simulation model.

The preliminary OSUBTM results show that the overall process has to import significant amount of power. This is in contrast with the PNNL Reference Design where the indirectly-heated biomass gasification system actually produced a HP steam and it was used to generate power for in-plant use by using a steam turbine generator. For the OSUBTM design, a natural gas fired steam boiler is needed to generate steam for its in-plant use, as reflected in its BFD of Figure 5-2. The balance of plant system consists of gas-fired steam boiler, boiler feed water treatment and a cooling tower system.

5.2.1 Process Description

Most of the processing plants/units are similar to that of the PNNL Reference Design; hence the process descriptions are mostly the same. One exception is with the BTS gasification plant.

Feed Handling and Preparation - Within the feed handling and preparation step, biomass is assumed to be received with a moisture content of 50 wt%. The OSUBTM design is based on a feed rate of 2,200 dry tons per day (2,000 dry metric tons per day) of wood chips. Pre-drying the wood chips as received feedstock is required before feeding it to the gasifier. For the OSUBTM design, the wood chips were dried to a moisture content of 12 wt% before feeding it to the gasifier, as in the PNNL Reference Design. The heat from the hot BTS system oxidizer spent air and the steam boiler flue gas were used to dry the biomass.

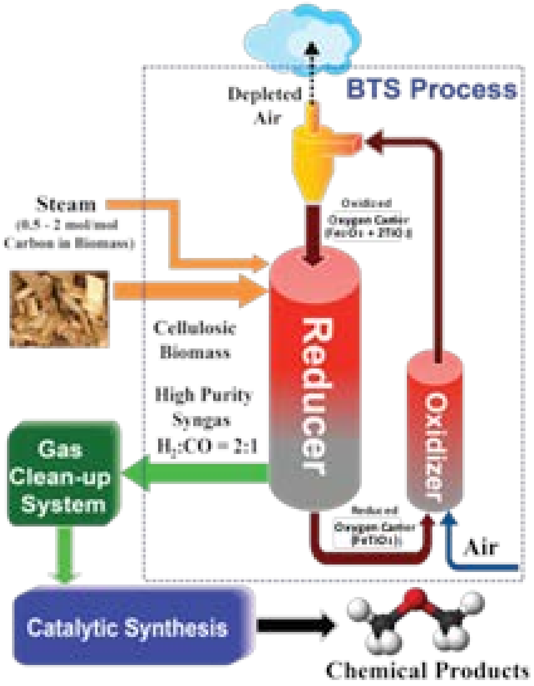
OSU Biomass to Syngas (BTS) Gasification – the technology was developed by the Ohio State University (OSU) using the iron-titanium composite metal oxide (ITCMO) chemical looping process to provide an alternative to the conventional biomass gasification processes. The main reactions involved in the BTS process are:

Reducer: $CH_xO_y + Fe_2O_3 + TiO_2 + H_2O \rightarrow CO + H_2 + FeTiO_3$

Oxidizer: $FeTiO_3 + O_2\,(Air) \rightarrow Fe_2O_3 + TiO_2$

It allows for the conversion of biomass to a tar free syngas with H_2/CO ratio of 2 in a single step without the need for an air separation unit (ASU) to provide the molecular oxygen, water gas shift (WGS) unit to adjust the H_2/CO ratio and a tar reforming unit to convert the tar to CO and H_2. A simplified schematic diagram of the BTS gasification process is shown in Figure 5-1.

Figure 5-1 OSU BTS Gasification



For this preliminary TEA analysis, OSU modeled the BTS gasifier using an ASPEN adiabatic reactor model at operating conditions developed by OSU. It is understood that the model results are preliminary and the process is not fully optimized via laboratory testing. Nexant performed the integration of the BTS gasification into the overall OSUBTM plant using the OSU syngas composition and conditions developed from this preliminary ASPEN model.

The BTS gasification process is consisted of a reducer and an oxidizer with cyclones for vapor/solid separation. Dried wood chips are fed into the reducer where steam is used to supply a portion of the heat required for gasification and for adjusting the syngas H_2/CO ratio. The reducer is operated at 1,480 ^{o}F (804 ^{o}C) and 23 psia. Heat is supplied to the reducer by circulating the oxygen carrier ITCMO between the reducer and the oxidizer.

A series of cyclone separators is used to remove particulates and ash from the raw syngas leaving the reducer. In the reducer cyclone the raw syngas is separated from the ITCMO and sent to downstream syngas cooling and scrubbing to remove hydrogen chloride and ammonia. The ITCMO is regenerated in the oxidizer using 1,010 ^{o}F (543 ^{o}C) preheated air. The flue gas from the oxidizer at 1,922 ^{o}F (1,050 ^{o}C) is sent to the oxidizer cyclone separator where the ITCMO are separated and returned to the reducer. The hot flue gas containing residual fines is sent to a secondary oxidizer cyclone separator for final recovery and disposal. The hot flue gas is sent onto Feed Handling and Preparation plant for biomass drying.

Syngas Compression and Cleanup – The syngas leaving the scrubber is cooled and compressed to about 450 psia using a centrifugal compressor with interstage cooling. The compressed gas is desulfurized to remove the H_2S and other organic sulfur (COS, CS_2, mercaptans) that will poison the methanol synthesis catalyst. Bulk of the H_2S is removed in a Lo-Cat sulfur removal unit, followed by final polishing in a ZnO catalyst bed.

Lo-Cat is a liquid-phase oxidation bulk sulfur removal process, commercially available and licensed by Merichem. Lo-Cat was selected for H_2S removal because it can handle the low H_2S concentration in a biomass generated syngas. In this process, H_2S is absorbed from a gas stream in a liquid solution and directly oxidized to sulfur in the solution. The syngas is fed to the Lo-Cat unit at 120 ^oF (49 ^oC) where the H_2S is absorbed and converted to elemental sulfur in an iron chelate-based solution. For the TEA, the Lo-Cat process is assumed to remove the sulfur in the syngas to a concentration of 10 ppm H_2S.

Following the Lo-Cat process, a fixed-bed ZnO desulfurization unit is used to polish the syngas by reducing the sulfur level to less than 1 ppm. ZnO sulfur polishing is a commonly used commercial process for protecting reformer and methanol synthesis catalysts. H_2S is captured by the ZnO sorbent as shown in the following reaction:

$$ZnO + H_2S \rightarrow ZnS + H_2O$$

In this study, both the Lo-Cat and ZnO sulfur polishing process were modeled as a component separation block.

CO_2 Removal – The syngas leaving the sulfur removal unit is cooled and then sent onto an amine unit to remove the CO_2 before compressing it to a higher pressure required for methanol synthesis. As in the PNNL Reference Design, it is assumed that the design is based on 90% CO_2 removal with the methyldiethanolamine (MDEA) CO_2 capture process of which the technology is commercially available and in wide practice. After CO_2 removal, the syngas is compressed and sent onto methanol synthesis. The MDEA CO_2 removal process is modeled as a component separation block.

Methanol Synthesis – The compressed syngas, combined with the internal recycled stream from the methanol reactor, is sent onto the methanol synthesis plant of which the principle reactions are:

$$CO + 2H_2 \rightarrow CH_3OH$$

$$CO_2 + 3H_2 \rightarrow CH_3OH + H_2O$$

The first reaction is the main methanol synthesis reaction. There are minor side reactions leading to the formation of higher alcohols and dimethyl ether as byproducts. These reactions include:

$$nCO + 2nH_2 \rightarrow C_nH_{2n+1}OH + (n-1)H_2O$$

$$2CH_3OH \rightarrow H_2O + CH_3OCH_3$$

Syngas based methanol production is a well-developed commercial process. In the PNNL study, methanol synthesis is assumed to take place in the gas phase at 500 °F (260 °C) and at 850 psia (59 bar), using a fixed-bed reactor, packed with ZnO/CuO catalyst. Waste heat is recovered via generating a medium or low pressure steam. The vapor phase methanol product is cooled by heat exchange with the compressed syngas feed and then further cooled by air and cooling water. When cooled, the product stream is sent to a HP and a MP flash tank where the liquid raw methanol product is separated from the non-condensable gases. Methanol synthesis has a relatively low per pass conversion, recycling of the unconverted syngas is required in order to achieve a reasonable yield for the process. It was stated in the PNNL study that ninety-five percent of the vapor phase from the product stream is compressed and recycled back to the methanol reactor.

The methanol synthesis unit was modeled as a two stage synthesis reactor (REquil) with interstage cooling. The reactor effluent from the second stage was cooled to 130 °F (54 °C) and flashed at high pressure of 795 psia (55 bar) followed by medium pressure flash at 415 psia (29 bar). The flashed vapor is compressed and recycled as reactor feed. About 5% of the flashed gas is purged to remove the inert. The high pressure purge gas contained appreciable fuel value. It was sent through a power recovery turbine before used as fuel for the steam boiler. The liquid from the medium pressure flash was collected as crude methanol product.

Further methanol refining by distillation would be required if chemical grade methanol product is desired. The PNNL study included a methanol product purification step to produce a high purity chemical grade methanol. This was deemed not necessary for the current TEA study, as the objective is to produce a raw methanol as a fuel blending stock. Both the OSUBTM and the PNNL Reference Design produce a raw methanol as the final product.

Balance of Plants – Different pressure levels of steam were generated and consumed throughout the OSUBTM process. These were balanced and controlled via a utility header within a steam system. Due to the relatively low gasifier syngas temperature 1,480 °F (804 °C) and high MP/LP steam demand for the gasification and CO_2 removal processes, there was not enough high level heat available for high pressure steam generation.

A steam boiler is needed to supplement the heat available from the syngas and flue gas sources to provide for various feed preheating and steam generation requirements. The primary fuel for the steam boiler is from the methanol synthesis purge gas. Natural gas is required to supplement the purge gas for fueling the steam boiler.

Figure 5-2 Block Flow Diagram – OSU Chemical Looping Biomass Gasification to Methanol Plant

