Project Title: Chemical Looping Gasification for Hydrogen Enhanced Syngas Production with In-Situ CO_2 Capture

Final Technical Report

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Abstract

This document is the final report for the project titled “Chemical Looping Gasification for Hydrogen Enhanced Syngas Production with In-Situ CO_2 Capture” under award number FE0012136 for the performance period 10/01/2013 to 12/31/2014.This project investigates the novel Ohio State chemical looping gasification technology for high efficiency, cost efficiency coal gasification for IGCC and methanol production application. The project developed an optimized oxygen carrier composition, demonstrated the feasibility of the concept and completed cold-flow model studies. WorleyParsons completed a techno-economic analysis which showed that for a coal only feed with carbon capture, the OSU CLG technology reduced the methanol required selling price by 21%, lowered the capital costs by 28%, increased coal consumption efficiency by 14%. Further, using the Ohio State Chemical Looping Gasification technology resulted in a methanol required selling price which was lower than the reference non-capture case.

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Executive summary

Task 1.0: Project Management and Planning

A. Summary of Project Management Activities

The success of the project revolved around the proper management of resources, time, and people. Professor L.-S. Fan, the principal investigator (PI) of the project, was responsible for supervising research activities as well as ensuring the progress of the project. The project manager, Dr. Statnick, assisted the PI to coordinate collaborations among the participants and to ensure that progress was updated to the USDOE in a timely manner. The research team at OSU was responsible for conducting the experimental work, compiling the progress reports, and submitting the reports. The individuals in the research team maintained a log of all the activities taking place regarding the project and the log was checked by the project manager periodically. OSU was responsible for the creation and maintenance of operating procedures, safety and emergency manual. WorleyParsons Inc. was responsible for techno-economic analysis for various coal to high hydrogen content syngas using chemical looping configurations. The research team at OSU and the project manager worked closely with WorleyParsons to provide technical direction, based on which a comprehensive techno-economic analysis on the coal to methanol production using chemical looping technology was completed. Clariant consulted on the oxygen carrier development aspect of the technology and synthesized optimized particle compositions for OSU testing. A time-only non-cost extension was sought with USDOE'S approval for the period from 9/30/14 to 12/30/14 for completing techno-economic analysis and experimental studies for cold-flow model and bench-scale studies. Overall, the five quarterly reports, three presentations and regular updates were provided to the USDOE on the status of the project. For purposes of the final report, the techno-economic analysis performed by WorleyParsons Inc. was integrated as a separate attachment at the end of the OSU experimental work.

Task 2.0: Oxygen Carrier Particle Improvement

A. Goal and Objectives of the Task

OSU has previously synthesized a working oxygen carrier particle for Chemical Looping Combustion applications. This project work focused on developing a highly reactive oxygen carrier for Chemical Looping Gasification (CLG) applications using coal as the fuel. Systematic studies on particle optimization, including reactivity, selectivity and recyclability, were completed. The ultimate goal of this task was to develop a highly selective oxygen carrier particle which can be tested in a bench-scale reactor for syngas production and will potentially perform well in larger-scale pilot and commercial demonstrations. The objectives of the task were as follows:

1. Establish a scientific and well-directed particle development procedure with a series of logical experiments
2. Screen, test, and confirm oxygen-carrier candidates with good syngas production potential in various lab-scale apparatus in order to demonstrate high-quality syngas production at bench-scale level and potentially at pilot and commercial scale in future
3. Investigate the performance of various chemical properties of the oxygen carrier to provide direction to future research.

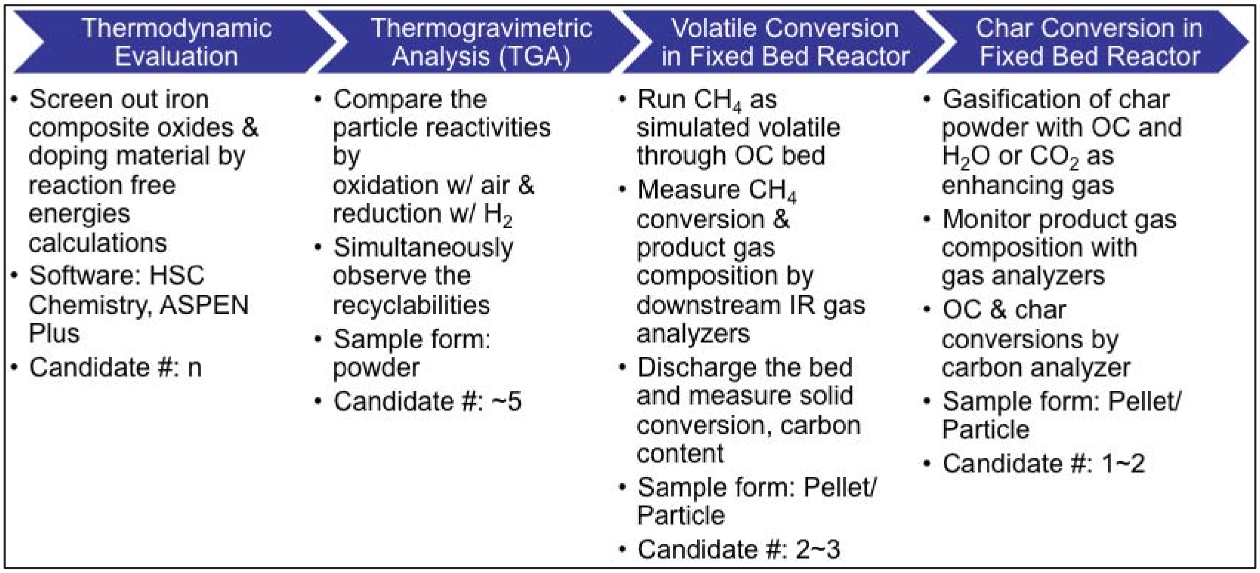
B. Experimental Methods

A standard screening/testing procedure was constructed to develop oxygen carriers for Coal Gasification of Chemical Looping.

Various material combinations based on thermodynamic calculations and guidance from previous experimental work and industrial partners was used to propose possible candidates for evaluation.

The preliminary testing screened out formulations based on the reactivity and recyclability criteria. These screening tests were carried out in the TGA with a combination of H_2 and air redox cycles, two promising formulations were selected. The performance of syngas production from coal volatile and coal char was tested with the chosen oxygen carrier formulations. The tests were carried out in a fixed bed reactor system in which the gas outlet compositions were monitored by IR gas analyzers. From the compositions, the oxygen carrier performances were calculated and quantified.

Exhibit 1: Particle development plan for testing solid fuel conversion



Thermodynamic Evaluation	Thermogravimetric Analysis (TGA)	Volatile Conversion in Fixed Bed Reactor	Char Conversion in Fixed Bed Reactor
• Screen out iron composite oxides & doping material by reaction free energies calculations • Software: HSC Chemistry, ASPEN Plus • Candidate #: n	• Compare the particle reactivities by oxidation w/ air & reduction w/ H_2 • Simultaneously observe the recyclabilities • Sample form: powder • Candidate #: ~5	• Run CH_4 as simulated volatile through OC bed • Measure CH_4 conversion & product gas composition by downstream IR gas analyzers • Discharge the bed and measure solid conversion, carbon content • Sample form: Pellet/ Particle • Candidate #: 2~3	• Gasification of char powder with OC and H_2O or CO_2 as enhancing gas • Monitor product gas composition with gas analyzers • OC & char conversions by carbon analyzer • Sample form: Pellet/ Particle • Candidate #: 1~2

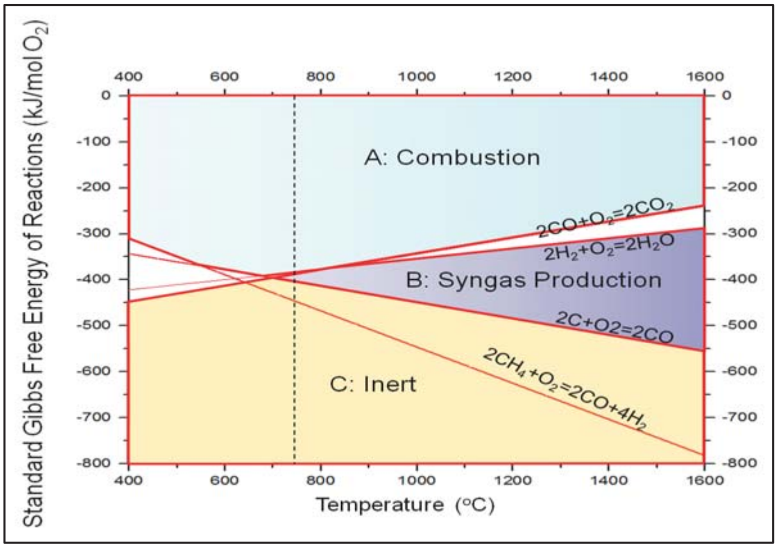
C. Results and Discussion

An Ellingham Diagram was constructed based on the species related to coal gasification. The diagram consisted of CH_4, CO, CO_2, H_2O, H_2 and pure carbon (C). The CH_4 represented the coal volatile, and C represented the coal char. With reaction temperature being the x-axis and reaction being the y-axis, the oxidation reactions of CO, C, H_2 and CH_4 drew out four curves, dividing the domain into three sections. Thermodynamically speaking, the higher the curve, the less spontaneous the reaction is. Therefore, if the oxidation curve of a metal oxide lays under the CO the H_2 oxidation curve, it indicates that the metal oxide tends to be oxidized by CO_2 and H_2O. As a result, the equilibrium gas concentration is CO/H_2 rich and favors syngas production. On the other hand, it is advantageous for the metal oxide curve to lie above the $2C + O_2 \rightarrow 2CO$ curve so that the equilibrium favors the carbon gasification instead of the reversed carbon deposition reaction. To meet these criteria, the reaction curve of desired oxygen carrier should lie within a certain triangular zone as shown in Exhibit 2.

After scanning through several elements including Al, Cr, Mn, Mg, V, and Zn combined with Fe were found to be promising. According to OSU’s extensive research on oxygen carrier development for chemical looping combustion, the material addition to Fe was expected to be a strong supporting material for the iron-based the particle. Considering the strength, heat capacity and cost, some were selected for further testing.

Eight formulations were tested in the TGA. All the Fe_2O_3 and support combinations show around 88% oxygen donation based solids conversion. This indicated that all the chosen oxygen carrier compositions had good reduction and oxidation reactivity. Certain cases showed a solid conversion based on iron to be higher than 100% which was because the support material participated in the oxygen donation process boosting the oxygen carrying capacity of the oxygen carrier material. The maximum time required to attain 50% solids conversion was around 1.3 minutes and the rates for all the compositions were comparable. Based on the heat transfer requirement for an auto-thermal chemical looping operation, the oxygen carrying capacity of the particles and ease of particle synthesis, the 50Fe50A and 50Fe50Al compositions were selected for recyclability testing. The change in reactivity over the 50 cycles was not significant.

Exhibit 2: Ellingham Diagram for selection of potential iron-based metal-oxide complexes



The two formulations were then tested in the fix bed setups. In the differential packing fixed bed, which simulated the solid conversion profile of a moving bed reactor, the oxygen carrier particles converted model coal volatile compound (CH_4) to high purity syngas ($CO:CO_2$ = 10:1) with above 90% CH_4 conversion. This indicated the oxygen carriers are capable of converting the coal volatile into syngas. Coal and coal char gasification were studied in fixed bed reactors. 50Fe50A and 50Fe50Al formulations gasified the solid fuels at similar reaction rates. However, under the same reaction conditions and amount of particle used, 50Fe50A showed a higher oxygen donation capacity for converting the coal to syngas and hence we chosen as the candidate for bench-scale testing studies.

D. Conclusions

The Ellingham diagram based thermodynamic calculations suggested that various iron-based metal oxides have the potential of generating syngas using the chemical looping concept. From the TGA studies 8 compositions, the 2 formulations were selected for particle synthesis and fixed bed experiments. The result of fix bed tests showed that both 50Fe50A and 50Fe50Al are both good candidates for Chemical Looping Coal Gasification, while 50Fe50A might be of slight advantage because the higher solid conversion potentially reduces the required amount of particle.

Task 3.0: Modification and Operation of the Bench-Scale Unit

A. Goal and Objectives of the Task

With the successful development of an oxygen carrier material suitable for partial oxidation applications, the OSU chemical looping gasification technology was tested in a bench scale moving bed reducer reactor. The goal of this task was to generate high purity syngas from various feedstock including coal volatile, sub-bituminous coal, and bituminous coal using the iron based oxygen carrier material in a co-current moving bed reactor. The composition (H_2: CO ratio) was controlled by co-injection of natural gas (methane- CH_4) and/or steam. The objectives of this task included:

1. Modifying the bench scale moving bed reactor system to facilitate the syngas generation tests
2. Achieving high purity syngas generation with >90% coal volatile conversion and >90% coal char
3. Demonstrating the syngas generation operation for more than 50 hours cumulative

B. Experimental Methods

The bench scale moving bed system was modified to enable co-current gas-solid contact pattern in the reactor. Apparatus/equipment for steam injection, gas sampling, ash removal, gas conditioning, and gas analysis were setup to monitor the product composition in the experiments.

CH_4 was used as the model compound for studying coal volatiles conversion in OSU chemical looping gasification process. CH_4 was injected into the bench scale moving bed reactor system as the feedstock co-currently with the oxygen carrier particles. Gas samples were obtained from different locations of the reactor to determine the concentration profile in the reducer. Sub-bituminous coal (PRB coal) and bituminous coal (Illinois #6 coal) were tested for syngas generation in the bench scale moving bed reactor system. Different feedstock combinations including a coal-only injection, coal co-injection with CH_4, and coal co-injection with CH_4 and steam were tested to generate syngas with different H_2:CO ratios. Feedstock combination determined by Aspen simulation in Task 5 was tested to verify the simulation result and give greater confidence to the economic simulations performed.

C. Results and Discussions

The modified bench scale moving bed system successfully demonstrated the establishment of the co-current gas-solid contact pattern and high purity syngas generation from various carbonaceous fuels. In coal volatile conversion tests, CH_4 was converted into a syngas with 90% dry basis purity and a H_2:CO ratio close to 2.0. The system also successfully converted coal into high purity syngas with a coal conversion of greater than 90%. Syngas with a H_2:CO ratio of 0.6, 1, and 2 was generated under the feedstock combinations of coal-only, coal co-injection with CH_4, and coal co-injection with CH_4 and steam, respectively.

D. Conclusions

The syngas compositions were close to the results of thermodynamic predictions. It can be concluded that the OSU chemical looping gasification technology with co-current moving bed reducer can effectively convert coal into high purity syngas. H_2-rich syngas with various H_2:CO ratio can be generated with the co-injection of CH_4 and/or steam. The syngas composition can be predicted with thermodynamic models.

Task 4.0: Cold Model Design and Operation

A. Goal and Objectives of the Task

A non-mechanical integrated reactor design is desired for the OSU CLG system to ensure proper and reliable residence times for the fuel, oxygen carrier and appropriate thermodynamic contact. The goal of this task was to initially design a cold flow model based on scaling factors to ensure proper sustained reaction scheme is feasible. The second goal was to construct and demonstrate good solids circulation, efficient fine separation and effective non-mechanical solids and gas flow control.

B. Experimental Methods

The process to obtain a design was based on assuming the realistic reactions feasible, a certain support weight fraction suitable for maintaining an auto-thermal operation and assumptions on residence time requirements as summarized in Exhibit 3 below:

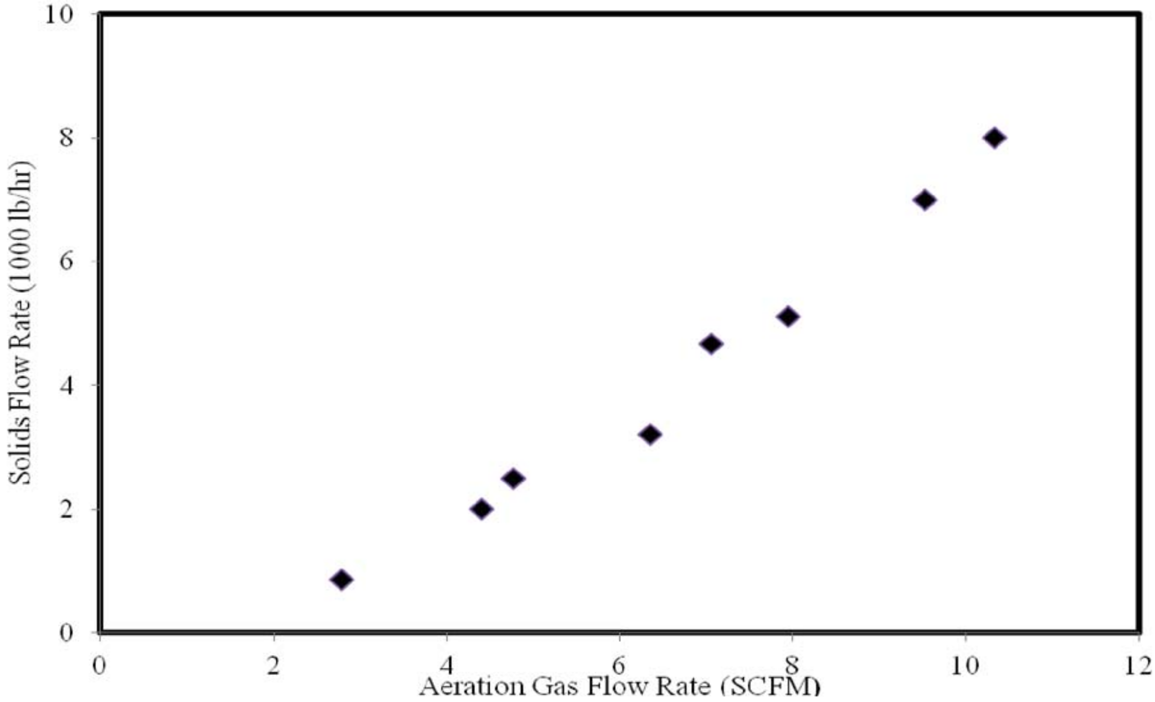
Exhibit 3: Scaling parameters for the cold-flow model based on design hot conditions

Parameters	Hot Unit	Cold Flow Model	Units
Capacity	25	25	kw
Gas Volume Flow Rate	20.51	54.70	slpm
CH_4 To OC Molar Ratio	1	1	-
Gas Molar Weight	16	29	g/mol
Oxygen Carrier Molar Weight	160	160	g/mol
Support To Oxygen Carrier Ratio	1	1	-
Mass Ratio	0.13	0.13	-
Pressure	1	1	atm
Temperature	800	300	K
Gas Density	0.24	1.17	kg/m3
Oxygen Carrier Density	2500	2500	kg/m3
Oxygen Carrier Volume Fraction	0.63	0.63	-
Gas Superficial Velocity	0.2	0.2	m/s
Reducer Height	1.5	1.5	m
Gas Residence Time	2.775	2.775	s
Solids Residence Time	22	22	min
Reactor Diameter	3	3	in
Minimum Fluidization Velocity	1.27	0.87	m/s
Combustor Gas Velocity	2.5	2.5	m/s
Combustor Diameter	3	3	in
Residence Time	5	5	min
Combustor Height	0.55	0.55	m

C. Results and Discussions

The system was assumed to use L-valve type design as a non-mechanical valve system. The hydrodynamic behavior of the L-valve design is as illustrated in Exhibit 4 below.

Exhibit 4: Relationship between aeration gas flow and solids circulation rate



When aeration gas is added to the L-valve, solid particles do not begin to flow until a minimum threshold amount of gas flow rate is added. When the aeration gas flow rate is low, the produced frictional force is not enough to initiate solids flow. Above the minimum threshold amount of gas flow, increasing the aeration gas flow rate causes the solids flow rate through the L-valve to increase. By controlling the aeration gas flow rate introduced to the L-valve, the solids flow rate of the chemical looping system can be controlled. Understanding the hydrodynamic behavior of the L-valve operation was an important milestone completed as a part of this project. Based on this understanding, a comprehensive scheme for sealing gases and the overall pressure balance was developed resulting in an integrated system design ready for further scale-up.

D. Conclusions

A detailed system design was completed based on hot to cold scaling factors. The L-valve design was analyzed in detail and the hydrodynamics for this specific oxygen carrier composition and flow demands were characterized. Based on the designed system, an integrated cold-model unit was developed and demonstrated. The system pressure balance was characterized for various zone seals injections and a scheme and design using non-mechanical valves was completed using the CLG system.

Task 5.0: Techno-economic Analysis

Syngas produced from coal gasification is currently used for integrated gasification combined cycle (IGCC) power generation and the production of chemical products including natural gas and methanol. IGCC power production provides a potential route for reducing greenhouse gas emission through pre-combustion carbon capture. The use coal gasification allows for the use domestic coal in chemical manufacturing and provides an opportunity for producing these commodities in the United States to provide supply security and an opportunity to export to overseas markets.

While coal costs tend to be low, the capital costs for gasification are high as a result of the gasification equipment and air separation unit (ASU) used for oxygen production. To address these high capital costs, Ohio State University (OSU) has developed a chemical looping gasification technology (CLG) to replace the gasifier and ASU components. In the OSU CLG process, an oxygen carrier (OC), iron oxide in the OSU approach, is circulated between an oxidizing and reducing reactor. In the oxidizing reactor, the OC is reacted with air to oxidize the iron. In reducing reactor, the OC is reduced by coal or natural gas to produce a syngas.

This techno-economic analysis investigates incorporating the OSU CLG technology into IGCC power generation and crude methanol manufacturing processes. Both of these processes require the production of a high hydrogen syngas. The technical options developed herein are based on and referenced to the Department of Energy National Energy Technology Laboratory (DOE/NETL) Cost and Performance Baseline Reports and the Baseline Analysis of Crude Methanol Production from Coal and Natural Gas.

A. Approach for completing the Techno-economic analysis

The approach to develop the assessment of the OSU CLG technology for high hydrogen syngas production included:

- Developing an Aspen model for the OSU CLG technology.
- Incorporating the OSU CLG model into an Aspen based model for power generation and methanol production.
- Determining the equipment size based on the process flows determined from the Aspen modeling.
- Developing capital and operating and maintenance (O&M) cost estimates based on the performance modeling
- Performing an economic assessment with the plant performance and cost estimating results. The economic assessment determined:
    - Cost of electricity (COE)
    - Methanol required selling price (RSP)
    - Cost of CO_2 captured
- Sensitivity studies around economic and process parameters

B. Summary of Findings

The initial process modelling of the OSU CLG technology using the Fe_2O_3-$FeAl_2O_4$ chemistry showed that a suitable syngas composition for methanol production could be achieved. The approach for producing H_2 in the oxidizer using the Fe_2O_3-$FeAl_2O_4$ chemistry has unfavourable thermodynamics for H_2 production, hence only the modelling of the methanol manufacturing process was performed in this study.

The methanol manufacturing process options compared in this study are summarized in Exhibit 5 below.

Exhibit 5: Methanol Manufacturing Options Compared in this Study

Option Label	**Feedstock**	**Syngas Production Method**	**Carbon Capture**	**Reference/ Basis**
MBL-1	Coal	Gasifier	No	NETL MBL Report
MBL-2	Coal	Gasifier	Yes	NETL MBL Report
MBL-3	Natural Gas	Reformer	Yes	NETL MBL Report
OSU-1	Coal	OSU CLG	Yes	This Study
OSU-2	Coal/Natural Gas	OSU CLG	Yes	This Study

A summary of the methanol production performance, the capital costs, and economic results are provided in Exhibit 6. A breakdown of the methanol RSP is provided in Exhibit 7. The OSU CLG technology provides several savings compared to the state-of-the-art coal base methanol production related to lower capital costs and higher feed stock utilization efficiencies. In comparing the MBL-2 and OSU-1 options, which are both based on a coal only feed and have CO_2 capture, the following benefits are realized:

- A lower methanol RSP by $0.37/gal, a 21% decrease.
- Lower capital costs by 28%
- Higher efficiency based 14% in coal consumption
- A methanol RSP lower than the reference non-capture case, which results in CO_2 capture cost less than 0.

Under the reference economic conditions, the methanol RSP of the OSU-CLG based systems were competitive when compared against the reference MBL-1 and MBL-2 which use coal as feedstock. A fair comparison with MBL-3 can be made by developing a study of the OSU-CLG system with natural-gas as the only feedstock.

Exhibit 6: Performance and Cost Results for Methanol Production Option

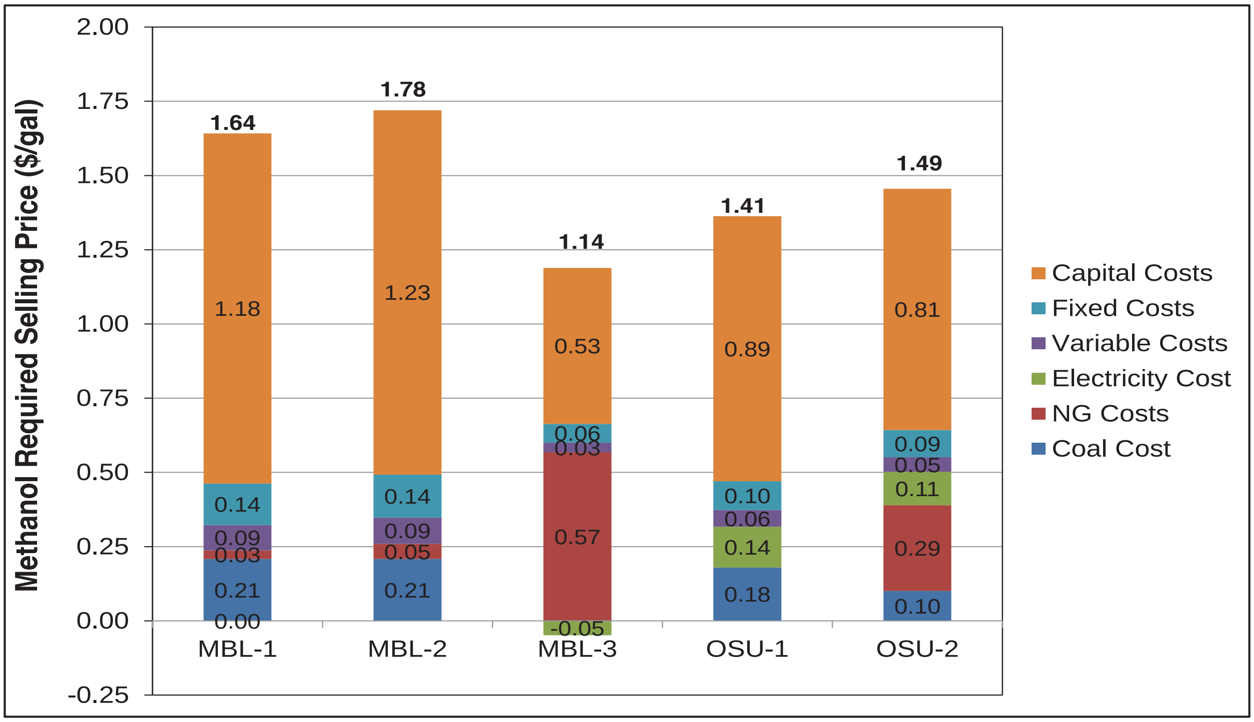
	Case				
	MBL-1	MBL-2	MBL-3	OSU-1	OSU-2
Performance Summary					
As Received Coal (lb/hr)	1,618,190	1,618,190	NA	1,395,457	718,631
Natural Gas to Reformer, OSU CLG (lb/hr)	NA	NA	583,677	NA	272,290
Crude Methanol (lb/hr)	941,823	941,823	940,989	918,582	918,582
Captured CO_2 (lb/hr)		1,569,410	235,808	1,302,138	663,393
Capital and Operating Cost Summary					
Total Plant Costs (2011 MM$)	4,586	4,775	2,028	3,497	2,996
Total Overnight Cost (2011 MM$)[N1]	5,572	5,802	2,485	4,236	3,634
Total As Spent Capital (2011 MM$)[N1]	6,580	6,852	2,935	5,003	4,291
Annual Fixed Operating Costs (x1000 $/yr)	156,650	162,051	70,644	108,291	94,034
Variable Operating Costs ($/gal)	0.085	0.088	0.032	0.056	0.049
Economic Comparison					
Methanol RSP ($/ton, 2011$)	495.27	535.58	346.56	425.106	446.69
Methanol RSP ($/Gal, 2011$)	1.64	1.78	1.15	1.41	1.48
Cost of CO_2 Captured	NA	15.66	NA	NA[N2]	NA[N2]

Notes:

[N1] The Total overnight and total as spent capital presented here for the MBL-1, MBL-2, and MBL-3 cases are different from those presented in the Crude Methanol Baseline report. This difference is related to using a consistent methodology for the owners' cost prepared for this study. The owners' costs methodology for the Crude Methanol Baseline study is not provided in that report.

[N2] The methanol required selling price for the OSU cases is less than that of the reference case, MBL-1. This is a result the OSU cases higher efficiency of the process and lower capital costs. Using MBL-1 as a reference, results in a negative carbon capture cost. Implications of a negative cost in this case are not clear.

Exhibit 7: Breakdown of Methanol Required Selling Price (RSP)



C. Sensitivity studies and recommendations for future work

Sensitivity studies were completed to identify parameters that have a significant impact on the production costs and thus identify opportunities to focus on for future cost reductions and parameters that could change and result in a change in the most favorable option. The results of this sensitivity study are:

- Feedstock Costs:
    - For coal based options (MLB-1, MLB-2, OSU-1 and OSU-2), the methanol RSP sensitivity to feedstock cost is significantly less than options that use only natural gas as a feedstock (MLB-3); approximately 10% with doubling coal costs and 30% with doubling natural gas costs.
    - With electricity cost increases from the reference price of $60/MWh to $130/MWh, the methanol RSP increases by 12% for the OSU CLG options. With the high electricity costs $130/MWh, the OSU options methanol RSPs are less than the RSP for the MBL-1 and MBL-2 options.
- OSU CLG Reactor Performance and Costs:

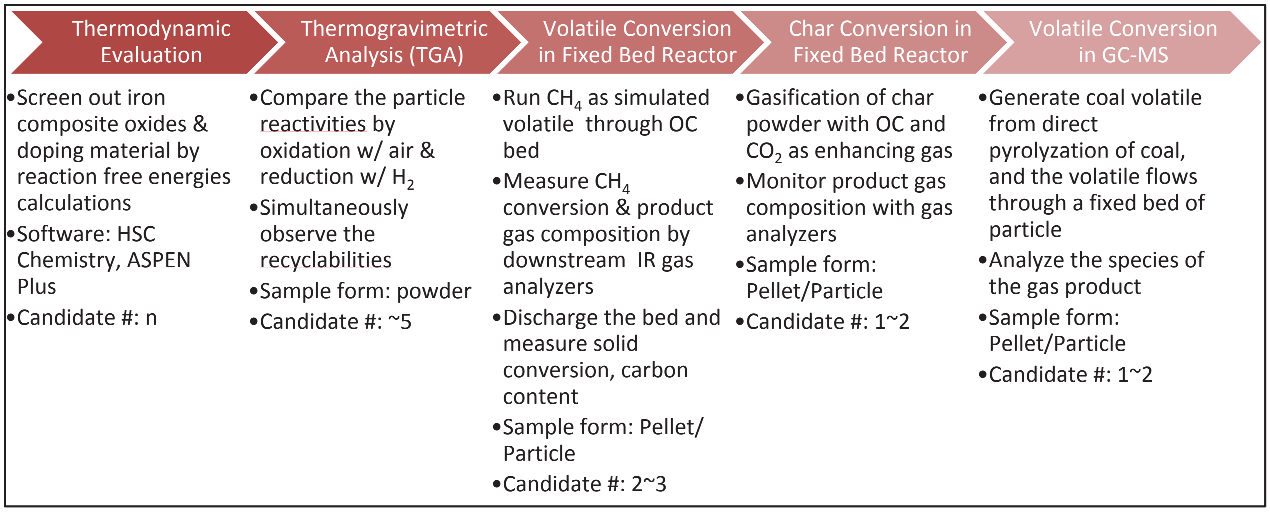
  - Faster reaction kinetics in the CLG will result in shorter residence times and larger reactor size. With a 50% decrease of the reactor residence time, the methanol RSP decreases by 4%. Future work should be performed to determine routes to increase the reaction kinetics between the coal and OC.
  - A 50% decrease the OSU CLG technology equipment costs results in a 5% decrease in the methanol RSP.
  - With a 50% increase in the OC cost, there is less than 1% increase in the methanol RSP. Therefore, uncertainty in the OC costs should not be a negative economic concern with regards to the further development of the OSU CLG technology.
  - Doubling the OC attrition rate, thus doubling the OC consumption rate during operations, increases the methanol RSP by less than 1%. This finding allows for greater flexibility in the improved OC materials. That is, materials that offer improved kinetics while attriting at a greater rate would be acceptable.
- CO_2 Emissions and Selling Price
  - The OSU CLG based options incorporate CO_2 capture and therefore essentially independent of CO_2 emission taxes. The methanol RSP for the OSU CLG options were lower than that of the MBL coal based option without CO_2 capture.
  - With CO_2 revenue up to \$50/ton CO_2, the methanol RSP decreases by 15% and 8% for the OSU-1 and OSU-2 options respectively.

I. Experimental Methods and Approach

A. Particle studies

The development of an appropriate metal oxide oxygen carrier with support material for particle manufacture is critical for a pilot and/or commercial scale demonstration of the Coal Gasification Chemical Looping (CLG) process. Therefore, a rigorous investigation was performed in this project. In general, there were five categories for particle screening and testing. First, multiple formulation candidates were proposed based on the thermodynamic principle of reaction Gibbs free energy (ΔG) and temperature of operation. A thermogravimetric analyzer (TGA) was used to investigate the particle reactivity and recyclability. A differential packing fixed bed was designed to simulate syngas generation from coal volatiles in moving bed mode. A standard fixed bed was used to demonstrate coal and coal char gasification. In addition, gas chromatograph-mass spectrometer (GC-MS) was used to qualitatively investigate the volatile cracking capabilities of the oxygen carriers. Exhibit 8 shows the flow of oxygen carrier development procedure.