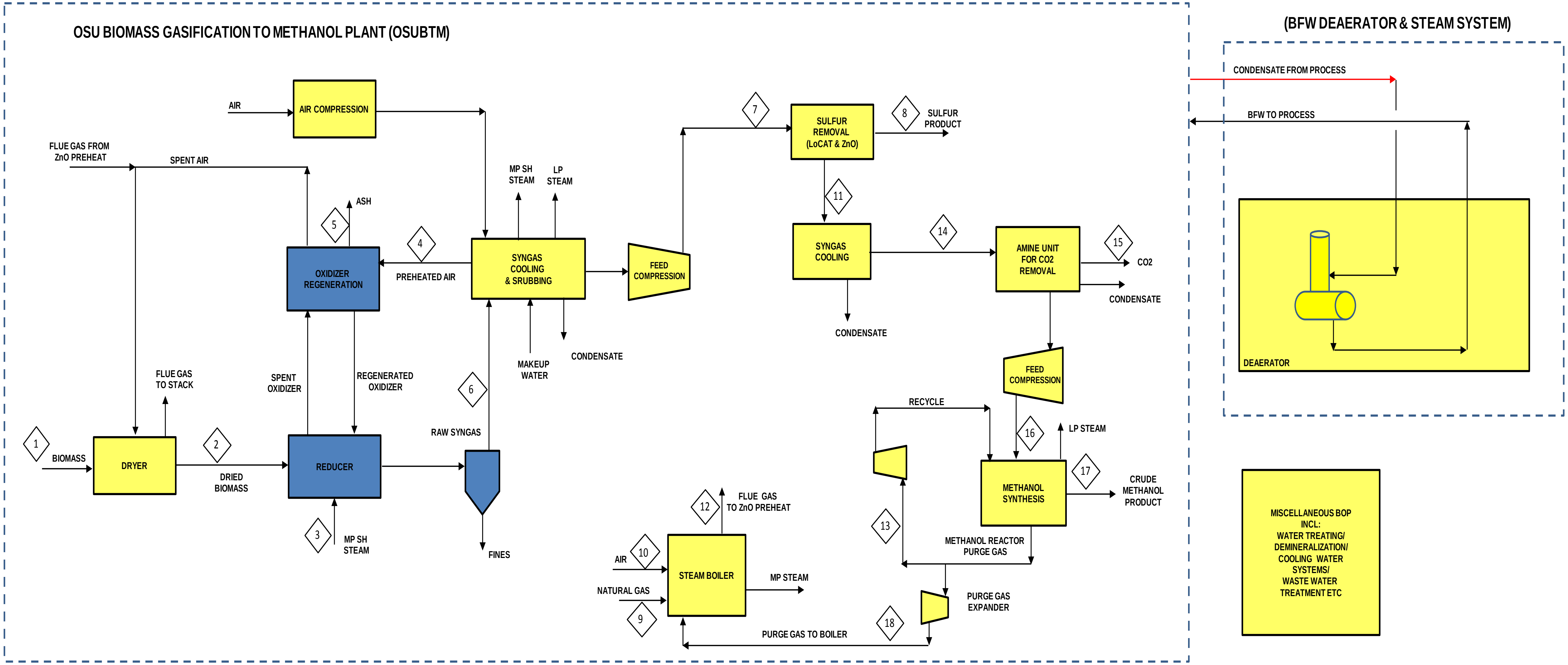


Table 5-1
OSUBTM Process Stream Flows

Stream No.		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18
Description		As Received Biomass	Dried Biomass	Steam Injection to Gasifier	Air to Oxidizer	Ash from Gasifier	Raw Syngas from Gasifier	Syngas to Sulfur Removal	Sulfur Product	Natural Gas to Steam Boiler	Air to Steam Boiler	Clean SG to Cooling	Flue Gas to ZnO Preheat	MeOH Flashed Gas Recycle	Cooled SG to CO2 Removal	CO2 Vent	Methanol Synthesis Feed Gas	Crude Methanol Product	Methanol Purge Gas to Fuel
Vapor/Liquid Flows, lbmoles/hr	Mol Wt																		
AR	39.95	-	-	-	141.84	-	-	-	-	-	110.96	-	110.96	-	-	-	-	-	-
CH4	16.04	-	-	-	-	-	1.12	1.12	-	698.25	-	1.12	0.00	19.83	1.12	-	1.12	0.07	1.05
CO	28.01	-	-	-	-	-	3,274.27	3,274.27	-	-	-	3,274.27	0.00	9,247.28	3,274.23	-	3,274.23	4.62	489.74
CO2	44.01	-	-	-	5.02	-	4,464.28	4,464.15	-	7.50	3.93	4,464.15	1,457.28	2,999.49	4,464.10	4,017.69	446.41	86.60	159.47
COS	60.07	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H2	2.02	-	-	-	-	-	6,924.95	6,924.95	-	-	-	6,924.95	0.00	14,427.32	6,924.88	-	6,924.88	1.87	762.37
H2O	18.02	-	-	9,039.83	151.36	-	9,041.84	37.69	-	-	118.41	37.69	2,431.00	8.68	37.69	37.69	-	201.34	0.46
H2S	34.08	-	-	-	-	-	1.13	1.13	-	-	-	-	-	-	-	-	-	-	-
N2	28.01	-	-	-	11,860.14	-	39.98	39.98	-	12.00	9,277.80	39.98	9,328.00	732.33	39.98	-	39.98	1.12	38.86
NH3	17.03	-	-	-	-	-	0.06	0.01	-	-	-	0.01	0.00	0.03	0.01	-	0.01	0.01	0.00
O2	32.00	-	-	-	3,181.65	-	-	-	-	-	2,488.90	-	301.45	-	-	-	-	-	-
SO2	64.06	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
S	32.06	-	-	-	-	-	-	-	1.13	-	-	-	-	-	-	-	-	-	-
HCN	27.03	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
C	12.01	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CL2	70.91	-	-	-	-	-	-	-	-	-	-	-	0.00	-	-	-	-	-	-
NO	30.01	-	-	-	-	-	-	-	-	-	-	-	1.32	-	-	-	-	-	-
NO2	46.01	-	-	-	-	-	-	-	-	-	-	-	0.01	-	-	-	-	-	-
HCL	36.46	-	-	-	-	-	0.52	0.52	-	-	-	0.52	0.26	4.90	0.52	-	0.52	0.26	0.26
CH3OH	32.04	-	-	-	-	-	-	-	-	-	-	-	-	405.69	-	-	-	2,955.66	21.57
C2H2	26.04	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
C2H4	28.05	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
C2H5OH	46.07	-	-	-	-	-	-	-	-	-	-	-	-	0.14	-	-	-	1.44	0.01
C3H8O-01	60.10	-	-	-	-	-	-	-	-	-	-	-	-	0.00	-	-	-	0.01	0.00
C2H6	30.07	-	-	-	-	-	-	-	-	24.00	-	-	-	-	-	-	-	-	-
C3H8	44.10	-	-	-	-	-	-	-	-	5.25	-	-	-	-	-	-	-	-	-
C4H10-1	58.12	-	-	-	-	-	-	-	-	3.00	-	-	-	-	-	-	-	-	-
DIMET-01	46.07	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
BENZE-01	78.11	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
NAPHT-01	128.17	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
FE2O3	159.70	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
FE3O4	231.55	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
FE0.947O	52.89	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
FE	55.85	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
FETIO3	151.75	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
TIO2	79.90	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Solid Flows, lbs/hr																			
Biomass		367,437	208,771	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ash		-	-	-	-	3,546	-	-	-	-	-	-	-	-	-	-	-	-	-
Total lbmoles/hr		-	-	9,039.83	15,340.02	-	23,748.15	14,743.81	1.13	750.00	12,000.00	14,742.68	13,630.27	27,845.68	14,742.53	4,055.38	10,687.15	3,253.00	1,473.79
Total lbs/hr		367,437	208,771	162,855	442,667	3,546	466,233	304,014	36	12,996	346,284	303,976	383,367	454,285	303,972	177,497	126,476	102,387	24,087
Mole Wt		-	-	18.02	28.86	-	19.63	20.62	32.07	17.33	28.86	20.62	28.13	16.31	20.62	43.77	11.83	31.47	16.34
Temperature, degF		60	243	572	1,011	1,922	1,479	110	300	60	59	427	100	130	110	110	297	131	63
Pressure, psia		25.00	23.00	470.00	21.00	14.70	23.00	465.00	439.00	25.00	14.70	445.00	15.00	795.00	411.50	411.50	905.00	415.00	25.00
Enthalpy, MMBtu/hr		(1,673.71)	(573.42)	(907.65)	87.25	0.44	(1,562.27)	(912.40)	0.09	(25.02)	(14.52)	(874.48)	(305.49)	(974.56)	(912.24)	(684.88)	(214.27)	(343.24)	(52.36)
H2/CO		-	-	-	-	-	2.11	2.11	-	-	-	2.11	1.06	1.56	2.11	-	2.11	0.40	1.56
H2+CO+CO2, lbmols/hr		-	-	-	5.02	-	14,663.50	14,663.36	-	7.50	3.93	14,663.36	1,457.28	26,674.09	14,663.21	4,017.69	10,645.52	93.08	1,411.58
CH4, mol%		-	-	-	-	-	0.0	0.0	-	93.1	-	0.0	0.0	0.1	0.0	-	0.0	0.0	0.1
C4 and Heavier HC, mol%		-	-	-	-	-	-	-	-	4.3	-	-	-	-	-	-	-	-	-
CO2, mol%		-	-	-	0.0	-	18.8	30.3	-	1.0	0.0	30.3	10.7	10.8	30.3	99.1	4.2	2.7	10.8
H2S, ppmV		-	-	-	-	-	48	77	-	-	-	-	-	-	-	-	-	-	-
HCl, mol%		-	-	-	-	-	0.002	0.004	-	-	-	0.004	0.002	0.018	0.004	-	0.005	0.008	0.018
									Note 2		Note 1			Note 3		Note 1		Note 3	

Notes:
1) P & T at atmospheric condition
2) Molten sulfur
3) P& T from MP flash.

5.3 OSUBTM PLANT PERFORMANCE RESULTS

5.3.1 Overall Performance Summary

Table 5-2 shows the overall performance summary of the OSUBTM plant modeling results. The PNNL Reference Design data, as modeled by the Nexant IHGBTM case study that was described in Section 4, is also included for comparison. The OSUBTM plant consumes the same amount of biomass feed (at 2,205 tons/day) and it produces approximately 20% more crude methanol, at 102,387 lbs/hr, as compared to a rate of 85,026 lbs/hr for the PNNL Reference Design. This higher carbon conversion efficiency (36% vs. 33%) is due to the higher amount of syngas that can be generated by the BTS gasification process; by about 12% increase in total H_2 and CO mole flow sending onto methanol synthesis.

The higher carbon efficiency for the OSUBTM plant negatively resulted in less energy available to support its overall in-plant steam utility and power consumption requirements. And, as a result, it has a lower overall plant thermal efficiency of 43% versus 51%, as compared to the PNNL Reference Design. The major contributing factors for a lower thermal efficiency for the OSUBTM plant are due to a combination of higher in-plant power consumption and lower power generation:

- The OSUBTM plant generates only 2.1 MWe from the methanol plant purge gas expander. For the PNNL Reference Design, the gasifier/hot flue gas system generated a HP superheated steam that was sent onto a steam turbine producing an additional 15.6 MWe; resulting in a total of 18.3 MWe of power generation for its in-plant use.

- The OSUBTM plant consumes significantly more power. Auxiliary power consumption is estimated at 47.6 MWe versus 36.3 MWe for the PNNL Reference Design. The major culprit is the power requirement for the scrubber syngas compressor, which is higher because of the increase in syngas flow within the OSUBTM plant. This higher in-plant power consumption resulted in a higher net power import for the plant, at 45.5 MWe, instead of 18.0 MWe, comparing to the PNNL Reference Design.

The following sections provide more breakdown on the energy efficiency for the two plant designs.