Exhibit 8: Particle development flow diagram



Thermodynamic Evaluation

- Screen out iron composite oxides & doping material by reaction free energies calculations
- Software: HSC Chemistry, ASPEN Plus
- Candidate #: n

Thermogravimetric Analysis (TGA)

- Compare the particle reactivities by oxidation w/ air & reduction w/ H_2
- Simultaneously observe the recyclabilities
- Sample form: powder
- Candidate #: ~5

Volatile Conversion in Fixed Bed Reactor

- Run CH_4 as simulated volatile through OC bed
- Measure CH_4 conversion & product gas composition by downstream IR gas analyzers
- Discharge the bed and measure solid conversion, carbon content
- Sample form: Pellet/ Particle
- Candidate #: 2~3

Char Conversion in Fixed Bed Reactor

- Gasification of char powder with OC and CO_2 as enhancing gas
- Monitor product gas composition with gas analyzers
- Sample form: Pellet/Particle
- Candidate #: 1~2

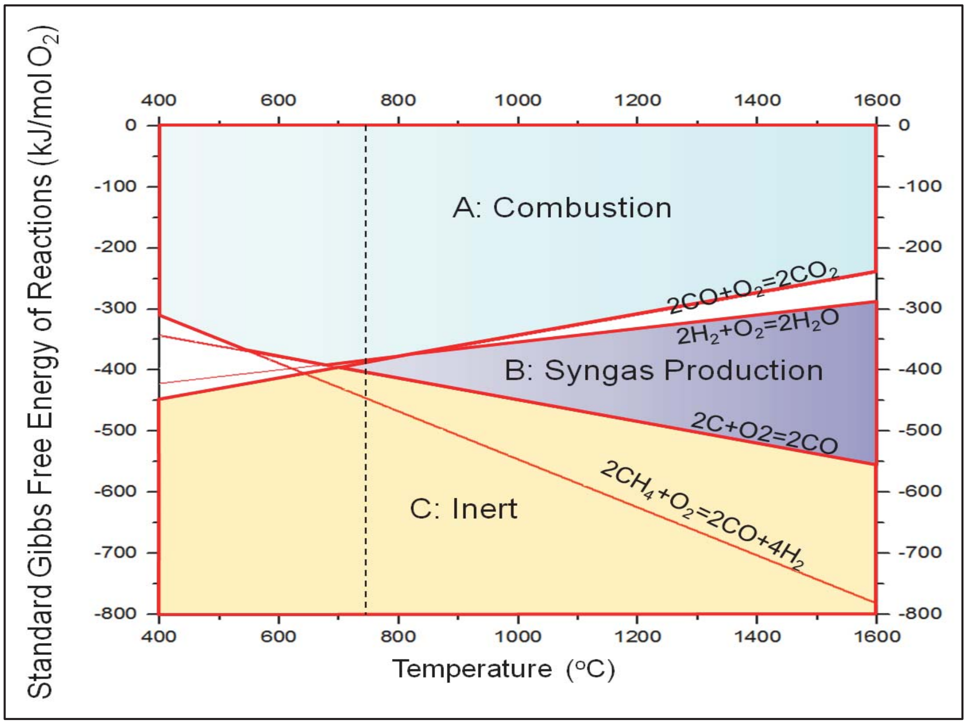
Volatile Conversion in GC-MS

- Generate coal volatile from direct pyrolyzation of coal, and the volatile flows through a fixed bed of particle
- Analyze the species of the gas product
- Sample form: Pellet/Particle
- Candidate #: 1~2

1. Thermodynamic Evaluation

The reduction behavior of oxygen carrier significantly affects the performance of CLG systems. In the reduction step, oxygen carrier should provide with a suitable oxidation capacity to partially convert carbonaceous fuels to syngas while preventing full oxidation to CO_2 and H_2O. Incomplete fuel conversion would require additional post-treatment systems, which decreases the overall process efficiency and increases the production costs. The reduction extent of the oxygen carrier also determines the solids circulation rate, which has a direct positive influence on the scale-up of the chemical looping process. The reduction properties of oxygen carrier largely depend on the thermochemical properties of the active metal oxides. Thus, the oxygen carrier material selection should focus on metal oxides that pose favorable equilibria. Such material selection can be conducted with the assistance of an adapted Ellingham diagram as shown in Exhibit 9.

Exhibit 9: Ellingham Diagram for Oxygen Carrier Material Selection



The Ellingham diagram illustrates the dependency of the Gibbs free energy changes (ΔG) of oxidation reactions with temperatures. It is a powerful thermodynamic phase diagram to evaluate the reduction and oxidation potentials of oxygen carrier materials. It follows the second law of thermodynamics:

$$\Delta G = \Delta H - T\Delta S$$

Where ΔH is the heat of reaction, and ΔS is the entropy change.

ΔG-T plots in the Ellingham diagram are generally linear, except when phase transitions occur. Phase transitions could increase or decrease entropy, which is directly reflected on the slope change. For example, when gas (in a disordered state) reacts with solid (in an ordered state) forming another solid, the entropy usually decreases, which leads to a positive slope in the Ellingham diagram. Two exceptions are the C and CH_4 lines, as their associated reactions are gas volume-increasing and iso-volumetric reactions, respectively.

An adapted Ellingham diagram, as shown in Exhibit 9, is constructed based on ΔG of the oxidation reaction of oxides from a lower valence state with 1 mole O_2 at 1 atm using HSC

Chemistry software. Ellingham diagram has been widely used in metallurgic studies as a useful tool to predict the favorable temperature for metal oxide reductions. This diagram can be further developed into a powerful method to evaluate the potential of metal oxides as oxygen carrier materials for chemical looping processes. Materials are sorted into several major zones according to their thermodynamic feasibility for different chemical looping processes. It should be noted that since most chemical looping processes are operated at temperatures higher than 750˚C, thermodynamic behaviors of metal oxides at temperatures lower than 750˚C are not discussed in this work. These material zones are outlined by the following critical reactions 1-3:

Reaction line 1: $2CO+O_2=2CO_2$
Reaction line 2: $2H_2+O_2=2H_2O$
Reaction line 3: $2C+O_2=2CO$

Zone A: metal oxides in this region lie above reaction line 1 and reaction line 2. They are able to work as oxygen carriers for both full oxidation and partial oxidation. Metal oxides in this region include NiO, CuO, CoO, Fe_2O_3, Fe_3O_4, etc.

Zone B: metal oxides in this region lie above the reaction line 3 and lie below reaction line 1 and reaction line 2. They are able to work as oxygen carriers for partial oxidation but not full oxidation. Metal oxides in this region include CeO_2.

Zone C: metal oxides in this region lie above reaction line 3. They are not able to work as oxygen carriers for fuel oxidation and are considered as inert materials. Metal oxides in this region include Cr_2O_3, SiO_2, etc.

From this diagram, it is noted that CH_4 is not stable at temperatures higher than 750˚C and CH_4 decomposition to C and H_2 is a spontaneous reaction. Thus, the reaction line of $2CH_4+O_2=2CO+ 4H_2$ does not play a role in determining zones of oxygen carriers. Further, Ellingham diagram only provides a thermodynamic indication on the possibility of metal oxides to work as oxygen carriers. However, the reaction kinetics, reactant stoichiometry, flow dynamics and reactor designs play important roles. For oxygen carriers in Zone A, which are applicable for both partial oxidation and combustion, it is the control on the amount of oxygen transferred from oxygen carrier that determines whether the product is partial oxidation to produce syngas (CO, H_2) or full-oxidation to produce CO_2 and H_2O. Partial oxidation of fuel requires less oxygen transferred from the metal oxides than in full combustion for a given feedstock. There are two approaches to achieve this target in the current CLG processes. The first is to control the oxidation states of oxygen carriers entering the reducer from the combustor, while maintaining the oxygen carrier flow rate constant. For metal oxides in Zone A, the syngas yield increases initially until the stoichiometric ratio for CH_4 partial oxidation is exceeded. The syngas yield decreases afterwards due to the oxidation of syngas to CO_2 and H_2O. For metal oxides in Zone B, syngas yield increases linearly, reaching maximum at the stoichiometric ratio for partial oxidation and remains constant. For metal oxides that have multiple oxidation states, such as iron oxides, the syngas yield line shows an integrated pattern of individual syngas lines of Fe_2O_3 (Zone A), Fe_3O_4 (Zone A) and FeO (Zone B). The material screening study was conducted in the second quarter of the project.

2. Particle Preparation

Two primary oxygen carrier synthesis methods have been used for this study. The first one is dry mechanical mixing. In this method appropriate weight fractions of Fe_2O_3 and the supports in their powdered forms are mechanically mixed in a rotary drum. To ensure homogenous mixing, glass and/or metal beads are added to the drum during the mixing process. On completion, the glass/metal beads are separated and the mixture is sintered in a furnace at different sintering temperatures. Sintering promotes the metal oxide and support binding, which is believed to increase the mechanical strength of the oxygen carriers. In the second synthesis method, the Fe_2O_3 and the support powders are mixed in the rotary drum in a slurry form. Here again, glass beads are added to enhance the mixing. This method is called the wet mixing method and this method is predicted to yield a more homogenous mixture as compared to the dry mechanical mixing method. On completion of the mixing, the slurry is dried and sintered in a furnace.

The oxygen carriers prepared by both these methods have been tested in their powered forms as well as in the form of pellets. These pellets were then subsequently sintered in the furnace. The sintering procedure and temperature were optimized so that the sintered particle has both good reactivity and physical strength.

Six support materials have been studied in this quarter – Al_2O_3, MgO, SiO_2, Bentonite, Kaoline and Support A. The support materials chosen have good porosity and maintain a high surface area within the particle on sintering. They also have the potential to react with iron oxides forming complex compounds that can increase the syngas yield and selectivity. Exhibit 10 has a list of the oxygen carrier compositions tested. Henceforth in this document, the oxygen carrier compositions will be referred to with their respective code names as listed in Exhibit 10.

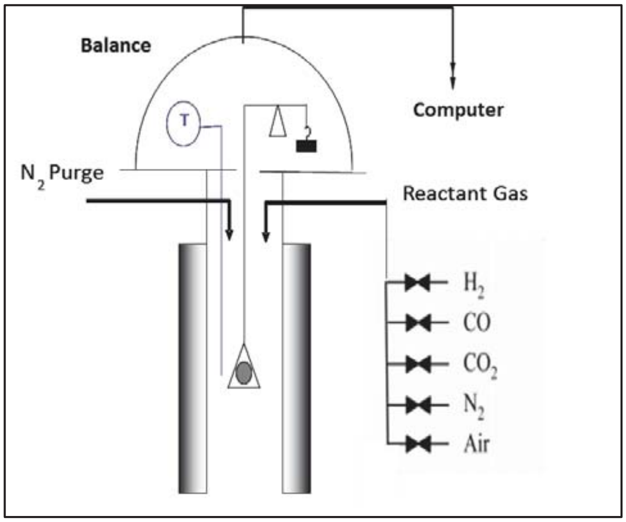
Exhibit 10: List of oxygen carrier compositions tested and their codes used for the study

Support	Metal Oxide:Support Weight Ratio	Code
A	60:40	60Fe40A
A	50:50	50Fe50A
Al_2O_3	60:40	60Fe40Al
Al_2O_3	50:50	50Fe50Al
MgO	60:40	60Fe40Mg
SiO_2	60:40	60Fe40Si
Bentonite	60:40	60Fe40BE
Kaolin	60:40	60Fe40KN

3. Reactivity/Recyclability Test in Thermogravimetric Analyzer (TGA)

Reactivity and recyclability tests were performed in a TGA. Exhibit 11 shows the TGA set-up with gas mixing panel.

Exhibit 11: TGA schematic used in Temperature Programmed Reduction Studies



A Setaram SETSYS Evolution Thermogravimetric Analyzer (TGA) was used to measure the reactivity of the oxygen carrier mixtures over a set number of redox cycles at 900°C. A gas mixing panel and a control sequence were used to control the mass flow rates of various gas species and altering the reducing, oxidizing and flushing gases to the TGA, respectively. The temperatures used in this screening were chosen to mimic the conditions in the larger-scale chemical looping system.

The first set of tests involved screening the compositions for their reactivities over three redox cycles. Each test started with a N_2 flushing step to ensure an inert atmosphere while the TGA was being heated to the reaction temperature. Each redox cycle involved a 20 minute H_2 reduction step, followed by a 5 minute N_2 flushing, and then a 20 minute Air oxidation step. The flushing step was interjected between the reduction and oxidation steps to ensure that there is absolutely no mixing between H_2 and air in the gas lines. At the end of the third cycle, there was a final flushing step with N_2 as the TGA cooled down.

After the preliminary reactivity screening tests, two oxygen carrier compositions were selected for long term recyclability studies based on their reactivity performance and other criteria essential for a commercial oxygen carrier particle. These selected compositions were further tested in the TGA over 10 and 50 redox cycles to study their long term stability, recyclability and redox profiles over large number of cycles. The reaction conditions for these tests were maintained the same as the 3 cycle tests. Since the 50 cycle test was only to investigate the recyclability of the oxygen carrier, the reduction and oxygen times were shortened to 10 minutes each to make efficient use of the instrument.

4. Differential-packing Fixed Bed for Volatile Conversion

The oxygen carrier particles were tested in a fixed bed reactor for coal volatiles conversion. The volatile matter in the coal is a mixture of organic and inorganic compounds. CH_4 was chosen as a model compound for fixed bed testing. The solid conversion profile of a moving bed reducer was simulated by a differential packing of the composite particles at different conversion values. The differential packing was achieved by packing 15 grams of reduced particles (50 % reduced) in the bottom section of the fixed bed. Another 15 grams of fresh particles highest oxidation state) were packed on the top of the earlier layer. The fixed bed reactor is made of steel and has an inner diameter of 1 inch. This is located in a vertical electric furnace with multiple thermocouples located along the axis for temperature control. The reacting gases were fed from the top of the reactor. The gas flow rates were controlled by mass flow controllers. The product gas from the fixed bed reactor passes through a $CaSO_4$ desiccant bed for conditioning before being analyzed by a CAI Infrared gas analyzer. The components that were quantified include CH_4, CO, and CO_2.

After the reactor was filled with particles, leak-tested and purged with N_2, it was heated to 990°C in the furnace. After the temperature was stabilized, a gas flow of 50% CH_4 (balanced by N_2) was introduced from the top of the reactor at 100 ml/min. This flow-rate was later lowered to 60 ml/min to study the effect of residence time. The gas product is diluted by a N_2 flow of 400 ml/min (later adjusted to 240 ml/min after adjusting CH_4 flow rate) before entering the CAI IR gas analyzer. This dilution is necessary based on the minimum flow rate requirement of the analyzer. The following reactions were expected to occur during the test:

$$CH_4 + FeO_x \rightarrow CO + H_2 + FeO_y \quad (x > y)$$
$$CO + FeO_x \rightarrow CO_2 + FeO_y \quad (x > y)$$
$$H_2 + FeO_x \rightarrow H_2O + FeO_y \quad (x > y)$$

5. Fixed Bed for Coal/Coal Char Conversion

A fixed bed reactor setup was used to compare the char gasification performances. The experiments were carried out using Illinois #6 coal and its processed coal char as the solid fuel. The fixed bed reactor is a stainless steel cylindrical chamber with a 1-inch inner diameter, placed vertically in a tubular furnace as shown in Exhibit 12. The chamber has a lock hopper on the top to isolate the gases while loading the solid mixture bed into the chamber. An enhancing gas stream consisting of 0.3 L/min carbon dioxide and 0.7 L/min N_2 was used throughout both oxygen carrier composition tests to accelerate the char gasification reaction. The individual flow rates of CO_2 and N_2 are determined based on the maximum detection limits of the Siemens Infrared gas analyzers downstream. The gas flow rates were controlled by ALICAT mass flow controllers. The product gas from the fixed bed reactor passes through an ash-removal water trap and a cobalt-based desiccant bed for conditioning. The components of the product gas stream were analyzed by Siemens Infrared gas analyzers in which the concentrations of H_2, CH_4, CO, and CO_2 were recorded.

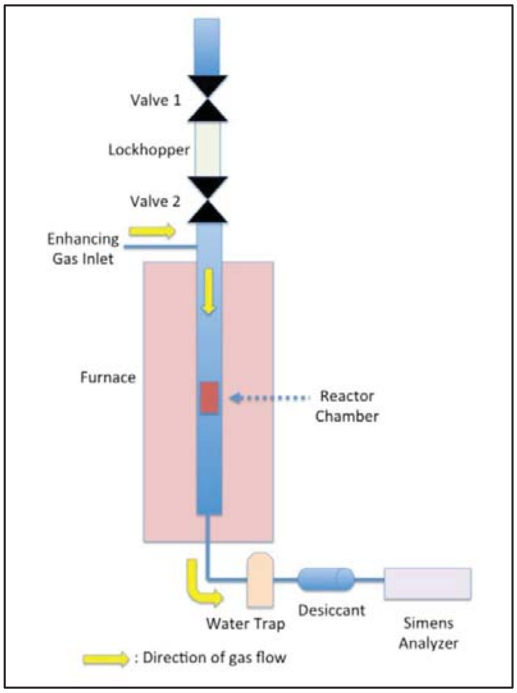
The experimental steps are as follow:

i. The empty reactor chamber is heated up to 1050℃ with the 0.3 L/min CO_2 and 0.7 L/min N_2 continuously flushing. Valve 1 and 2 are both closed.

ii. When the temperature and the CO_2 concentration reading are stabilized at 1050℃ and 30%, valve 1 is opened to fill the materials of the fixed bed into the lock hopper. The fixed bed consisted of two distinct sections. The top section is a mixture of 10 grams of composite metal oxide particles and 1.6 grams of coal powder. The bottom section is 10 grams of composite metal oxide particles which are at the highest oxidation state. Such a setup is to create more contact between the gasification products generated from the top section and the oxygen carriers at the bottom section to maximize the fuel conversion.

iii. After the fixed bed materials are filled into the lock hopper, valve 1 is closed. Then, valve 2 is opened, discharging the fixed bed materials into the chamber. The logging of the concentration readings on the analyzers is immediately started, with the discharge time logged as 0:00:00.

Exhibit 12: Fixed bed setup for analyzing coal and coal-char gasification in presence of the oxygen carrier particles



6. Volatile Cracking Test in Gas Chromatography–Mass Spectrometry (GC-MS)

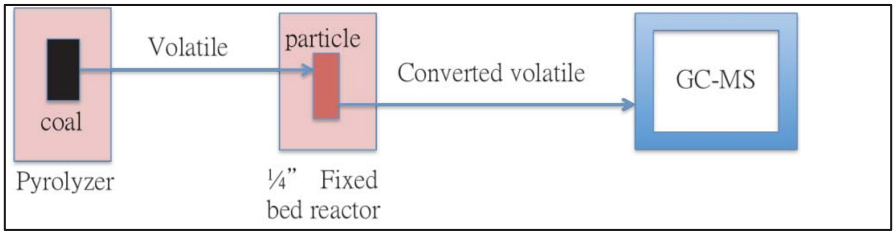
The GC-MS (Agilent Technologies, 7890B GC system/5977A-MSD) setup as shown in Exhibit 13 was used to study the volatile cracking capability of the oxygen carrier particles. The procedure of the experiment is as follow:

i. A small amount coal is placed in a pyrolyzer continuously flushed by a stream of carrier gas (50ml/min He). Then, the pryrolyzer heats up to the devolatilization temperature 900˚C, generating various types of coal volatiles.

ii. The carrier gas carries the volatiles and passes through a quarter-inch-diameter reactor. The reactor contains either (a) no particle, (b) a 50Fe50Al fixed bed, or (c) a 50Fe50A fixed bed. The volatile is expected to be thermally-cracked or cracked by the particles.

iii. The carrier gas further carries the cracked volatile into the GC column at where the various species separate due to different travelling speeds of the molecules. Then, the MS measures the mass-to-charge the ratio of the species. Based on information obtained, the exact formula of the volatile species can be determined. Noted that the lighter hydrocarbons and carbonaceous species such as CH_4, CO and CO_2 cannot be detected by GC-MS.

Exhibit 13: GC-MS set-up with coal pyrolyzer and a fixed bed set-up to study the tar cracking capacity of the oxygen carrier

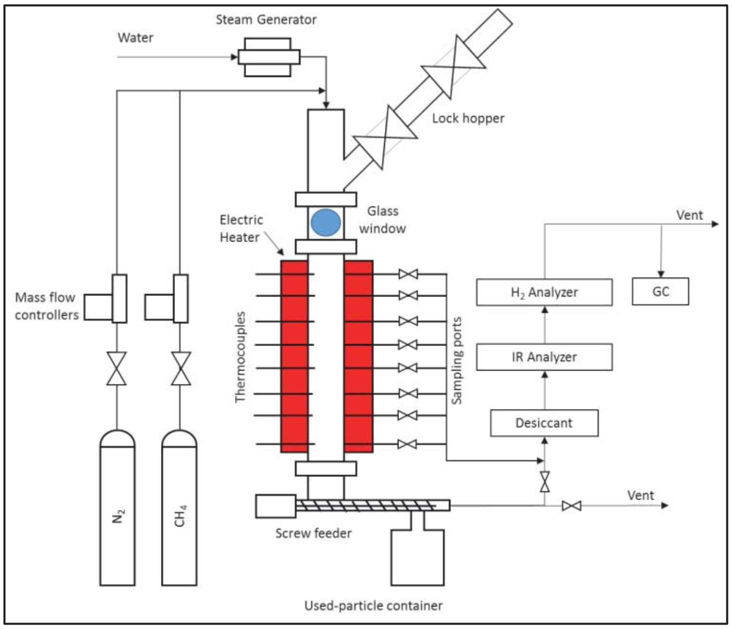


B. Laboratory Bench Unit Prototype Studies

1. Bench Scale System Modification

The bench scale moving bed reactor system is modified to facilitate CLG experiments as shown in Exhibit 14. On a high level, the system consists of a gas mixing panel, a moving bed reactor, and a gas analysis system.

Exhibit 14: Bench scale moving bed system



The gas mixing panel enables the injection of inert or reactive gases, including N_2, steam and CH_4, into the reactor. N_2 is used as an inert gas to flush out air from the reactor, as well as an internal standard for determining the flowrate of the syngas generated from gasification reactions. CH_4 and steam will be injected to enhance H_2 production in the reactor. N_2 and CH_4 are supplied by gas tanks, and their flowrate is controlled by mass flow controllers. In order to inject steam, a steam generator is installed on top of the reactor. The steam generator consists of a pump, which is capable of providing adjustable water flow and a cavity heated externally to vaporize water. All the gases are mixed and injected from top of the reactor.

The moving bed reactor consists of a 2 inch (5 cm) ID steel tube heated externally by electric heaters along with a screw feeder installed at the bottom of the reactor. Solid particles are constantly removed from the reactor by the screw feeder and fall into a container connected to the outlet of the screw feeder. Gaseous products flow out to the gas analysis system. A tilted pipe with a lock hopper is installed at the top of the reactor, from which pulverized coal and oxygen carrier particles can be added to the reactor. A glass window is located below the solid feeding pipe to monitor the solid level. Eight thermocouples are installed along the reactor height to measure the temperature profile. Opposite to the thermocouples, eight gas sampling ports are set to enable the sampling of solid and gas from different parts of the reactor.

Gaseous product from the reactor is sent to a gas analysis system. The gas is cooled using a condenser and is dried using a desiccant bed. Afterwards, the gas is sent through two gas analyzers and sampled by a micro-GC (gas chromatograph) before it is vented. The two gas analyzers include an IR analyzer for CO, CH_4, CO_2, and O_2, and a TCD analyzer for H_2. The data collected by the analyzers is sent to a computer and logged automatically.

2. Moving Bed Reducer Testing

The 50Fe50A oxygen carrier particles, which showed better kinetic performance from Task 2, were tested in the bench scale moving bed reducer for converting coal volatiles. CH_4 was used to mimic coal volatiles similar to the fixed bed test. The bench unit was setup and modified as described in the previous section. The oxygen carrier particles used in the moving bed reactor was spherical particles of 1.5mm diameter. The moving bed reactor was purged with N_2 and heated to a certain temperature before the injection of fuel gas. The solid flow rate in the reactor was set to be 20g/min, and a gas mixture of 90% CH_4 and 10% N_2 was introduced in a flow of 2 SLPM.

To test the conversion of coal to syngas, PRB coal was tested in the bench scale moving bed reducer for syngas generation. The proximate and ultimate analysis of a sample of the PRB coal is shown in Exhibit 15. The coal was pulverized and sieved to under 100 mesh (<150 micron). CH_4 and/or steam were co-injected with coal in order to adjust the composition of syngas produced, namely, increase H_2 content in syngas. The coal contribution in terms of fuel value was around 50% HHV as specified to be the maximum limit in attachment 2 of the RFP. The conditions of testing were carefully chosen to ensure a high syngas yield and a decent H_2:CO ratio by adjusting temperature, manipulating the solids flow, enhancing gas and the residence time.

Exhibit 15: Proximate and ultimate analysis of PRB coal

Proximate	%	
Moisture	13.505	as received
Ash	7.91	dry
Volatile Matter	41.29	dry
Fixed Carbon	50.80	dry
Ultimate	%	
Carbon	71.25	dry
Hydrogen	4.54	dry
Nitrogen	1.13	dry
Ash	7.91	dry
Oxygen (DIFF)	14.68	dry

Three sets of experiments were conducted to study the conversion of coal to syngas. In the first set of experiments, only coal was introduced with the oxygen carrier particles. The mass ratio between the oxygen carrier particle and coal was around 5. In the second set of experiments, CH_4 was co-injected into the reactor with coal and oxygen carrier particles. The amount of CH_4 was chosen to adjust the H_2:CO ratio of the syngas produced to 1. In the third set, CH_4 and water were co-injected with coal in order to further increase the H_2:CO ratio to above 1.8. In this case, the molar ratio among the reactants was about $C_{coal}:CH_4:H_2O:Fe_2O_3 = 4:6:5:7$, which was predicted by process simulations in Task 5 to sustain an auto-thermal operation of the integrated system.

A bituminous coal, Illinois #6, was also tested in the bench-scale moving bed system for syngas generation. As was done for PRB coal, the coal sample was pulverized and sieved. The size of the coal powder was under 100 mesh (<150 micron). CH_4 was co-injected with coal into the reactor in order to increase the H_2 concentration in the syngas produced. The proximate and ultimate analysis of the coal sample is shown in Exhibit 16.

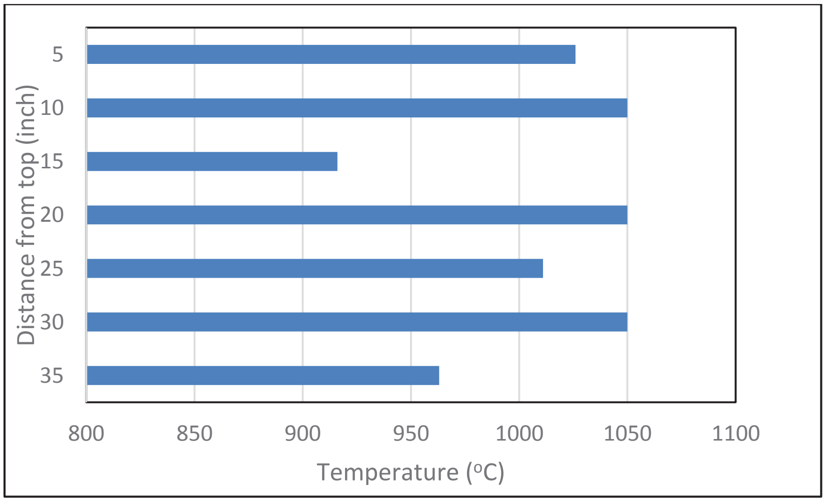
Exhibit 16: Proximate and ultimate analysis of Illinois #6 coal

Proximate	%	
Moisture	4.2	as received
Ash	11.1	dry
Volatile Matter	38.5	dry
Fixed Carbon	50.4	dry
Ultimate	%	
Carbon	70.2	dry
Hydrogen	4.8	dry
Nitrogen	0.9	dry
Oxygen (DIFF)	9.9	dry

In all the tests, the bench scale reducer reactor was heated the desired temperature, ranging from 950°C to 1000 °C, before the injection of fuel and/or water. Pulverized coal was mixed with the oxygen carrier particles before being added into the reactor from the lock hopper. After the experiment, the reduced oxygen carrier particles were examined in a carbon analyzer, where the particles were heated to 900°C in flowing oxygen. The effluent gas was titrated in an electrolytic cell, which determined the cumulative CO_2 content.

It should be noted that although the reactor temperature was set to a certain value, the actual temperature in the reactor was unevenly distributed due to the structure of the electric heater. The range of variation was to the tune of 100°C from the heater set-point. An exemplary temperature profile is shown in Exhibit 17. The reactor temperatures in the rest of this section refer to the peak temperature or the temperature set point of the reactor.

Exhibit 17: Exemplary temperature distribution in bench scale reactor



C. Integrated Cold Flow model studies

The CLG system, consists of the co-current moving bed reducer, the fluidized bed combustor, and other components that integrate the system and ensure the system operated as expect, including riser, cyclone, non-mechanical valve, gas sealing devices, and the gas product separation device.

1. Design Basis Assumptions

The conceptual design of the CLG process for scaling the cold flow model is based on the mass balance of the system. The design is based on the following assumptions:

1. The chemical reactions taking place in the chemical looping gasification process are equivalent to
   a. Reducer: $CH_4 + 2Fe_2O_3 = CO + 2H_2 + 2FeO$

   It is recognized that other reactions are taking place in the reducer, for example, coal de-volatilization, char gasification, and further oxidization of carbon/hydrogen to CO_2/H_2O and etc. However, to simplify the scaling process for the cold flow model, we assume the equivalent composition of coal is CH_4 and the conversion and selectivity are 100% without other products.

   b. Combustor: $4FeO + O_2 = 2Fe_2O_3$
2. The oxygen carrier has a composition of Fe_2O_3: support = 1:1. With the optimization of oxygen carrier particles is still ongoing, the detailed composition of the oxygen carrier may change. The composition will change the required solids circulation rate of the system. The higher weight fraction of the supporting material leads to a higher solids circulation rate. With the change of the composition, the detailed design of the system may change. However, for the purpose of verification of the system integration, it is acceptable to assume a composition of the oxygen carrier. And the assumed ratio is a reasonable value.
3. The particle density is assumed to be 2500 kg/m^3 with a particle size of 1.5 mm. Based on the extensive operational experience for similar processes developed in the Ohio State University, i.e., the counter-current Syngas Chemical Looping combustion system and the Coal Direct Chemical Looping combustion system, the selected particle properties ensure a continuous and stable operation of the integrated system.
4. It is assumed that the temperature in the system is constant at 800 K. Although for a commercial chemical looping gasification system, the temperature in the reducer may

decrease from the solids inlet to the solids outlet due to the endothermic reactions taking place in the process. However, in this study, to simplify the scaling process of the cold flow model, the temperature is treated at constant. It is expected that this assumption will have minor influence on the hydrodynamic characteristics of the reducer.

5. It is assumed that the residence time for the gas in the reducer should be larger than 2 seconds, and the solids residence time shall be over 20 minutes.
6. To simplify the scaling process, it is assumed that the gas properties are not changed through the reducer. Although for the real reducer reactor, the gas composition will change with the gasification process. Due to the chemical reaction, the gas volume will change, and the viscosity will change. However, this will be considered as the detailed design of the reducer, and can only be quantified when a workable operational condition is specified.
7. It is assumed that the required residence time of the oxygen carrier particles in the combustor is 5 minutes so that they can be fully regenerated.
8. The process gas in the cold flow model is air at room temperature of 300 K.

2. Design Parameters: Hot to cold model scale factors:

Based on the above assumptions, the conceptual design of a 25 kW_{th} chemical looping gasification system and its corresponding cold flow model are determined. Exhibit 18 shows key parameters for the reducer and the combustor for both hot unit and cold flow model.

With the parameters for the combustor being initially determined, the diameter of the riser is then calculated. This calculation is based on the requirement that there is enough gas velocity in the riser to entrain solids particles from the air reactor and transport them to the cyclone inlet at the top of the system. The determination of the riser diameter for the cold flow model is slightly different from that of the hot unit. For the hot unit, the gas in the riser is the oxygen-depleted air from the combustor, which is less than the air provided to the combustor. However, in the cold flow model, the air is not consumed, and thus the air flow rate in the riser is the same as that in the combustor.

Exhibit 18: Scaling parameters for the cold-flow model based on design hot conditions

	Hot Unit	**Cold Flow Model**	
Capacity	25	25	kW
Gas Volume Flow Rate	20.51	54.70	slpm
CH_4 To OC Molar Ratio	1	1	-
Gas Molar Weight	16	29	g/mol
Oxygen Carrier Molar Weight	160	160	g/mol
Support To Oxygen Carrier Ratio	1	1	-
Mass Ratio	0.13	0.13	-
Pressure	1	1	atm
Temperature	800	300	K
Gas Density	0.24	1.17	kg/m3
Oxygen Carrier Density	2500	2500	kg/m3
Oxygen Carrier Volume Fraction	0.63	0.63	-
Gas Superficial Velocity	0.2	0.2	m/s
Reducer Height	1.5	1.5	m
Gas Residence Time	2.775	2.775	s
Solids Residence Time	22	22	min
Reactor Diameter	3	3	in
Minimum Fluidization Velocity	1.27	0.87	m/s
Combustor Gas Velocity	2.5	2.5	m/s
Combustor Diameter	3	3	in
Residence Time	5	5	min
Combustor Height	0.55	0.55	m

D. Techno-economic analysis design basis document

A design basis document was generated to serve as a starting point for the comprehensive techno-economic analysis. The design basis document is included below:

1. General Design Basis

The OSU proposed a novel CLG system to selectively produce H_2 for efficient electricity generation or H_2-rich syngas for crude methanol production, both with minimal carbon emissions. The proposed project fell under the Area of Interest (AOI) 3, High Hydrogen Production, of the U.S. Department of Energy Nation Energy Technology Laboratories (DOE/NETL) Funding Opportunity Number DE-FOA-0000784. The following Design Basis sets forth the minimum criteria to be used in the Techno-economic Analyses Deliverables required by AOI 3. This AOI required:

1. At least one reference IGCC case using conventional technology and one IGCC case using the advanced technology, and
2. One reference gasification-based liquids production case using conventional technology and one gasification-based liquids production case using the advance technology.

For investigating OSU's CLG technology, crude methanol production was selected as the liquids production case. For methanol production, the syngas composition is controlled to facilitate methanol synthesis using state-of-the-art methods. The nominal production rate for the methanol facility is 50,000 barrels per day (bpd) of product (includes all liquid by-products). For electricity generation, a power plant with a nominal output of 500 to 550MW will be used as the basis. The OSU CLG will be developed to produce hydrogen to be fired in combustion turbine with subsequent heat recovery and steam generation (Rankine Cycle). Both processes will be designed to meet the DOE performance targets of 90% CO_2 capture, with a CO_2 purity greater than 95%.

The OSU CLG technology relies on employing a solid iron-based oxygen carrier to react with coal char in order to selectively generate a hydrogen-rich syngas, with near-zero CO_2 emission, which can then be used for power generation or crude methanol production. OSU has developed two distinct system configurations to be investigated which are described below:

1. Two-reactor system which uses natural gas co-feeding to increase the hydrogen content in the syngas (Refer to Exhibit 19).
2. A three-reactor system which inherently captures carbon from coal (Refer to Exhibit 20). In the liquid production case, steam and carbon dioxide are injected into the oxidizer in order to produce syngas required for methanol production. For electricity generation, only steam is injected into the oxidizer to produce hydrogen, which is then fed to the combine cycle power generation portion of the plant.