Table 5-2 Overall Performance Comparison

Performance Summary BTM Plant	OSU Chemical Looping BTM	Nexant IHGBTM Model
Feed		
Dry Wood Chips, Tons/D	2,205	2,205
AR Wood Chips (50% Moisture), Tons/D	4,409	4,409
Natural Gas, SCFH	284,625	0
Products		
Crude Methanol, lbs/hr	102,387	85,026
Tons/D	1,229	1,020
MMGal/Year	136	113
Methanol (100%), lbs/hr	94,706	81,507
Power Generation/Consumption, kW		
Gas Turbine Power	0	0
Gas Expander Power	2,103	2,679
Steam Turbine Power	0	15,633
TOTAL POWER, kWe	**2,103**	**18,313**
AUXILIARY LOAD SUMMARY, kWe		
Feed Handling & Preparation		
Biomass Handling & Preparation	742	742
Gasification		
Oxidizer Air Blower/Char Combustion Air Blower	4,266	4,324
Steam Boiler FG Blower/Char Comb. FG Blower	1,465	914
Syngas Cleanup & Compression		
Steam Boiler Air Compressor/Reformer Air Compressor	1,734	1,315
Flue Gas Blower/Reformer FG Blower	0	662
Scrubber Water Makeup Pump	63	63
Scrubber Condensate Transfer & Recirculation Pump	9	13
Scrubber Syngas Compressor	29,037	19,731
Miscellaneous Syngas Cleanup (Air Fans & Misc. Pumps)	330	204
Acid Gas Removal & Methanol		
Amine AGR / Lo-Cat /ZnO	1,717	778
Lo-Cat Regeneration Air Compressor	28	28
Methanol Feed Compressor	4,421	4,096
Methanol Purge Gas Recycle Compressor	1,789	2,510
Power Generation		
Steam System Auxiliaries	216	370
Miscellaneous Balance of Plant (incl. CW Pumps)	1,751	546
Total Auxiliary Loads, kWe	**47,568**	**36,294**
Net Power, kWe	**-45,464**	**-17,981**
Thermal Efficiency, % LHV (Crude Methanol)	**43.0%**	**50.7%**
Carbon Efficiency, % (Crude Methanol)	**36%**	**33%**
Water Demand/Discharge		
Water Demand/Discharge, gpm		
Cooling Tower Makeup	1,075	705
Boiler Feed Water Makeup	336	351
Scrubber Water Makeup	198	198
Total Water Demand	1,609	1,254
Wastewater, gpm	269	176

5.3.1.1 Thermal Efficiency Breakdown

Table 5-3 shows the breakdown of the calculated conversion and thermal efficiency for the OSUBTM plant, as compared to the PNNL Reference Design as modeled by the Nexant IHGBTM model. As stated earlier, because of the in-plant steam requirements for the BTS gasification plant and the amine CO_2 recovery unit, a steam boiler and steam distribution system is required as part of the balance of plants design. The boiler is fired by the purge gas from the methanol synthesis plant, along with natural gas which has to be imported. Calculations accounted for the imported natural gas.

Table 5-3 Thermal Efficiency Comparison

Thermal and Conversion Efficiencies	**OSU Chemical Looping BTM**	**Nexant IHGBTM Model**
Lower Heating Values (LHV), MMBtu/hr		
Feeds		
Biomass (BM)	1,281	1,281
Natural Gas (NG)	264	0
Products		
Methanol	820	711
Biomass to Methanol Conversion Efficiency	64%	55%
$LHV_{Methanol} / (LHV_{BM})$		
Carbon Efficiency		
$C_{methanol} / (C_{BM} + C_{NG} + C_{air/water\ feeds})$	36%	33%
Net Power Import, MW	-45	-18
MMBtu/hr	-155	-61
Overall Thermal Efficiency	43%	51%
$LHV_{Methanol} / (LHV_{BM} + LHV_{NG} + \text{Power Import})$		

5.3.1.2 Power Generation and Consumption

Table 5-4 compares the power consumptions from the various sources grouped into several major process areas.

Table 5-4 Power Generation & Consumptions

POWER Generation & Consumptions	OSU Chemical Looping BTM	Nexant IHGBTM Model
Power Generation, MW		
STG	0.0	15.6
Purge Gas Expander	2.1	2.7
Total Generation, MW	2.1	18.3
Power Consumption, MW		
Gasification	6.5	6.0
SG Cleanup & Compression	31.2	22.0
AGR	1.7	0.8
Methanol Feed Compression & Synthesis	6.2	6.6
Steam System & Balance of Plant	2.0	0.9
Total Consumption, MW	47.6	36.3
Net Import/Export, MW	-45.5	-18.0

1. Feed Handling, Preparation and Gasification – Estimated consumption is comparable. The primary power consumers in this section are the air and flue gas blowers. Estimated consumption is 0.5 MWe higher for the OSUBTM plant.
2. SG Cleanup and Compression – The main contributor in power consumption in this section is for syngas compression. Table 5-5 tabulated the ASPEN syngas stream flows from the different processing areas within both the OSUBTM and the PNNL Reference Plant. As shown in Table 5-5, The compressed syngas flow rate is approximately 1.5 times higher for the OSUBTM than the PNNL Reference Design (3,230 ACFM vs 2,163 ACFM), resulting in the 9.2 MWe higher power consumption in this process section.
3. AGR and CO_2 Removal – Major power consumption is from the amine plant circulation pumps. While it is small in comparison with syngas compression, consumption for the OSUBTM is almost twice as much, in comparison with the PNNL Reference Design. This is due to a higher CO_2 flow in the syngas generated from the BTS gasification process, resulting in the requirement of a larger amine plant. The higher CO_2 content for the OSUBTM plant (4,464 vs 2,142 lbmols/hr as shown in Table 5-5) increases not only the recirculation pump power consumption, but also the amine plant reboiling and cooling requirements. These, in turn, resulted in a higher reboiler steam generation requirement and cooling tower duty (hence makeup water need) for the OSBTM plant.
4. Methanol Feed Compression and Synthesis – Methanol syngas feed for the OSUBTM plant is approximately 10% higher (Table 5-5). This increases the methanol feed compressor load by 0.3 MWe. The higher methane purity in the OSUBTM methanol feed (1.1 lbmols/hr vs 352 lbmols/hr CH_4), however, results in lower feed inert content and hence its purge rate; and thus reducing the methanol recycle compressor load by 0.7

MWe. The net effect is a reduction of 0.4 MWe in comparison with the PNNL Reference Design.

The lower purge gas rate and the lower CH_4 content in the purge gas has a negative impact on the overall energy import as will be discussed later. The OSUBTM purge gas LHV is only 145 MMBtu/hr compared to 279 MMBtu/hr for the PNNL Reference Design primarily due to the lower CH_4 content.

5. Balance of Plants – BOP mainly consists of a steam boiler, steam distribution and cooling tower system; and power consumption was estimated based on ISBL process flow requirements.

Table 5-5 Syngas and Feed Stream Comparisons

	Gasifier Syngas		Compressed Syngas		Feed to Amine CO2 Removal		Methanol Synthesis Feed	
Case Description	**OSU Chemical Looping BTM**	**Nexant IHGBTM Model**	**OSU Chemical Looping BTM**	**Nexant IHGBTM Model**	**OSU Chemical Looping BTM**	**Nexant IHGBTM Model**	**OSU Chemical Looping BTM**	**Nexant IHGBTM Model**
Gas Compositions (lbmols/hr)								
H2	6,925	1,427	6,925	4,524	6,925	6,228	6,925	6,228
CO	3,274	2,607	3,274	2,171	3,274	2,915	3,274	2,915
H2+CO	10,199	4,034	10,199	6,695	10,199	9,142	10,199	9,142
H2/CO	2.11	0.55	2.11	2.08	2.11	2.14	2.11	2.14
H2O	9,042	5,562	38	25	38	32	-	-
CH4	1.12	992	1.12	964	1.12	352	1.12	352
CO2	4,464	828	4,464	2,142	4,464	2,011	446	201
% CH4 (dry)	0.01%	16%	0.01%	10%	0.01%	3%	0.01%	4%
% CO2 (dry)	30%	13%	30%	22%	30%	17%	4%	2%
Tar, lbmols/hr	-	349	-	-	-	-		
% Other Inerts (dry)	0.3%	-	0.3%	0.4%	0.3%	0.3%	0.4%	0.4%
	-	-						
Stream Conditions	-	-						
Total lbmols/hr	23,748	11,848	14,744	9,869	14,743	11,577	10,687	9,735
Total lbmols/hr (Dry)	14,706	6,285	14,706	9,844	14,705	11,545	10,687	9,735
Total lbs/hr	466,235	241,582	304,014	181,324	303,972	190,030	126,476	109,804
Mol Wt	20	20	20.62	18.37	20.62	16.41	11.83	11.28
Temp. degF	1,480	1,598	110	110	110	110	297	300
Psia	23	23	465	465	412	407	905	905
Total SCFM	150,207	74,936	93,255	62,420	93,247	73,223	67,596	61,573
Total ACFM	358,156	189,505	3,232	2,163	3,651	2,902	1,599	1,462

5.3.1.3 Steam Generation and Consumption

The overall thermal efficiency for the OSUBTM plant is lower than the PNNL Reference Design due to (1) its higher power import requirement as shown in the previous section, and (2) the need for a supplemental fuel to generate steam for its in-plant use.

The overall process steam and reheating requirements for the OSUBTM plant and the PNNL Reference Design are shown in Table 5-6. The total process steam injection for syngas generation is about the same:

- 162,855 lbs/hr to the OSUBTM plant's BTS gasification plant, and
- 162,799 lbs/hr in combined flow to the indirectly-heated gasifier and the reformer for the PNNL Reference Design.

Both processes produce a syngas with a H_2:CO of 2.1 for downstream methanol synthesis. The BTS gasification process, at its current stage of development, produces more syngas which is good; but the syngas also contains a significant amount of CO_2 which had to be removed before it can be sent onto methanol synthesis. This requires a much larger amine plant for the OSUBTM design with a reboiling steam requirement that is about 2.2 times that for the PNNL Reference Design. The net effect is a significant increase in steam consumption for the OSUBTM plant of about 192,800 lbs/hr or 193 MMBtu/hr of LP steam.

Table 5-6 Process Steam Consumptions

Process Steam Consumptions	**OSU Chemical Looping BTM**	**Nexant IHGBTM Model**
Process Steam Requirement, lb/hr		
Gasifier	162,855	73,120
Sulfur Removal	850	850
SMR	0	89,679
AGR Reboiler	350,854	158,043
Total Process Steam Required. Lb/hr	514,559	321,692
Total Process Steam Duty Required, MMBtu/hr	529	336

In addition to generating the required LP steam for CO_2 removal, duties for preheating various process streams were estimated and these are summarized in Table 5-7.

Table 5-7 Biomass Drying & Process Preheating Duties

Process Preheat & Biomass Duties	**OSU Chemical Looping BTM**	**Nexant IHGBTM Model**
Preheat/Drying Duties , MMBtu/hr		
Biomass Drying	196	196
Oxidation Air Preheat	92	0
ZnO Feed Preheat	37	48
Makeup BFW Preheat	19	19
Expander Reheat	6	6
SMR Feed Preheat	0	34
Total Preheat Duty, MMBtu/hr	350	303

As shown, the estimated preheating duty for the OSUBTM plant is about 47 MMBtu/hr more than the PNNL Reference Design. The OSUBTM plant is able to reduce the ZnO feed preheating by 11 MMBtu/hr by feed/effluent heat exchange which is not available for the PNNL Reference Design. It also does not have a SMR unit and hence no SMR feed preheating. The combine heat savings is about 45 MMBtu/hr. However, the need to preheat the oxidation air to 1,010 ^{o}F for the oxidizer in the BTS gasification process consumes 92 MMBtu/hr of syngas cooling duty which negates the savings and increases the preheat duty by 47 MMBtu/hr.

The total steam and process preheating duties are summarized in Table 5-8. The OSUBTM plant requires 239 MM Btu/hr more duties than the PNNL Reference Design.

Table 5-8 Total Heating Duty Comparison

Steam Generation, Process Preheat and Biomass Drying Duties	**OSU Chemical Looping BTM**	**Nexant IHGBTM Model**
Steam Generation, MMBtu/hr	529	336
Process Preheat & Biomass Drying, MMBtu/hr	350	303
Total Required Duty, MMBtu/hr	879	639

The primary source of heat for steam generation and process heating are heat recovery from the syngas, the combustion flue gas and methanol reactor cooling. The available heat from these sources is summarized in Table 5-9.

Table 5-9 Available Syngas & Flue Gas Duties

Available Syngas & Flue Gas Duties	**OSU Chemical Looping BTM**	**Nexant IHGBTM Model**
Available Duties , MMBtu/hr		
Syngas	-292	-320
Flue Gas (Purge gas, Char or Spent Air)		
Char Combustor	0	-218
SMR Combustor (Purge Gas)	0	-127
Methanol Purge Gas	-145	
Spent Air	-134	0
Rx Cooling	-92	-68
Total Available Duty, MMBtu/hr	-663	-734

By comparing the total required duty in Table 5-8 and the total available duty in Table 5-9, the OSUBTM plant has a net shortage of 216 MMBtu/hr of heating duty which is made up by firing supplemental natural gas fuel. The PNNL Reference Design has an excess of 95 MMBtu/hr which is used to generate SH HP steam for STG power generation.

As discussed above, the higher power import and supplemental natural gas fuel are required to support the OSUBTM plant. Hence, the overall thermal efficiency for OSUBTM plant is lower than the PNNL Reference Design by 7%.

5.3.2 Overall Carbon Balances

Table 5-10 shows the carbon balance for the OSUBTM plant. The carbon input to the plant consists of carbon in the air and natural gas in addition to carbon in the biomass. Carbon leaves the plant as unburned carbon in the slag, CO_2 in the stack gas and methanol product.

Table 5-10 Overall Carbon Balance

Overall Carbon Balance, lb/hr	In	Out
Biomass Feed	92,962	
Supplemental Fuel (NG)	9,387	
Reboiler Condensate	0	
Combustion & Oxidizer Air	108	
Makeup Water	0	
MeOH Product		36,632
CO2 Removed		48,257
Combustion Flue Gas		17,564
Biomass Removed Water		0
Ash & Sulfur		0
LP Steam		0
Deaerator Vent		0
LoCAT Regenerator Vent		0
Treated Raw water to Cooling Tower	0	
CT Evaporation & Drift Losses		0
CT Blowdown		0
Convergence Tolerance		2
Total	**102,456**	**102,456**

5.3.3 Overall Sulfur Balances

Table 5-11 shows the sulfur balance for the OSUBTM plant. Sulfur input comes from the sulfur in the biomass and the supplemental fuel. Sulfur output includes the sulfur recovered in the LoCAT and ZnO plant and sulfur emitted in the stack gas. Sulfur in the slag is assumed to be negligible.

Table 5-11 Overall Sulfur Balance

Overall Sulfur Balance, lb/hr	**In**	**Out**
Biomass Feed	37	
Supplemental Fuel (NG)	0	
Reboiler Condensate	0	
Combustion & Oxidizer Air	0	
Makeup Water	0	
MeOH Product		0
CO2 Removed		0
Combustion Flue Gas		0
Biomass Removed Water		0
Ash & Sulfur		36
LP Steam		0
Deaerator Vent		0
LoCAT Regenerator Vent		0
Treated Raw water to Cooling Tower	0	
CT Evaporation & Drift Losses		0
CT Blowdown		0
Convergence Tolerance		1
Total	**37**	**37**

5.3.4 Cooling Water Duty and Cooling Tower Requirement

Table 5-12 shows the comparison of the cooling water duties for the OSUBTM vs. the PNNL Reference Design. Total OSUBTM plant cooling water duty is approximately 52% higher. This is primarily due to the following:

- Higher scrubber overhead condensation duty resulting from higher syngas flow to the scrubber and the syngas compressors in the OSUBTM plant. Table 5-5 shows the comparison of the gasifier syngas streams. The total dry gas flow for the OSUBTM plant is about 2.3 times that of the PNNL Reference Design (14,706 lbmols/hr vs 6,285 lbmols/hr). Therefore, the OSUBTM scrubber overhead stream carries more steam and as a result requires higher condensation duty. The OSUBTM CW load was reduced by air fan cooling and makeup water preheat.

- Higher amine plant regenerator overhead cooling duty is required because of the increased reboiling duty as discussed in section 5.3.1.3.

Table 5-12 OSUBTM Cooling Water Loads

Process Cooling Water Duties (MMBtu/hr)	**OSU Chemical Looping BTM**	**Nexant IHGBTM Model**
SG Cooling / Scrubbing	64	77
Scrubber Overhead SG Compression	116	79
LoCAT	0.1	0
Steam Boiler/SMR SG Cooling	5	9
Methanol Production	67	62
CO2 Stripper Overhead Condenser	303	136
Total Process Cooling Water Duty	555	364
STG Surface Condenser Duty *	0	0
Total CW & CT Loads, MMBtu/hr	555	364
* Air Cooled		

5.3.5 Overall Water/Steam System Flow Diagrams

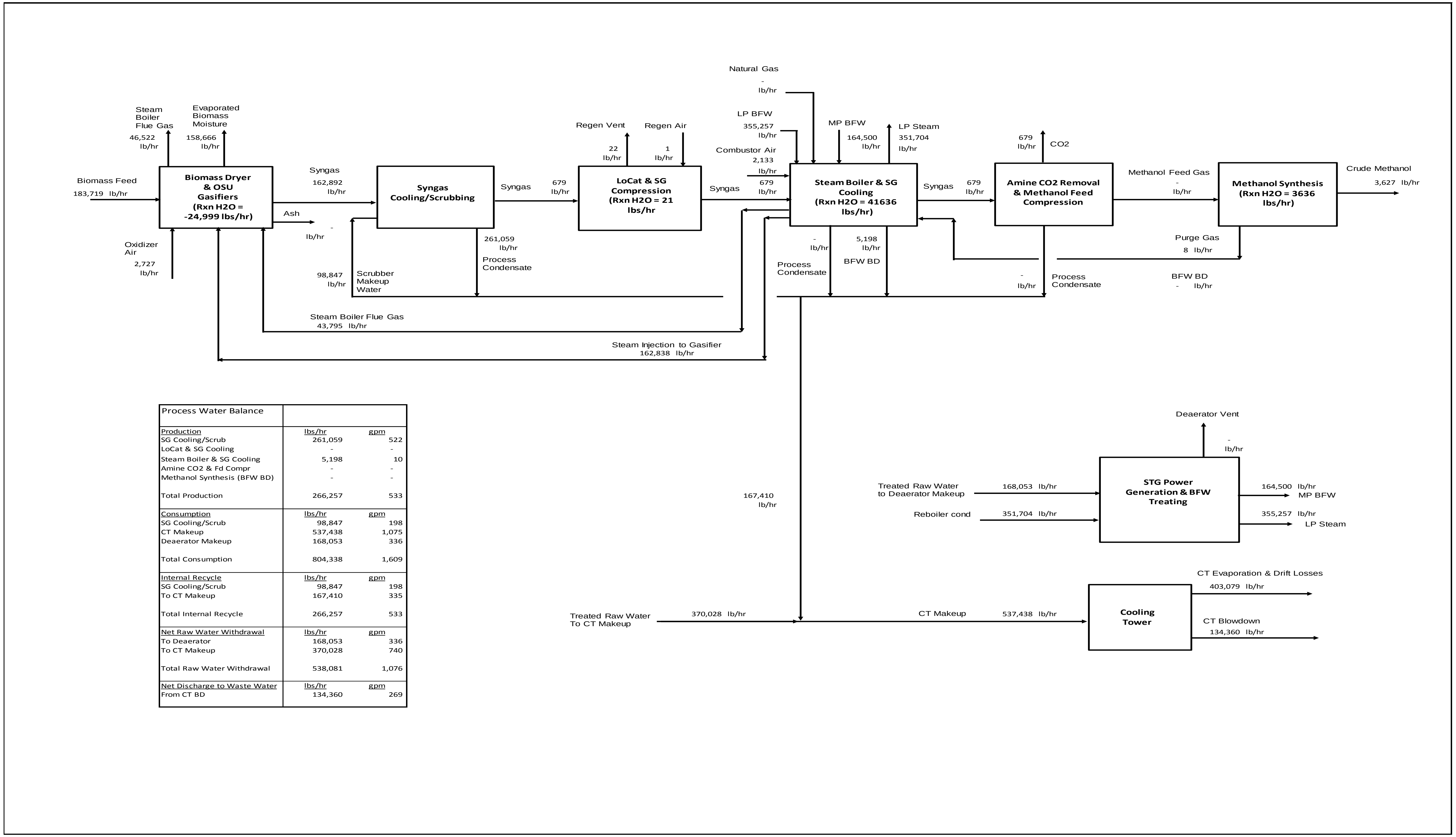
Figure 5-3 shows the overall water flows and balance for the OSUBTM Plant.

The water demand is mainly for the scrubber, deaerator and the cooling tower water makeup requirements. The total water demand is 1,609 gpm. Internal condensate and BFW blowdown are recycled to make up 533 gpm of the water demand. The balance of 1,076 gpm is supplied as makeup water to the plant. Process water discharge to waste treatment facility consists of 269 gpm of cooling tower blowdown.

Figure 5-4 shows the overall steam flow diagram for the OSUBTM Plant.

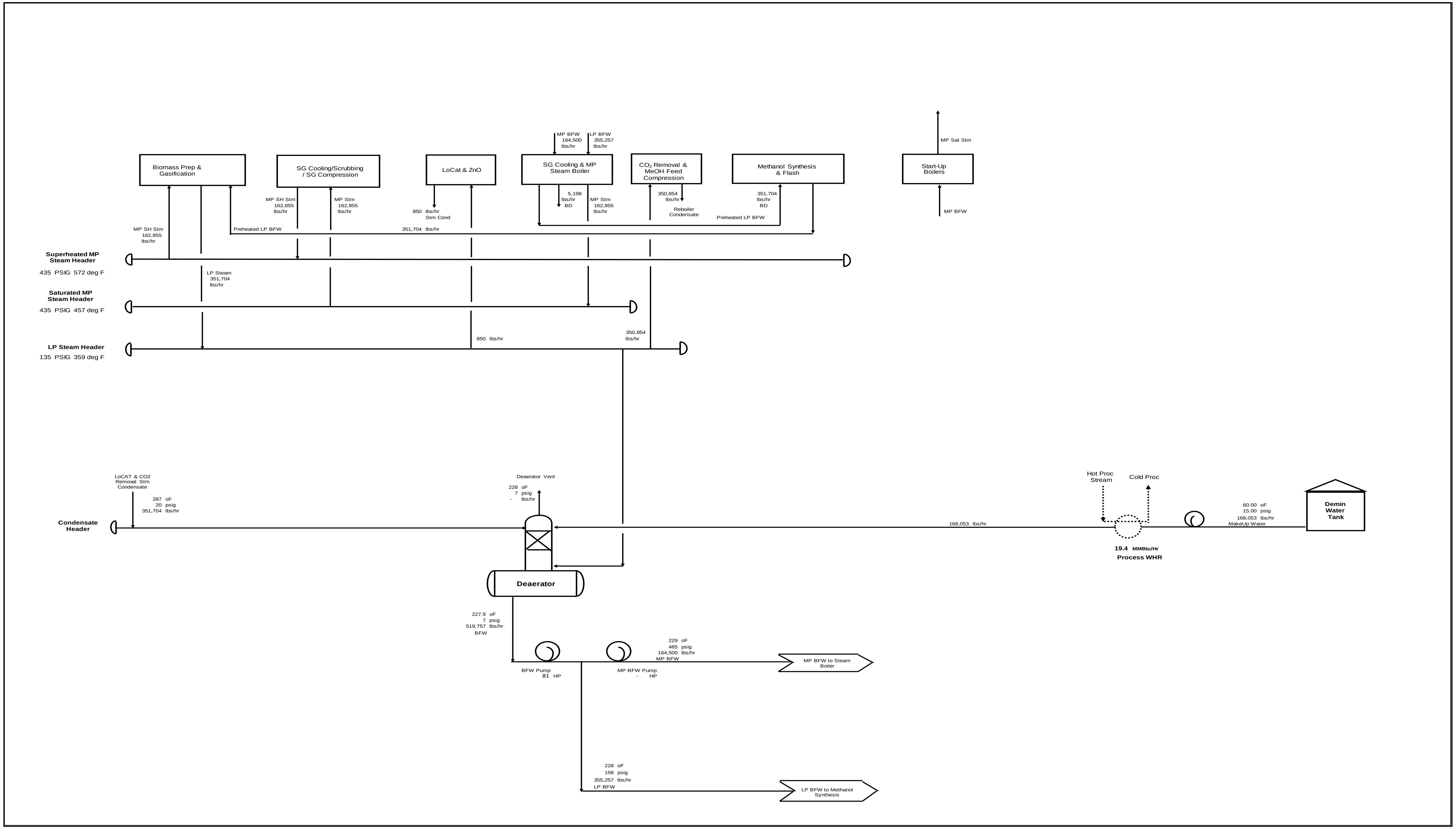
No high pressure steam is generated due to the insufficient heat available from BTS syngas, flue gas and reactor cooling. Medium pressure steam (435 psig, 572 ^{o}F) is generated by syngas heat recovery to provide MP steam injection for the reducer in the gasification plant. LP steam (135 psig, 359 ^{o}F) is generated by syngas heat recovery and supplemental natural gas fired steam boiler to provide LP steam for the amine CO_2 removal unit and other users. The process condensates and makeup water are sent to the deaerator where the mixture is deaerated by stripping with a small amount of steam and treated for boiler feed water (BFW) uses. Makeup water from storage is heated by recovering process heat before mixing with the condensates. The deaerator operates at 7 psig and 228 ^{o}F.