Exhibit 19: OSU CLG Two-reactor system

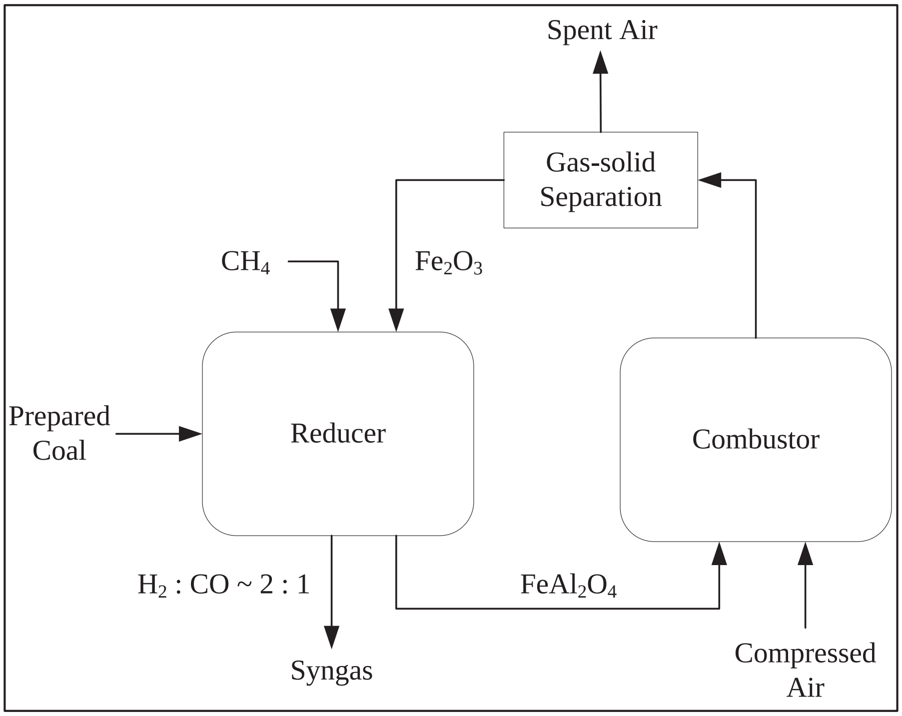
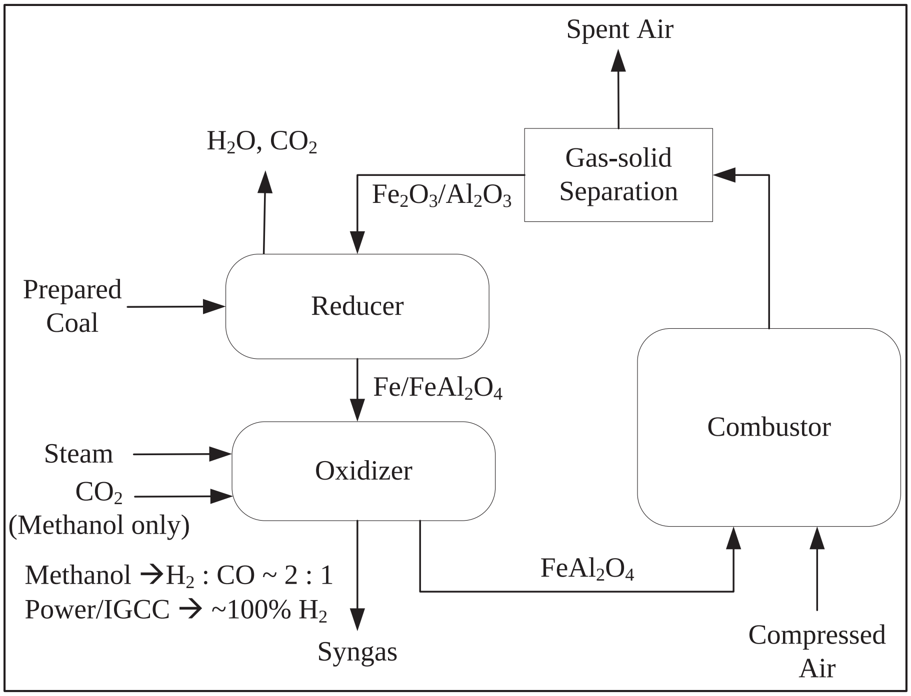


Exhibit 20: OSU CLG Three-reactor system



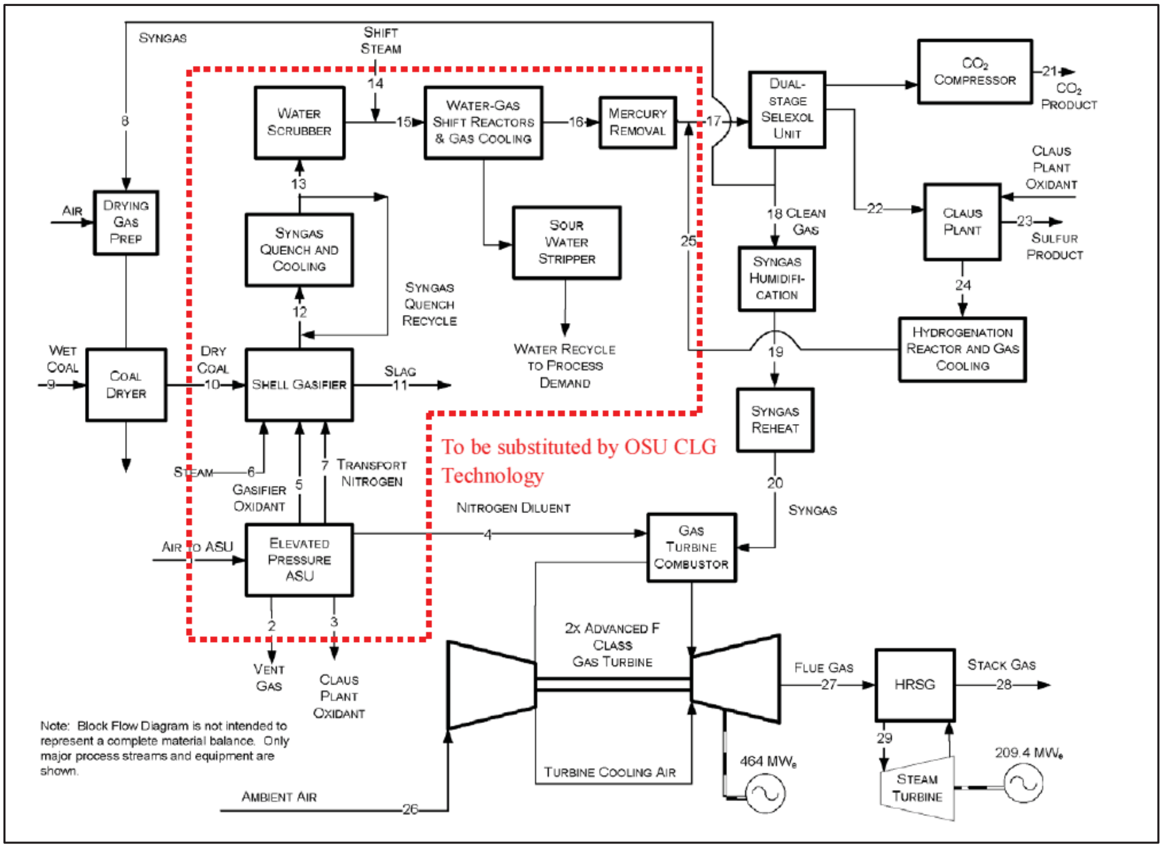
The selected cases that will be considered in the Techno-Economic Analysis are presented in Exhibit 21.

Exhibit 21: Selected Cases for Techno-economic Analysis

Case Description		Technology Status	Reference
Power Generation			
1.	Case 6 - IGCC case with Shell gasifier with no carbon capture/PRB Coal	Conventional	DOE/NETL Low Rank Coal Study
2.	Case 6 - IGCC case with Shell gasifier with carbon capture/PRB Coal	Conventional	DOE/NETL Low Rank Coal Study
3.	OSU CLG – IGCC case with carbon capture using 3-reactor system/PRB Coal	Advanced	N/A
Crude Methanol Production			
1.	Reference from DOE/NETL (pending)	Conventional	CTL Baseline
2.	OSU CLG – 2 reactor system	Advanced	N/A
3.	OSU CLG – 3 reactor system	Advanced	N/A

For power generation, the reference IGCC cases using conventional technologies were considered from the DOE/NETL Baseline Volume 3a study. The potential IGCC cases presented in this report utilize two advanced F turbines (232 MW each) and include gasification technologies offered by GE (radiant only), CoP E-Gas and Shell. Ultimately, the non-capture and capture cases based on the Shell gasifier, Cases S1A and S1B, of the aforementioned study was selected as the Shell gasifier utilizes a dry coal feed, in a similar manner than the OSU CLG technology. A process flow comparison between the proposed OSU CLG system and reference Case 6 can be observed in Exhibit 22.

Exhibit 22: Bituminous Baseline study – Case S1A IGCC with CO_2 capture BFD (Shell gasifier)



For crude methanol production the reference case using a conventional technology is pending the NETL's publication of CTL Baseline report.

2. NETL Publications

The following publications are referenced throughout the document and will serve as a basis for the TEA:

- NETL's "*Cost and Performance Baseline for Fossil Energy Plants*", referred to as "Baseline Studies"
  - "*Volume 3a: Low-Rank Coal to Electricity*" (May 2011)
  - "*Volume 4: Bituminous Coal to Liquid Fuels with Carbon Capture*" (pending), referred to as the "CTL Baseline"
- NETL's "*High Value Gasification Products: Crude Methanol Cases*", referred to as the "Crude Methanol study"

- NETL's Series of Quality Guidelines for Energy Systems Studies (QGESS).

3. Site Characteristics

The plant in this study is assumed to be located at a generic plant site in the Midwestern U.S. as specified in the Bituminous Baseline study. The site specification and ambient conditions are presented in Exhibit 23 and Exhibit 24. The ambient conditions are the same as the International Standards Organization (ISO) conditions.

Exhibit 23: Site Characteristics

Site Characteristics	**Midwest ISO**
Topography	Level
Size, acres	300
Transportation	Rail
Ash/Slag Disposal	Offsite
Water	50% Municipal and 50% Ground water
Location	Montana, USA
CO_2 Storage	Compressed to 15.3 MPa (2,215 psia), transported 80 km (50 mi) and sequestered in a saline formation at a depth of 1,239 m

Exhibit 24: Site Conditions

Site Conditions	**Midwest (ISO)**
Elevation, m (ft)	1,036 (3,400)
Barometric Pressure, MPa (psia)	0.09 (13.0)
Design Ambient Dry Bulb Temperature, °C (°F)	5.6 (42)
Design Ambient Wet Bulb Temperature, °C (°F)	2.8 (37)
Design Ambient Relative Humidity, %	62
Cooling Water Temperature, °C (°F)	15.6 (60)
Air composition based on published psychrometric data, mass %	
H_2O	0.616
Ar	1.280
CO_2	0.050
O_2	22.999
N_2	75.055
Total	100.00

The following evaluation considerations are site-specific, and will not be quantified for this study. Allowances for normal conditions and construction will be included in the cost estimates. Typically the considerations of these factors do not have a significant impact on the cost unless the site specific situation is unusual or extreme.

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

4. Fuel, Oxygen Carrier, CO_2 specifications

Coal and potentially natural gas are required by the OSU CGL processes. This section contains characteristics of both the coal and natural gas to be used in the techno-economic analysis.

Coal

The design coal in the project will be PRB coal with characteristics listed in Exhibit 25:

Exhibit 25: Design Coal Characteristics

Proximate Analysis	**Dry Basis, %**	**As Received, %**
Moisture	0.0	25.77
Ash	11.04	8.19
Volatile Matter	40.87	30.34
Fixed Carbon	48.09	35.70
Total	100.0	100.0
Ultimate Analysis	**Dry Basis, %**	**As Received, %**
Carbon	67.45	50.07
Hydrogen	4.56	3.38
Nitrogen	0.96	0.71
Sulfur	0.98	0.73
Chlorine	0.01	0.01
Ash	11.03	8.19
Moisture	0.00	25.77
Oxygen[1]	15.01	11.14
Total	100.0	100.0
Heating Value	Dry Basis	As Received
HHV, kJ/kg	26,787	19,920
HHV, Btu/lb	11,516	8,564
LHV, kJ/kg	25,810	19,195
LHV, Btu/lb	11,096	8,252
Hardgrove Grindability Index		57
Ash Mineral Analysis		%
Silica	SiO_2	38.09
Aluminum Oxide	Al_2O_3	16.73
Iron Oxide	Fe_2O_3	6.46
Titanium Dioxide	TiO_2	0.72
Calcium Oxide	CaO	16.56
Magnesium Oxide	MgO	4.25
Sodium Oxide	Na_2O	0.54
Potassium Oxide	K_2O	0.38
Sulfur Trioxide	SO_3	15.08
Phosphorous Pentoxide	P_2O_5	0.35
Barium Oxide	Ba_2O	0.00
Strontium Oxide	SrO	0.00
Unknown	---	0.84
Total	---	100.0
Trace Components		ppmd
Mercury[2]	Hg	0.081

[1] *By difference*

[2] *Mercury value is the mean plus one standard deviation using EPA's ICR data*

Natural Gas

The characteristics of the natural gas to be used in the techno-economic are listed in Exhibit 26:

Exhibit 26: Design Natural Gas Characteristics

Component		**Volume Percentage**
Methane	CH_4	93.1
Ethane	C_2H_6	3.2
Propane	C_3H_8	0.7
n-Butane	C_4H_{10}	0.4
Carbon Dioxide	CO_2	1.0
Nitrogen	N_2	1.6
	Total	100.0
Units	LHV	HHV
kJ/kg	47,454	52,581
MJ/scm	34.71	38.46
Btu/lb	20,410	22,600
Btu/scf	932	1,032

Non-Fuel Feedstock Characteristics

Limestone is a potential consumable that may be used for desulfurization. The characteristics of the limestone are shown in Exhibit 27.

Exhibit 27: Greer Limestone Characteristics

Component	**Dry Basis %**
Calcium Carbonate, $CaCO_3$	80.40
Magnesium Carbonate, $MgCO_3$	3.50
Silica, SiO_2	10.32
Aluminum Oxide, Al_2O_3	3.16
Iron Oxide, Fe_2O_3	1.24
Sodium Oxide, Na_2O	0.23
Potassium Oxide, K_2O	0.72
Balance	0.43
Total	100.00

Oxygen Carrier

The oxygen carrier to be used in the techno-economic assessment will be explored by OSU during the initial portion of the project. This exploration will define the parameters of the oxygen carrier to be used including:

- Oxygen loading
- Specific heat

CO_2 Specifications

The CO_2 is to be transported and injected as a supercritical fluid in order to avoid two-phase flow and to reach maximum efficiency. CO_2 is supplied to the pipeline at the plant fence line at a pressure of 15.3 MPa (2,215 psia). The CO_2 product gas composition varies, but is expected to meet the specification from the DOE/NETL Baseline studies summarized in Exhibit 28. If required, a glycol dryer located near the mid-point of the compression train is used to meet the moisture specification.

Exhibit 28: CO_2 Pipeline Specification

Parameter	Units	Parameter Value
Inlet Pressure	MPa (psia)	15.3 (2,215)
Outlet Pressure	MPa (psia)	10.4 (1,515)
Inlet Temperature	°C (°F)	35 (95)
N_2 Concentration	Ppmv	< 300
O_2 Concentration	Ppmv	< 40
Ar Concentration	Ppmv	< 10
H_2O Concentration	Ppmv	< 150

5. Balance of Plant Requirements, Plant Capacity Factor and Sparing Philosophy

Assumed balance of plant requirements are listed in Exhibit 29.

Exhibit 29: Balance of plant design requirements

Feedstock and Other Storage	
Coal	30 days
Slag	30 days
Sulfur	30 days
Natural Gas	Pipeline delivery (no on site storage)
Plant Distribution Voltage	
Motors below 1 hp	110/220 volt
Motors 250 hp and below	480 volt
Motors above 250 hp	4,160 volt
Motors above 5,000 hp	13,800 volt
Steam and Gas Turbine Generators	24,000 volt
Grid Interconnection Voltage	345 kV
Water and Waste Water	
Cooling system	Recirculating, Evaporative Cooling Tower
Makeup water	The water supply is 50 percent from a local POTW and 50 percent from groundwater, and is assumed to be in sufficient quantities to meet plant makeup requirements. Makeup for potable, process, and de-ionized (DI) water is drawn from municipal sources.
Feed water	Process water treatment is included and will produce boiler feed quality water for the gasification, natural gas and steam cycle systems.
Process Wastewater	Process wastewater and storm water that contacts equipment surfaces will be collected and treated for recycle. Selected blowdown will be discharged through a permitted discharge permit.
Water Discharge	Most of the wastewater is to be recycled for plant needs. Blowdown will be treated for chloride and metals, and discharged.
Sanitary Waste Disposal	Design will include a packaged domestic sewage treatment plant with effluent discharged to the industrial wastewater treatment system. Sludge will be hauled off site.
Solid Waste	Gasifier slag is assumed to be a solid waste that is classified as non-hazardous. An offsite waste disposal site is assumed to have the capacity to accept waste generated throughout the life of the facility. Solid waste sent to disposal is at an assumed nominal fee per ton, even if the waste is hauled back to the mine. Solid waste generated that can be recycled or reused is assumed to have a zero cost to the technology.

Plant Capacity Factor

For the coal to methanol cases, the plant capacity factor to be used will be 90%, per Attachment 2 of the RFP and consistent with the CTL Baseline, the CBTL study and the Crude Methanol study.

For the power generation case, the plant capacity factor to be used will be 80%, per Attachment 2 and consistent with the Baseline Studies.

Sparing Philosophy

For the equipment common to the reference methanol production plant and the power generation facility, the sparing philosophy used in the reference plant will be followed. For the equipment related to the OSU CLG technology guidance from OSU will be followed to ensure the design, including the proposed maintenance schedule, is capable of achieving the target capacity factor. Discussion of multiple trains and other potential redundancies need to be reviewed. There is no redundancy other than normal sparing for rotating equipment.

6. Cost Estimation Methodology

WorleyParsons will develop conceptual level capital and operation and maintenance (O&M) cost estimates for methanol production and power generation facilities incorporating the OSU CLG technology. These cost estimates will follow the Cost Estimation Methodology for NETL Assessments of Power Plant Performance and support the comparative economic analysis of the OSU processes with the DOE/NETL reference cases.

Capital Costs

Capital cost estimates will be developed using various models available to WorleyParsons for conceptual level capital cost estimating, including in-house proprietary parametric models and ICARUS from Aspen Tech. For equipment unique to the OSU CLG technology, ICARUS will be used for developing costs for reactor vessels, absorbers, and other specialized process equipment based on the equipment size, basic design, and materials of construction information developed through the process modeling or provided by the OSU. Factored estimates for equipment such as pumps, compressors, turbines, etc. will be developed using in-house proprietary models. For components common to those in the DOE reference cases, the capital costs provided in the DOE Report, escalated to June 2011 dollars will be used along with the developed capital costs to estimate a total plant cost.

O&M Costs

O&M costs estimates will be based on the methodologies provided in the Baseline Study Reports. For the reference non-capture and capture cases, the costs presented will be updated to the June 2011 dollars where required. For the CLG technology, costs will be developed based on inputs provided from the process modeling and guidance from OSU. The O&M accounts considered are listed and described in the following sub sections. The identification of O&M cost type, either fixed or variable is also provided in this list. The cost type indicates how the account is treated in the economic analysis.

Operating Labor (Fixed Cost)

The manpower estimates to operate and maintain the facility will be developed by the OSU with guidance from WorleyParsons using the manpower basis presented in the NETL/DOE Cost and Performance Baseline Reports as basis. Labor rates presented in the Cost and Benefit Reports and escalated to June 2011 will be used.

Administrative Labor and Support Labor (Fixed Cost)

Labor administration and overhead charges are assessed at a rate of 25 percent of the burdened O&M labor.

Maintenance Material and Labor (Fixed Cost)

Maintenance costs are evaluated on the basis of relationships of maintenance cost to initial capital cost. This represents a weighted analysis in which the individual cost relationships were considered for each major plant component or section.

Consumables (Variable Cost)

For general consumables such as limestone and water, costs provided by the DOE report escalated to June 2011 dollars will be used in the analysis. For consumables specific to the OSU CLG technology, the process modeling and guidance from OSU will be used to develop the consumption rates and costs. Further, input from OSU and catalyst vendors will be used for developing the oxygen carrier costs.

Waste Disposal (Variable Cost)

Waste quantities and disposal costs were determined similarly to the consumables. Ash and spent oxygen carrier from the OSU CLG technology will be treated in similar manner as fly ash and bottom ash from the PC cases in the NETL/DOE Cost and Performance Baseline Reports. Disposal costs from these studies will be escalated from to the current study reference, June 2011.

Co-Products and By-Products (Other than CO_2) (Variable Cost)

Co-Products Costs will be assessed according to Attachment 2 of the RFP. For the methanol production cases, net power imports or exports shall be priced at $60/MWh. Liquid co-product prices shall be indexed to the primary product (methanol) using the same ratios in the corresponding NETL studies.

By-product quantities will be determined similarly to the consumables. However, due to the questionable marketability of these by-products (bottom ash; fly ash co-mingled with FGD products, carryover from the OC material) no credit will be taken for potential saleable value.

CO_2 Transport, Storage and Monitoring (Variable Cost)

The CO_2 transportation, storage and monitoring costs are defined by the Attachment 2 of the RFP. Transport and storage (T&S) cost for CO_2 shall be based on a Midwest location at \$10/ton CO_2.

Economic Analysis

Upon the completion of capital and O&M cost estimates, WorleyParsons will perform a comparative financial analysis to compare the methanol production and power generation facilities incorporating the OSU CLG technology to state-of-the-art facilities. For the methanol production cases the first year selling costs will be determined. For the power generation cases, the analysis will estimate and compare the first year cost of electricity (COE) and the cost of captured CO_2 emissions using the plant performance and capital and O&M costs developed in the earlier tasks. The methodology for this analysis will follow that outlined in the DOE Report and Cost Estimation Methodology for NETL Assessments of Power Plant Performance. The global economic assumptions identified as the basis for the financial analyses in the DOE Report will be used in the analysis.

Sensitivity of required selling price (COE or methanol) will be performed on the following parameters:

1. Critical advanced technology performance parameters
2. Capital cost of advanced technology
3. Non-coal fuel prices
4. Sales of CO_2 at plant gate prices of \$0-60/ton
5. Cost of CO_2 emissions of \$0-60/ton
6. For liquids cases:
    - Power price for net imports/exports at \$60-135/MWh
    - Finance structure by assessing capital charge factors of 0.12-0.25

II. Results and Discussion

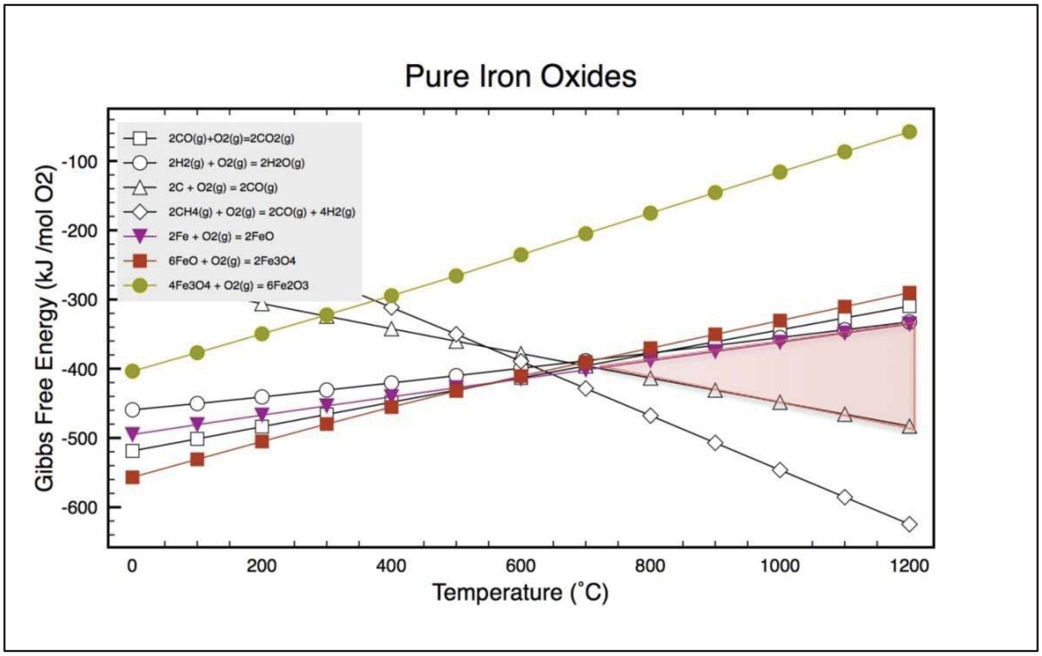
A. Oxygen Carrier Particle development and optimization

1. Material Search

The thermodynamic analysis guided the screening for a possible oxygen carrier composition. The compositions were used as a guide for determining the formula of the active material, support and dopant of the oxygen carrier. The maximum temperature limit was set at 1200 ℃, based on previous development work and material cost considerations. The Ellingham diagram is used for analyzing the choice of metal-oxide from a Gibbs free-energy minimization stand-point. The reaction curves are plotted for deriving the metal-oxide from its reduced state differing by a single mole of oxygen. This helps in a direct comparison on the position of the Ellingham diagram line for the particular metal-oxide. For two reactions on the same abscissa, the reaction with a lower ordinate is considered as favorable. As a result, if the two reactions are combined by eliminating the oxygen in the equations, the reaction with lower Gibbs free energy will be favored. Since the goal of this project is to generate H_2 and CO without over-oxidizing them to H_2O and CO_2, the target oxidation zone should be located above carbon gasification ($2C + O_2 \rightarrow 2CO$) line, and CH_4 partial oxidation ($2CH_4 + O_2 \rightarrow 2CO + 4H_2$). The favorable zone would also include a location below full-oxidation of syngas ($H_2 + O_2 \rightarrow 2H_2O$, $2CO + O_2 \rightarrow 2CO_2$). The identical set of 4 black curves in each Exhibit from 30 to 36 represented the four reactions mentioned above.

Iron oxides are known to be a good candidate for chemical looping combustion (CLC), which can convert carbonaceous fuel to energy while producing a high purity stream of CO_2. As shown in Fig. 13, pure iron oxide does not serve the purpose of this project for partial oxidation producing syngas, as the Fe_3O_4/Fe_2O_3 and FeO/Fe_3O_4 redox pairs lie above the syngas generation zone (shaded area), and Fe/FeO pair barely passes through.

Exhibit 30: Ellingham Diagram analysis using Iron-oxide as the metal carrier



However, a good selection of candidates was found with the combinations of iron and other metal elements. These metal elements include elements like chromium, manganese, magnesium, vanadium, tungsten and zinc, are analyzed below. The individual lines passing through the syngas generation zone indicate that the composite materials can thermodynamically reduce carbonaceous fuel to syngas without Carbon deposition. The synergistic effect brought by the addition of another metal to form an iron-oxide composite helps prevent Carbon-deposition.

Exhibit 31: Ellingham Diagram analysis using $FeAl_2O_4$ composite as the oxygen carrier

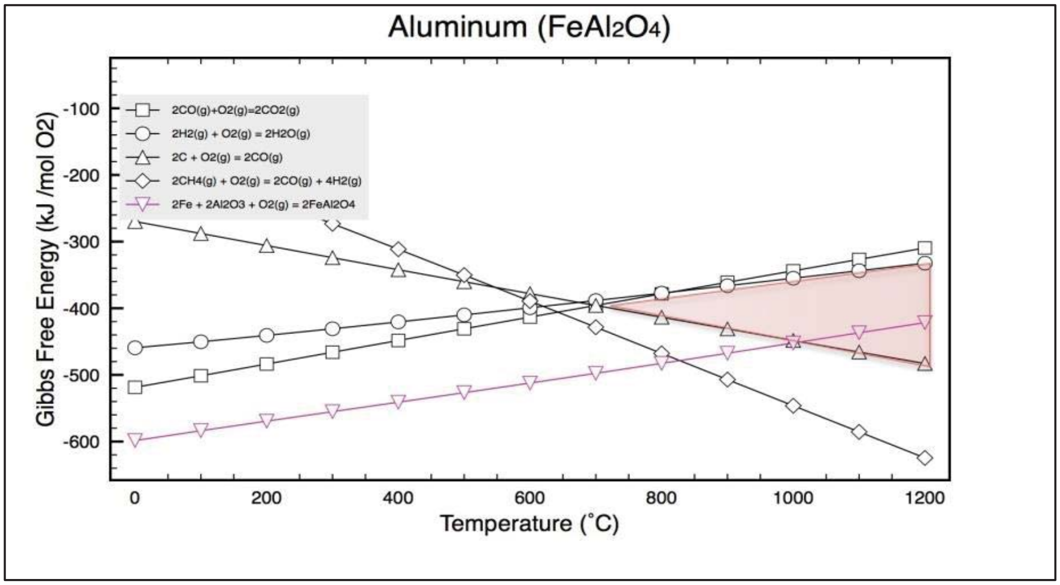


Exhibit 32: Ellingham Diagram analysis using Cr_2FeO_4 composite as the oxygen carrier

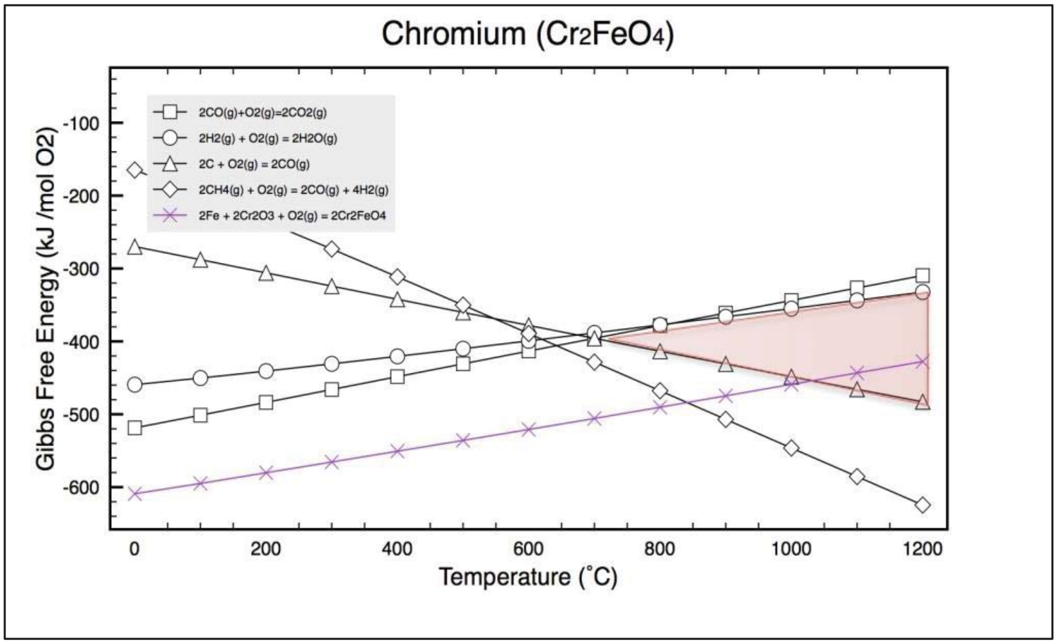


Exhibit 33: Ellingham Diagram analysis using Fe_2MgO_4 composite as the oxygen carrier

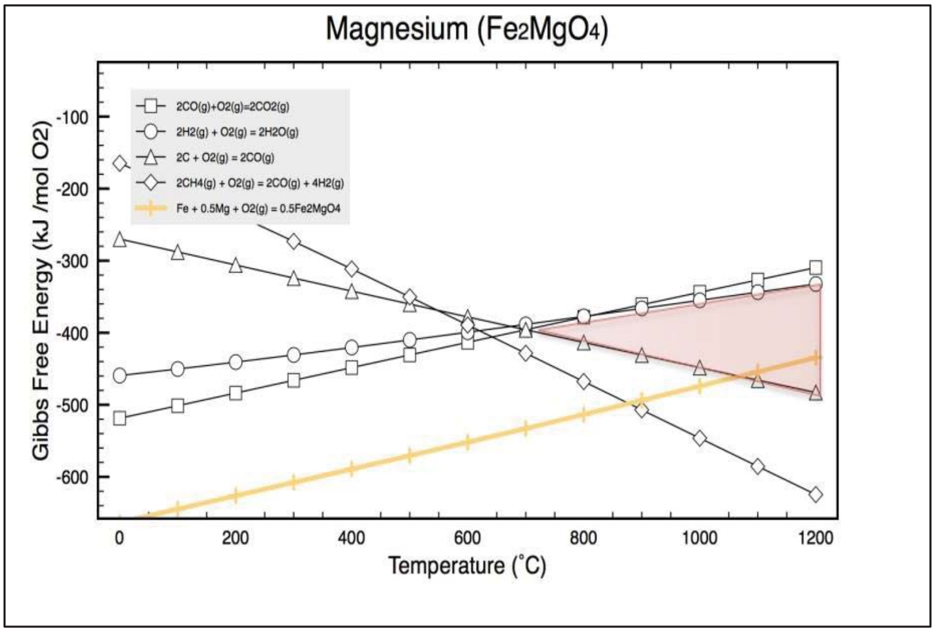


Exhibit 34: Ellingham Diagram analysis using Fe_2MnO_4 composite as the oxygen carrier

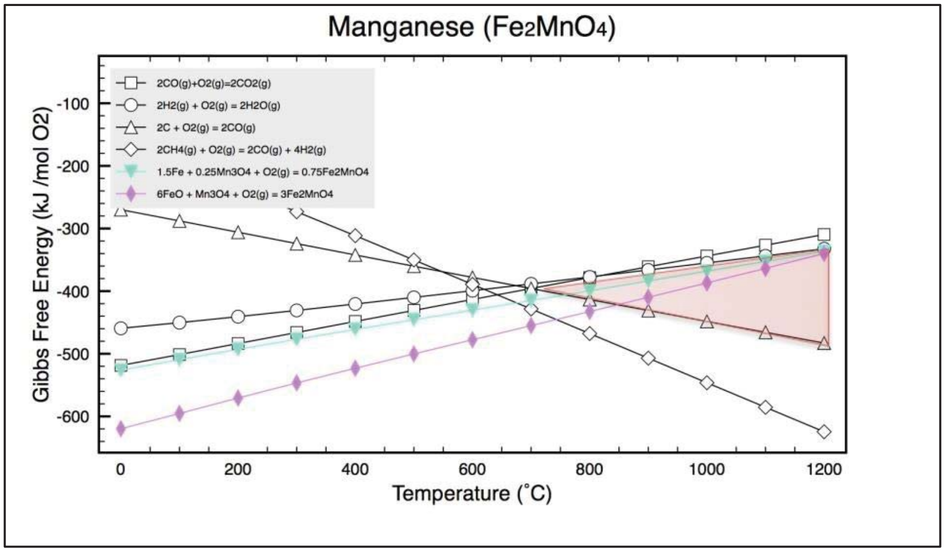


Exhibit 35: Ellingham Diagram analysis using FeV_2O_4 composite as the oxygen carrier

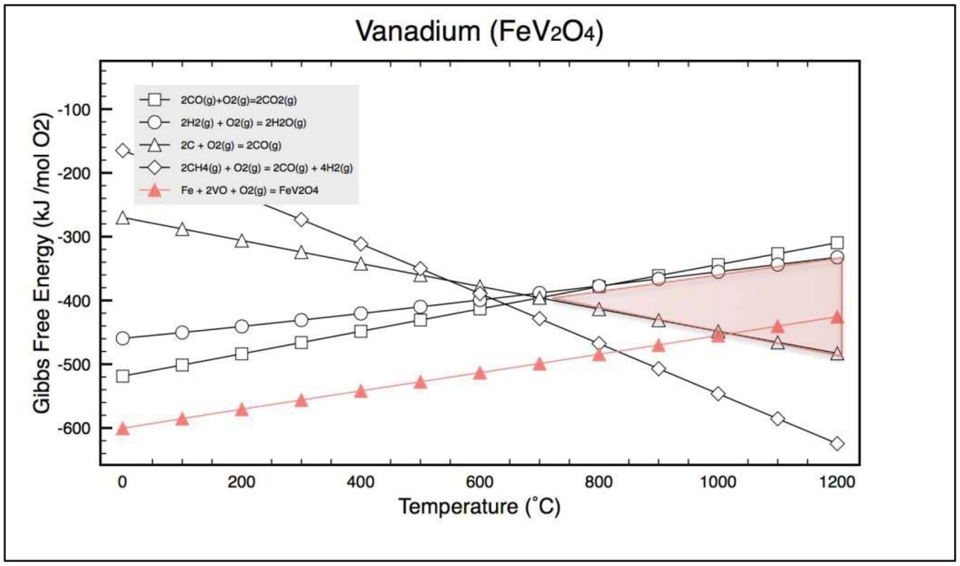
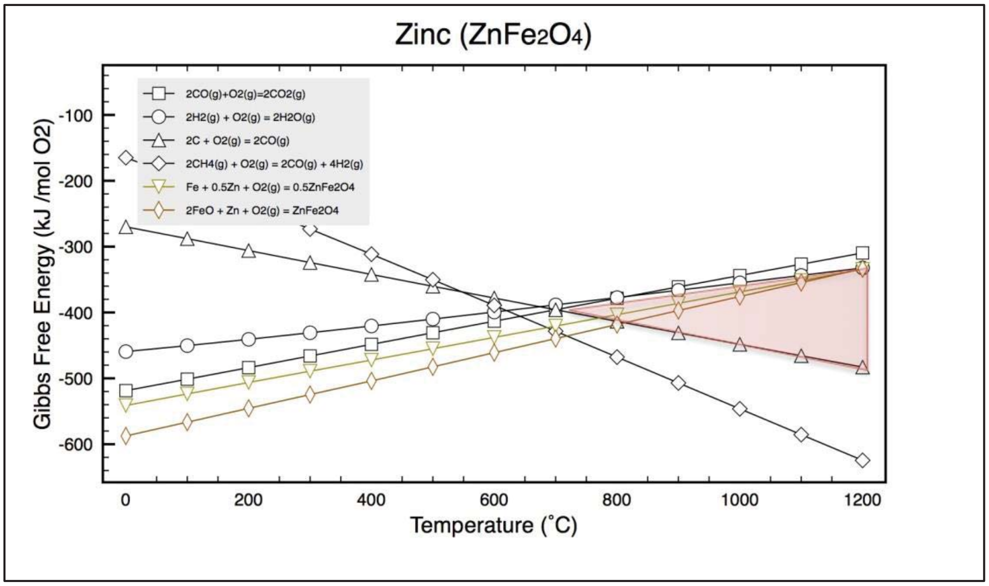


Exhibit 36: Ellingham Diagram analysis using $ZnFe_2O_4$ composite as the oxygen carrier



The thermodynamic analysis identifies a set of potential combinations of active material, support and dopant to provide a desirable equilibrium syngas composition. No kinetic data is considered in the Ellingham diagram analysis. For every composition, a robust reactivity is needed in order to quickly establish the steady state equilibrium to give the high syngas ratio. Therefore several accessible materials mentioned further preceded to the reactivity tests in a TGA.