Figure 5-3 OSUBTM Overall Water Flow Diagram



Process Water Balance		
Production	lbs/hr	gpm
SG Cooling/Scrub	261,059	522
LoCat & SG Cooling	-	-
Steam Boiler & SG Cooling	5,198	10
Amine CO2 & Fd Compr	-	-
Methanol Synthesis (BFW BD)	-	-
Total Production	266,257	533
Consumption	lbs/hr	gpm
SG Cooling/Scrub	98,847	198
CT Makeup	537,438	1,075
Deaerator Makeup	168,053	336
Total Consumption	804,338	1,609
Internal Recycle	lbs/hr	gpm
SG Cooling/Scrub	98,847	198
To CT Makeup	167,410	335
Total Internal Recycle	266,257	533
Net Raw Water Withdrawal	lbs/hr	gpm
To Deaerator	168,053	336
To CT Makeup	370,028	740
Total Raw Water Withdrawal	538,081	1,076
Net Discharge to Waste Water	lbs/hr	gpm
From CT BD	134,360	269

Figure 5-4 OSUBTM Steam Flow Diagram



5.4 OSUBTM PLANT COST ESTIMATE

5.4.1 CAPEX

Estimated capital costs are summarized in tables 5-13, 5-14 and 5-15.

Table 5-13 shows the Total Project Investment (TPI) summary for the OSUBTM plant which was developed from the total purchased equipment costs (TPEC) per methodology as described in section 3.1. The cost basis year is 2011, as with the PNNL Reference Design. Costs are broken down and presented in a format that is similar to what DOE had done in their 2014 "Baseline Analysis of Crude Methanol Production from Coal and Natural Gas" study report. In this format, the relative cost contributions to the overall methanol production can be separated out and compared. The estimated Total Project Investment (TPI) for the OSUBTM plant is $347 MM.

For the OSUBTM plant, OSU provided an estimated TPI cost for the BTS gasification plant system at $26.8 MM that is based on an estimated TPEC of $10.9 MM. It replaces the costs for the indirectly heated biomass gasifier and the tar and steam reforming units in the PNNL Reference Design. The combined gasifier/tar reformer/steam reformer TIC for the Nexant Reference Design is ~$31MM:

- Gasification - $15 MM
- Tar Reforming – $7 MM
- SMR – $9 MM

For the rest of the process plants that are common for both designs (e.g., amine CO_2 removal, LoCat and ZnO polishing, etc), their costs were estimated by capacity factor from that of the PNNL Reference Design. Others, such as gas compressors, purge gas expander, STG and steam boiler and cooling tower system, are sized and costed as individual equipment.

Table 5-14 shows a comparison of the TPI cost for the OSUBTM plant with the PNNL Reference Design. The estimated TPI cost for the OSUBTM plant is $48 MM higher, and most of the cost increase is due to the enlarged amine CO_2 removal plant ($145MM vs $89MM).

Per methodology used by the DOE to calculate the required sales price (RSP) for methanol, Table 5-15 shows the addition of the various owner's costs to determine the total overnight cost (TOC) for the OSUBTM plant. Owner's costs include preproduction costs, inventory capital, initial costs for catalyst and chemicals, land and financing costs; and some of which are estimated as a function of the project TPC, as per DOE's 2011 "Quality Guidelines For Energy System Study: Cost Estimation Methodology for NETL Assessments of Power Plant Performance" . The estimated TOC for the OSUBTM plant is $443 MM.

Table 5-13 OSUBTM Total Project Investment (TPI) Cost Estimate Summary

Cost Basis – Year 2011

Acct No.	Item/Description	Totral Purchased Equipment Cost (TPEC)	Bulk*	Purchased Equipment Installation	Total Installed Cost (TIC)	Eng'g CM H.O., Fees	Contingencies	TOTAL PROJECT INVESTMENT	
		$1,000	$1,000	$1,000	$1,000	$1,000	$1,000	$1,000	$/TPD MeOH
1	BIOMASS PREP & DRYING	$11,400	$12,313	$4,446	$28,159	$10,146	$4,218	$42,524	$34,610
2	GASIFICATION WITH TAR REFORMING/HEAT RECOVERY/SCRUBBING								
2.1	Gasification	$10,868	$11,737	$4,238	$26,843	$9,672	$4,021	$40,537	$32,993
2.2	Syngas Cooling & Scrubbing	$2,703	$2,919	$1,054	$6,677	$2,406	$1,000	$10,083	$8,207
2.3	Feed Compression (to 460 psia)	$4,964	$5,361	$1,936	$12,261	$4,418	$1,837	$18,515	$15,070
2.4	Sulfur Removal (LoCAT & ZnO)	$672	$726	$262	$1,660	$598	$249	$2,507	$2,040
2.5	MP Steam Boiler	$2,796	$3,020	$1,091	$6,907	$2,489	$1,035	$10,430	$8,489
	SUBTOTAL 2.	$22,003	$23,768	$8,581	$54,348	$19,583	$8,141	$82,072	$66,799
3	SYNGAS CLEANUP & COMPRESSION								
3.1	Aminne CO2 Recovery	$38,898	$42,010	$15,170	$96,078	$34,619	$14,392	$145,089	$118,089
3.2	Methanol Feed Compression	$5,276	$5,698	$2,058	$13,032	$4,696	$1,952	$19,680	$16,018
	SUBTOTAL 3.	$44,174	$47,708	$17,228	$109,110	$39,315	$16,344	$164,769	$134,107
4	METHANOL SYNTHESIS & PURIFICATION								
4.1	Methanol Synthesis (Crude Methanol)	$11,405	$12,317	$4,448	$28,170	$10,151	$4,220	$42,541	$34,624
4.2	Crude Methanol Distillation	$0	$0	$0	$0	$0	$0	$0	$0
4.3	Purge Gas Expander	$822	$888	$321	$2,031	$732	$304	$3,068	$2,497
	SUBTOTAL 4.	$12,227	$13,206	$4,769	$30,202	$10,882	$4,524	$45,608	$37,121
5	STEAM SYSTEM & POWER GENERATION	$796	$860	$310	$1,966	$708	$294	$2,969	$2,416
6	BALANCE OF PLANT & OFFSITES (OSBL)	$2,388	$2,579	$931	$5,898	$2,125	$883	$8,906	$7,249
	CALCULATED TOTAL COST	$92,989	$100,432	$36,266	$229,682	$82,760	$34,406	$346,848	$282,302

* Bulk includes Instrumentation and Controls, Piping, Electrical Systems, Buildings and Yard Improvement

Table 5-14
OSUBTM vs PNNL Reference Design Total Project Investment (TPI) Cost

Acct No.	Item/Description	OSU Chemical Looping BTM	Nexant IHGBTM Model
	Total Project Investment (Yr 2011)	$MM	$MM
1	BIOMASS PREP & DRYING	$42.52	$42.52
2	GASIFICATION WITH TAR REFORMING/HEAT RECOVERY/SCRUBBING		
2.1	Gasification	$40.54	$23.00
2.2	SG Cooling & Scrubbing (TAR Reformer & Scrubbing for IHGBTM)	$10.08	$17.65
2.3	SG Compression (to 460 psia)	$18.52	$14.55
2.4	Sulfur Removal (LoCAT & ZnO)	$2.51	$2.51
2.5	MP Steam Boiler (SMR for IHGBTM)	$10.43	$13.25
	SUBTOTAL 2.	$82.07	$70.97
3	SYNGAS CLEANUP & COMPRESSION		
3.1	Amine CO2 Recovery	$145.09	$89.41
3.2	Methanol Feed Compression	$19.68	$18.41
	SUBTOTAL 3.	$164.77	$107.81
4	METHANOL SYNTHESIS & PURIFICATION		
4.1	Methanol Synthesis (Crude Methanol)	$42.54	$37.63
4.2	Crude Methanol Distillation	$0.00	$0.00
4.3	Purge Gas Expander	$3.07	$3.37
	SUBTOTAL 4.	$45.61	$41.00
5	STEAM SYSTEM & POWER GENERATION	$2.97	$28.09
6	BALANCE OF PLANT & OFFSITES (OSBL)	$8.91	$8.24
	CALCULATED TOTAL PROJECT INVESTMENT	$347	$299

Table 5-15
OSUBTM Total Overnight Cost Summary

Owner's Costs	$1,000
Preproduction Costs	
6 months All Labor (notes 1 & 2)	$5,189
1 Month Maintenance Materials (Note 2)	$462
1 Month Non-Fuel Consumables	$3,520
1 Month Waste Disposal	$62
25% of 1 Months Fuel Cost at 100% CF	$208
2% of TPI	$6,937
Subtotal	$16,378
Inventory Capital	
60 day supply of fuel at 100% CF	$10,569
60 day supply of non-fuel consumables at 100% CF	$533
0.5% of TPI (spare parts)	$1,734
Subtotal	$12,836
Initial Cost for Catalyst and Chemicals	$4,896
Land	$900
Other Owner's Cost (note 3)	$52,027
Financing Costs	$9,365
Total Owner's Cost	**$96,403**
Total Project Investment (TPI)	**$346,848**
Total Overnight Costs (TOC)	**$443,250**

Notes:
1) Includes operating and administrative labor plus maintenance labor which is 40% of the maintenance material + labor
2) Maintenance material + labor is 2.4% of TPI
3) Other owner's cost is15% of TPI including preliminary studies, front end engineering, owner's engineers,
owner's management reserve and legal & permitting costs

5.4.2 ANNUAL O&M COST ESTIMATE

Table 5-16 shows the annual O&M cost estimate breakdown for the OSUBTM plant, and Table 5-17 compares the annual O&M cost between the OSUBTM plant and the PNNL Reference Design.