2. TGA Reactivity/Recyclability Testing

For comparison, conversions during the 3rd redox cycle have been considered for each test. Exhibit 37 shows the percentage solid reduction attained in the 3rd reduction step during the 3 redox cycle tests. For the ease of calculation, the following assumptions have been made:

a. The oxygen carriers react uniformly in the TGA during the redox cycles
b. The weight change observed during the redox cycles were attributed only to the active metal oxide
c. 100% solid reduction corresponds to reduction of the active metal oxide to its lowest oxidation state

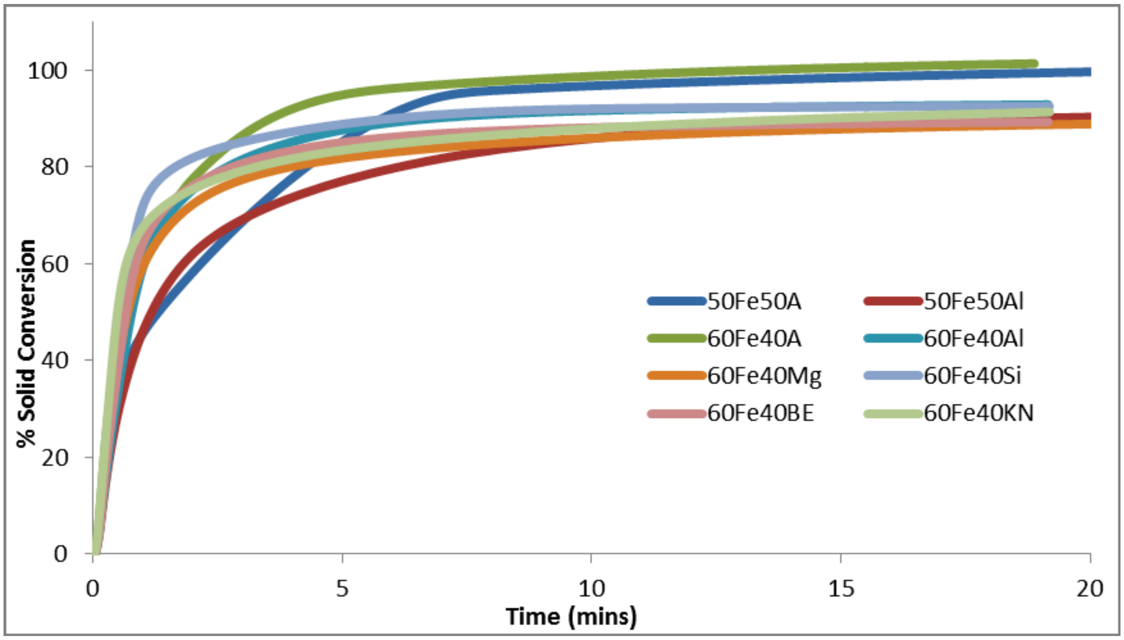
The % solid reduction has been calculated using the following formula:

$$\%\ \mathrm{Solid\ Reduction} = \left(\frac{\mathrm{W_i} - \mathrm{W_t}}{\mathrm{W_i} - \mathrm{W_{red}}}\right) * 100$$

Where W_i is the initial weight of the oxygen carrier, W_t is its instantaneous weight and W_{red} is the calculated weight corresponding to 100% reduction of the oxygen carrier. The reduction step has been considered for comparing the reactivity of the oxygen carrier compositions because it is the rate limiting step; the oxidation reaction is much faster.

As seen from the % solid reduction data given in Exhibit 37, all the Fe_2O_3 and support combinations show solid conversion over 88%. This means that all the oxygen carrier compositions tested have good reduction and oxidation reactivity. Certain cases show a solid conversion higher than 100% which could be because the support material is not completely inert and it boosts the oxygen carrying capacity of the oxygen carrier material.

Exhibit 37: % Solid reduction vs time in the third reduction step during the 3 redox cycle tests



Another possible reason for the solid reduction above 100 % could be incomplete mixing of the metal oxide and support During operation of a commercial scale chemical looping unit, optimization suggests that for complete fuel conversion and high syngas selectivity in the reducer the solid conversion required is between 30-50%. If the oxygen carrier conversion is lower than 30%, the CO_2 and H_2O concentrations in the syngas will increase. If the oxygen carrier conversion is higher than 60%, excess iron might be formed to expedite carbon deposition. It can be also concluded from Exhibit 38 that most the oxygen carriers show decreased reaction rates after they are reduced by 60%. Thus, the oxygen carrier conversion should not be too high in order to reduce the residence time and reactor volume. Exhibit 38 shows the time required for 50% reduction of the oxygen carriers in the third reduction step during the 3 redox cycle tests. The rates of reaction for all the compositions are quite fast. The maximum time required to attain 50% solid conversion is about 1.3 minutes and the rates for all the compositions are comparable to each other.

In a chemical looping unit, it is extremely essential to have effective heat transfer and oxygen carriers serve the purpose of heat transfer materials. Optimized fuel conversions, solid circulation rates and heat energy balance of the system suggest that oxygen carriers with higher inert support loading act as better heat transfer material. However, increasing the support loading means decreasing the active metal oxide content, which subsequently reduces the oxygen carrying capacity of the oxygen carrier particles. Hence there is a tradeoff. Based on the heat transfer requirement for an auto-thermal chemical looping operation, the oxygen carrying capacity of the particles and ease of particle synthesis, the 50Fe50A and 50Fe50Al compositions were selected for recyclability testing.

Exhibit 38: 50% Solid reduction v.s. time in the third reduction step during the 3 redox cycle tests

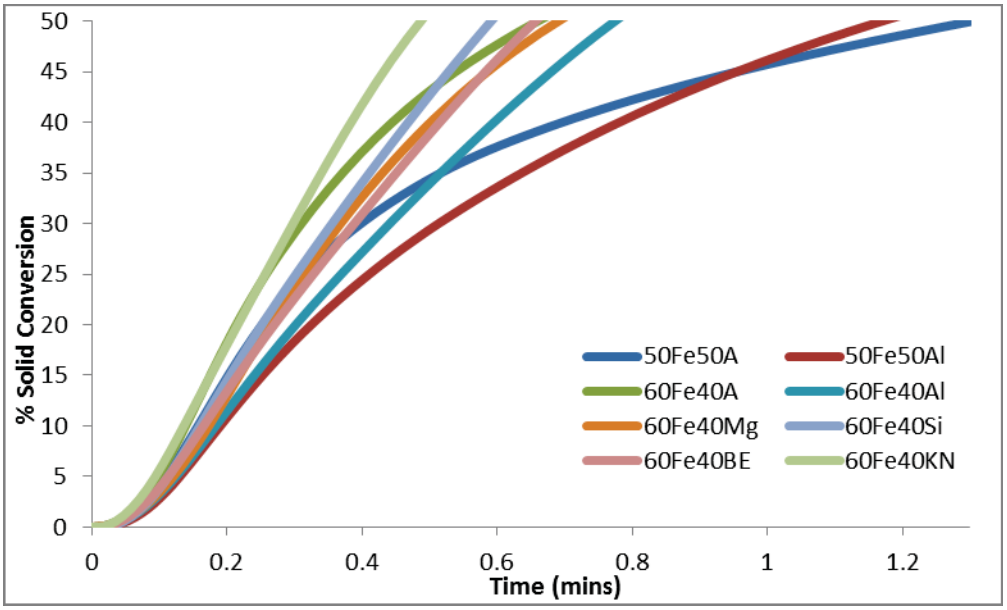


Exhibit 38 show results from the 10 redox cycle tests on 50Fe50A and 50Fe50Al. The percentage solid reduction during the reduction step of each cycle has been plotted against the cycle number. As seen from the results, both the oxygen carrier compositions perform very well as far as recyclability is concerned. The solid conversion remains more or less constant over the period of 10 cycles for both the oxygen carrier compositions. These two oxygen carrier samples were further tested for 50 redox cycles and neither showed any loss in reactivity over the period of 50 cycles. Exhibit 39 is a sample TGA curve showing 50 redox cycles on the oxygen carrier sample 50Fe50A. As seen from the exhibit, the oxygen carrier gets slightly activated over the first 15 cycles, which results in increased reactivity. After about 25 cycles, there is some deactivation observed. But it can be safely said that the change in reactivity over the 50 cycles is not very significant. After the TGA testing, the 50Fe50A and 50Fe50Al compositions were selected for subsequent syngas generation testing in the fixed bed setup. This work helped us optimize the metal-oxide composition with the identification of optimal temperature and pressure conditions for minimal or no carbon deposition.

Exhibit 39: %Solid Reduction vs cycle number for the 10 redox cycle tests

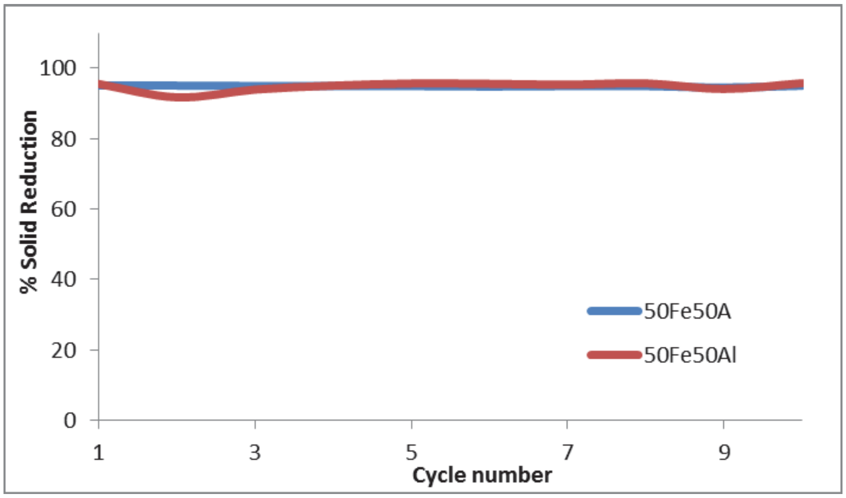
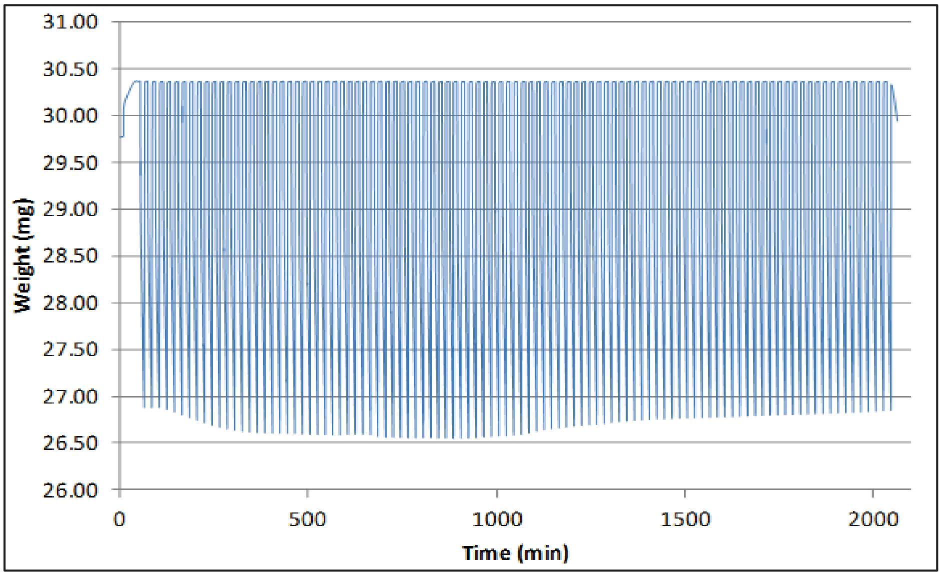


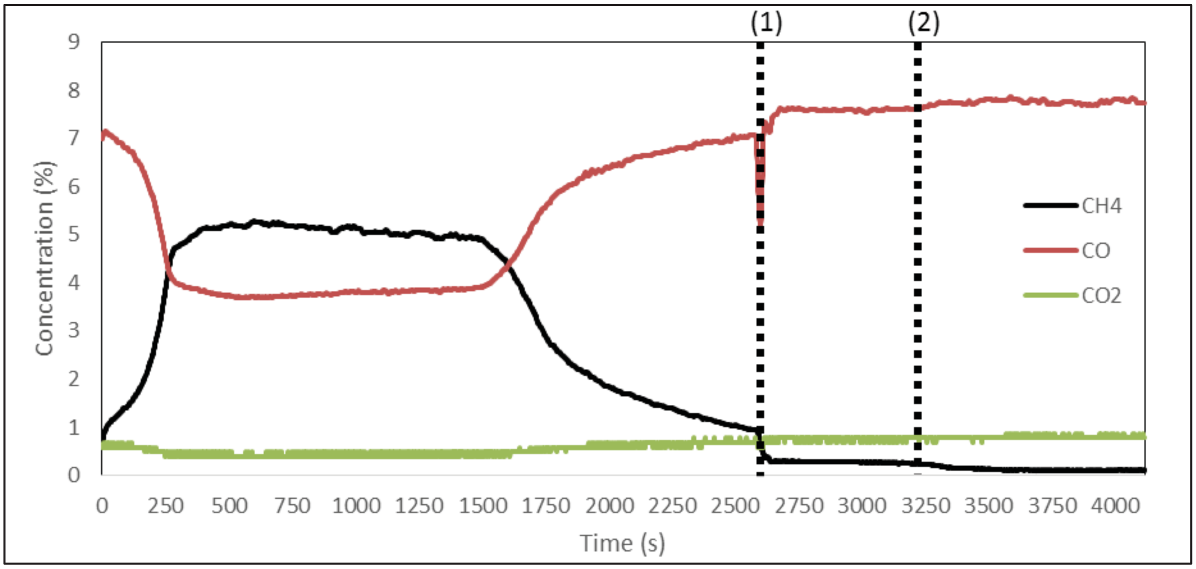
Exhibit 40: Sample TGA curve for 50 redox cycles



3. Differential-packing Fixed Bed for Volatile Conversion

The concentration of carbonaceous species in the gaseous product of the fixed test is shown in Exhibit 41. It was observed that the conversion of CH_4 decreases from about 90% to lower than 50%, and remained low for about 20 min before CH_4 concentration started to decrease significantly.

Exhibit 41: Concentration of carbonaceous species in the product of the fixed bed reactor



(1) Adjusted gas flow rate to 60 ml/min; (2) Increased temperature to 1050˚C

After the conversion of CH_4 gradually increased to a value of 90%, the flow rate of CH_4 was decreased to 60 ml/min. This was done to order to provide a longer residence time. An immediate increase in CH_4 conversion was observed. A steady production of syngas was maintained for 10 min. At the end of 10 min, the temperature of the reactor was increased to 1050°C to further improve the kinetics of the reaction. A slight improvement in conversion was achieved. It should be noted that during the period of syngas generation, the ratio between CO and CO_2 was mildly fluctuating around 10, indicating a high quality of syngas due to the thermodynamic property of the oxygen carrier material.

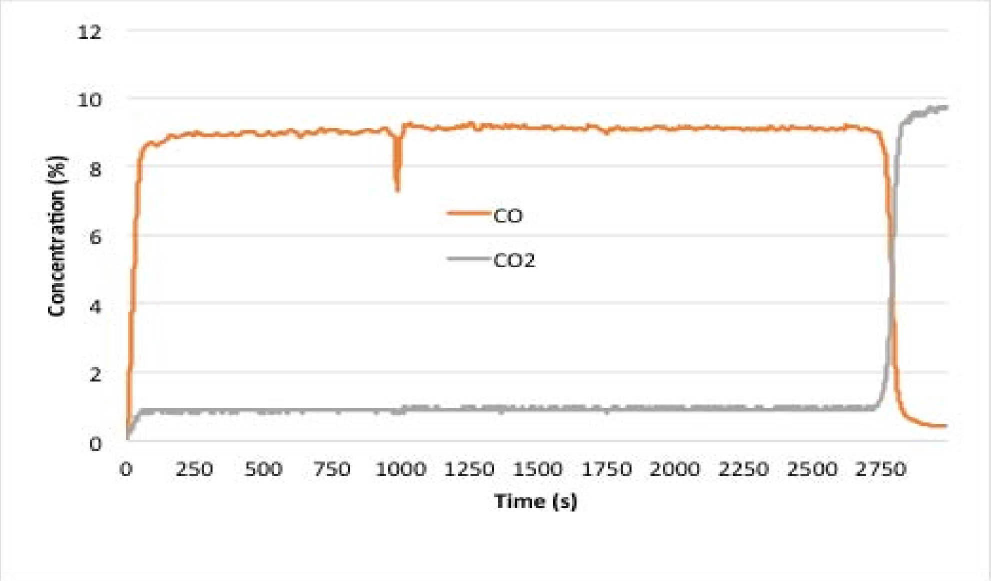
Thermodynamic analysis shows that CH_4 can be fully oxidized by iron oxide at the operating temperature of the fixed bed experiments. However, the unconverted CH_4 from the reactor and the improvement in conversion after gas residence time was shortened or after temperature was elevated point to the existence of a kinetic restriction over the reaction. The kinetic nature was further exemplified by a fixed bed experiment in which the reduced iron oxide from the previous experiment was subjected to the oxidation by CO_2.

A stream of 50% CO_2 (balanced by N_2) was introduced to the reactor at 60 ml/min, and the product was diluted by 240 ml/min N_2 before entering the gas analyzer. To examine the effect of kinetic factors, the input gas was switched to a pure CO_2 stream at the same flow rate, and diluted by 540 ml/min N_2 before entering the gas analyzer. The following reactions occurred:

$$Fe + CO_2 = FeO + CO$$
$$FeO + CO_2 = Fe_3O_4 + CO$$

The concentration of carbonaceous species in the gas product is shown in Exhibit 42. After the CO_2 concentration was increased to 100%, the concentration of gaseous product did not change. It can be inferred that, unlike the oxidation of CH_4, the change in kinetic condition had little, if any, effect on the product concentration. The rate limiting step in simulated coal volatiles conversion was determined to be the reaction between the CH_4 and the oxygen carrier particles, instead of the reaction between CO/H_2 and the particles.

Exhibit 42: Concentration of products of the fixed bed reactor-Switched input gas to pure CO_2



A notable change in the product composition occurred after a certain period of oxidation. It can be inferred that the extent of the reaction in the fixed bed was determined by thermodynamic factors rather than kinetic factors. The flip-over of the concentration of CO and

CO_2 illustrated the dependence of this reaction on solid composition: when the two solid phases at the exit are Fe and FeO, the concentration of CO is way higher than that of CO, and syngas could be generated with high quality. In cases when the two phases are FeO and Fe_3O_4, the CO in syngas will mostly be oxidized to CO_2.

4. Coal/Coal Char Conversion in Fixed Bed

a. Coal Conversion

The expected reactions in the chamber were as follow:

$C+CO_2 \rightarrow 2CO$ (char gasification)

$CH_4 + FeO_x \rightarrow CO+H_2 + FeO_y\ (x>y)$ (volatile conversion)

$CO+FeOx \longleftrightarrow CO_2+FeO_y\ (x>y)$

$H_2+FeOx \longleftrightarrow H_2O+FeO_y\ (x>y)$

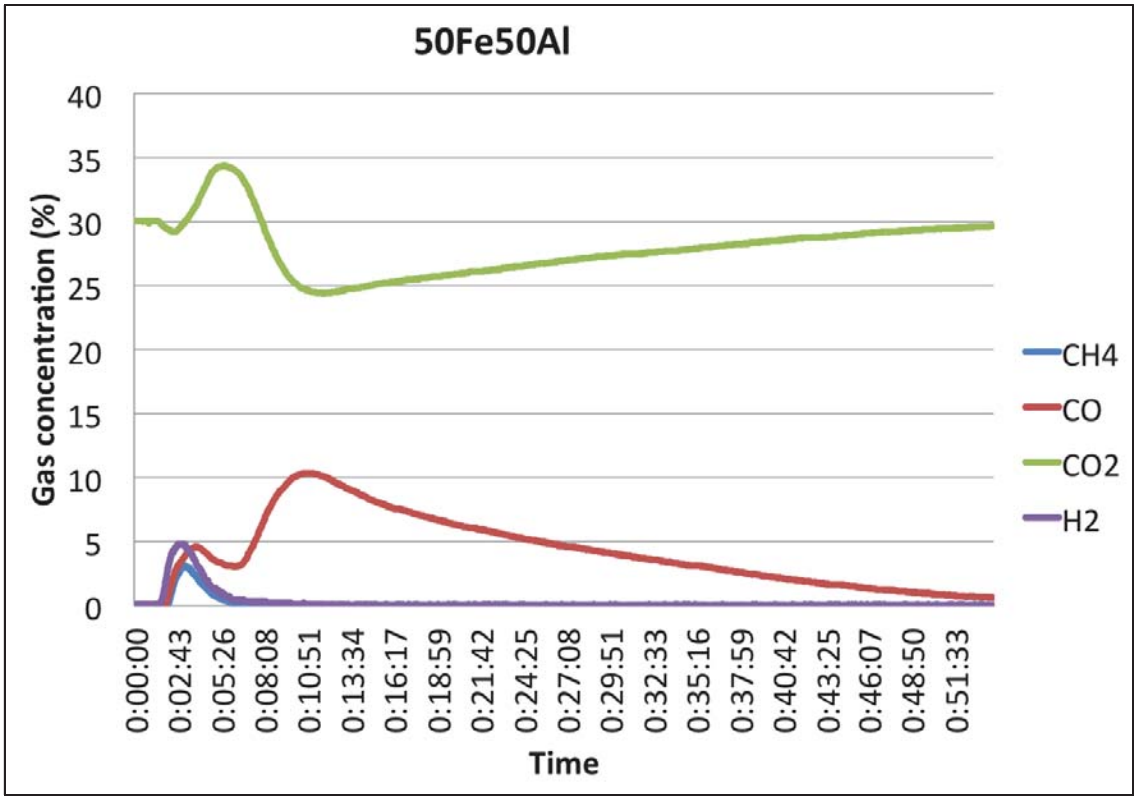
A typical gas concentration at the product gas outlet of a fixed bed test is shown in Exhibit 43. The time 0 in Exhibit 43 represents the coal and particle mixture was introduced into the fixed bed. The initial concentration shows the constant 30% CO_2 in the enhancing gas. After around 90 seconds, a set of concentration peaks including CH_4, CO and H_2 appeared along with a slight concentration drop of the CO_2. The concentration peaks are attributed to the de-volatilization of coal. The CO_2 drop is related to the concentration rising of the other carbonaceous species. The individual production rate of each species can be calculated based on the constant 0.7 L/min N_2 flow:

$$Total\ gas\ flow\ rate = \frac{0.7}{(100\% - CO\% - H_2\% - CO_2\% - CH_4\%)}$$

$$Q_i = Total\ gas\ flow\ rate\ \times\ M\%$$

where ‘i' being CO_2, CO, H_2 or CH_4. The assumption here is that the product gas stream consisted only of N_2, CO_2, CO, H_2 and CH_4.

Exhibit 43: Concentration profile of the outlet gas product of 50Fe50Al



The individual flow rates are shown in Exhibit 44 and 45. The $CO+CO_2+CH_4$ curve represents the total carbon flow rate detected at the outlet. The time when $CO+CO_2+CH_4$ curve decreases to 0.3 L/min represents the period when all carbonaceous species were released from the coal gasification in the reactor as 0.3 L/min represents the enhance gas flow rate. As shown in Exhibit 44, after 9 minutes from the beginning of the test, the CO_2 flow rate at the outlet dropped below the enhancing gas inlet (0.3 L/min), while the $CO+CO_2+CH_4$ value was still above 0.3 L/min. This indicates at that moment, the reactions occurring in the reactor were a combination of char gasification and the reverse reaction of Eq (16) between the enhancing CO_2 and composite metal oxide particles.

Exhibit 44: Individual flow rates of the outlet gas products of 50Fe50Al

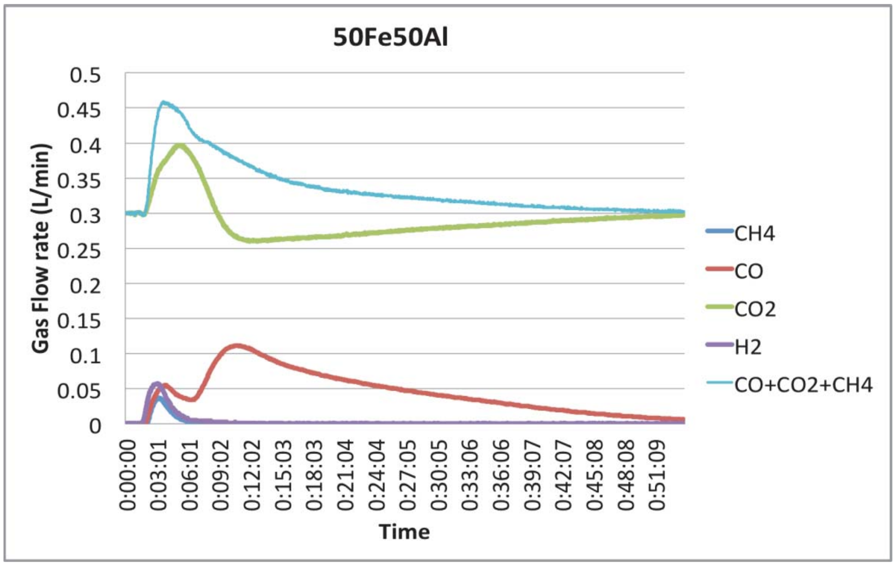
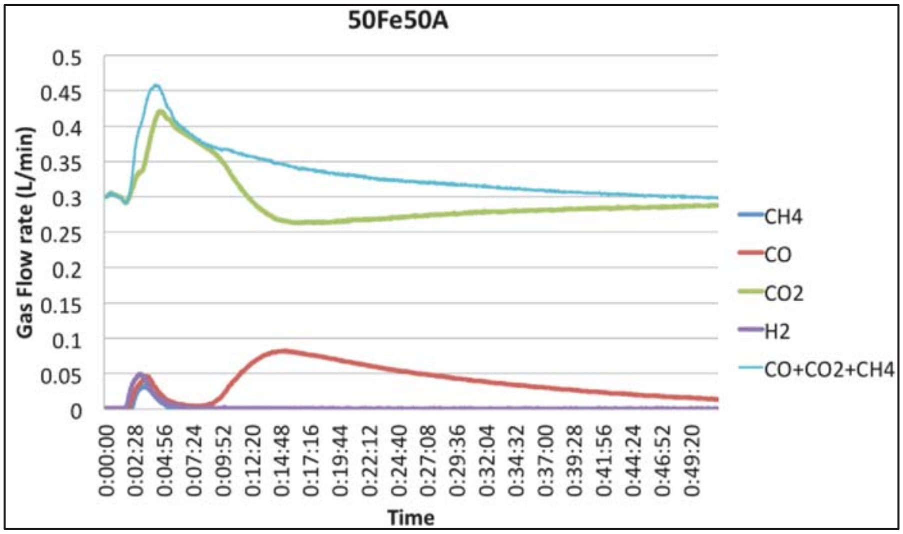
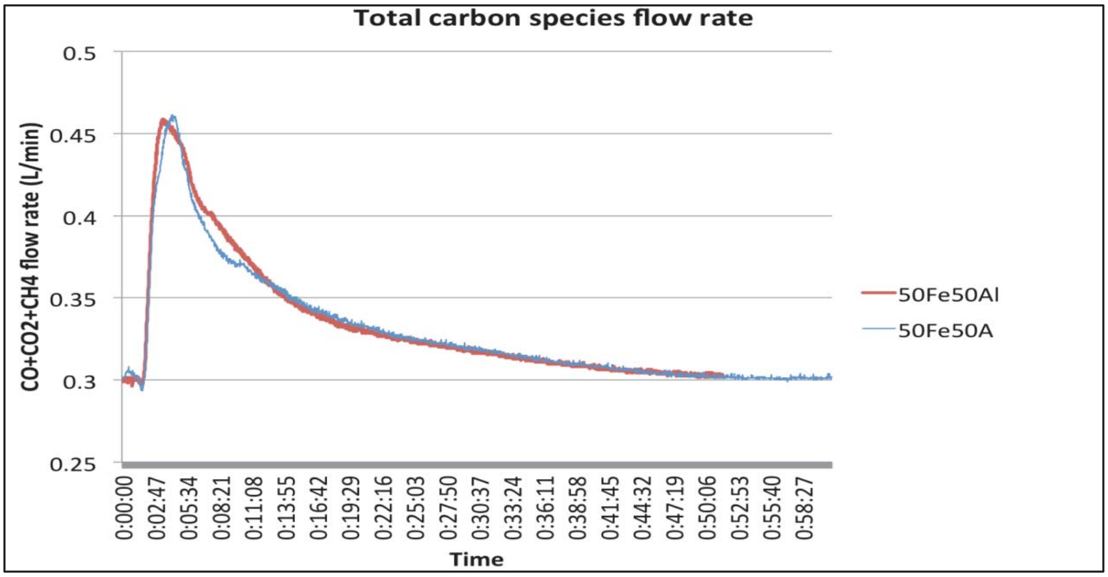


Exhibit 45: Individual flow rates of the outlet gas products of 50Fe50A



To compare the coal conversion capabilities of the various particles, the total carbon species flow rates ($CO+CO_2+CH_4$) of both compositions are shown in Exhibit 46. The result shows that the two compositions are comparable for gasifying the solid fuel.

Exhibit 46: Total flow rates of the carbonaceous gas detected at the gas outlet

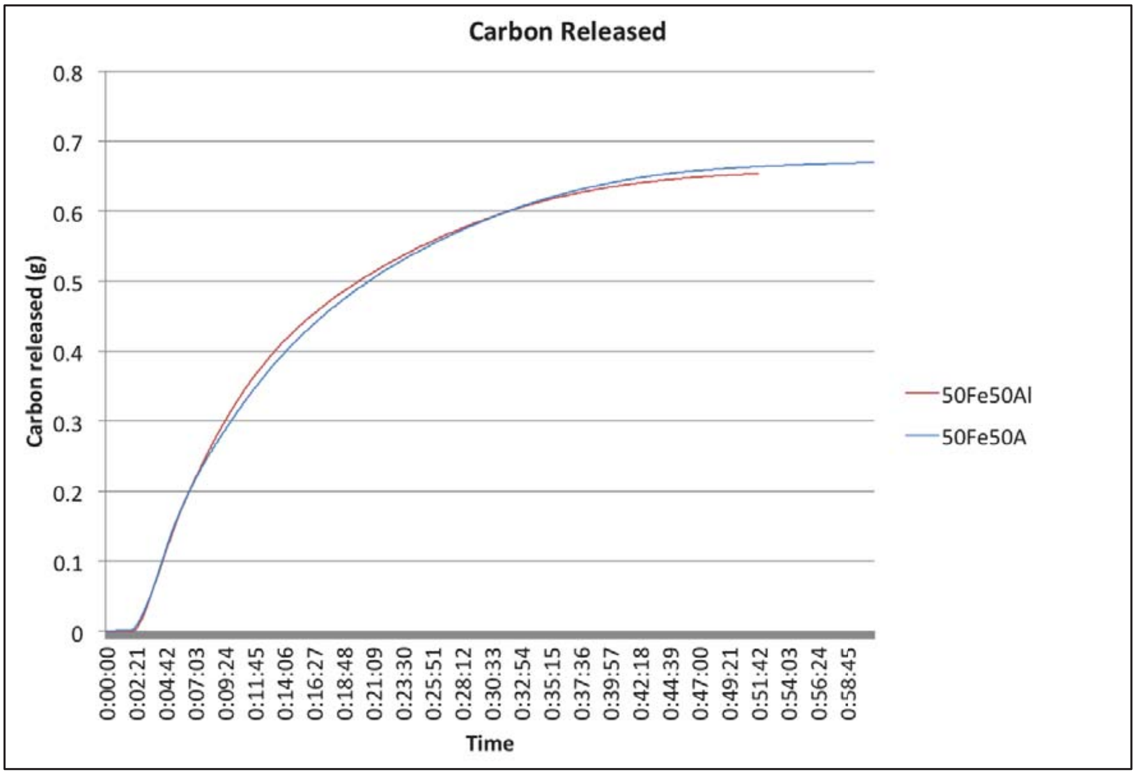


The total weight of molecular carbon generated by coal gasification can be calculated from carbon balance by integrating the total carbon species flow rates over time:

$$Total\ carbon\ weight = \frac{12\ (\frac{g}{mol})}{22.4} \int_0^{t'} (Q_{CO} + (Q_{CO2} - 0.3) + Q_{CH4})dt$$

The time dependent results of the integrations are shown in Exhibit 47. The total carbon released through the gasification was similar between the two compositions (0.65 and 0.67 grams for 50Fe50Al and 50Fe50A, respectively). In Illinois #6 coal, the fixed carbon wt% is around 63.75 wt%, which corresponds to 1.02 grams of carbon in each of the fixed bed test. The ~35% differences between the theoretical value and the integrated values might be caused by the unconverted heavier hydrocarbon species such as aromatics and long chains. In the actual reactor system, the residence time for these hydrocarbons in contact with the oxygen carrier material is longer and a full conversion to CO and H_2 is expected.

Exhibit 47: Amount of carbon released versus reaction time

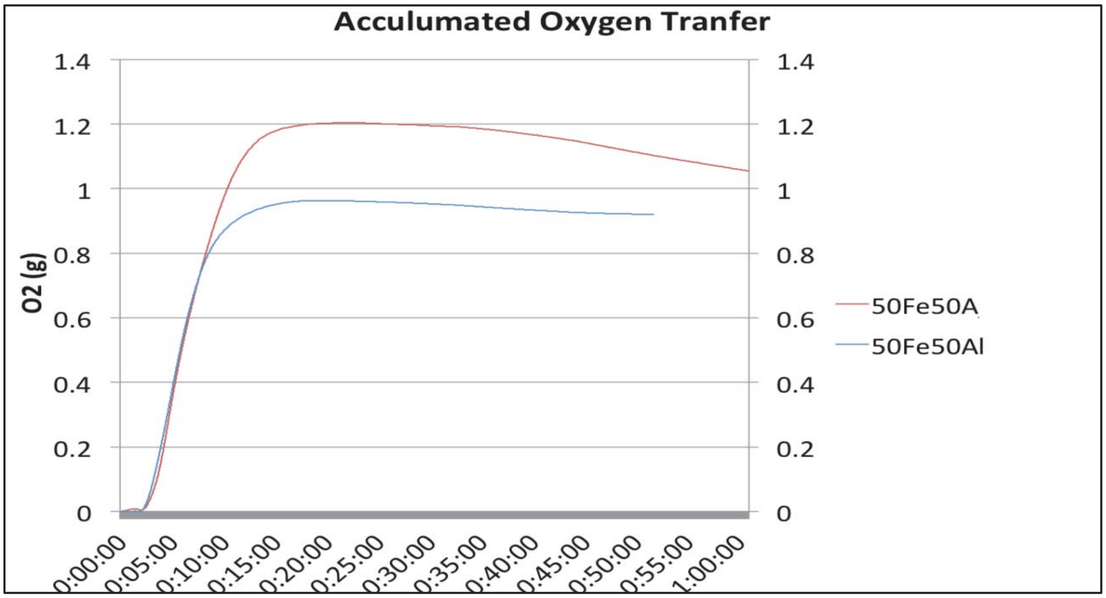


The oxygen transfer capacities were calculated by an elemental oxygen balance as shown below:

$$\text{Accumulated Transferred Oxygen} = \frac{16\ (\frac{\text{g}}{\text{mol}})}{22.4} \int_0^{t'} (Q_{CO} + 2(Q_{CO_2} - 0.3))dt$$

The results are shown in Exhibit 48. The two curves represent the amount of oxygen transferred by the particles. The decreasing trends of oxygen transferred after 17 minutes indicates that the reduced metal oxides were partially re-oxidized by the CO_2 enhancing gas. A higher utilization of the oxide oxygen content was observed in the test of 50Fe50A. This is a more desirable characteristic as it indicates faster reaction kinetics. For a moving bed reactor design, faster kinetics can reduce the particle inventory requirement; potentially improving the system efficiency and economics.

Exhibit 48: Amount of oxygen transferred versus reaction time



The carbon residue and the % solid conversion were analyzed by the carbon analyzer as shown in Exhibit 49.

Exhibit 49: Summary of the gasification parameters

Formula	**Carbon residue**	**% solid reduction**	**% solid reduction calc. by Eq (9)**	**Duration upon coal gasification finished**	**Total gaseous carbon release**
50Fe50A	<5%	34.0	35.0	60 mins	0.67 g
50Fe50Al	<5%	27.9	30.7	53 mins	0.65 g

For both compositions, the amount of carbon residue are less than 5% and do not affect the measurement of % solid reduction. The % solid reduction can also be calculated by the accumulated transferred oxygen:

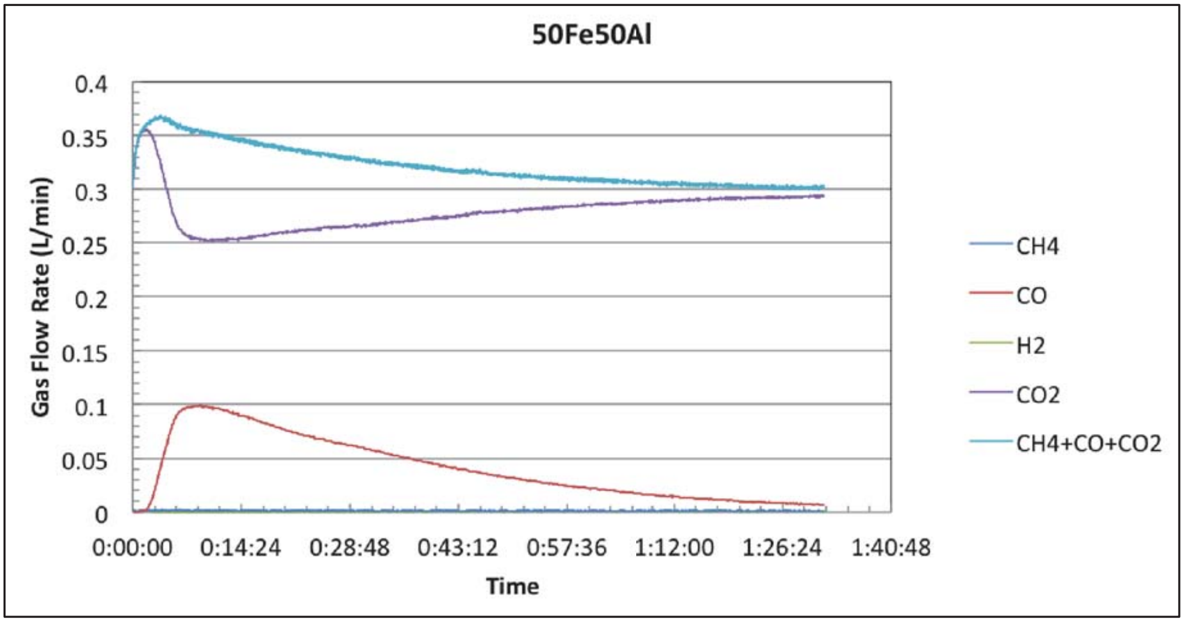
$$\%\, solid\ reduction = \frac{Accumulated\ Transferred\ Oxygen}{Total\ Active\ Oxygen\ content}$$

In this study, the total active oxygen content for both compositions is 3 grams. The two sets of % solid reduction are in good agreement to each other, confirming the mass balance and the accuracy of conclusions from this task. The two oxygen carrier compositions converted coal into gaseous species at a similar rate. However, in the case of 50Fe50A, more coal conversion occurred because of more interaction between the composite metal oxide and coal, instead of relying on additional enhancing gas to trigger the gasification reaction. Based on the result of the fixed bed study, 50Fe50A is a more favorable composition for chemical looping gasification.

b. Char Conversion

The estimated gas flow rates at the gas outlet are shown in Exhibit 50. The origin of the x-axis represents the start of enhancing gas injection (0.7 L/min N_2, 0.3 L/min). Compared to the coal conversion tests performed in Q3, the major difference was that no CH_4 or H_2 was observed. The absence of these two species indicated that the residue volatile content in the char samples were minimal.

Exhibit 50: Gas flow rates at the outlet of the fix bed consisted of 50Fe50Al particle and char



The same mass balance calculations for carbon and oxygen as stated in Q3 were again performed in this char conversion study. The results are summarized in Exhibit 51.

Exhibit 51:Summary of particles performance

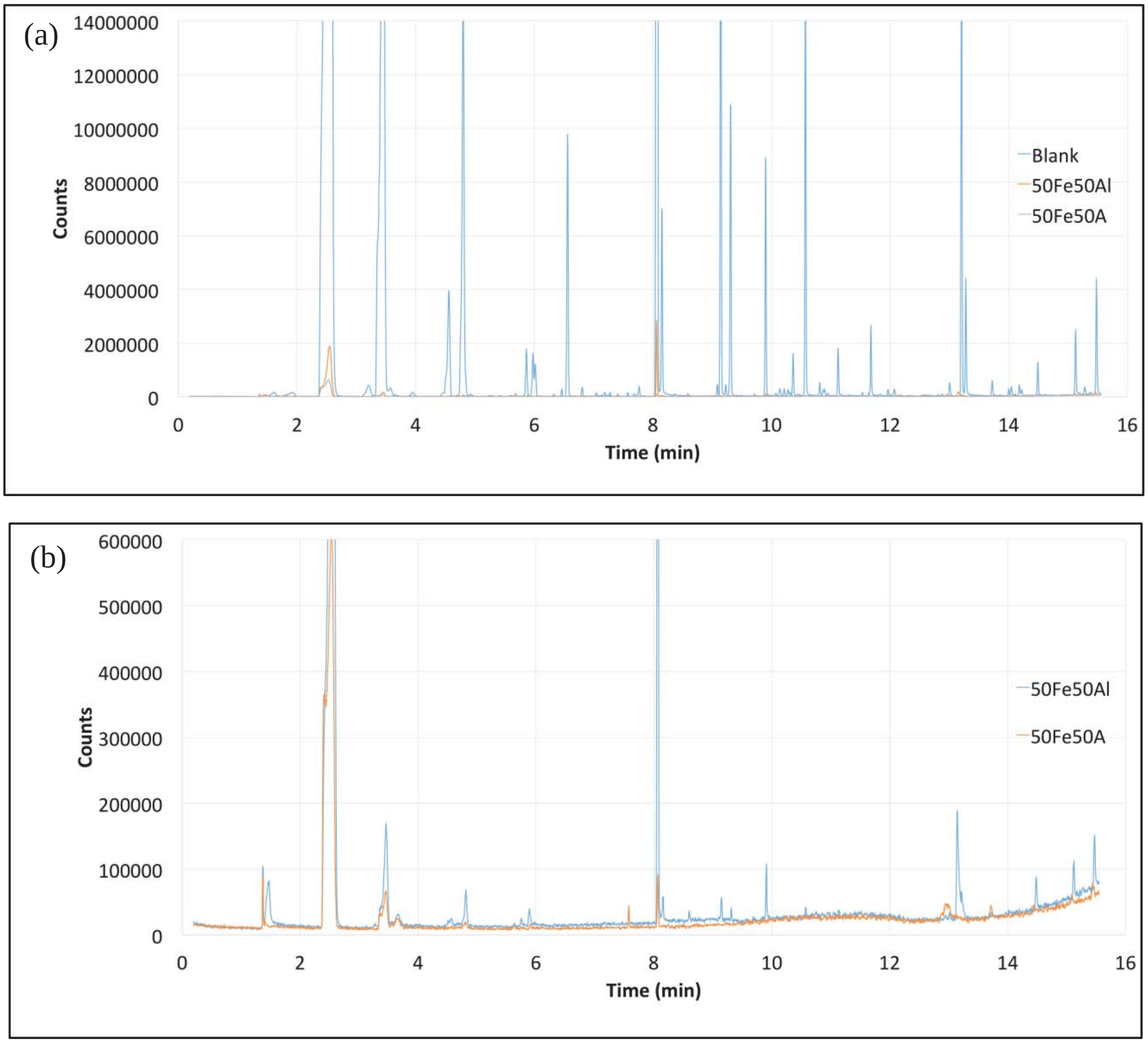
Particle formula	Carbon released	%Carbon conversion	Maximum oxygen transferred	Maximum % solid reduction
50Fe50Al	1.00 g	72.5%	0.552 g	18.4%
50Fe50A	1.06 g	76.8%	0.368 g	12.7%

The results are in good agreement to the conclusion reached in Q3. For the two formulas, while the gasification capabilities of the carbon species in char are comparable, 50Fe50A has a superior oxygen transfer capability, which is a desirable property for oxygen carrier.

5. Volatile Conversion in GC-MS

The GC-MS spectra of the three different fixed bed materials are shown in Exhibit 52 (a) and (b). The x-axis represents the time of detection, while the y-axis represents the detection counts/intensity. As each peak being one specific species, the area under each peak is proportionally related to the amount of the species. The lighter species are usually detected earlier and vice versa. As shown in Exhibit 52 (a), the reactor with no particle (blank) retained significantly more volatile than the other two cases with particles. Most of species observed are aromatics, which are relatively difficult to crack. The species range from single aromatics (benzene, toluene), double aromatics (naphthalene, bi-phenol), to the heavier poly-aromatics (fluoranthene, pyrene) . Compared to the blank case, the particles cracked considerable amounts of volatiles in the latter two cases. Exhibit 52 (b) is a close-up plot for the two particle cases. The intensity of the two cases cannot be directly compared because the amount of coal used might not be exactly the same. However, the performances of 50Fe50Al and 50Fe50A can still be qualitatively compared. 50Fe50Al has more significant peaks towards the right of the plot, especially between the 8-10 minute window, whereas the peaks on the left are comparable to the ones from 50Fe50A. This indicates that in the 50Fe50A case, there might be more double aromatics being cracked into single aromatics. This again 50Fe50A might be a better candidate for the oxygen carrier in the chemical looping gasification process.

Exhibit 52: (a), (b) GC-MS spectra of cracked volatiles



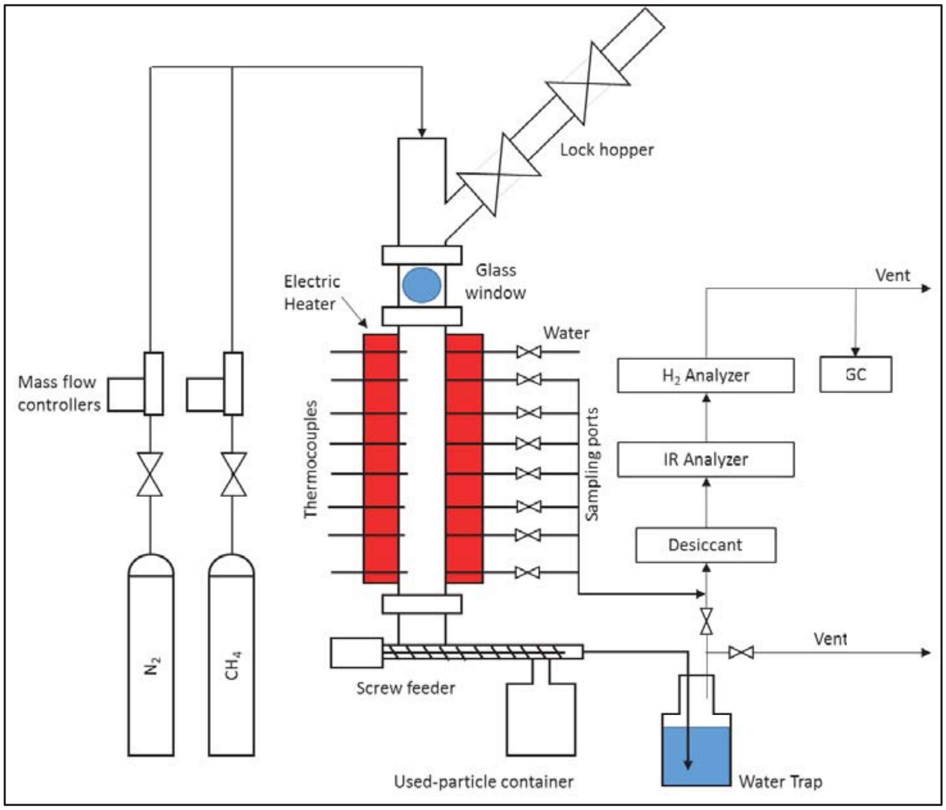
B. Laboratory Bench-scale prototype reducer results

1. Bench Scale System Modification

The bench scale moving bed reducer was modified for effective water/steam injection. Originally, steam was injected from the top injection port along with N_2 and/or CH_4. However, it was found that the steam was cools and was condensed by the low temperature solid inventory on top of the heated reactor section. As a result, a significant delay was observed between the steam injection and the corresponding response observed in the gas composition. To reduce the delay, steam was injected from the topmost gas sampling port. The steam generator was removed as water can be vaporized easily in the new injection port, which is in the heated section.

The oxygen carrier particles and the coal ash are separated based on the significant size difference. This was confirmed by the observation that during the coal conversion tests, ash was carried out from the reactor by the effluent gas. A water trap was added at the gas outlet to remove the ash. The modified bench scale moving bed system is shown in Exhibit 53.

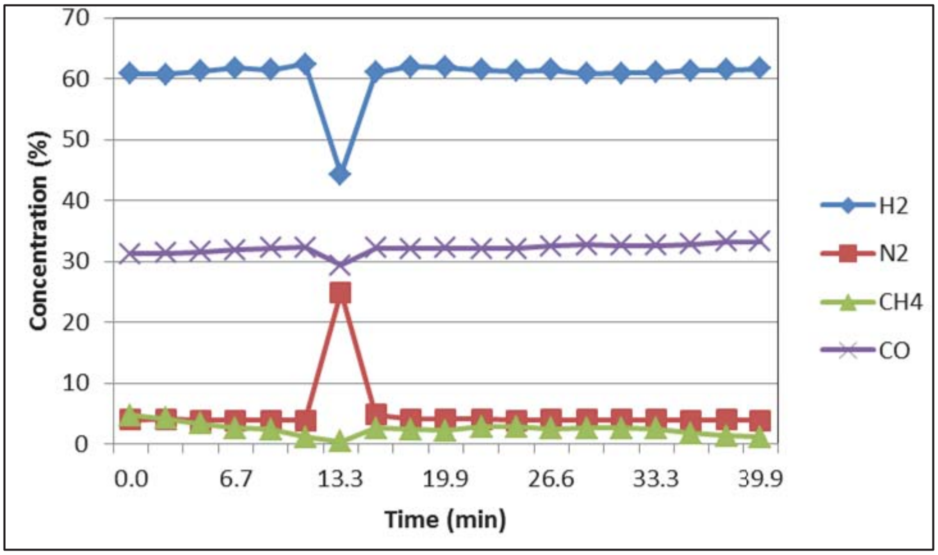
Exhibit 53: Modified bench scale moving bed system



2. Moving Bed Reducer Testing: Coal volatile conversion in moving bed reducer

It is shown that at steady state, the moving bed reactor could effectively convert coal volatile (CH_4) into high quality syngas. The gas composition (dry basis) at the reactor outlet is shown in Exhibit 54. When the temperature was 1040℃, the syngas produced in the reactor consisted of approximately 60% H_2 and 30% CO, and about 3% of unconverted CH_4. The CH_4 conversion was estimated to be around 95%, and syngas purity, defined by the total percentage of H_2 and CO in effluent gas (wet base, N_2 free), was higher than 85%. At temperature of 1050℃, no CH_4 was detected in the product stream, and a full conversion of feedstock was achieved.

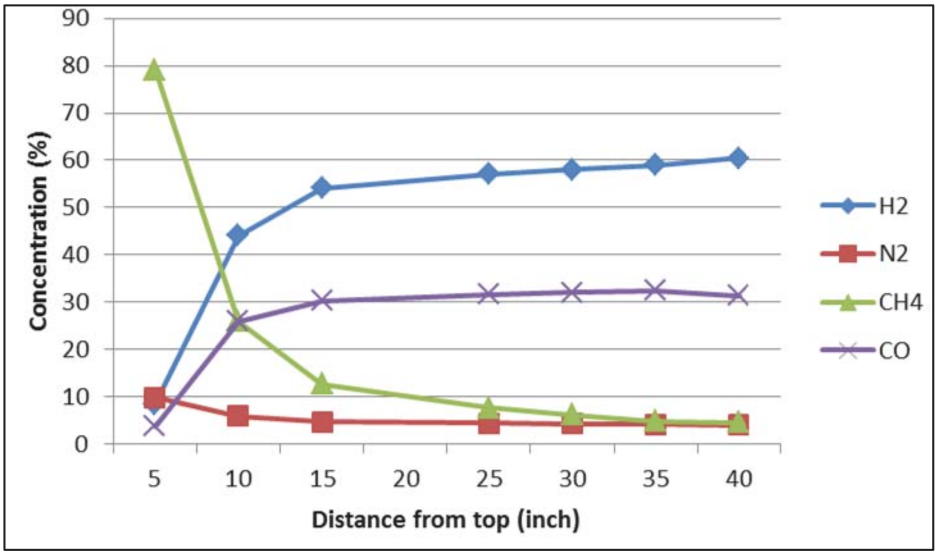
Exhibit 54: Reactor outlet gas composition at 1040°C. The interruption at 13.3min was due to gas analysis system adjustment



The analysis on reduced particles showed no carbon deposition detected on the particles, which was expected as the moving bed reactor was employed and the stoichiometry ratio between CH_4 and Fe_2O_3 was controlled.

The gas composition variation along the moving bed reactor was studied by taking samples from the gas sampling ports. The composition profile at peak temperature of 1020°C is shown in Exhibit 55. It is expected that the volatile conversion would not be significantly increased by a longer reactor, corresponding to a longer residence time, as the gas composition profile became “flat” near the lower portion of the moving bed reactor.

Exhibit 55: Gas composition at different location of the reactor at 1020℃

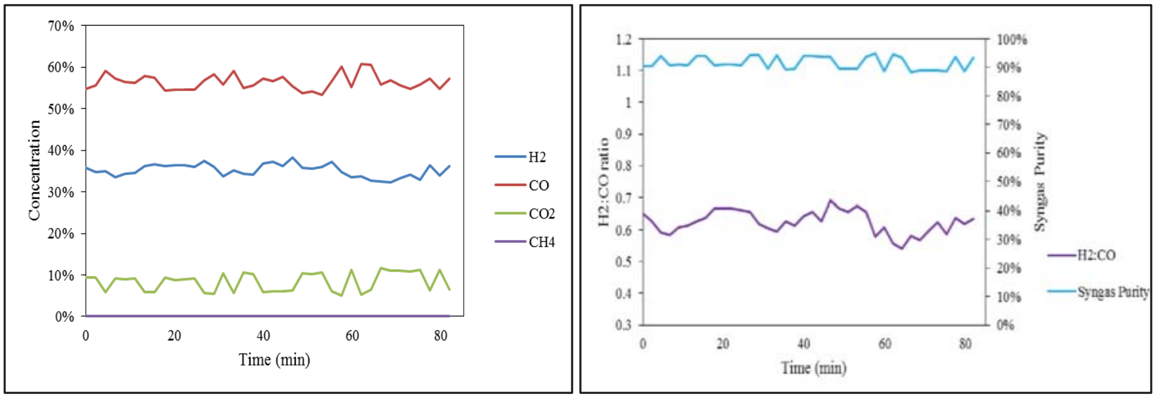


At higher temperature, this profile was contracted towards the inlet of the reactor, and the gas and solid composition reached the final value sooner. Thereby, higher volatile conversion and/or shorter residence time could be achieved.

3. Moving Bed Reducer Testing: Coal Conversion in moving bed reducer

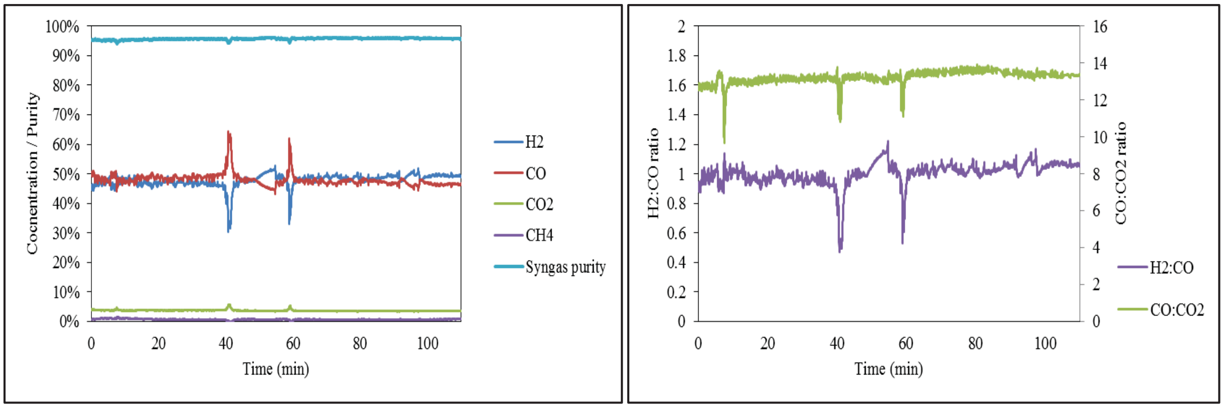
The syngas composition (gas chromatography data) from the first set of experiments (PRB coal only) is shown in Exhibit 56. After a start-up and transition period, the PRB coal was steadily converted into syngas. The H_2:CO ratio was fluctuating around 0.65, which was consistent with the elementary composition of the coal. The CO:CO_2 ratio was around 7, and the syngas purity (dry base) was well above 88%. Minimal CH_4 was observed at the gas outlet, indicating a near complete conversion of coal volatile. Carbon conversion in coal was estimated to be about 93%, corresponding to a 90% char conversion given a full conversion of coal volatiles.

Exhibit 56. Syngas composition (N_2 and H_2O free) in syngas generation with PRB coal only



In the second set of experiments, CH_4 was co-injected with coal to increase the H_2 content in syngas. The amount of CH_4, calculated with thermodynamics, was chosen so that the H_2: CO ratio of the product syngas was 1:1. It can also be determined proximately by letting the ratio of C: H in all fuels to be 1:2. The syngas composition (gas analyzer data) from the second set of experiments (PRB coal) is shown in Exhibit 57. The concentrations of H_2 and CO were almost identical, and the H_2: CO ratio was successfully controlled to approximately 1. The CO: CO_2 ratio was about 13, which was very close to the thermodynamic limit. A syngas purity of about 95% (dry base) with CH_4 less than 1% was achieved in the operation. The spikes in the gas composition were caused by air/oxygen seepage into the reactor when oxygen carrier particles and coal were being filled during the operation. Molecular oxygen in the reactor oxidized the char rapidly and resulted in a rapid generation of CO and CO_2. The result confirms that the chemical looping process is capable of controlling the composition of syngas precisely and steadily.

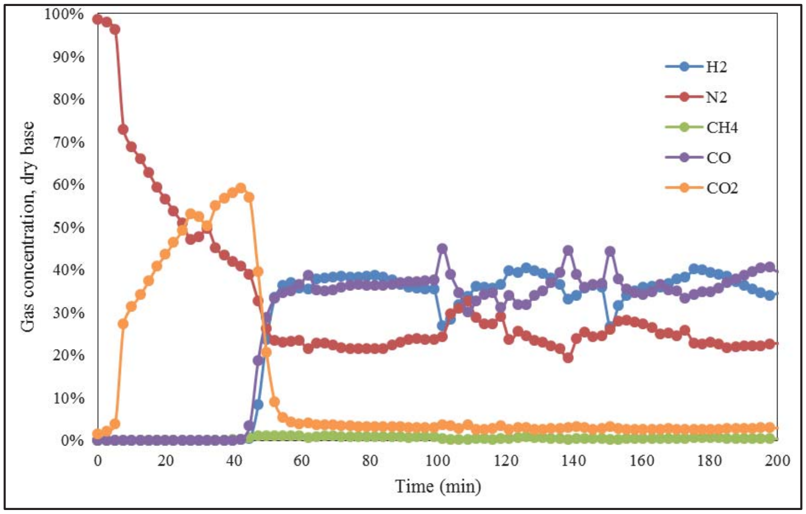
Exhibit 57: Syngas composition (N_2 and H_2O free) in syngas generation with PRB coal and CH_4 co-injection



Similar experiments were also conducted with a bituminous coal (Illinois #6). The mass ratio between oxygen carrier particles and coal was 9:1. CH_4 was sent into the reactor from the top of the reactor at the flow rate of 0.75 standard liters per minute (SLPM). The ratio between coal and CH_4 was chosen to produce a syngas stream with H_2:CO ratio of 1. During the entire period of the experiment, a N_2 flow of 1.5 SLPM was maintained.

The composition of product gas (gas chromatography data, dry base) from the reactor outlet is shown in Exhibit 58.

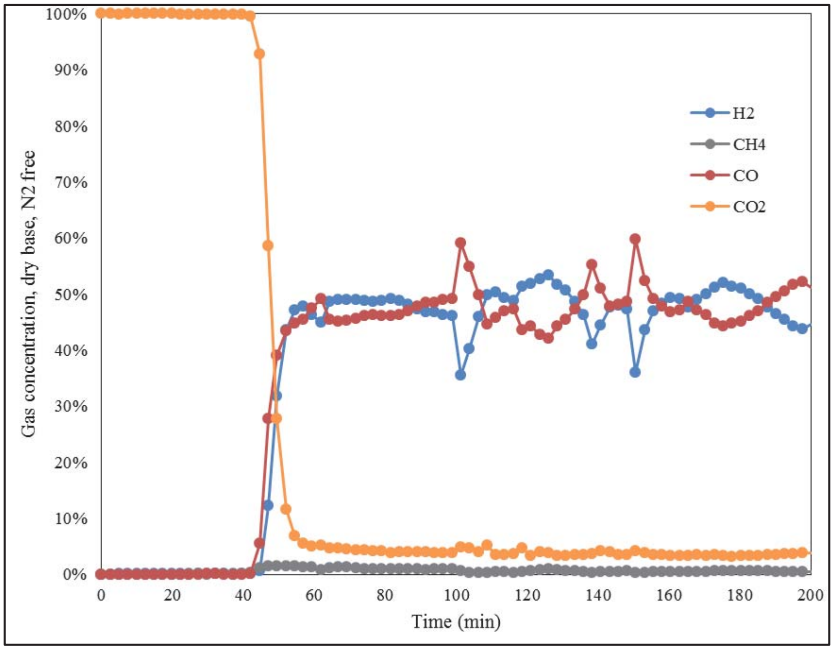
Exhibit 58: Syngas composition (dry base) in syngas generation test with Illinois #6 coal and CH_4 co-injection



In the first 40 minutes of the test, the only product generated from the reactor was CO_2. The concentration of CO_2 was increasing, showing an increasing carbon conversion rate. As the coal/char moved downwards with the oxygen carrier particles, the carbon holdup in the reactor was increasing, resulting in an increasing carbon conversion rate.

After the 40-minute start-up period in which all fuel was converted to CO_2, a sharp change in gas composition occurred. The concentration of CO_2 sharply dropped to about 3%. At the same time, the concentration of H_2 and CO went up to about 37%.

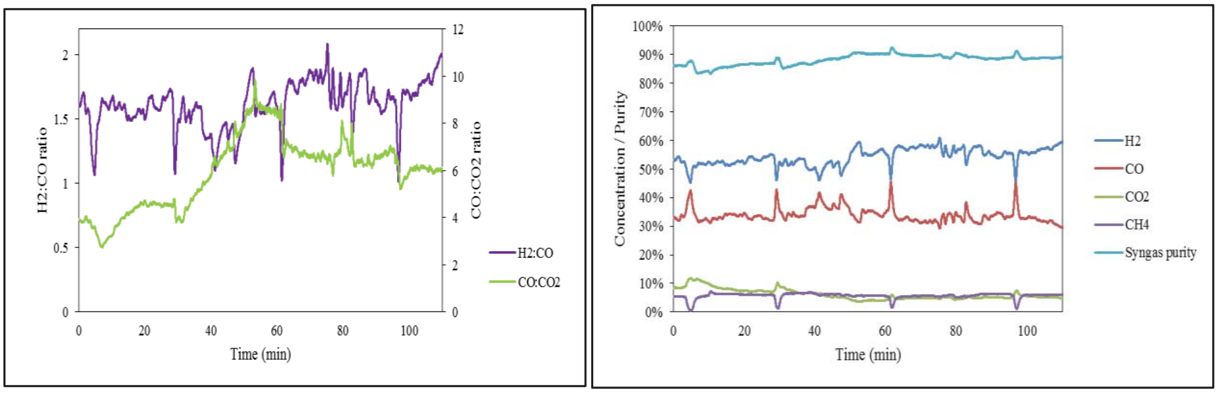
Exhibit 59: Syngas composition (dry base, N_2 free) in syngas generation test with Illinois #6 coal and CH_4 co-injection



The syngas composition without N_2 is shown in Exhibit 59. As mentioned in the previous reports, the spikes in the gas composition were a result of air seepage into the reactor during particle makeup. It was observed that syngas with a H_2: CO ratio of about 1 was produced. The CO_2 concentration was about 4% in total dry syngas (N_2 free), and the CO: CO_2 ratio was greater than 13. The experiments confirmed that the co-current moving bed reducer system combined with the iron-based oxygen carrier particle is capable of converting different types of coal into high purity syngas with adjustable composition. It should be noted that bituminous coal is usually difficult to process in combustion or gasification systems due to its high tendency of caking or agglomeration. However, the use of bituminous coal in the bench scale unit tests has not cause any flow issue in the moving bed reactor. Future experimental effort for chemical looping gasification with moving bed reducer will focus on the scale-up testing of the system.

In the third set of experiments, CH_4 and H_2O were co-injected into the reactor. The ratio of the species was determined by process simulations in Task 5. Thermodynamic calculation showed that at the reactor outlet, the oxygen carrier is reduced to FeO, and the syngas has a H_2:CO ratio of about 1.8, and M value of about 1.5. The syngas composition (gas analyzer data) from the bench scale reactor test is shown in Exhibit 60. The H_2: CO ratio of the syngas was about 1.7, and the M value was 1.4. The test result was close to the thermodynamic calculation except that about 5% of CH_4 appeared at the reactor outlet.

Exhibit 60: Syngas composition (N_2 and H_2O free) in syngas generation with PRB coal, CH_4, and H_2O co-injection



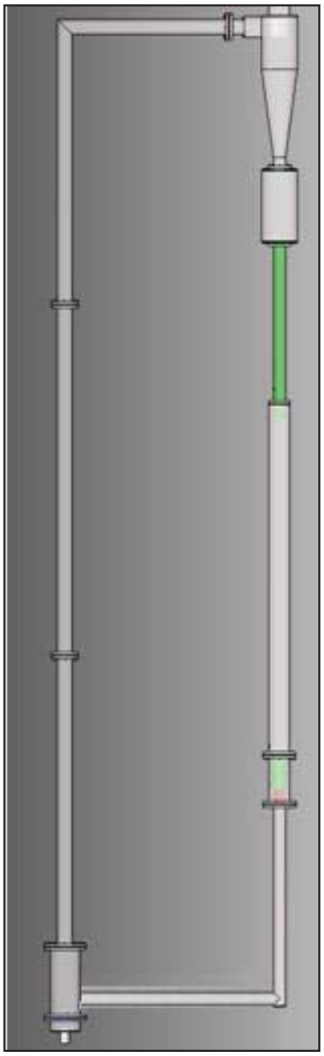
The deficit is explained by considering the kinetic difference between FeO/Fe and Fe_3O_4/FeO. As discussed in the previous reports, the reaction between CH_4 and Fe_3O_4/FeO is slower than that between CH_4 and FeO/Fe. In the third set of experiments, the oxygen carrier was only reduced to FeO, and CH_4 was reacting with the oxygen carrier particles with the slowest kinetics. Therefore, less H_2 was produced in the reactor rendering lower H_2:CO ratio and M value. This problem can be easily overcome with a pressurized reactor system with better kinetics and/or a longer reactor than the bench-scale unit.

C. Cold Model Operation Results

L-valve is used in the system to non-mechanically transport oxygen carrier particles from the fuel rector to the air reactor with controllable solids flow rate. The solids flow control device is an important component of particulate processes. Mechanical devices such as butterfly valves, rotary disc valves and screw feeder generally control the solids flow through the use of moving components and are usually used under low temperature/pressure conditions, where the risks of sealing and mechanical problems are minimal. Non-mechanical devices are normally the choice for applications under more stringent conditions, as they control the solids flow rate through the use of external aeration gas. The non-mechanical devices contain no moving parts and thus, are of reduced risks in wear and seizure, and low costs in operation, maintenance and replacement. Commonly used non-mechanical devices include seal pots, loop seals, J-valves, L-valves and V-valves. Among these devices, the L-valve is the most commonly used in both the dense-phase and dilute-phase flow conditions as in pneumatic transport lines, bubbling fluidized beds and circulating fluidized beds. A gas is introduced from the vertical section of L-valve to drive the solids flowing from reducer to the combustor. The solids flow rate is controlled by the gas flow rate introduced.

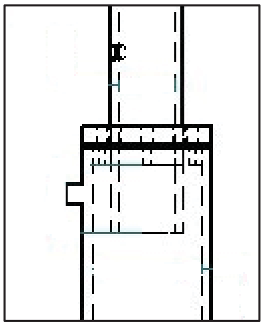
Other components are designed according to the parameters of the reducer and the combustor. Cyclone is used to separate regenerated oxygen carrier particles from oxygen-depleted air. The gas-solids inlet of the cyclone is connected to the top of the riser. Its solids outlet is connected to the top of a solids receiver. The solids receiver stores the regenerated particles from the cyclone and ensures a constant solids level in the reducer. Standpipes are used between solids receiver and the top of the reducer, and between the bottom of the reducer and the L-valve to ensure gas sealing between the two reactors. By injecting enough gas at the middle of the standpipe and keeping the pressure at the gas injection point higher than the two ends of the standpipe, the gas leakage from either end is prevented. The overall system configuration of the cold flow model is shown in Exhibit 61.

Exhibit 61: Cold Flow Model Assembly for CLG process



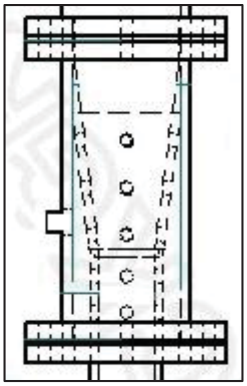
The Fuel Reactor, which is a 3-inch cylindrical column with a height of 60 inches, has a gas inlet near the top of the side wall and a dipleg from the top for solids inlet. The location of the gas inlet is higher than the bottom tip of the solids inlet so that the gas inlet is above the solids level in the fuel reactor and thus local fluidization of the solids particles can be avoided. A detailed drawing of the top section of the fuel reactor is shown in Exhibit 62.