Table 5-17 shows total fixed annual operating and maintenance costs (FXOPEX) which includes the annual operating labor, maintenance and administration labor costs, property taxes and insurance. Although the annual operating labor cost is assumed to be the same for the OSUBTM and the Nexant Reference Design, the maintenance, administrative and support labor cost plus the property tax and insurance are function of the TPI. Therefore, with the higher TPI, the OSUBTM plant FXOPEX is $1.5 MM higher than the PNNL Reference Design

The annual variable operating cost including feed and fuel is estimated at $101.5 MM vs $70.9 MM for the PNNL Reference Design. The non-fuel portion is $43.7 MM which includes maintenance material cost, water, catalyst and chemicals, waste disposal and power import costs.

The biomass feed cost is $48.9 MM annually based on $67.5/ton of dry biomass feed on year 2011 basis. The natural gas supplemental fuel cost is estimated at $9 MM, based on $4/1000 SCF of natural gas.

Table 5-16
OSUBTM Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES

Case:	OSU Chemical Looping Gasification BTM			
Plant Size, Crude Methanol	102,387	lbs/hr	**Cost Basis (Year)**	2011
	1,229	Tons/D	**Capacity Factor (%)**	90
	136	MMgal/year		
Primary/Secondary Fuel:	Biomass/Natural Gas			

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (base):	$39.70	$/hr
Operating Labor Burden:	30.00	% of base
Labor Overhead Charge	25.00	% of labor

Operating Labor Requirements per Shift	**units/mod**	**Total Plant**
Skilled Operator	2.0	2.0
Operator	6.0	6.0
Foreman	1.0	1.0
Lab Tech's etc	2.0	2.0
TOTAL Operating Jobs	11.0	11.0

	Annual Cost $
Annual Operating Labor Cost	$4,973,140
Maintenance Labor Cost	$3,329,738
Administration & Support Labor	$2,075,719
Property Taxes and Insurance	$6,936,954
TOTAL FIXED OPERATING COSTS	**$17,315,551**

VARIABLE OPERATING COSTS

Maintenance Material Cost					**$4,994,607**

Consumables	Consumption Initial	Consumption /Day	Unit Cost	Initial Fill Cost	
Water(/1000 gallons)	0	2,317	3.50	**$0**	**$2,662,099**
Chemicals					
MU & WT Chem (lbs)	0	93	1.22	$0	$37,424
Oxygen Carrier Makeup (metric ton)	4,960	24	610.74	$3,029,490	$4,776,899
Amine Solvent (lbs)	0	215	3.06	$0	$216,379
Tar Reforming Catalyst (lbs)	0	0	8.61	$0	$0
Steam Reforming Catalyst (lbs)	0	0	27.08	$0	$0
ZnO Catalyst (lbs)	91,504	706	5.88	$538,002	$1,364,310
Methanol Synthesis Catalyst(lbs)	127,953	168	10.38	$1,328,484	$573,344
LoCAT Chemicals (ton)	0	1	457.88	$0	$219,671
Subtotal Chemicals				**$4,895,976**	**$7,188,028**
Other					
Supplemental Electricity (MWh consumed)	0	1,091	78.58	$0	$28,166,832
Gases, N2 etc.(/100scf)	0	0	0.00	$0	$0
LP Steam (/1000 lbs)	0	0	0.00	$0	$0
Subtotal Other				**$0**	**$28,166,832**
Waste Disposal:					
Flyash (ton)	0	0	0.00	$0	$0
Slag (ton)	0	44	46.62	$0	$666,438
Subtotal Waste Disposal				**$0**	**$666,438**
By-products & Emissions					
Sulfur (ton)	0	0	0.00	$0	$0
Supplemental Electricity (MWh generated)	0	0	-78.58	$0	$0
Subtotal By-Products				**$0**	**$0**
TOTAL VARIABLE OPERATING COSTS				**$4,895,976**	**$43,678,003**
Biomass (T/D, Dry)	0	2205	67.51	**$0**	**$48,892,032**
Natural Gas (1000 SCF)	0	6831	4.00	**$0**	**$8,975,934**

Table 5-17
OSUBTM vs. PNNL Reference Design Operating Cost

OPERATING COSTS, 2011 $MM/YR	OSU Chemical Looping BTM	Nexant IHGBTM Model
FIXED OPERATING COSTS		
Annual Operating Labor Cost	$5.0	$5.0
Maintainence Labor Cost	$3.3	$2.9
Administration & Support Labor	$2.1	$2.0
Property Taxes and Insurance	$6.9	$6.0
TOTAL FIXED OPERATING COSTS	$17.3	$15.8
VARIABLE OPERATING COSTS		
NON-FUEL VARIABLE OPERATING COSTS		
Maintenance Material Cost	$5.0	$4.3
Water	$2.7	$1.6
Chemicals		
MU & WT Chemicals	$0.04	$0.04
Chemicals & Catalysts	$7.2	$4.1
Supplemental Electricity	$28.2	$11.1
Solid Waste Disposal	$0.7	$0.8
TOTAL NON_FUEL VARIABLE OPERATING COSTS	$43.7	$22.0
FUEL		
Biomass	$48.9	$48.9
Natural Gas	$9.0	$0.0
TOTAL VARIABLE OPERATING COSTS	$101.5	$70.9
TOTAL OPERATING COSTS	**$118.9**	**$86.7**

5.4.3 Methanol Required Selling Price (RSP)

Table 5-18 shows a summary comparison of the estimated capital expenditure (CAPEX), operating expenditure (OPEX), power output, and methanol required sales price (RSP) for the OSUBTM and the Nexant Reference Design. The OSUBTM plant estimated RSP is $1.38/gal of crude methanol compared to $1.28/gal for the PNNL Reference Design. A breakdown of the cost components for the RSP is shown in Table 5-19. The power import cost is a major contributor to the higher RSP for the OSUBTM plant.

Table 5-18
OSUBTM Plant Performance and Economic Summary

Cost & Performance	OSU Chemical Looping BTM	Nexant IHGBTM Model
CAPEX, $MM		
Total Installed Cost (TIC)	$230	$198
Total Project Investment (TPI)	$347	$299
Total Overnight Cost (TOC)	$443	$383
OPEX, $MM/yr		
Fixed Operating Cost (OC_{Fix})	$17	$16
Variable Operating Cost Less Fuel (OC_{VAR})	$44	$22
Fuel Cost (OC_{Fuel})	$58	$49
Power Production, MWe		
Expander	2.1	2.7
Steam Turbine	0.0	15.6
Total Power Output	2.1	18.3
Auxiliary Power Consumption	47.6	36.3
Net Power Output	-45.5	-18.0
Power Import/Export, MWh/yr (MWH)	-398,266	-157,513
Thermal Efficiency,% LHV	43.0%	50.7%
Crude Methanol Production		
Tons/D	1,229	1,020
MMgal/Yr	136	113
RSP, $/gal MeOH	**$1.38**	**$1.28**

Table 5-19
RSP Cost Components

RSP Cost Components	OSU Chemical Looping BTM	Nexant IHGBTM Model
CAPEX	$0.40	$0.42
Fixed OPEX	$0.14	$0.16
Variable OPEX	$0.13	$0.11
Biomass & NG Cost	$0.48	$0.48
Power Import Cost	$0.23	$0.11
RSP Total ($/gal methanol)	**$1.38**	**$1.28**

5.5 RSP SENSITIVITY STUDY

A sensitivity analysis was carried out to look at the effect of various cost parameters on the methanol RSP. It is performing by varying a single variable while holding the others constant. Table 5-20 shows the baseline value used, upper and lower limit for each of the parameters of interest.

Table 5-20
OSUBTM Plant Sensitivity Parameters and Limits

Parameter	**Baseline Value**	**Upper Limit**	**Lower Limit**
TPEC of the OSU BTS chemical looping system	100%	+30%	-30%
OSU BTS ITCMO cost (affects initial fill cost and O&M costs)	100%	+50%	-50%
Capacity factor	90%	100%	70%
Feedstock price	$67.51/dry ton	$90/dry ton	$40/dry ton
Supplemental natural gas price	$4/1000 SCF	$9/1000 SCF	$2/1000 SCF
Power price for net export/import	$0.078/kWh	$0.12/kWh	$0.4/kWh

5.5.1 RSP Sensitivity to OSUBTM BTS Gasification Plant Cost

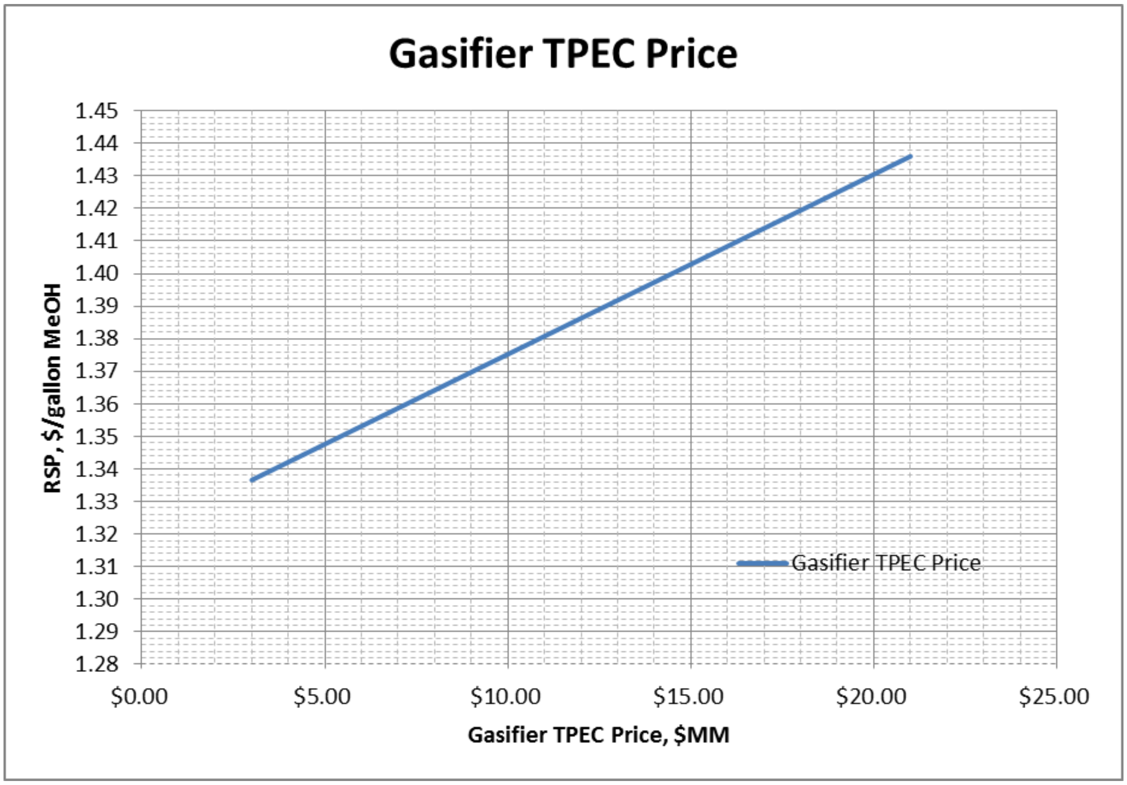
Figure 5-5 shows the sensitivity of the OSUBTM RSP to the estimated purchased equipment cost (TPEC) of the BTS gasification plant. The RSPs are estimated based on the following baseline parameters:

Capacity Factor	90%

Biomass Price	$67.51/dry ton
Natural Gas Price	$4/1000 SCF
Cost of Electricity Import	$0.078/kWh

The OSUBTM BTS gasification TPEC is varied from $3 to $21MM. The baseline TPEC is $10.9 MM. RSP increases by $0.027/gal methanol for every $5 MM increase in the TPEC.