Exhibit 62: Detailed drawing of top section of fuel reactor



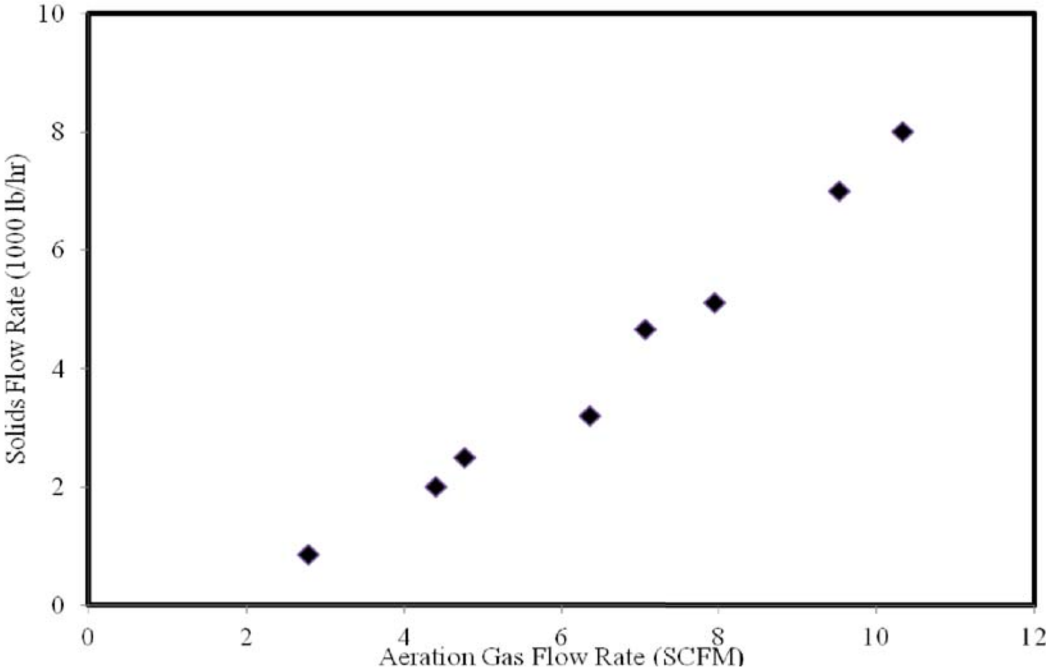
The bottom section of the fuel reactor is connected to a product gas separator, which is used to separate gas from the solids in the co-current moving bed without fluidizing or entraining solids particles. As seen in Exhibit 63, the product gas separator has a cone-shape inner wall and a cylindrical outer wall. The inner cone is used to connect the 3” fuel reactor and 1.5” zone seal so that the solids particles can smoothly move from the fuel reactor to the downstream. Multiple layers of small holes with porous plugs filled in them were drilled on the cone so that the product gas can be separated from moving bed solids particles without disturbing solids flow. A gas outlet on the cylindrical outside wall allows the product gas to be transport to downstream for further process.

Exhibit 63: Product gas separator



The relationship between the aeration gas flow rate and the solids flow rate through the L-valve for the oxygen carrier particles which is in the range of Geldart group D under ambient conditions was obtained, as shown in Exhibit 64. When aeration gas is added to the L-valve, solid particles do not begin to flow until a minimum threshold amount of gas flow rate is added. When the aeration gas flow rate is low, the produced frictional force is not enough to initiate solids flow. Above the minimum threshold amount of gas flow, increasing the aeration gas flow rate causes the solids flow rate through the L-valve to increase. By controlling the aeration gas flow rate introduced to the L-valve, the solids flow rate of the chemical looping system can be controlled. Understanding the hydrodynamic behavior of the L-valve operation was an important milestone completed as a part of this project.

Exhibit 64: Relationship between aeration gas flow rate and solids circulation rate



The pressure differential gauges were installed to measure the pressure drops of the riser, the combustor, the fuel reactor, the L-valve, and the two zone seals. Pressure gauges were installed to measure the pressures at the two outlets. The fluidization of the combustor under different gas velocities was observed. It was found that the combustor starts to fluidize when the gas velocity in the combustor reaches to the minimum fluidization velocity, then changes to slugging bed when the gas velocity is increased. The combustor operates at turbulent fluidized bed condition when the gas velocity reaches to about 3 times of the minimum fluidization velocity of the solids particles.

The riser operates under dilute phase gas-solids flow condition when aeration gas is introduced to the L-valve to start solids circulation of the system. Solids are entrained from the surface of fluidized bed of the combustor and carried to the inlet of the cyclone with some back mixing near the wall of the riser.

With the increase of the solids flow rate by increasing the aeration gas of the L-valve, the operation of the riser and the combustor is pretty smooth with no noticeable pressure drop change for the combustor and increasing pressure drop through the riser, which are expected.

Different gas flow rates were introduced to the gas inlet of the fuel reactor. The pressure drop through the fuel reactor increased with increasing gas flow rate. The operation of the fuel reactor was smooth, keeping under moving bed condition with different gas flow rates introduced.

Performance of the zone seal was tested during the system operation. Under different operational conditions, zone seal gases were introduced to the two zone seals, trying to prevent gas mixing from different reactors. Pressure drops were measured for different sections of the zone seals. With enough gas flow through zone seal gas injection point, a relatively higher pressure point than those at the both end of the zone seal can be generated, which indicates the zone seal gas splits and flows in both upwards and downwards directions, thus gas sealing reaches.

Pressures and pressure drops of the reactors and parts of the system were recorded under all kinds of operational conditions. A typical pressure drops in the reactors and key components during the operation are listed in Table 3.

Exhibit 65: Pressure drops in the reactors/key components of the CLG cold flow model.

Components	**Pressure Drop, in.H_2O**
Combustor	14
Riser	3
L-valve	4
Reducer	23

D. Techno-Economic Analysis Results

The detailed techno-economic analysis results are presented in a separate attachment at the end of the document.

III. Conclusions

Elemental screenings based on thermodynamic calculations suggested that various iron-based metal composites have potential of generating syngas using the chemical looping concept. The results showed that Al, Cr, Mn, Mg, V, and Zn were the possible options as the second metallic element for oxygen carrier materials. With Fe_2O_3 being the active component in all the formulation tested, the reactivities are all comparable to each other, according to the TGA tests. Based on the heat transfer requirement for an auto-thermal chemical looping operation, the oxygen carrying capacity of the particles and ease of particle synthesis, 50Fe50A and 50Fe50Al are selected for further testing. The result of fix bed tests showed that both 50Fe50A and 50Fe50Al are both good candidates for Chemical Looping Coal Gasification, while 50Fe50A might be of slight advantage because the higher solid conversion potentially reduces the required amount of particle. Bench-Scale studies using the optimized particle composition were completed using a co-current downward moving bed reactor. A variety of syngas compositions were obtained in-line with the thermodynamic predictions for coal-volatile model compounds, coal-char, coal and steam injection. Importantly, a high degree of controllability on the syngas composition was demonstrated for the co-current moving bed reducer reactor using the oxygen carrier particle developed in the screening study. An integrated cold-model system was designed using the data from bench-scale experiments using non-mechanical valves and hot to cold scaling factors. A comprehensive pressure balance study using non-mechanical valves and zone-seals for specific residence times and thermodynamic contact mode was completed. The cold-model study coupled with the particle development study and the bench-scale study prove the technological feasibility of the technology and direct towards a larger scale integrated hot-unit demonstration.

The current TEA investigated incorporating the OSU chemical looping gasification technology into coal fired IGCC power generation and coal based methanol production facilities at a conceptual level. One of the primary goals was to develop approaches the OSU CLG technology to produce high hydrogen content syngas required for IGCC power generation with pre-combustion CO_2 capture and control over the syngas composition for methanol production. While sufficient control over the syngas composition was achieved for syngas production, a syngas with a sufficiently high hydrogen content to reach 90% CO_2 capture was not. To achieve this level of hydrogen, reducing the iron in the oxygen carrier to the metallic state without the formation of elemental carbon is required. In the process models developed, conditions to achieve this objective could not be identified.

Two methanol production options incorporating the OSU CLG technology were developed for this, one with 100% coal as the feedstock to the process (OSU-1) and the other with 50% coal and 50% natural gas as the feedstock (OSU-2). A summary of the performance, capital and operating costs, and the economics of coal to methanol process developed under this project are provided in Table 15. In this exhibit the results are compared to the results from the draft version of the DOE/NETL *Baseline Analysis of Crude Methanol Production from Coal and Natural Gas*. From the DOE/NETL study, case MBL-1 is based on a coal feed and gasifier without CO_2 capture, case MBL-2 is based on a coal feed and gasifier with CO_2 capture, and case MBL-3 is based on natural gas feed and a reformer with CO_2 capture

The OSU CLG technology provides several savings compared to the state-of-the-art coal base methanol production related to lower capital costs and higher efficiencies. In comparing the MBL-2 and OSU-1 options, which are both based on a coal only feed and have CO_2 capture, the following benefits are realized:

- A lower methanol RSP by \$0.37/gal, a 21% decrease.
- Lower capital costs by 28%
- Higher efficiency based 14% in coal consumption
- A methanol RSP lower than the reference non-capture case, which results in CO_2 capture cost less than 0.

Under the reference economic conditions, the methanol RSP of the OSU-CLG based systems were not found to economically favorable compared to the natural gas fed MBL-3 option. The low capital costs of the reformer based facility were a significant economic advantage all of the coal fired cases that could not be overcome for the reference natural gas cost of \$6.13/MMBtu.

Exhibit 66: Performance and Cost results for Methanol Production Options

	Case				
	MBL-1	**MBL-2**	**MBL-3**	**OSU-1**	**OSU-2**
Performance Summary					
As Received Coal (lb/hr)	1,618,190	1,618,190	NA	1,395,457	718,631
Natural Gas to Reformer, OSU CLG (lb/hr)	NA	NA	583,677	NA	272,290
Crude Methanol (lb/hr)	941,823	941,823	940,989	918,582	918,582
Captured CO_2 (lb/hr)		1,569,410	235,808	1,302,138	663,393
Capital and Operating Cost Summary					
Total Plant Costs (2011 MM$)	4,586	4,775	2,028	3,497	2,996
Total Overnight Cost (2011 MM$)	5,572	5,802	2,485	4,236	3,634
Total As Spent Capital (2011 MM$)	6,580	6,852	2,935	5,003	4,291
Annual Fixed Operating Costs (x1000 $/yr)	156,650	162,051	70,644	108,291	94,034
Variable Operating Costs ($/gal)	0.085	0.088	0.032	0.056	0.049
Economic Comparison					
Methanol RSP ($/ton, 2011$)	495.27	535.58	346.56	425.106	446.69
Methanol RSP ($/Gal, 2011$)	1.64	1.78	1.15	1.41	1.48
CO2 TS&M Costs	0.00	0.06	0.01	0.05	0.03
Coal Cost	0.21	0.21	0.00	0.18	0.10
NG Costs	0.03	0.05	0.57	0.00	0.29
Electricity Cost	0.00	0.00	-0.05	0.14	0.11
Variable Costs	0.09	0.09	0.03	0.06	0.05
Fixed Costs	0.14	0.14	0.06	0.10	0.09
Capital Costs	1.18	1.23	0.53	0.89	0.81
Cost of CO_2 Captured	NA	15.66	NA	NA*	NA*

Notes:
*The methanol RSP for the OSU CLG based systems is less than the reference non CO_2 capture case, MBL-1. This results in a negative cost of CO_2 capture which does not have a clear interpretation.

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OHIO STATE UNIVERSITY

CONCEPTUAL DESIGN AND COST ESTIMATE OF CHEMICAL LOOPING GASIFICATION BASED IGCC AND METHANOL PRODUCTION

Techno-Economic Analysis

Final Report

108101-06559 – PCS-0-LI-011-0001-R0

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SYNOPSIS

This report documents the Techno Economic analysis of the Ohio State University's Chemical Looping Gasification technology for IGCC power generation and Methanol production.

Disclaimer

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PROJECT 108101-06559 - TECHNO-ECONOMIC ANALYSIS

REV	DESCRIPTION	ORIG	REVIEW	WORLEY-PARSONS APPROVAL	DATE	CLIENT APPROVAL	DATE
A	Draft for Review	JCS J.C. Simpson			March 9, 2015		
0	Final	JCS J.C. Simpson			March 31, 2015		

CONTENTS

LIST OF EXHIBITS

EXECUTIVE SUMMARY

Syngas produced from coal gasification is currently used for integrated gasification combined cycle (IGCC) power generation and the production of chemical products including natural gas and methanol. IGCC power production provides a potential route for reducing greenhouse gas emission through pre-combustion carbon capture. The use coal gasification allows for the use domestic coal in chemical manufacturing and provides an opportunity for producing these commodities in the United States to provide supply security and an opportunity to export to overseas markets.

While coal costs tend to be low, the capital costs for gasification are high as a result of the gasification equipment and air separation unit (ASU) used for oxygen production. To address these high capital costs, Ohio State University (OSU) has developed a chemical looping gasification technology (CLG) to replace the gasifier and ASU components. In the OSU CLG process, an oxygen carrier (OC), iron oxide in the OSU approach, is circulated between an oxidizing and reducing reactor. In the oxidizing reactor, the OC is reacted with air to oxidize the iron. In reducing reactor, the OC is reduced by coal or natural gas to produce a syngas.

This techno-economic analysis investigates incorporating the OSU CLG technology into IGCC power generation and crude methanol manufacturing processes. Both of these processes require the production of a high hydrogen syngas. The technical options developed herein are based on and referenced to the Department of Energy National Energy Technology Laboratory (DOE/NETL) Cost and Performance Baseline Reports and the Baseline Analysis of Crude Methanol Production from Coal and Natural Gas.

Approach

The approach to develop the assessment of the OSU CLG technology for high hydrogen syngas production included:

- Developing an Aspen model for the OSU CLG technology.
- Incorporating the OSU CLG model into an Aspen based model for power generation and methanol production.
- Determining the equipment size based on the process flows determined from the Aspen modeling.
- Developing capital and operating and maintenance (O&M) cost estimates based on the performance modeling
- Performing an economic assessment with the plant performance and cost estimating results. The economic assessment determined:
    - Cost of electricity (COE)
    - Methanol required selling price (RSP)
    - Cost of CO_2 captured
- Sensitivity studies around economic and process parameters

Summary of Findings

The initial process modelling of the OSU CLG technology using the Fe_2O_3-$FeAl_2O_4$ chemistry showed that a suitable syngas composition for methanol production could be achieved. The approach for producing H_2 in the oxidizer using the Fe_2O_3-$FeAl_2O_4$ chemistry has unfavourable thermodynamics for H_2 production, hence only the modelling of the methanol manufacturing process was performed in this study.

The methanol manufacturing process options compared in this study are summarized in Exhibit ES-0-1.

Exhibit ES-0-1: Methanol Manufacturing Options Compared in this Study.

Option Label	Feedstock	Syngas Production Method	Carbon Capture	Reference/ Basis
MBL-1	Coal	Gasifier	No	NETL MBL Report
MBL-2	Coal	Gasifier	Yes	NETL MBL Report
MBL-3	Natural Gas	Reformer	Yes	NETL MBL Report
OSU-1	Coal	OSU CLG	Yes	This Study
OSU-2	Coal/Natural Gas	OSU CLG	Yes	This Study

A summary of the methanol production performance, the capital costs, and economic results are provided in Exhibit ES-2. A breakdown of the methanol RSP is provided in Exhibit ES-3. The OSU CLG technology provides several savings compared to the state-of-the-art coal base methanol production related to lower capital costs and higher feed stock utilization efficiencies. In comparing the MBL-2 and OSU-1 options, which are both based on a coal only feed and have CO_2 capture, the following benefits are realized:

- A lower methanol RSP by $0.37/gal, a 21% decrease.
- Lower capital costs by 28%
- Higher efficiency based 14% in coal consumption
- A methanol RSP lower than the reference non-capture case, which results in CO_2 capture cost less than 0.

Under the reference economic conditions, the methanol RSP of the OSU-CLG based systems were competitive when compared against the reference MBL-1 and MBL-2 which use coal as feedstock. A fair comparison with MBL-3 can be made by developing a study of the OSU-CLG system with natural-gas as the only feedstock.

Exhibit ES-2: Performance and Cost results for Methanol Production Options.

	Case				
	MBL-1	**MBL-2**	**MBL-3**	**OSU-1**	**OSU-2**
Performance Summary					
As Received Coal (lb/hr)	1,618,190	1,618,190	NA	1,395,457	718,631
Natural Gas to Reformer, OSU CLG (lb/hr)	NA	NA	583,677	NA	272,290
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Capital and Operating Cost Summary					
Total Plant Costs (2011 MM$)	4,586	4,775	2,028	3,497	2,996
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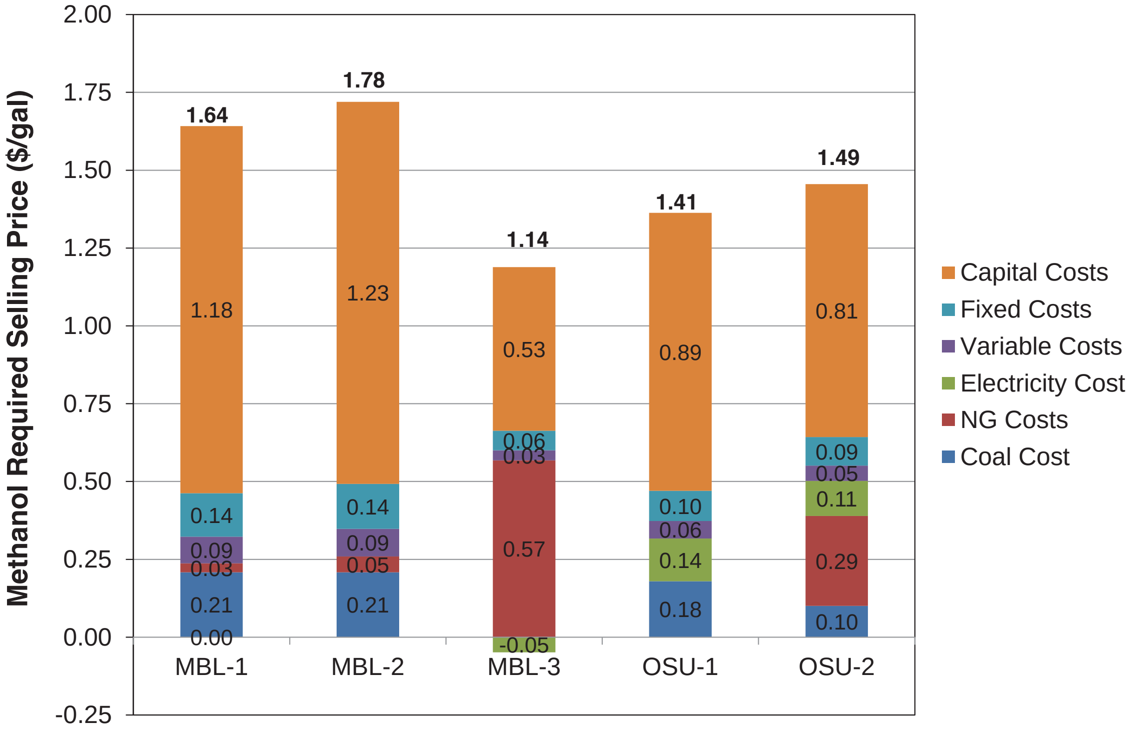
Total As Spent Capital (2011 MM$)[N1]	6,580	6,852	2,935	5,003	4,291
Annual Fixed Operating Costs (x1000 $/yr)	156,650	162,051	70,644	108,291	94,034
Variable Operating Costs ($/gal)	0.085	0.088	0.032	0.056	0.049
Economic Comparison					
Methanol RSP ($/ton, 2011$)	495.27	535.58	346.56	425.106	446.69
Methanol RSP ($/Gal, 2011$)	**1.64**	**1.78**	**1.15**	**1.41**	**1.48**
Cost of CO_2 Captured	**NA**	**15.66**	**NA**	**NA[N2]**	**NA[N2]**

Notes:

[N1] The Total overnight and total as spent capital presented here for the MBL-1, MBL-2, and MBL-3 cases are different from those presented in the Crude Methanol Baseline report. This difference is related to using a consistent methodology for the owners' cost prepared for this study. The owners' costs methodology for the Crude Methanol Baseline study is not provided in that report.

[N2] The methanol required selling price for the OSU cases is less than that of the reference case, MBL-1. This is a result the OSU cases higher efficiency of the process and lower capital costs. Using MBL-1 as a reference, results in a negative carbon capture cost. Implications of a negative cost in this case are not clear.

Exhibit ES-3: Breakdown of Methanol Required Selling Price.



Sensitivity studies were completed to identify parameters that have a significant impact on the production costs and thus identify opportunities to focus on for future cost reductions and parameters that could change and result in a change in the most favorable option. The results of this sensitivity study are:

- Feedstock Costs:
  - For coal based options (MLB-1, MLB-2, OSU-1 and OSU-2), the methanol RSP sensitivity to feedstock cost is significantly less than options that use only natural gas as a feedstock (MLB-3); approximately 10% with doubling coal costs and 30% with doubling natural gas costs.
  - With electricity cost increases from the reference price of $60/MWh to $130/MWh, the methanol RSP increases by 12% for the OSU CLG options. With the high electricity costs $130/MWh, the OSU options methanol RSPs are less than the RSP for the MBL-1 and MBL-2 options.
- OSU CLG Reactor Performance and Costs
  - Faster reaction kinetics in the CLG will result in shorter residence times and larger reactor size. With a 50% decrease of the reactor residence time, the methanol RSP decreases by 4%. Future work should be performed to determine routes to increase the reaction kinetics between the coal and OC,
  - A 50% decrease the OSU CLG technology equipment costs results in a 5% decrease in the methanol RSP.

  - With a 50% increase in the OC cost, there is less than 1% increase in the methanol RSP. Therefore, uncertainty in the OC costs should not be a negative economic concern with regards to the further development of the OSU CLG technology.
  - Doubling the OC attrition rate, thus doubling the OC consumption rate during operations, increases the methanol RSP by less than 1%. This finding allows for greater flexibility in the improved OC materials. That is, materials that offer improved kinetics while attriting at a greater rate would be acceptable.
- CO_2 Emissions and Selling Price
  - The OSU CLG based options incorporate CO_2 capture and therefore essentially independent of CO_2 emission taxes. The methanol RSP for the OSU CLG options were lower than that of the MBL coal based option without CO_2 capture.
  - With CO_2 revenue up to $50/tonne CO_2, the methanol RSP decreases by 15% and 8% for the OSU-1 and OSU-2 options respectively.

Recommendations for Future Evaluations

The production of a very high hydrogen syngas from the OSU CLG technology would be very beneficial. Based on the findings of this study, the following recommendations are suggested:

1. While there is not a strong sensitivity to OSU CLG capital costs, as illustrated in this study, further work to substantiate these costs should be performed. This work should include the more detail design and sizing of the reactors and equipment requirements to incorporate that OSU CLG technology into the process.
2. Explore a OSU CLG configuration using natural gas as the only feedstock to the OSU CLG process
3. Explore the configurations to investigate the possibility of power generation as part of the OSU CLG system design
4. Incorporate sensitivity study findings in experimental work to direct future research to verify the OC costs and the reactor residence time assumptions at a larger integrated system scale
5. Explore chemistries beyond the Fe_2O_3-$FeAl_2O_4$ cycle for high hydrogen syngas compositions for IGCC power generation applications

1. INTRODUCTION

The overall goal of the project is to evaluate the Ohio State University (OSU) Chemical Looping Gasification (CLG) technology for the production of high hydrogen syngas compositions for integrated gasification combined cycle (IGCC) power generation and crude methanol production. Based challenges with the producing syngas compositions suitable for 90% carbon capture from IGGC power generation and technical direction from OSU, the study was limited to the production of methanol from configurations with coal only feedstock and coal and natural gas feedstock. The change in the technical direction was based on preliminary results from the Aspen modeling of the CLG system.

The overall project objectives for project period are to perform a techno-economic analysis (TEA) of applying OSU CLG technology to IGCC power generation with a net generation of 550 MWe and crude methanol production facility with a 10,000 tonne/day capacity. For both applications the primary fuel/feedstock is subbituminous Powder River Basin (PRB) coal.

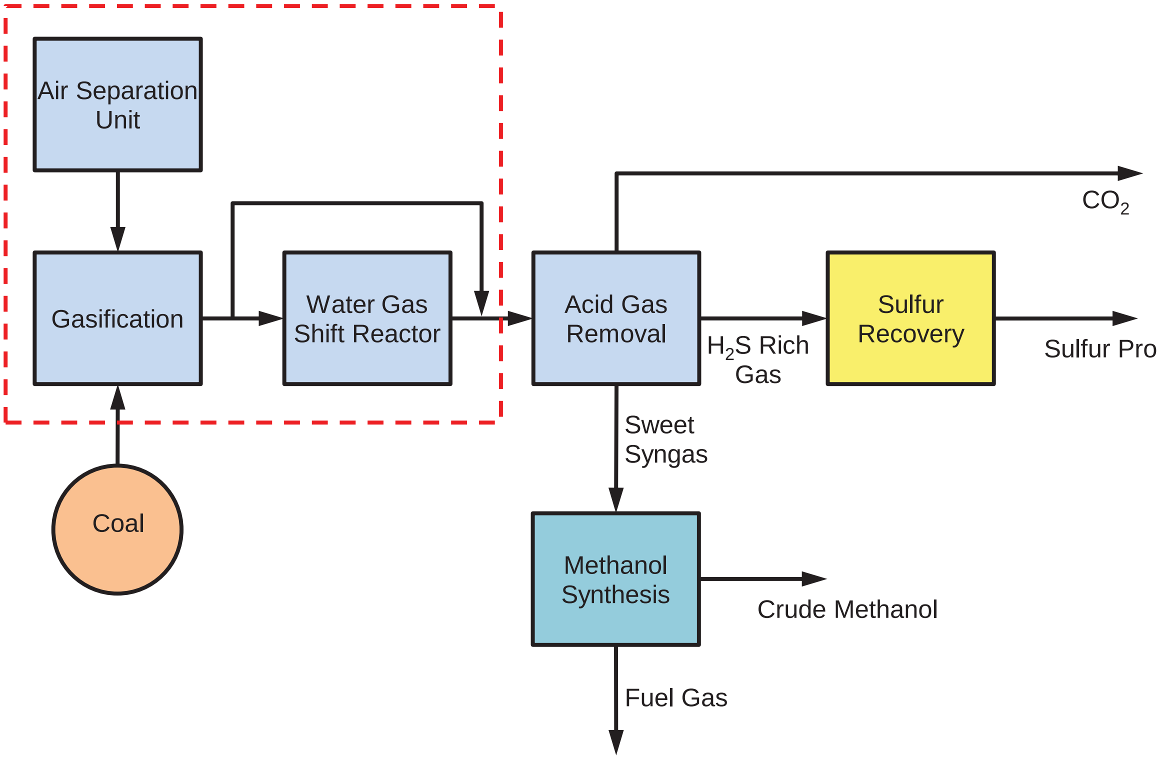
The preliminary TEA is conducted in accordance with the guidelines provided in SOPO [1].

The evaluation scope includes:

- Developing an Evaluation Basis Document that defines essential technical and functional requirements in establishing a conceptual design based on a nominal 550 MW (net), PRB fired greenfield IGCC power plant for the pre combustion capture technologies, similar to that used for Cases S1A and S1B, PRB fired Shell gasifier IGCC without and with CO_2 capture, of the Cost and Performance Baseline for Fossil Energy Plants, Volue 3a: Low Rank Coal to Electricity, IGCC Cases [2] and the Baseline Analysis of Crude Methanol Production from Coal and Natural Gas[3].
- Modeling and optimizing IGCC and methanol processes that utilize the OSU CLG technology for the production of the required high hydrogen syngas and capture of CO_2,
- Sizing the major equipment in the processes,
- Estimating the equipment costs
- Performing a lifecycle cost analysis to determine the unit production costs for power and methanol and the cost of CO_2 capture, and
- Comparing the unit costs to those presented in the Department of Energy/National Energy Technology Laboratory baseline studies.
- Analyzing sensitivity of key parameters impacting cost of methanol production.

In industry, methanol production methods, from syngas to methanol, vary significantly. To minimize performance and cost variations from this component, and isolate the differences to the syngas production method, the syngas from the CLG was matched to the syngas composition in the methanol baseline report and the equipment sized to the production rate. Additionally, the cost provided in the methanol production equipment was scaled to account slight differences in methanol production rates. Exhibit 1-1 further illustrates this approach with a high level block diagram of the coal to methanol process as described in the methanol baseline report. The dashed red box identifies that components replace by the OSU CLG technology, specifically the air separation unit, gasifier, and water shift reactor.

Exhibit 1-1 Block Diagram of Reference Coal Based Methanol Production Configuration



Adapted from NETL[3]

2. EVALUATION BASIS

2.1 Engineering/Technical Design Specifications

The essential technical and functional requirements used as a basis in establishing conceptual designs for this study are provided in the following sections.

2.1.1 Site Conditions

The IGCC plant in this study is assumed to be located at a generic plant site in the Montana while the coal to methanol plant is assumed to be located in the Midwest. These site selections provide minimum variation of the cases developed from the reference cases presented in the baseline studies. The site specification and ambient conditions are presented in Exhibit 1-1 and Exhibit 1-2. The ambient conditions are the same as ISO conditions.

Exhibit 2-1 Site Characteristics

Site Characteristics	
Topography	Level
Size, acres	300
Transportation	Rail
Ash/Slag Disposal	Offsite
Water	50% Municipal and 50% Ground water
Location	Greenfield
CO_2 Storage	Compressed to 15.3 MPa (2,215 psia), transported 80 km (50 mi) and sequestered in a saline formation at a depth of 1,239 m

Exhibit 2-2 Site Conditions

Site Conditions	Montana Site[2]	Midwest Site
Elevation, m (ft)	1,036 (3,400)	0 (0)
Barometric Pressure, MPa (psia)	0.09 (13.0)	0.101 (14.7)
Design Ambient Dry Bulb Temperature, °C (°F)	5.6 (42)	15 (59)
Design Ambient Wet Bulb Temperature, °C (°F)	2.8 (3.7)	10.8 (51.5)
Design Ambient Relative Humidity, %	62	60
Cooling Water Temperature, °C (°F)	2 (36)	15.6 (60)
Cooling Water Range, °C (°F)	11 (20)	8.3 (15)
Air composition based on published psychrometric data, mass %		
H_2O	0.616	
Ar	1.280	
CO_2	0.050	
O_2	22.999	
N_2	75.055	
Total	100.000	

The following evaluation considerations are site-specific, and are not be quantified for this study. Allowances for normal conditions and construction are included in the cost estimates. Typically the considerations of these factors do not have a significant impact on the cost unless the site specific situation is unusual or extreme.

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

2.1.2 Coal Characteristics and Pricing

The design coal in the project is Powder River Basin (PRB) subbituminous Coal. The coal characteristics are listed in Exhibit 2-3.

Exhibit 2-3 Design Coal Characteristics

Rank	**Sub-bituminous**	
Seam	**Rosebud PRB, Area D**	
Sample Location	**Montana**	
Proximate Analysis (weight %)		
	AR	**Dry**
Moisture	25.77	0
Ash	8.19	11.04
Volatile Matter	30.34	40.87
Fixed Carbon (BD)	35.70	48.09
HHV, kJ/kg	19,920	26,787
HHV, Btu/lb	8,564	11,516
LHV, kJ/kg	19,195	25,810
LHV, Btu/lb	8,252	11,096
Ultimate Analysis (weight %)		
	AR	**Dry**
Moisture	25.77	0
Carbon	50.07	67.45
Hydrogen	3.38	4.56
Nitrogen	0.71	0.96
Chlorine	0.01	0.01
Sulfur	0.73	0.98
Ash	8.19	11.03
Oxygen (BD)	11.14	15.01
Sulfur Analysis (weight %)		
	AR	**Dry**
Pyritic	-	0.63
Sulfate	-	0.01
Organic	-	0.34
Trace Components (ppmd)		
Mercury	-	0.081

2.1.3 Natural Gas

The characteristics of the natural gas used in the techno-economic are listed in Exhibit 2-4.

Exhibit 2-4 Design Natural Gas Characteristics

Component Volume	Percentage	
Methane, CH_4	93.1	
Ethane, C_2H_6	3.2	
Propane, C_3H_8	0.7	
n-Butane, C_4H_{10} [1]	0.4	
Carbon Dioxide, CO_2[2]	1.0	
Nitrogen, N_2[2]	1.6	
Total	100.0	
	LHV[3]	**HHV**[3]
MJ/scm	34.71	38.46
kJ/kg	47,454	52,581
Btu/scf	932	1,032
Btu/lb	20,410	22,600

Notes:
1. The reference data reported the mean volume percentage of higher hydrocarbons (C_4+) to be 0.4%. For simplicity, the above composition represents all the higher hydrocarbons as n-butane (C_4H_{10}).
2. The reference data reported the mean volume percentage of CO_2 and N_2 (combined) to be 2.6%. The above composition assumes that the mean volume percentage of CO_2 is 1.0%, with the balance (1.6%) being N_2.
3. LHV = lower heating value; HHV = higher heating value

2.2 CO_2 Specifications

The captured CO_2 is transported and injected as a supercritical fluid in order to avoid two-phase flow and to reach maximum efficiency. CO_2 is supplied to the pipeline at the plant fence line at a pressure of 15.3 MPa (2,215 psia). The CO_2 product gas composition varies, but is expected to meet the specification from the DOE/NETL Baseline studies summarized in Exhibit 2-5. If required, a glycol dryer located near the mid-point of the compression train is used to meet the moisture specification.

Exhibit 2-5 CO_2 Pipeline Specification

Parameter	Units	Value
Inlet Pressure	MPa (psia)	15.3 (2,215)
Outlet Pressure	MPa (psia)	10.4 (1,515)
Inlet Temperature	°C (°F)	35 (95)
N_2 Concentration	ppmv	< 300
O_2 Concentration	ppmv	< 40
Ar Concentration	ppmv	< 10
H_2O Concentration	ppmv	< 150

2.3 Plant Capacity Factor

For the coal to methanol cases, the plant capacity factor is 90%, per Attachment 2 of the RFP and consistent with the CTL Baseline, the CBTL study and the Crude Methanol study.

For the power generation case, the plant capacity factor is 80%, per Attachment 2 and consistent with the Baseline Studies.

2.4 Sparing Philosophy

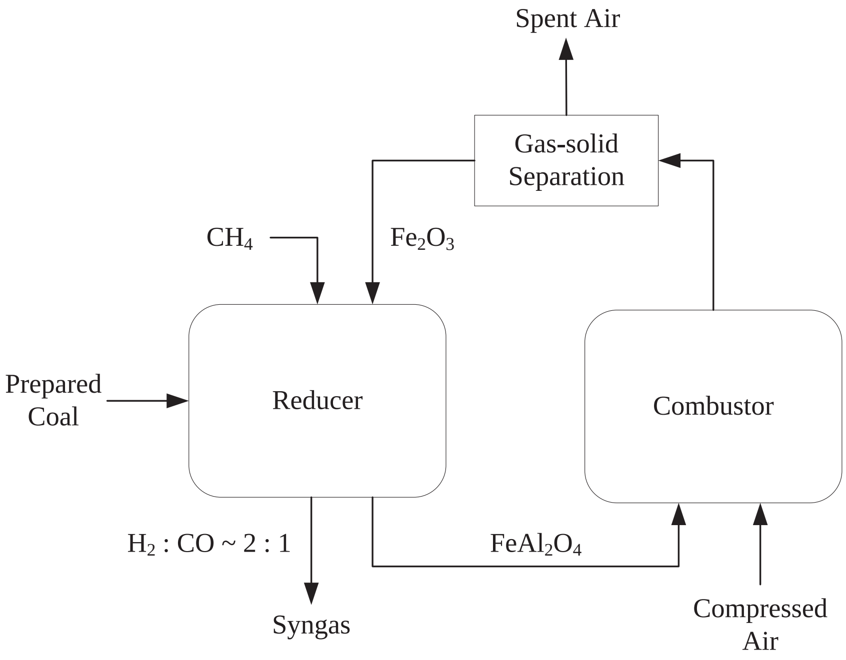
For the equipment common to the reference methanol production plant and the power generation facility, the sparing philosophy used in the reference plant is followed. For the equipment related to the OSU CGL technology, guidance from OSU is followed to ensure the design, including the proposed maintenance schedule, is capable of achieving the target capacity factor. Discussion of multiple trains and other potential redundancies need was reviewed. There is no redundancy other than normal sparing for rotating equipment.

2.5 Design Cases

The OSU CLG technology relies on employing a solid iron-based oxygen carrier to react with coal in order to selectively generate a hydrogen-rich syngas, with near-zero CO_2 emission, which can then be used for power generation or crude methanol production. OSU has developed two distinct system configurations which are described below:

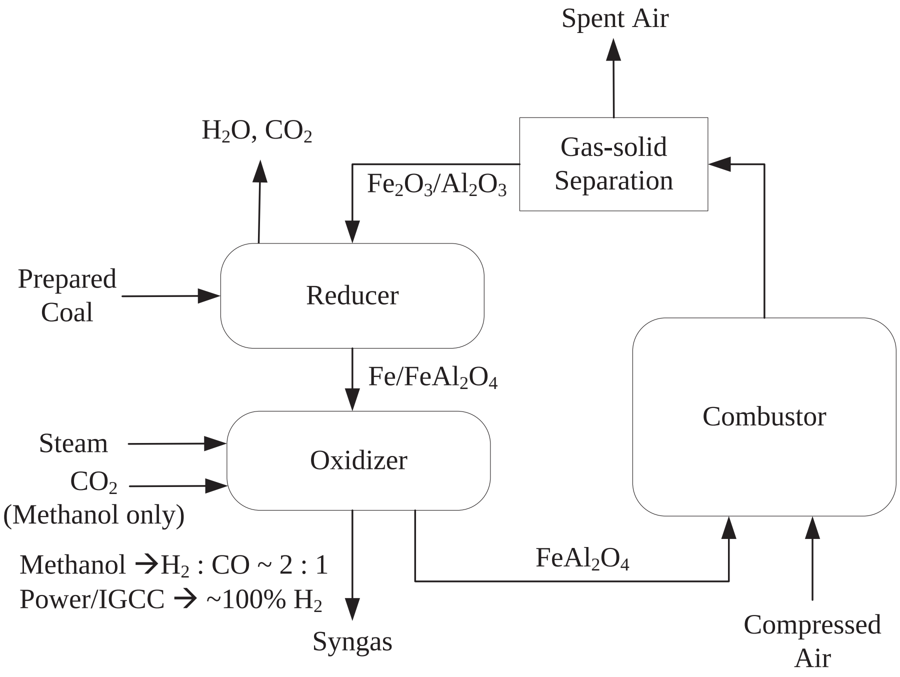
A) Two-reactor system which uses natural gas co-feeding to increase the hydrogen content in the syngas (Refer to Exhibit 2-6).

Exhibit 2-6 OSU CLG Two-reactor system



B) A three-reactor system which inherently captures carbon from coal (Refer to Exhibit 2-7). In the liquid production case, steam and carbon dioxide are injected into the oxidizer in order to produce syngas required for methanol production. For electricity generation, only steam is injected into the oxidizer to produce hydrogen, which is then fed to the combine cycle power generation portion of the plant.

Exhibit 2-7 OSU CLG Two-reactor system



The selected cases considered in this techno-economic Analysis are presented in Exhibit 2-8. Due to modelling difficulties encountered with the three reactor system only the two reactor options were pursued.

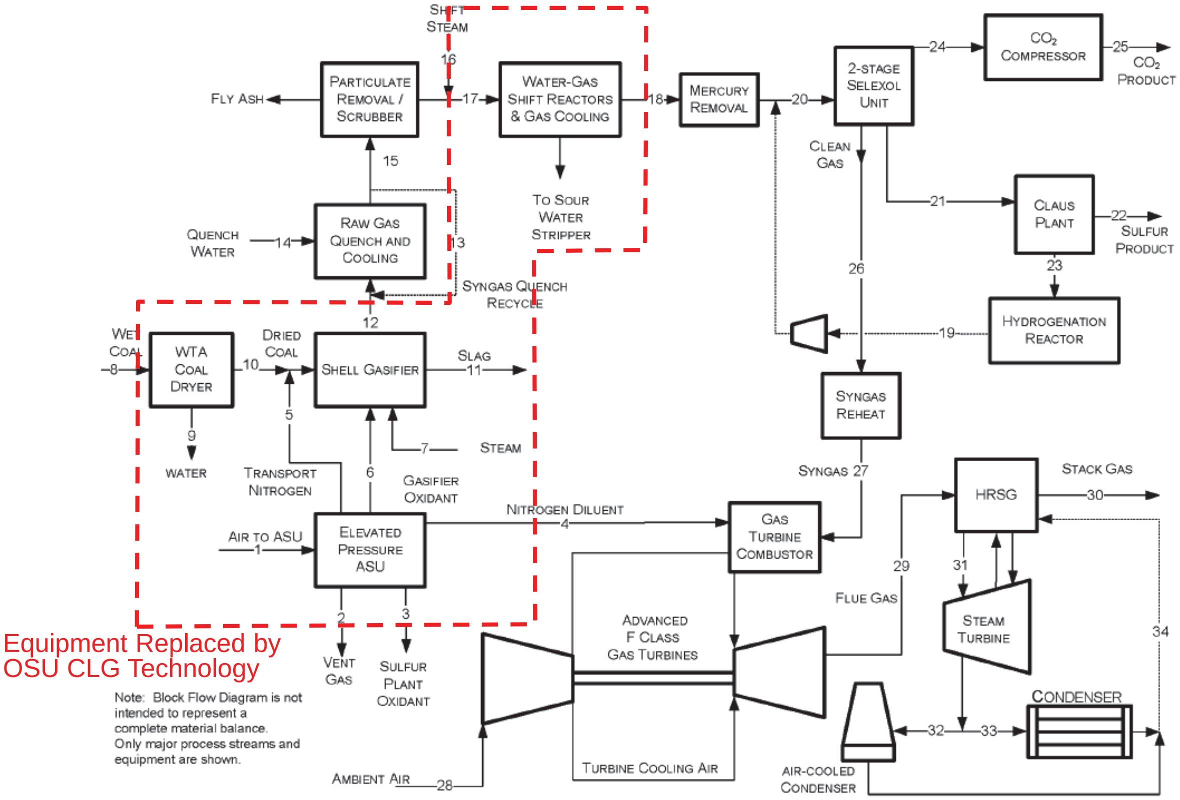
Exhibit 2-8 Selected Cases

Case Description		Technology Status	Reference
Power Generation			
IGCC 1	Case S1A - IGCC case with Shell gasifier, with no carbon capture	Conventional	Low Rank Coal Baseline study
IGCC 2	Case S1B - IGCC case with Shell gasifier with carbon capture	Conventional	Low Rank Coal Baseline study
IGCC 3	OSU CLG – IGCC case with carbon capture using 3-reactor system	Advanced	N/A
Crude Methanol Production			
MBL-1	Reference from DOE/NETL (pending) coal feedstock without CO_2 capture	Conventional	NETL MBL Report
MBL-2	Reference from DOE/NETL (pending) coal feedstock with CO_2 capture	Conventional	NETL MBL Report
MBL-3	Reference from DOE/NETL (pending) natural gas feedstock with CO_2 capture	Conventional	NETL MBL Report
OSU-1	OSU CLG – 2 reactor system – coal only	Advanced	N/A
OSU-2	OSU CLG – 2 reactor system – coal and natural gas	Advanced	N/A
OSU-3	OSU CLG – 3 reactor system	Advanced	N/A

For power generation, the reference IGCC cases using conventional technologies were considered from the DOE/NETL Baseline Volume 3a study. The potential IGCC cases presented in this report utilize two advanced F turbines (232 MW each) and include gasification technologies offered by GE (radiant only), CoP E-Gas and Shell. Ultimately, the non-capture and capture cases based on the Shell gasifier, Cases S1A and S1B, of the aforementioned study was selected as the Shell gasifier utilizes a dry coal feed, in a similar manner than the OSU CLG technology. A process flow comparison between the proposed OSU CLG system and reference Case S1B can be observer in Exhibit 2-9.

For crude methanol production the reference case using a conventional technology is pending the NETL's publication of CTL Baseline report. The block diagram for this process, as presented in a draft version of this report is presented in Exhibit 1-1.

Exhibit 2-9 Low Rank Coal Baseline study – Case S1B IGCC with CO_2 capture BFD (Shell gasifier)



Note: Block Flow Diagram is not intended to represent a complete material balance. Only major process streams and equipment are shown.

3. METHANOL PRODUCTION FACILITY WITH OSU CLG TECHNOLOGY EVAULATION METHODOLOGIES

A conceptual methanol plant with CO_2 capture design and costs were developed to compare the OSU CLG to state-of-the-art methanol production from coal without and with CO_2 Capture. The state-of-the-art methanol production from coal, and as illustrated in Exhibit 1-1, consist of an air separation unit (ASU) for the production of oxygen that is used to gasify coal in a gassifier to produce a syngas. This syngas is then conditioned through removing the CO_2 and sulfur and adjusting the H_2 to CO ratio to approximately 2. The conditioned syngas is then fed into a methanol production system for the production of crude methanol. As illustrated in Exhibit 1-1, the OSU CLG technology replaces the ASU and gasifier blocks in the state-of-the-art coal to methanol process. The methodologies used in this evaluation include:

- Methanol production performance modeling
- Equipment sizing
- Capital and O&M cost estimates
- Economic analysis

Results from the OSU process development activities and the methodologies for the system modelling, equipment sizing, cost estimating, and economic analyses are provided in the following subsections. Details of the overall coal to methanol process are provided in section 4.

3.1 Performance Modeling Methodology

Aspen performance modeling executed during this study focused on understanding and controlling the syngas composition from the OSU CLG technology and integrating the CLG technology into the coal to methanol process IGCC power production. The following subsections address these two steps.

3.1.1 OSU CLG Syngas Composition

The objectives of Aspen OSU CLG syngas composition modeling were directed at:

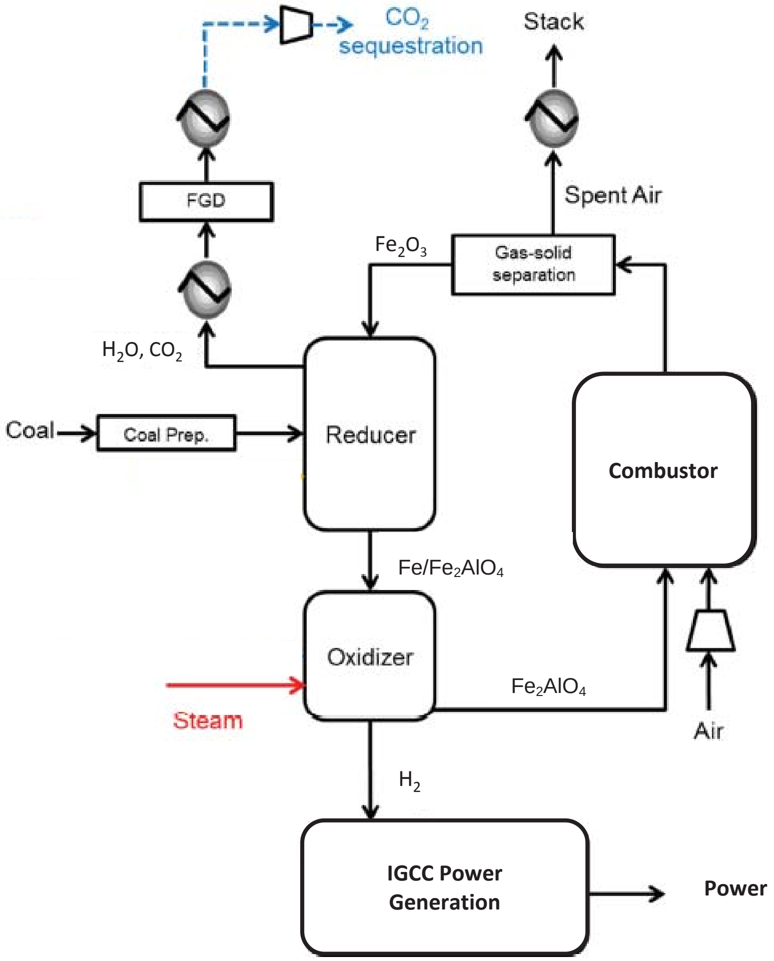
1. Producing a suitable gas composition for either methanol or power production,
2. Maximize efficiency (minimize feed stock consumption),
3. Provide sufficient heat to maintain temperatures through reactor, and
4. Providing a high level of carbon capture.

IGCC Power Production

For IGCC power generation, 95% of the carbon in the fuel must be convered to CO_2 while the bulk of the hydrogen remains as H_2 to be used as fuel in the combustion turbine. OSU's goal in this project was to provide carbon capture through the OSU CLG equipment without relying on traditional acid gas separation equipment such as Rectisol. OSU's approach was based on a three reactor system as illustrated in Exhibit 3-1. In the first reactor reducing reactor, the C in the coal is completely oxidized to produce CO_2 which is then purified, compressed and sent to storage. Ideally, the process is operating under strong reducing conditions near the solids exit of the reducing reactor so that the iron is reduced to the FeO and the Fe states. The resulting OC are then transported to the first oxidizer where they react with steam to produce hydrogen and partially oxidized oxygen carrier. The hydrogen is then fed to the IGCC unit for powere production while the OC moves to the send oxidizer where it reacts with air to produce completely oxidized OC.

During the initial Aspen modeling of this system, it was found that OC reduced to the point where metalic iron was present (Fe), solid carbon would also be present in the OC. This solid carbon would then be transported with the OC to the first oxidizing reactor where it would be oxidized to either CO or CO_2 and mixed with the fuel to the IGGC unit and result in CO_2 emissions. Resolutions to these CO_2 emissions were not identified by the OSU/WorleyParsons team. Therefore, based on these results, an IGCC power generation option was not developed during this study.

Exhibit 3-1 Proposed Three Reactor System for H_2 Generation



Methanol Production

For methanol production, the metric commonly used to compare syngas compositions is the M value, defined as:

$$M = \frac{H_2 - CO_2}{CO + CO_2}$$

Where H_2, CO_2, and CO are the mole fraction percentages of syngas components

For methanol production, the M value is typically on the order of 2. From the gas compositions in the DOE/NETL Crude Methanol Report, this value is calculated as 1.70. To limit the changes to the methanol synthesis process following the OSU CLG gasification system from that in the Crude Methanol Report, the composition of the sweet syngas, as characterized by the M-value, from the OSU CLG was matched to that of the Crude Methanol Report.

The initial configurations developed attempted to avoid the use of CO_2 capture to achieve the desired M value through adjusting the steam and OC flow rates and temperatures to the OSU reducing reactor. The Aspen modeling performed on these configurations indicated that achieving this M value while obtaining sufficient heat from the reactions was difficult. To maintain temperatures in the reactors, the fuel is required to be oxidized so that sufficient amounts of CO_2 are produce and the target M value could not achieved. Additionally, the syngas contained significant amounts of H_2S which would potential poison catalysts in the methanol production block and/or become an impurity in the final product. Therefore, an acid gas removal system was integrated into the system to mitigate these problems. To limit differences between the Crude Methanol Report configuration and the configurations developed in the current project Rectisol was selected for the AGR.

The optimization of the OSU CLG was performed with an Aspen model that focused on the CLG reactors, neighboring equipment and the inputs to these systems. The block diagram in Exhibit 3-2 illustrates the components considered in this optimization process. The oxidizer and reducer were treated as RGibbs reactors in the Aspen model. The model was exercised through varying the reactor operating conditions and reactor inputs and monitoring the syngas composition and the duty of the reducing reactor. The optimization objective was to achieve an H_2 to CO molar ratio of 1.7 and reducing reactor heat duty of zero. Exhibit 3-3 illustrates the syngas composition versus Fe_2O_3 to C ratio scan used to determine the optimal composition for the development of the methanol synthesis process.

Exhibit 3-2 Block Diagram for Optimization of CLG System

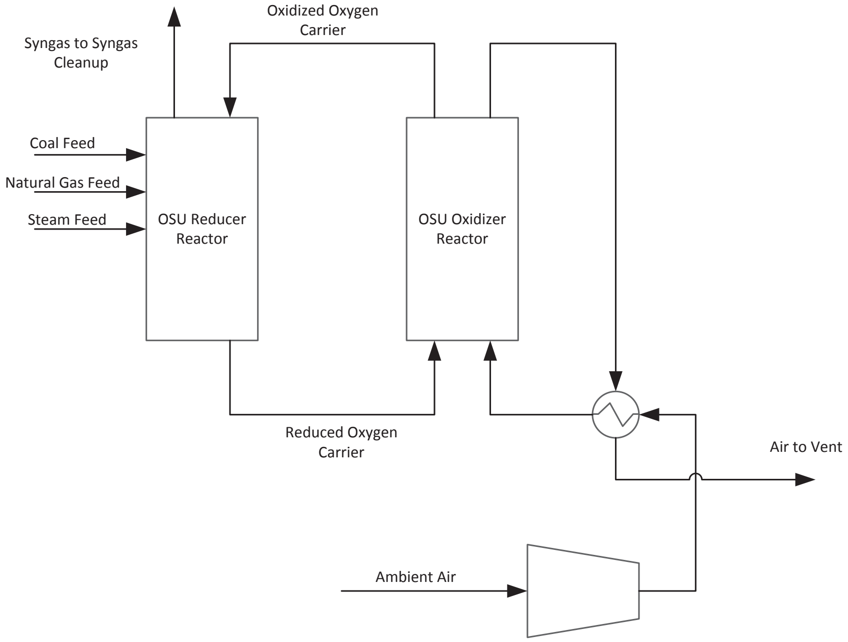
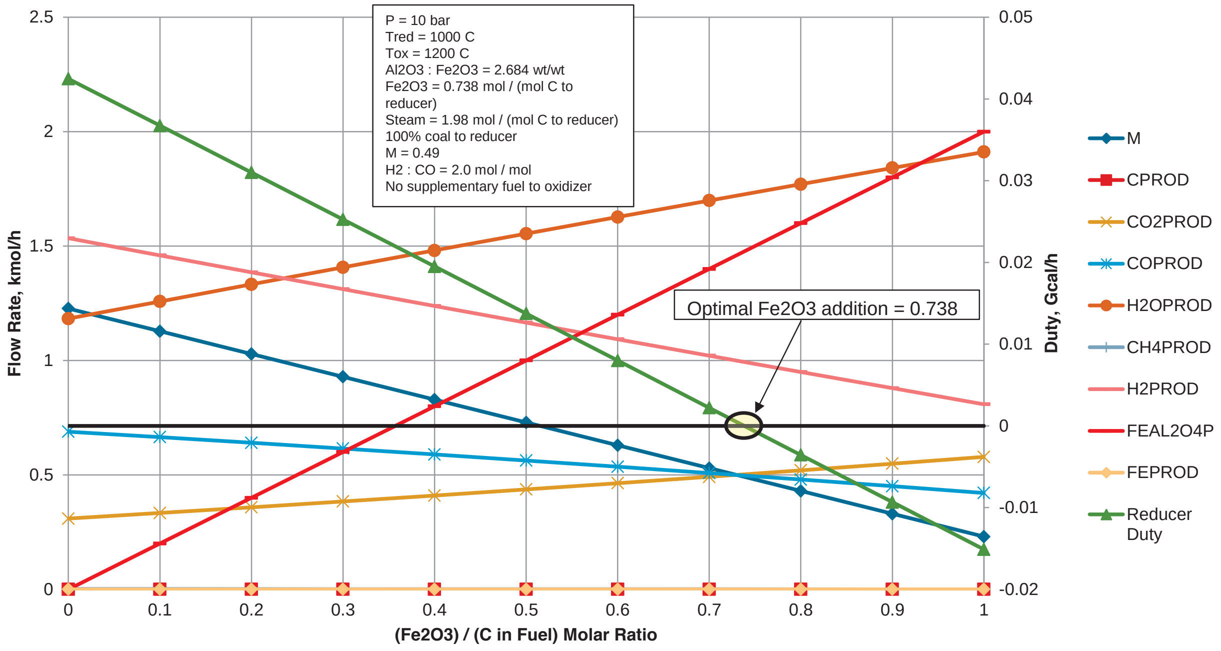


Exhibit 3-3 Optimization Scan for OSU CLG System



3.1.2 Overall Process Integration

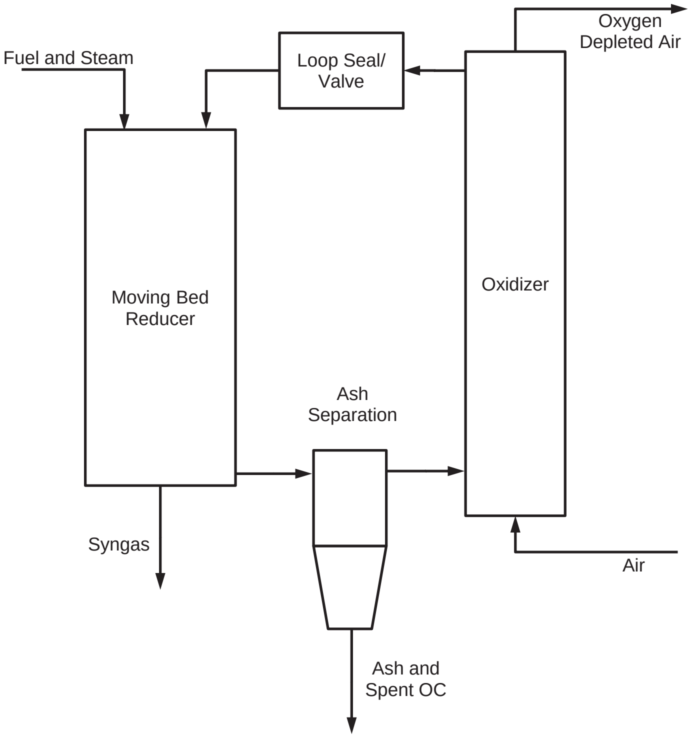
The modeled OSU CLG system was integrated in the complete methanol production process through a second aspen model. The focus of this second model was to balance the steam generation against steam loads in the system. Primary heat sources through the system are the Claus plant, gas cooling, the purge gas combustion, and the methanol synthesis process. The heat generated from the methanol synthesis process was determined through modeling a process similar to that presented in the Crude Methanol Baseline report. The steam loads included the process steam sent to the OSU CLG equipment and the steam to the steam turbine for power generation. The steam system was designed such that the purge gas boiler was used as an evaporator to generate saturated steam that was subsequently superheated in through process cooling. This superheated steam was then used as process steam. Additional superheating from the boiler was used to generate steam for the steam turbine and power generation.

3.2 Equipment Sizing Methodology

For equipment outside of the OSU-CLG block, the resulting mass and energy balance data from the simulation models were used to size major pieces of equipment of the plant. To maintain consistency for the methanol synthesis process between the Crude Methanol Baseline report and the current study, the sizing of this equipment was assumed to be the same for the same methanol production rate. Additionally for this equipment, sizing and sparing philosophies consistent with the Methanol Baseline Report were used.

The OSU CLG components considered in the cost estimate are illustrated in Exhibit 3-4. The reducer and oxidizer size are derived from the residence time required for the reaction between the coal and the oxygen carrier and the oxygen carrier and air. From OSU experimental work, residence times for the complete reaction of carbon in the coal were on the order of 20 minutes. The reducer reactor design was assumed to be a moving bed reactor, with a conservative residence time of 30 minutes. For the reducer, the larger diameter of the two primary reactors, the dimensions were assumed to be 20 feet in diameter with 12 inches of refractor liner and 50 feet tall. The diameter of the oxidizer reactor was determined from the oxidizer to reducer residence time ratio and the OC bulk density ratio in the reactors. The height of the oxidizer reactor was assumed to the same as the reducer, 50 feet. The bulk ratio density between the reactors was assumed to be 0.6 to 0.4 based on a moving bed for the reducer reactor and a bubbling fluidized bed for the oxidizer reactor. The resulting oxidizer outside diameter is 8 feet which includes 12 inches of refractory lining.

Exhibit 3-4 OSU CLG Components Considered for Sizing and Cost Estimating



3.3 Cost Estimating Methodology

3.3.1 Capital Cost Estimates

The total capital cost estimates include the cost of equipment, freight, bulk materials and labor (direct and indirect) for equipment installation and erection; materials and labor for construction of buildings, supporting structures, and site improvements; engineering, construction management, and start-up services (Professional Services); and process and project contingency. The estimate excludes owner's costs and is provided as "overnight" costs; that is, escalation to period of performance is excluded.

Equipment Costs

Costs for other equipment and balance of plant items were developed via scaling and/or parametric modelling based on key project and equipment parameters. Where possible, costs were scaled from the Crude Methanol Baseline Study. This approach was used when determining the methanol synthesis equipment costs to maintain a focus on the cost difference between the gasifier and CLG systems. For other equipment costs, capital costs were developed using a combination of commercial capital cost estimating software, factored equipment estimates, vendor information and WorleyParsons in-house parametric models supplemented by WorleyParsons' extensive in-house equipment cost database.

The capital costs for the OSU-CLG equipment were developed based on vendor quotes for steel and refractory for the vessel sizes determined from the equipment sizing activities. A factor approach was then utilized to determine the materials and labor cost components, 0.5 and 1.0 of the equipment costs, associated with the bare erected costs. These cost factors are based on WorelyParsons' previous experience in developing costs for large refractory line steel pressure vessels.

Engineering, Construction Management, Home Office and Fees

Home office expenses and other owner's costs were based on an allocation included in the Methanol Baseline report. For the OSU CLG equipment, rates similar to those applied to the gasification equipment in the Crude Methanol Baseline report were used.

Process Contingency

Process contingency is typically applied technologies that are not commercially proven to cover costs required to install an operating unit. These costs are associated with technical unknowns at various points of the development cycle. Typical process contingencies per QGESS are as follows:

Commercial:	0%-10%
Modules Operated at Scale:	5%-20%
Bench Scale:	30%-70%
New Concept w/ Limited Data:	40%+

The OSU CLG technology has been developed and tested at the Bench Scale in the laboratories at the OSU and is currently undergoing pilot scale testing at the National Carbon Capture Center in Wilsonville Alabama. The status of the OSU CLG technology can be considered between the Bench Scale and Modules Operated at Scale. Based on this status, a process contingency of 15% was used in the cost estimates.

Project Contingency

The project contingency covers costs that will be incurred in installing the project but are not accounted for in the level of cost estimating performed at this time. As a technology or project progress, and more detailed engineering is performed, the project contingency will decrease. For the equipment common between the Methanol Baseline report and the current study, the contingency rates were held constant. For the OSU CLG equipment the project contingency rates were set at 15%.

3.3.2 Operating Cost Estimates

The operating costs and related maintenance expenses (O&M) pertain 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 Labor Support
- Consumables
- Waste Disposal
- Fuel

- Electricity
- Co-Product or By-Products credit (that is, a negative cost for any byproducts sold)

There are two components of O&M costs; fixed O&M, which is independent of power generation, and variable O&M, which is proportional to annual power generation. The fixed operating costs do not include the cost of capital. The variable O&M cost includes an estimate of fuel cost. The annual consumables costs include accounting for the annual capacity factor (90%); that is:

Annual Cost = Hourly Consumption Rate x 8760 hours/yr x 0.90 x Unit Cost.

A breakdown of the various cost components considered is provided in the following sub sections.

Operating Labor (Fixed Cost)

The manpower estimates to operate and maintain the facility were developed by WorleyParsons with input from OSU using the manpower basis presented in the NETL/DOE Cost and Performance Baseline Reports. Based on this review, the number of personal required to operate the OSU CLG based methanol plant is the same as the number of personal required for the gasifier/coal Methanol Baseline cases [3]. Therefore, the same operating labor costs are used in all cases in this assessment.

Administrative Labor and Support Labor (Fixed Cost)

Labor administration and overhead charges are assessed at a rate of 25 percent of the burdened O&M labor.

Maintenance Material and Labor (Fixed Cost)

Maintenance costs are evaluated on the basis of relationships of maintenance cost to initial capital cost. This represents a weighted analysis in which the individual cost relationships are considered for each major plant component or section.

Consumables (Variable Cost)

For general consumables such as limestone and water, costs provided by the DOE report escalated to June 2011 dollars are used in the analysis. For consumables specific to the OSU CLG technology, specifically the oxygen carrier (OC) costs OSU developed the consumption rates and costs. Based on OSU's assessment, which considered the cost of the alumina and iron oxides raw materials and the productions costs to produce a sintered ceramic body, the nominal cost for the OC are set at $600/t.

Waste Disposal (Variable Cost)

Waste quantities and disposal costs are determined similarly to the consumables. Ash and spent oxygen carrier from the OSU CLG technology are treated in similar manner as fly ash and bottom ash from the PC cases in the NETL/DOE Cost and Performance Baseline Reports

Co-Products and By-Products (Other than CO_2) (Variable Cost)

Co-Products Costs are assessed according to Attachment 2 of the RFP. For the methanol production cases, net power imports or exports are priced at $60/MWh.

By-product quantities are determined similarly to the consumables. However, due to the questionable marketability of these by-products (bottom ash; fly ash co-mingled with FGD products, carryover from the OC material) no credit is taken for potential saleable value.

CO_2 Transport, Storage and Monitoring (Variable Cost)

The CO_2 transportation, storage and monitoring costs are defined by the Attachment 2 of the RFP. Transport and storage (T&S) cost for CO_2 are be based on a Midwest location at \$11/tonne CO_2 as used in the Crude Methanol Baseline study.

3.3.3 Owners' Costs

The economic analysis accounts for the owner's costs associated with the facilities. For real world projects, these costs are strongly dependent on location and the owners involved in the project. For the current study the methodology and guidance regarding the basis and rates for the owners' costs are consistent with the DOE/NETL Baseline studies and are summarized in Exhibit 3-5

Exhibit 3-5 Owners' Costs Basis

Owner's Costs	Basis
Preproduction costs	
6 Months all labor	Sum of Operating, Maintenance and Administrative Labor
1 Month maintenance materials	Annual maintenance materials @ 85% capacity
1 Month non-fuel consumables	Annual consumables @ 85% capacity
1 Month waste disposal	OPEX disposal costs @ Capacity Factor (CF)=85%
25% of 1 months fuel cost at 100% CF	Annual fuel costs @ 85% capacity
2% TPC	TPC
Inventory Capital	
60 day supply of fuel and consumables at 100% CF	OPEX fuel and consumables
Spare parts	0.5% of TPC
Land	$3,000/acre, 300 acre for PC plants
Financing Costs	2.7% of TPC
Other Owner's Costs includes: • Preliminary feasibility studies, including Front-End Engineering Design (FEED) study • Economic development • Construction and/or improvement of roads and/or railroad spurs outside of site boundary • Legal Fees • Permitting costs • Owner's engineering • Owner's Contingency (Management reserve, funds to cover costs relating to delayed startup, fluctuations in equipment costs, unplanned labor incentives) Costs not included: • EPC risk premium • Transmission interconnection-cost of connecting to grid beyond plant busbar • Taxes on capital costs • Unusual site improvements	15% of TPC

3.4 Economic Analysis Methods and Background

3.4.1 Economic Analysis Metrics

The economic analysis uses the capital and O&M cost estimates along with global economic assumptions to determine the following economic metrics to compare the technologies:

- First-year required selling price breakdown including:
    - Capital
    - Fuel
    - Variable O&M
    - Fixed O&M
    - TS&M
- Cost of CO_2 captured

Methanol Required Selling Price (RSP)

The methanol RSP ($/gal) is calculated using the following equation adapted from the DOE/NETL Baseline reports.

$$RSP = \frac{\begin{matrix}\text{first year} \\ \text{capital charge}\end{matrix} + \begin{matrix}\text{first year} \\ \text{fixed operating} \\ \text{costs}\end{matrix} + \begin{matrix}\text{first year} \\ \text{variable operating} \\ \text{costs}\end{matrix}}{\text{Annual Net Methanol Production}}$$

$$RSP = \frac{(CCF)(TOC) + OC_{FIX} + (CF)(OC_{VAR})}{(CF)(PR)}$$

where:

RSP = Methanol Required Selling Price, revenue received by the producer ($/gal) during the power plant's first year of operation (expressed in base-year dollars) assuming that the COMP escalates thereafter at a nominal annual rate equal to the general inflation rate

CCF = capital charge factor based on financial structure, set to 0.237 as per Attachment 2. This factor takes into account the financial structure and construction period to distribute the costs of the plant operational life (unitless)

TOC = total overnight capital costs, expressed in base-year dollars ($)

OC_{FIX} = the sum of all fixed annual operating costs, expressed in base-year dollars ($)

OC_{VAR} = the sum of all variable operating costs (fuel and variable O&M costs), expressed in base-year dollars ($/gal)

CF = Capacity factor (unit-less)

PR = Total production from facility operating for 1 year, 8760 hours (gal).

Cost of CO_2 Captured

Cost of CO_2 captured ($/tonne CO_2) is calculated using the following equation:

$$CO_2\ Capture\ Cost = \frac{COE_{Capture} - COE_{No\,Capture}}{CO_2\ Captured_{Per\ Net\ Output}}$$

where:

CO_2 Captured$_{\text{Per Net Output}}$ = amount of CO_2 captured per unit of production (tonne CO_2/gal)

3.4.2 Assumption in DOE/NETL Methanol Baseline Report

The Methanol Baseline Study calculated the first year RSP for the methanol product on a $/gal basis. The first year RSP estimates the first year selling price required to match the return on equity to the internal rate of return for operations of 30 years and the assumed financial structure and escalation of the methanol selling price.

In the Methanol Baseline Study determined the RSP using financial structures representative of a commercial fuels project and one with loan guarantees or other government subsidies. The financial assumptions and structures used to estimate the RSPs are shown in Exhibit 3-6, Exhibit 3-7, and Exhibit 3-8. [4]

Exhibit 3-6 Financial assumptions for the DOE/NETL Methanol Baseline Study[5]

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	30 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%[a]
Distribution of Total Overnight Capital over the Capital Expenditure Period (before escalation)	5-Year Period: 10%, 30%, 25%, 20%, 15%
Working Capital	zero for all parameters
% of Total Overnight Capital that is Depreciated	100% *(this assumption introduces a very small error even if a substantial amount of TOC is actually non-depreciable)*
INFLATION	
RSP, O&M, Fuel Escalation (nominal annual rate)	3.0%[b] RSP, O&M, COE, Fuel

[a] A nominal average annual rate of 3.6 percent is assumed for escalation of capital costs during construction. This rate is equivalent to the nominal average annual escalation rate for process plant construction costs between 1947 and 2008, according to the Chemical Engineering Plant Cost Index.

[b] An average annual inflation rate of 3.0 percent is assumed. This rate is equivalent to the average annual escalation rate between 1947 and 2008 for the U.S. Department of Labor's Producer Price Index for Finished Goods, the so-called "headline" index of the various Producer Price Indices. (The Producer Price Index for the Electric Power Generation Industry may be more applicable, but that data does not provide a long-term historical perspective since it only dates back to December 2003.)

Exhibit 3-7 Financial structure for commercial fuels projects [6]

Type of Security	Percent of Total	Current (Nominal) Dollar Cost
Debt	50	8% (LIBOR=3.5% + 4.5%)
Equity	50	20%

Exhibit 3-8 Financial structure for loan guarantee projects [7]

Type of Security	Percent of Total	Current (Nominal) Dollar Cost
Debt	60	4.56% (CMT=4.34% + 0.22%)
Equity	40	20%

The capital charge factor (CCF), which is the portion of the total overnight capital costs to include in the annual production costs, is determined by the financial structures defined in the above exhibits. For the commercial fuel project financial structure, the CFF is 21.8% while for the projects with a loan guarantee financial structure, the CCF is 17.0%. These values are different than the value specified to be used in the current study, 23.7%, in Attachment 2 to the RFP. Based on the guidance provided in Attachment 2, 23.7% is used as the reference CCF for the current study. A summary of the cost estimating results and impact of the change in CCF is provided in Exhibit 3-9. Additionally, two other clarifications are:

1. There is a small discrepancy between the Owner's costs in the two studies, as illustrated in this exhibit. Details regarding the buildup of the Owner's costs are not provided in the Crude Methanol Baseline Study which resulted in difficulties in identifying the differences between the values. The revised analyses include the Owner's costs calculated according to the assumptions in Exhibit 3-5.
2. Both plants are sited in the Midwestern USA. The Methanol Baseline Study used $11/tonne for TS&M costs according to the National Energy Technology Laboratory (NETL). (2013). QGESS: Carbon Dioxide Transport and Storage Costs in NETL Studies.[8] From Attachment 2, the TS&M costs for a Midwest location are $10/tonne. To maintain consistency with the Crude Methanol Baseline Study, $11/tonne was used as the TS&M costs.