Figure 5-5 RSP Sensitivity to BTS Gasification Plant Cost



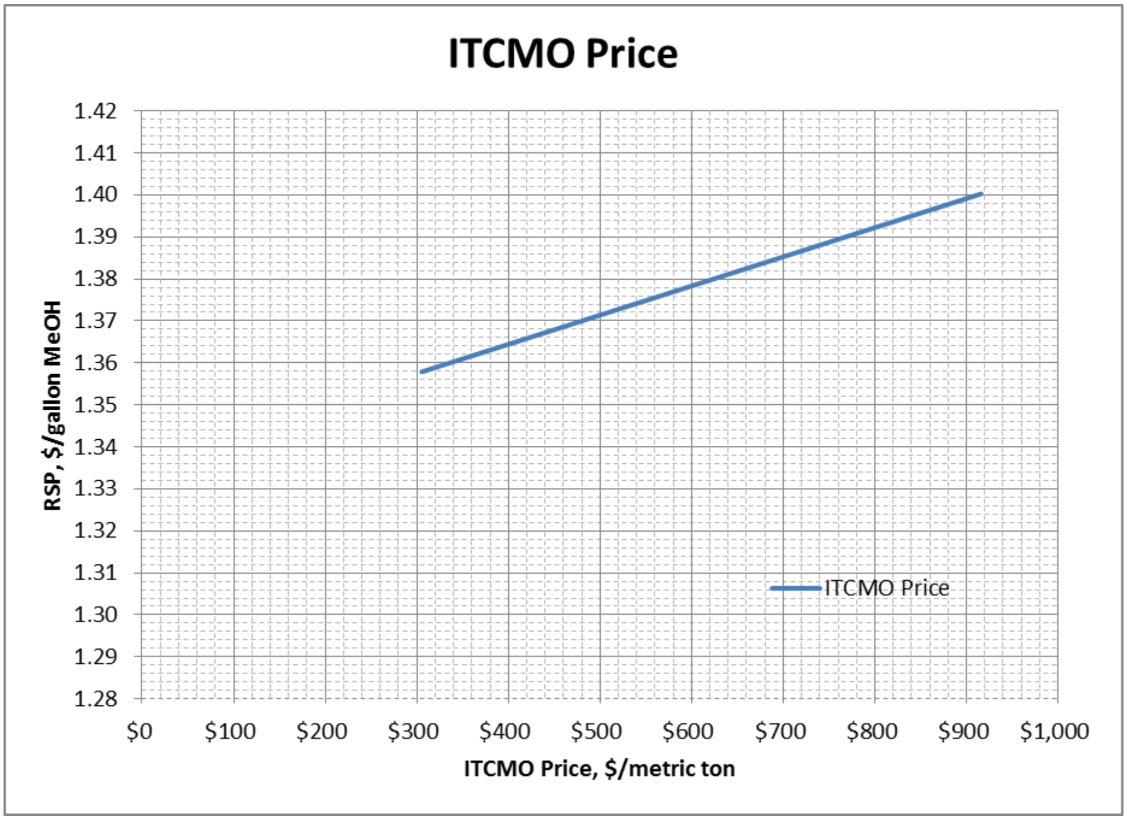
5.5.2 RSP Sensitivity to Iron Titanium Composite Metal Oxide (ITCMO) Price

Figure 5-6 shows the sensitivity of the OSUBTM RSP to the ITCMO price. The RSPs are estimated based on the following baseline parameters:

Capacity Factor	90%
Biomass Price	$67.51/dry ton
Natural Gas Price	$4/1000 SCF
Cost of Electricity Import	$0.078/kWh

The ITCMO price is varied from $300 to $900/metric ton of ITCMO. The estimated baseline ITCMO price is $611/metric ton. RSP increases by $0.008/gal methanol for every $100/metric ton increase in the ITCMO price.

Figure 5-6 RSP Sensitivity to ITCMO Catalyst Cost



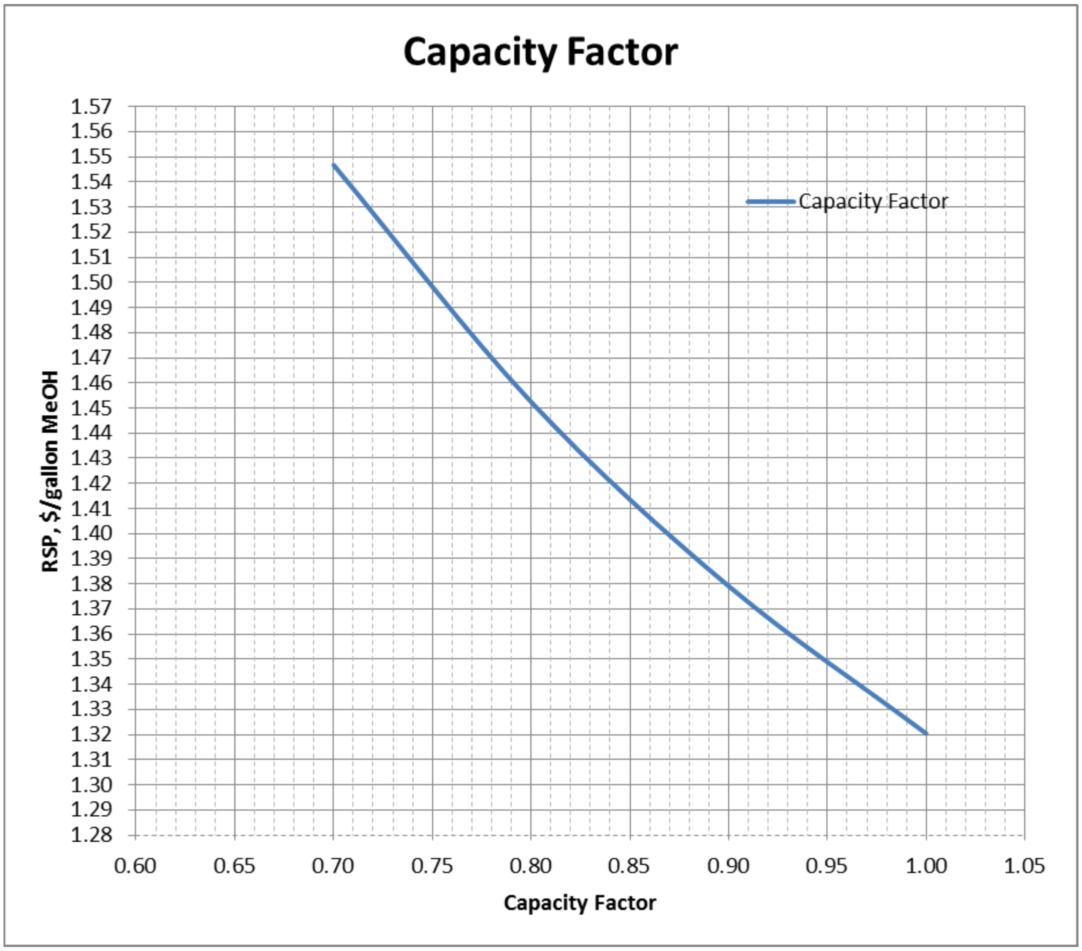
5.5.3 RSP Sensitivity to Capacity Factor

Figure 5-7 shows the sensitivity of the OSUBTM RSP to the plant capacity factor. The RSPs are estimated based on the following baseline parameters:

Capacity Factor	90%
Biomass Price	$67.51/dry ton
Natural Gas Price	$4/1000 SCF
Cost of Electricity Import	$0.078/kWh

The capacity factor is varied between 0.7 and 1.0. The capacity factor of 0.9 is the basis for the TEA study. The RSP decreases by average of $0.073/gal methanol for every 0.1 increase in the capacity factor.

Figure 5-7 RSP Sensitivity to Plant Capacity Factor



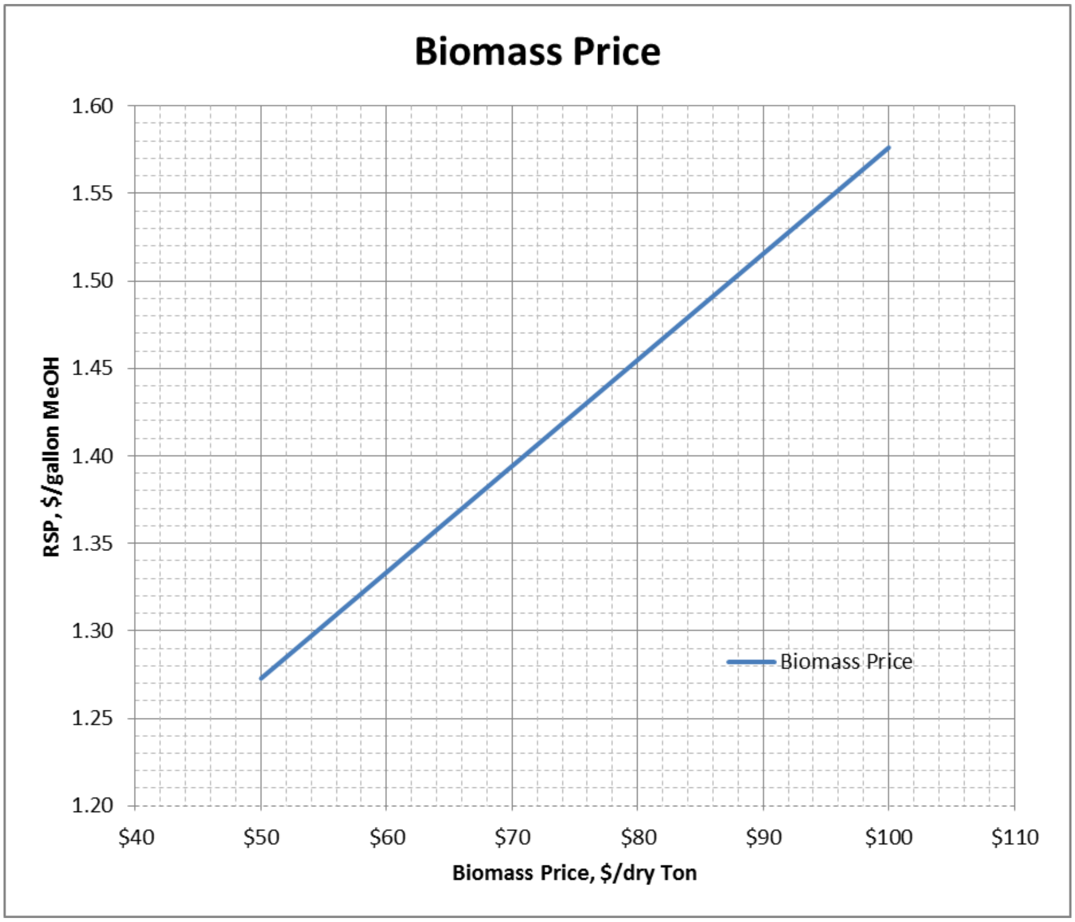
5.5.4 RSP Sensitivity to Biomass Price

Figure 5-8 shows the RSP sensitivity to biomass price. The RSPs are estimated based on the following baseline parameters:

Capacity Factor	90%
Biomass Price	$67.51/dry ton
Natural Gas Price	$4/1000 SCF
Cost of Electricity Import	$0.078/kWh

The biomass price is varied between $40/dry ton and $90/dry ton. The biomass price used for the study is $67.51/dry ton. Biomass feed price has a significant impact on the methanol RSP. The RSP decreases by $0.06/gal methanol for every $10/ton reduction in the biomass price.

Figure 5-8 RSP Sensitivity to Biomass Price



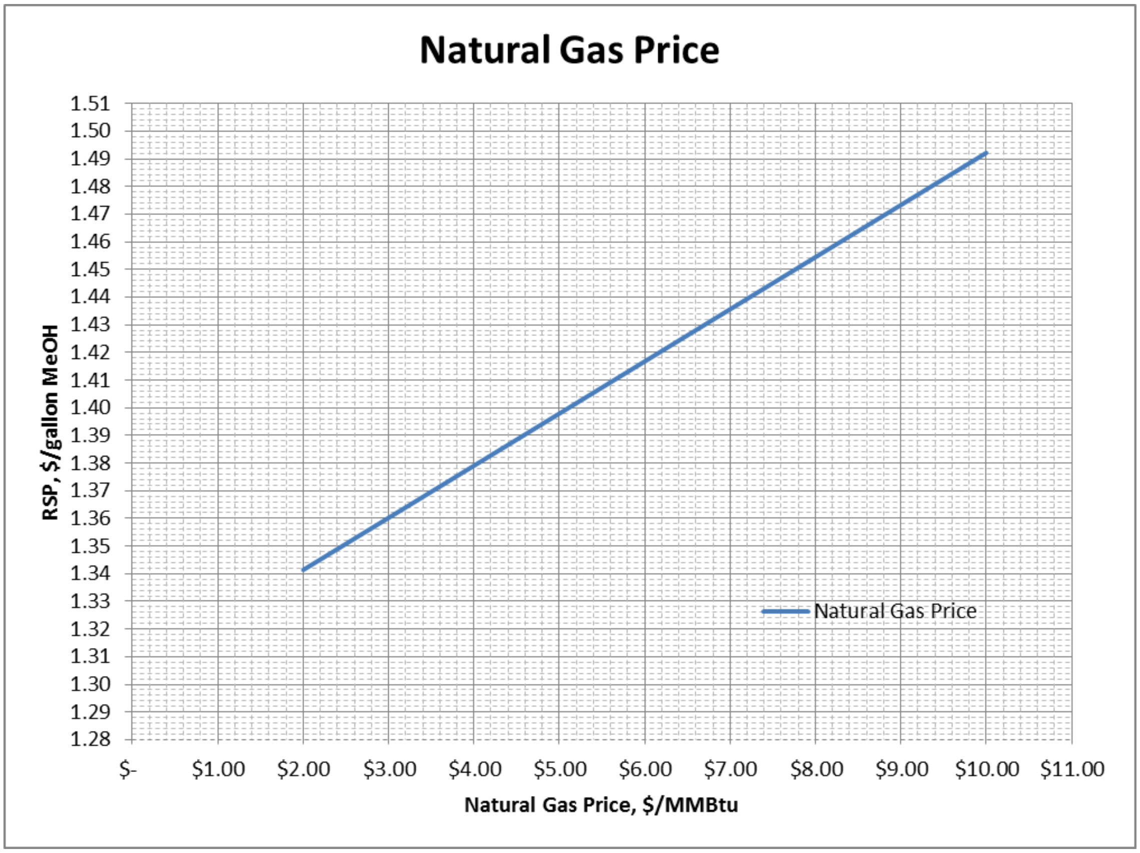
5.5.5 RSP Sensitivity to Natural Gas Price

Figure 5-9 shows the RSP sensitivity to natural gas price. The RSPs are estimated based on the following baseline parameters:

Capacity Factor	90%
Biomass Price	$67.51/dry ton
Natural Gas Price	$4/1000 SCF
Cost of Electricity Import	$0.078/kWh

The natural gas price is varied between $2/1000 SCF and $10/1000 SCF. The natural gas baseline price used is $4/1000 SCF. The RSP increases by $0.019/gal methanol for every $1/MMBtu increase in the natural gas price. Lowering the NG price alone will not achieve the targeted reference design RSP of $1.28/gal MeOH.

Figure 5-9 RSP Sensitivity to Natural Gas Price



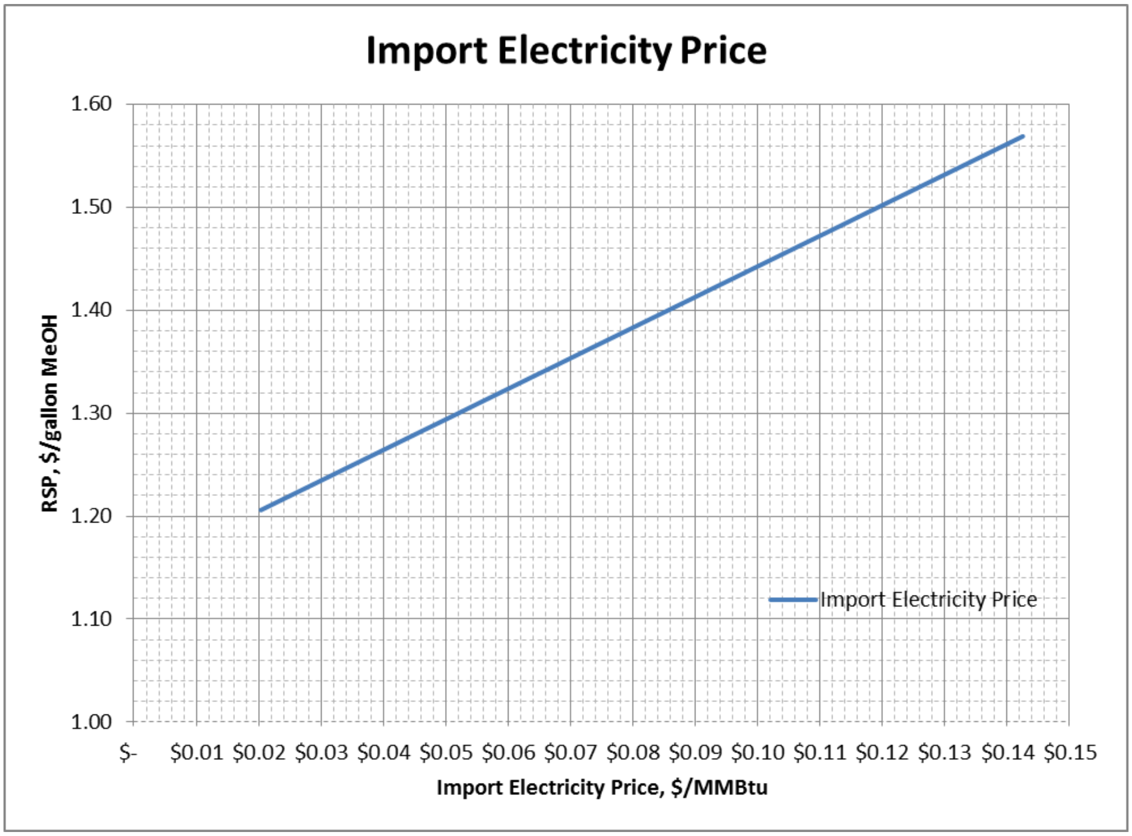
5.5.6 RSP Sensitivity to Import Electricity Price

Figure 5-10 shows the RSP sensitivity to import electricity price. The RSPs are estimated based on the following baseline parameters:

Capacity Factor	90%
Biomass Price	$67.51/dry ton
Natural Gas Price	$4/1000 SCF
Cost of Electricity Import	$0.078/kWh

The import electricity price is varied between $0.02/kWh and $0.14/kWh. The baseline electricity price used is $0.078/kWh. The RSP increases by $0.05/gal methanol for every $0.02 increase in the import electricity price. Lowering of the imported electricity price to $0.04/kWh will achieve the targeted reference design RSP of $1.28/gal MeOH.

Figure 5-10 RSP Sensitivity to Import Electricity Price



A preliminary techno-economic analysis (TEA) was performed to evaluate and compare the OSU's BTS gasification process against a conventional biomass gasification process, for fuel-grade methanol production. The reference case design representing a conventional indirectly-heated biomass gasification process was selected from a recent DOE report entitled "Techno-economic Analysis for the Thermochemical Conversion of Biomass to Liquid Fuels, June 2011, US DOE Pacific Northwest National Laboratory (PNNL-19009)." This report, along with its companion 2015 study "Process Design and Economics for the Conversion of Lignocellulosic Biomass to Hydrocarbons via Indirect Liquefaction, March 2015, NETL/TP-5100-62402, PNNL-23822," contain a detailed set of plant heat and material balances and cost estimates for the various process units within an overall biomass to methanol plant. These were used to develop the performance and cost estimates to evaluate a BTS process-based biomass to methanol design.

For this preliminary TEA study, Nexant first modeled the PNNL indirectly-heated gasification biomass to methanol (IHGBTM) reference design using ASPEN, and benchmarked the heat and materials balance results and cost estimates using the published data from the PNNL reports. This was done to establish a working model, allowing the balance of plants to be identified, utility consumptions and the overall plant performance determined. The benchmarked model was then used as a tool to estimate the performance of the overall OSU BTS gasification-based biomass to methanol (OSUBTM) process, of which the BTS gasification system performance data and cost estimates were provided by the OSU. It is understood that the BTS gasification system performance data provided at this point is only preliminary and not yet fully optimized. The current TEA study is to be updated with actual data obtained from the experimental portion of the project.

The plant performance of the BTS process-based OSUBTM design and its estimated total project investment cost are shown in Table 5-2 and 5-14 respectively. In comparison with the IHGBTM Reference Design, the OSUBTM plant performance shows:

- Higher carbon efficiency – it consumes the same amount of biomass feed but produces approximately 20% more crude methanol, as a result of a higher amount of syngas that is generated by the BTS gasification process, but

- Lower overall plant thermal efficiency - the higher carbon efficiency for the OSUBTM plant negatively resulted in less energy available to support its overall in-plant steam utility and power consumption requirements. Main reasons for its higher power and fuel requirements are due to:

  - Increased syngas flow - resulting in a higher syngas compression power consumption downstream of the BTS process unit.
  - Increased CO_2 content (hence flow) in the syngas stream – resulting in a larger amine plant for CO_2 removal plant; and hence its associated energy requirement. The BTS

syngas contains twice as much CO_2 as that in the IHGBTM Reference Design. This increases the amine plant utility requirements such as amine regeneration steam, condensing duty, and circulation load significantly. In the PNNL IHGBTM Reference Design, char and tar are formed in the gasifier. The char is carried by the circulating heat transfer medium to a combustor and is combusted to provide the heat for gasification and biomass drying. The CO_2 from char combustion is vented to the stack hence reducing the flow and CO_2 content of the syngas. Whereas in the OSUBTM design, all the biomass carbon is converted to carbon oxides in the gasifier and no char or tar is formed. The CO_2 from gasification is carried in the syngas and processed through heat recovery, syngas cleanup and compression before CO_2 removal.

- Steam and air stream preheating requirement for the BTS gasification process, and
- A lower methane content in the syngas feed to methanol synthesis resulting in a lower heating value purge gas to be used as supplementary fuel for in-plant use. The net requirement is the need for importing natural gas as fuel, along with additional power.

Section 5 of the report provides a more detailed account of the OSUBTM model balance of plants results. The OSUBTM design has a higher estimated TPI cost at $347MM, of which the increase in the amine CO_2 removal plant cost due to a larger plant size requirement is the major culprit. The net result in the estimated RSP of methanol for the OSUBTM is at $1.38/gal, about 7.8% higher than that estimated for the IHGBTM design.

Given the modeling results generated from this preliminary TEA study, the cause and effect highlighted for its energy deficiency, it is recommended that an optimization study (1) to be carried out within the BTS gasification process to minimize its CO_2 content in the produced syngas, or (2) to seek an alternative BTS gasification integrated process scheme for methanol production to minimize its overall energy efficiency disadvantage.