Exhibit 3-9 Comparison of Economic Analyses in MBL Report and Methodology Used in This Study

	As Reported in DOE/NETL MBL Report		Methodology Used in This Study	
DOE Crude Methanol Case	**MLB-1**	**MLB-2**	**MLB-1**	**MLB-2**
Total Plant Cost (2011 MM$)	4,586	4,775	4,586	4,775
Owner's Costs (2011 MM$)	1,029	1,069	987	1,027
Total Overnight Cost (2011 MM$)	5,615	5,844	5,572	5,802
RSP[A] Component Details ($/gal)				
Capital[B]	0.85 1.09	0.89 1.14	1.18	1.23
Fixed O&M	0.14	0.14	0.14	0.14
Variable O&M	0.09	0.09	0.09	0.09
Coal	0.21	0.21	0.21	0.21
Natural gas	0.03	0.05	0.03	0.05
CO_2 T&S	0	0.06	0.00	0.06
RSP[B] Total ($/gal)	**1.31 1.56**	**1.43 1.68**	**1.65**	**1.78**
RSP[E] Total ($/ton)	**396.70 469.29**	**432.81 508.37**	**496.50**	**537.23**
Costs of CO_2 captured[BC] ($/tonne)	**N/A**	**12.89 14.85**	**N/A**	**15.34**

[A] *Capacity factor assumed to be 90 percent.*

[B] *Values are shown for two financial structures.*
The first (lower) value is based on the loan guarantee finance structure.
The second (higher) value is based on the commercial fuels finance structure.

[C] *Excluded CO_2 T&S.*

[D] *Includes CO_2 T&S.*

[E] *Based on 332.6 gal/tonne or 301.73 gal/ton.*

4. OSU CLG DESIGN CASES

Two design cases are evaluated in this TEA for methanol production with the OSU CLG Technology:

Case OSU-1	Methanol production with coal as the feedstock to the CLG reactor. This case includes CO_2 capture and purchased electricity.
Case OSU-2	Methanol production with coal and natural gas as the feedstock to the CLG reactor. Similar to Case OSU-1, this case includes CO_2 capture and purchased electricity.

4.1 Process Description

A block flow diagram providing an overview of the OSU CLG based coal to methanol production configurations developed during the current study is provided in Exhibit 4-1. For Case OSU-1, the feedstock to the process is coal only, while for Case OSU-2, the feedstock is a combination of coal and natural gas (50%/50% on an HHV basis). In both cases CO_2 capture, conditioning, and compression for pipeline transportation are considered.

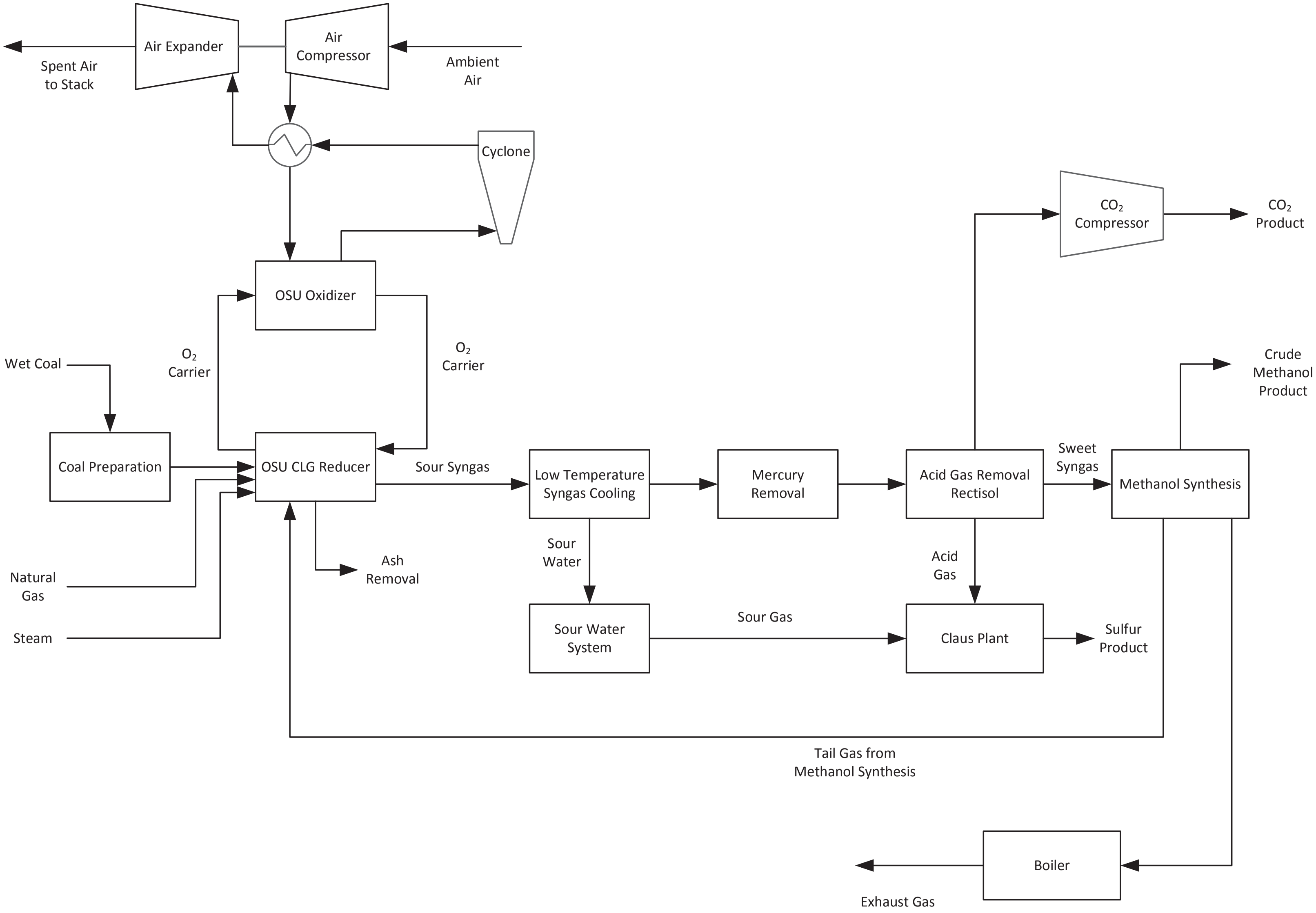
From a high level, coal to methanol process can be considered in two steps:

1. The conversion of the feedstock, coal and natural gas, to a syngas with a composition suitable for syngas production, and
2. The conversion of the syngas to the methanol product.

Aside from replacing the air separator unit and gasifiers with the OSU CLG technology, the other major changes from the Methanol Baseline Report are the elimination of the water shift reactors and power generation. For the OSU CLG cases, power for operating the equipment is assumed to be imported.

For the coal to methanol process illustrated in Exhibit 4-1, syngas is generated from the gasification of PRB coal in the OSU CLG reducer. The gasification occurs through the partial oxidation of the coal by oxygen from the oxygen carrier in the OSU CLG Reducer. In this process, the iron oxides are reduced to lower oxidation state. Additionally, steam is added to the reactor to enhance in the production of H_2. The reduced oxygen carrier is sent back to the OSU Oxidizer where it is oxidized with air. The syngas is cooled and then sent to acid gas scrubbers to remove the H_2S and 90% of the CO_2 a sweet syngas. Sweet syngas is used to produce methanol. Flash gas from the methanol process is recycled back to the reducer reactor and the purge gas is combusted in a boiler to produce steam. The following subsections provide details and process flow diagrams for the blocks in the system.

Exhibit 4-1 OSU Block Flow Diagram for Cases OSU CLG - 1 and Cases OSU CLG - 2



4.1.1 CLG Reactors

For syngas production, CLG technology replaces air separation unit, gasifiers, and water gas shift reactors in the reference coal to methanol process, see Exhibit 1-1. A process flow diagram of the equipment for OSU-1 and OSU-2, as modeled in Aspen, is provided in Exhibit 4-2. In this diagram, the syngas is produced through the partial oxidation of the fuel with steam and oxygen from the OC. The fuel is as received coal in case OSU-1 and as received coal and natural gas in OSU-2. The steam is added to the reactor to aid in the production of H_2. The heat for this reaction is supplied by the OC and the complete oxidation of a portion of the fuel to CO_2. In this process, the iron oxides are reduced to lower oxidation state. The extent of the oxidation is controlled through the OC to coal ratio to the reducer reactor. The syngas from this reactor is fed to the syngas cooling and conditioning block while the OC is set to the oxidizer.

Prior to the oxidizer, the OC passes through a particulate classifier to remove the fines which include the coal ash and OC attrition products. In the oxidizer the OC is reacted with oxygen in the compressed air to re-oxidize the OC. This process is exothermic which increase the temperature of the air and the OC. The hot air from the oxidizer is used to preheat the air coming into the system to maintain the temperature in the oxidizer. Further the compressed air leaving the system is passed through a particulate control device to remove any fines and an expander to recover a portion of the energy used to compress the air. This expander is used to drive the compressor. The remainder of the energy required for the compression process is provided by an electric motor.

4.1.2 Low temperature Syngas Cooling and Conditioning

As part of the syngas conditioning for the methanol synthesis, the syngas temperature from the reducer reactor is reduced in a syngas cooler and compressed. In the proposed a heat recovery unit is included to produce superheated steam. The resulting cooled gas is conditioned with a water scrubber and compressed to decrease downstream equipment sizes, enhance the CO_2 capture process, and meet the high gas pressures for methanol production. A common process flow diagram for the OSU-1 and OSU-2 options is provided in Exhibit 4-5. Syngas cooling and scrubbing mass balances for these options are shown in Exhibit 4-6 and Exhibit 4-7.

The additional compression step for the OSU options is a result of the lower operating pressures of the OSU CLG equipment with respect to the gasifier based coal to methanol processes. Operating at these low pressures would result in very large flue gas cleaning equipment (AGR) equipment. An additional consideration is that a syngas pressure of 700 psia is required for the methanol synthesis reactor. Therefore, a syngas compressor is added after the syngas cooler and prior to the mercury removal system. In the study systems, the pressure is increased from 140 psia to 500 psia. In addition to reducing the equipment size, the pressure increase also improves the efficiency of the Rectisol acid gas removal system.

4.1.3 Acid Gas Removal, Sulfur Recovery Unit, and CO_2 Compression

Removal of H_2S and CO_2 are required to produce a syngas composition for methanol synthesis. If introduced into the methanol synthesis process, the sulfur, as H_2S, would poison the catalyst and/or become an impurity in the final product. An acid gas removal system is typically added to clean and modify the syngas composition as required.

Based on the emphasis to maintain design continuity between the Methanol Baseline study and the current study, the Rectisol AGR technology was selected. The benefits of the Rectisol technology to the syngas conditioning are similar for both designs, specifically the ability to reduce H_2S below 100 ppbv in order to maintain an adequate catalyst lifetime through preventing catalyst poisoning.

A common process flow diagram for the Acid Gas Removal, Sulfur Recovery Unit, and CO_2 Compression systems for the OSU-1 and OSU-2 options are provided in Exhibit 4-8. Acid Gas Removal, Sulfur Recovery Unit, and CO_2 Compression mass balances for the OSU-1 and OSU-2 are provided in Exhibit 4-9 and Exhibit 4-10.

The Rectisol process is based on chilled methanol physical adsorption solvent. For physical adsorption, the solvent loading is enhanced through lower temperatures (-30 to -100°F) and increased pressure. In the configuration for this study, the methanol solvent from the absorber is stripped in two stages of flashing via pressure reduction. The acid gas leaving the first stage solvent regenerator is suitable for processing in a Claus plant. A detailed description of the Claus unit can be found in the Crude Methanol baseline report. The regenerated solvent from the first stage is virtually free of sulfur compounds but contains some CO_2. The second stage of absorption then removes the remaining CO_2 present. The rich solvent from the bottom of the second stage of the absorber is stripped in a steam-heated regenerator and returned to the top of the absorption column after cooling and refrigeration.

To provide CO_2 capture, the CO_2 stream from the Rectisol unit is compressed to 2,200 psig in a multiple-stage, intercooled compressor to supercritical conditions for pipeline transportation to the storage site.

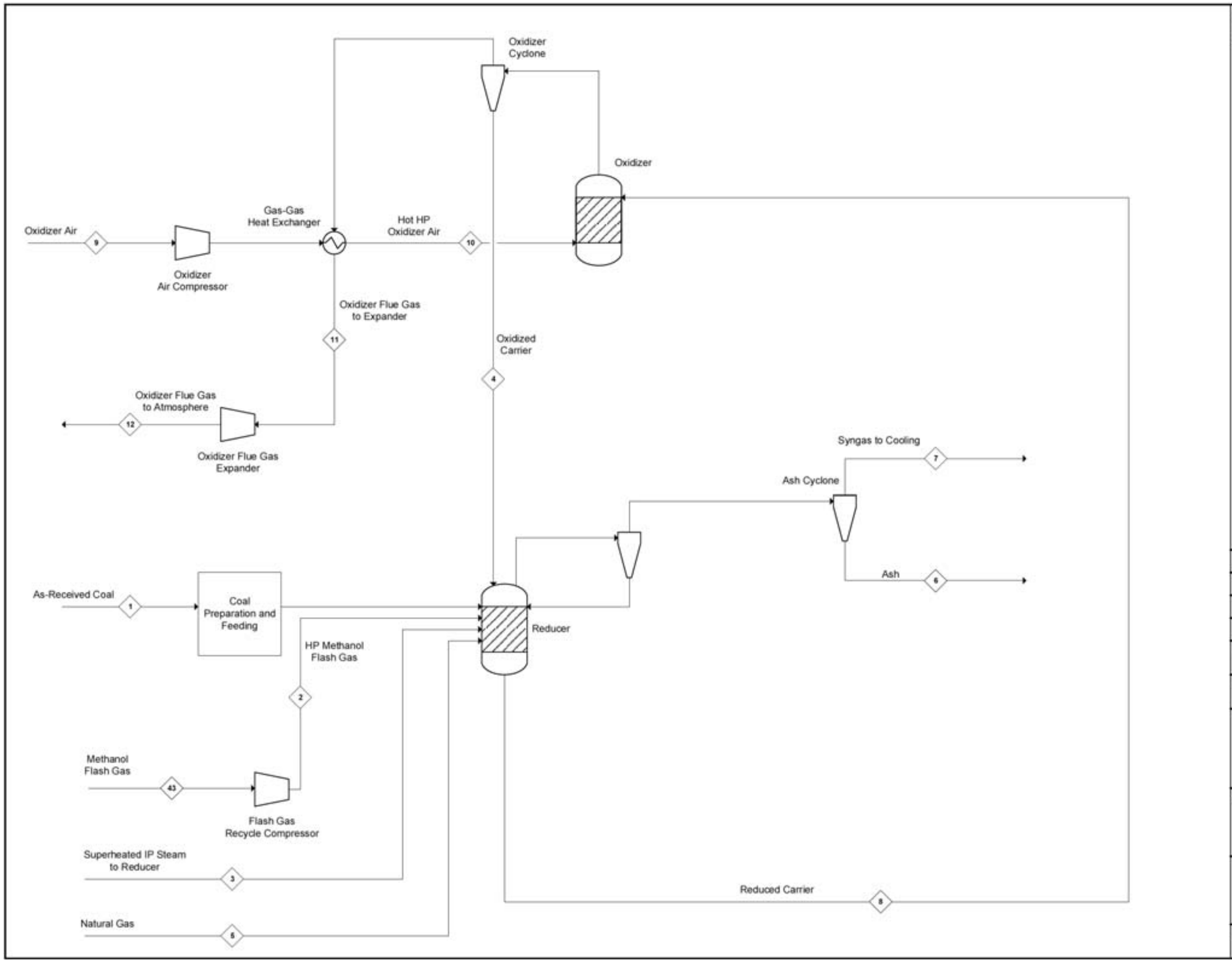
4.1.4 Methanol Synthesis

There are a variety of methanol synthesis processes that can be selected for the proposed facility. These processes can be split between vapor phase and liquid phase processes. To remain consistent with the Methanol Baseline report, the vapor-phase methanol process was selected. This selection is made so that the differences between the Methanol Baseline study and the current study are limited to changes in the gasification and air separation equipment. The methanol plant process flow diagrams for the OSU-1 and OSU-2 are provided in Exhibit 4-11. The corresponding mass diagrams are provided in Exhibit 4-12 and Exhibit 4-13.

The methanol reactor converts hydrogen and carbon monoxide to methanol in a catalytic packed-bed reactor. The primary side reactions produce ethanol, propanol, and formaldehyde. Acetone and acetaldehyde are also common impurities in the methanol product and are captured in this analysis.

In the process developed, CO_2-lean syngas with a H_2/CO ratio of 1.7:1 from the AGR process is compressed from 500 psia to the synthesis loop operating pressure of 755 psia in the syngas compressor. The compressed syngas is mixed with the recycled gas, heated to 400ºF, and routed to the methanol reactor. The reactor is steam cooled to facilitate near isothermal operation at 475ºF and 735 psia. In-line blowers, coolers, and knock-out drums are used within the synthesis loop to maintain pressure and remove crude methanol. To limit CO_2 emissions, the flash gas from the process is recycled back into the reducing reactor. The purge gas, which contains inserts such as N_2, is sent to a boiler where it is combusted to generated steam.

Exhibit 4-2 Process Flow Diagram for Chemical Looping System (OSU-1 and OSU-2)



NOTES:

1. Coal transport and drying methods to be determined later. Coal is processed as-received.

REV	DATE	DESCRIPTION
A	12-31-14	INITIAL ISSUE

ENGINEER/TECH SPECIALIST: E. Mace
DATE: December 31, 2014
CHECKED BY: J. Simpson
PROJECT MANAGER: J. Simpson

WorleyParsons
resources & energy

CLIENT/PROJECT TITLE
The Ohio State Univerisity
Chemical Looping Gasification to
Methanol

PROCESS FLOW DIAGRAM
Chemical Looping System
(Case 1 and Case 2)

WORLEYPARSONS DWG. NO.
REV A

Exhibit 4-3 Mass Balance for Chemical Looping System – Option OSU-1

STREAM	1	2	3	4	5	6	7	8	9	10	11	12
Description	As-Received Coal	HP Methanol Flash Gas	Superheated IP Steam to Reducer	Oxidized Carrier	Natural Gas	Ash	Syngas to Cooling	Reduced Carrier	Oxidizer Air	Hot HP Oxidizer Air	Oxidizer Flue Gas to Expander	Oxidizer Flue Gas to Atmosphere
VAPOR / LIQUID												
Mole Flow, kmole/hr												
Ar	0.00	0.34	0.00	0.00	0.00	0.00	0.34	0.00	540.51	540.51	540.51	540.51
CH_4	0.00	18.90	0.00	0.00	0.00	0.00	149.56	0.00	0.00	0.00	0.00	0.00
C_2H_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	7.13	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.00	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.00	1.15	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.00	3.18	0.00	0.00	0.00	0.00	11,430.88	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	387.91	0.00	0.00	0.00	0.00	15,222.66	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00	2.04	0.00	0.00	0.00	0.00	0.00
H_2	0.00	18.20	0.00	0.00	0.00	0.00	24,465.77	0.00	0.00	0.00	0.00	0.00
H_2O	0.00	9.60	40,897.84	0.00	0.00	0.00	35,754.23	0.00	574.07	317.23	317.23	317.23
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	141.56	0.00	0.00	0.00	0.00	0.00
HCl	0.00	0.00	0.00	0.00	0.00	0.00	1.33	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	0.00	13.41	0.00	0.00	0.00	0.00	173.56	0.00	43,981.30	43,981.30	43,978.19	43,978.19
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	1.72	0.00	0.00	0.00	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	6.17	6.17
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.04	0.04
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	11,799.54	11,799.54	558.75	558.75
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	0.00	459.84	40,897.84	0.00	0.00	0.00	87,343.65	0.00	56,895.42	56,638.57	45,400.89	45,400.89
Mass Flow, kg/hr												
Ar	0.00	13.64	0.00	0.00	0.00	0.00	13.64	0.00	21,592.15	21,592.15	21,592.15	21,592.15
CH_4	0.00	303.27	0.00	0.00	0.00	0.00	2,399.40	0.00	0.00	0.00	0.00	0.00
C_2H_6	0.00	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	228.51	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.00	0.36	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.00	53.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.00	89.00	0.00	0.00	0.00	0.00	320,184.00	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	17,071.95	0.00	0.00	0.00	0.00	669,946.00	0.00	0.00	0.00	0.01	0.01
COS	0.00	0.00	0.00	0.00	0.00	0.00	122.31	0.00	0.00	0.00	0.00	0.00
H_2	0.00	36.69	0.00	0.00	0.00	0.00	49,320.05	0.00	0.00	0.00	0.00	0.00
H_2O	0.00	172.90	736,786.00	0.00	0.00	0.00	644,123.00	0.00	10,342.12	5,714.92	5,714.92	5,714.92
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	4,824.68	0.00	0.00	0.00	0.00	0.00
HCl	0.00	0.00	0.00	0.00	0.00	0.00	48.32	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.10	0.00	0.00	0.00	0.00	0.00
N_2	0.00	375.56	0.00	0.00	0.00	0.00	4,862.01	0.00	1,232,070.00	1,232,070.00	1,231,980.00	1,231,980.00
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	29.34	0.00	0.00	0.00	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	185.18	185.18
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	2.03	2.03
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	377,571.00	377,571.00	17,879.42	17,879.42
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00
Total	0.00	18,344.93	736,786.00	0.00	0.00	0.00	1,695,872.88	0.00	1,641,575.27	1,636,948.07	1,277,353.71	1,277,353.71
SOLIDS												
Mass Flow, kg/hr												
Coal	632,975.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ash	0.00	0.00	0.00	0.00	0.00	51,825.25	0.00	0.00	0.00	0.00	0.00	0.00
C	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Fe_2O_3	0.00	0.00	0.00	3,589,130.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Fe_3O_4	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FeO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FeS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FeS_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Fe	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
$FeAl_2O_4$	0.00	0.00	0.00	0.00	0.00	0.00	0.00	7,812,770.00	0.00	0.00	0.00	0.00
Al_2O_3	0.00	0.00	0.00	9,459,590.00	0.00	0.00	0.00	4,876,370.00	0.00	0.00	0.00	0.00
Total	632,975.00	0.00	0.00	13,048,720.00	0.00	51,825.25	0.00	12,689,140.00	0.00	0.00	0.00	0.00
ALL PHASES												
Total Mole Flow, kmole/h	0.00	459.84	40,897.84	115,252.00	0.00	0.00	87,343.65	92,776.28	56,895.42	56,638.57	45,400.89	45,400.89
Total Mass Flow, kg/h	632,975.00	18,344.94	736,786.00	13,048,700.00	0.00	51,825.25	1,695,870.00	12,689,100.00	1,641,580.00	1,636,950.00	1,277,360.00	1,277,360.00
Temperature, C	15.00	107.16	250.00	1,050.00	37.78	841.00	841.00	841.00	15.00	800.00	206.59	94.90
Pressure, bar	1.01	10.67	10.00	10.00	31.03	10.00	10.00	10.00	1.01	10.00	9.31	3.01
Vapor Frac	0.00	1.00	1.00	0.00	1.00	0.00	1.00	0.00	1.00	1.00	1.00	1.00
Density, kg/m^3	1,411.38	13.79	4.29	4,273.83	22.11	3,486.88	2.09	4,934.22	1.22	3.23	6.54	2.77
Average MW	1.00	39.89	18.02	113.22	17.33	1.00	19.42	136.77	28.85	28.90	28.14	28.14

Exhibit 4-4 Mass Balance for Chemical Looping System – Option OSU-2

STREAM	1	2	3	4	5	6	7	8	9	10	11	12
Description	As-Received Coal	HP Methanol Flash Gas	Superheated IP Steam to Reducer	Oxidized Carrier	Natural Gas	Ash	Syngas to Cooling	Reduced Carrier	Oxidizer Air	Hot HP Oxidizer Air	Oxidizer Flue Gas to Expander	Oxidizer Flue Gas to Atmosphere
VAPOR / LIQUID												
Mole Flow, kmole/hr												
Ar	0.00	0.16	0.00	0.00	0.00	0.00	0.16	0.00	463.61	463.61	463.61	463.61
CH_4	0.00	88.10	0.00	0.00	6,636.11	0.00	748.36	0.00	0.00	0.00	0.00	0.00
C_2H_6	0.00	0.00	0.00	0.00	228.09	0.00	0.01	0.00	0.00	0.00	0.00	0.00
C_3H_8	0.00	0.00	0.00	0.00	49.90	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	28.51	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	5.53	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.00	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.00	0.91	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.00	15.27	0.00	0.00	0.00	0.00	12,922.25	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	242.04	0.00	0.00	71.28	0.00	7,696.37	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00	1.18	0.00	0.00	0.00	0.00	0.00
H_2	0.00	8.92	0.00	0.00	0.00	0.00	24,516.89	0.00	0.00	0.00	0.00	0.00
H_2O	0.00	8.53	17,463.47	0.00	0.00	0.00	16,015.07	0.00	492.41	272.10	272.10	272.10
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	72.77	0.00	0.00	0.00	0.00	0.00
HCl	0.00	0.00	0.00	0.00	0.00	0.00	0.68	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00
N_2	0.00	10.14	0.00	0.00	114.05	0.00	205.78	0.00	37,724.57	37,724.57	37,721.91	37,721.91
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	2.64	0.00	0.00	0.00	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	5.29	5.29
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.04	0.04
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	10,120.95	10,120.95	479.27	479.27
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	0.00	379.63	17,463.47	0.00	7,127.94	0.00	62,182.18	0.00	48,801.54	48,581.23	38,942.22	38,942.22
Mass Flow, kg/hr												
Ar	0.00	6.51	0.00	0.00	0.00	0.00	6.51	0.00	18,520.48	18,520.48	18,520.48	18,520.48
CH_4	0.00	1,413.33	0.00	0.00	106,462.00	0.00	12,005.69	0.00	0.00	0.00	0.00	0.00
C_2H_6	0.00	0.09	0.00	0.00	6,858.70	0.00	0.36	0.00	0.00	0.00	0.00	0.00
C_3H_8	0.00	0.00	0.00	0.00	2,200.22	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	1,657.20	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	177.15	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.00	0.62	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.00	42.14	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.00	427.81	0.00	0.00	0.00	0.00	361,957.00	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	10,652.15	0.00	0.00	3,136.99	0.00	338,716.00	0.00	0.00	0.00	0.01	0.01
COS	0.00	0.00	0.00	0.00	0.00	0.00	70.91	0.00	0.00	0.00	0.00	0.00
H_2	0.00	17.98	0.00	0.00	0.00	0.00	49,423.11	0.00	0.00	0.00	0.00	0.00
H_2O	0.00	153.65	314,609.00	0.00	0.00	0.00	288,516.00	0.00	8,870.86	4,901.92	4,901.92	4,901.92
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	2,480.12	0.00	0.00	0.00	0.00	0.00
HCl	0.00	0.00	0.00	0.00	0.00	0.00	24.88	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.38	0.00	0.00	0.00	0.00	0.00
N_2	0.00	284.04	0.00	0.00	3,194.85	0.00	5,764.59	0.00	1,056,800.00	1,056,800.00	1,056,720.00	1,056,720.00
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	44.98	0.00	0.00	0.00	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	158.84	158.84
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.74	1.74
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	323,858.00	323,858.00	15,335.91	15,335.91
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	0.00	13,175.50	314,609.00	0.00	123,509.97	0.00	1,059,010.53	0.00	1,408,049.34	1,404,080.40	1,095,638.89	1,095,638.89
SOLIDS												
Mass Flow, kg/hr												
Coal	325,969.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ash	0.00	0.00	0.00	0.00	0.00	26,688.92	0.00	0.00	0.00	0.00	0.00	0.00
C	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Fe_2O_3	0.00	0.00	0.00	3,078,550.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Fe_3O_4	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FeO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FeS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FeS_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Fe	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
$FeAl_2O_4$	0.00	0.00	0.00	0.00	0.00	0.00	0.00	6,701,330.00	0.00	0.00	0.00	0.00
Al_2O_3	0.00	0.00	0.00	8,113,880.00	0.00	0.00	0.00	4,182,660.00	0.00	0.00	0.00	0.00
Total	325,969.00	0.00	0.00	11,192,430.00	0.00	26,688.92	0.00	10,883,990.00	0.00	0.00	0.00	0.00
ALL PHASES												
Total Mole Flow, kmole/h	0.00	379.63	17,463.47	98,856.05	7,127.94	0.00	62,182.18	79,578.05	48,801.55	48,581.24	38,942.22	38,942.22
Total Mass Flow, kg/h	325,969.00	13,175.51	314,609.00	11,192,400.00	123,510.00	26,688.92	1,059,010.00	10,884,000.00	1,408,050.00	1,404,080.00	1,095,640.00	1,095,640.00
Temperature, C	15.00	109.29	250.00	1,050.00	37.78	841.00	841.00	841.00	15.00	800.00	206.59	94.90
Pressure, bar	1.01	10.67	10.00	10.00	31.03	10.00	10.00	10.00	1.01	10.00	9.31	3.01
Vapor Frac	0.00	1.00	1.00	0.00	1.00	0.00	1.00	0.00	1.00	1.00	1.00	1.00
Density, kg/m^3	1,411.38	11.86	4.29	4,273.83	22.11	3,486.88	1.84	4,934.22	1.22	3.23	6.54	2.77
Average MW	1.00	34.71	18.02	113.22	17.33	1.00	17.03	136.77	28.85	28.90	28.14	28.14

Exhibit 4-5 Process Flow Diagram for Syngas Cooling and Scrubbing (OSU-1 and OSU-2)

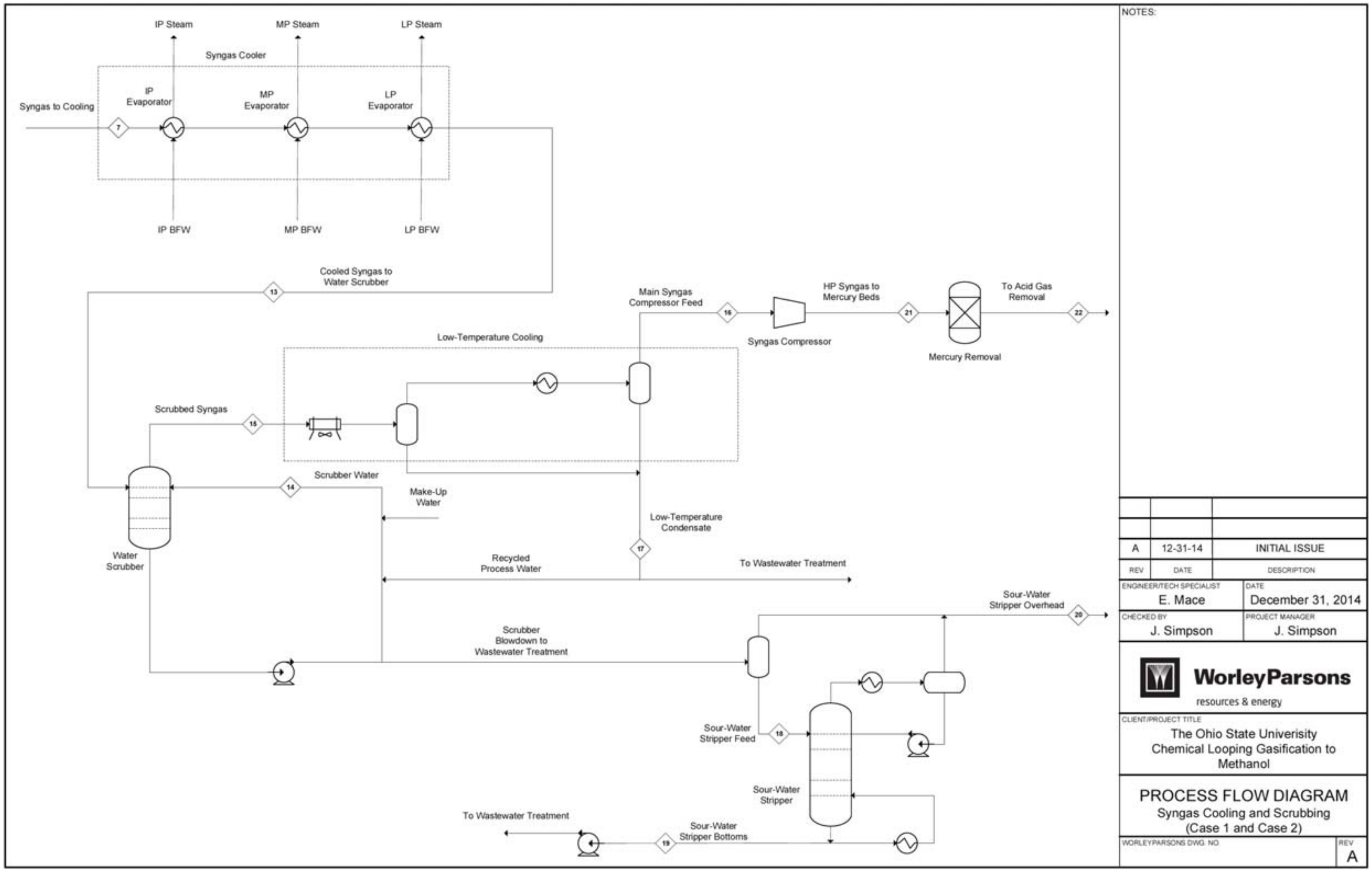


Exhibit 4-6 Mass Balance for Syngas Cooling and Scrubbing –Option OSU-1.

STREAM	13	14	15	16	17	18	19	20	21	22
Description	Cooled Syngas to Water Scrubber	Scrubber Water	Scrubbed Syngas	Main Syngas Compressor Feed	Low-Temperature Condensate	Sour-Water Stripper Feed	Sour-Water Stripper Bottoms	Sour-Water Stripper Overhead	HP Syngas to Mercury Beds	To Acid Gas Removal
VAPOR / LIQUID										
Mole Flow, kmole/hr										
Ar	0.34	0.00	0.34	0.34	0.00	0.00	0.00	0.00	0.34	0.34
CH_4	149.56	0.00	149.56	149.55	0.01	0.00	0.00	0.00	149.55	149.55
C_2H_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	11,430.88	0.00	11,430.64	11,429.95	0.68	0.00	0.00	0.24	11,429.95	11,429.95
CO_2	15,222.66	0.00	15,218.16	15,195.16	22.80	0.11	0.00	4.50	15,195.14	15,195.14
COS	2.04	0.00	2.04	2.03	0.00	0.00	0.00	0.00	2.03	2.03
H_2	24,465.77	0.00	24,465.07	24,463.54	1.53	0.00	0.00	0.70	24,463.54	24,463.54
H_2O	35,754.23	20,928.90	35,589.03	587.80	35,000.99	20,690.92	20,999.49	94.61	123.05	123.05
H_2S	141.56	0.00	141.41	140.73	0.68	0.01	0.00	0.15	140.72	140.72
HCl	1.33	0.00	0.00	0.00	0.00	1.33	1.33	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	173.56	0.00	173.56	173.55	0.01	0.00	0.00	0.00	173.55	173.55
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	1.72	0.00	0.18	0.00	0.00	1.51	1.35	0.20	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	87,343.65	20,928.90	87,169.99	52,142.65	35,027.13	20,693.88	21,002.17	100.40	51,677.88	51,677.88
Mass Flow, kg/hr										
Ar	13.64	0.00	13.64	13.64	0.00	0.00	0.00	0.00	13.64	13.64
CH_4	2,399.40	0.00	2,399.35	2,399.16	0.19	0.00	0.00	0.06	2,399.16	2,399.16
C_2H_6	0.01	0.00	0.01	0.01	0.00	0.00	0.00	0.00	0.01	0.01
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	320,184.00	0.00	320,177.00	320,158.00	19.11	0.01	0.00	6.85	320,158.00	320,158.00
CO_2	669,946.00	0.00	669,748.00	668,736.00	1,003.23	4.96	0.00	197.84	668,735.00	668,735.00
COS	122.31	0.00	122.31	122.22	0.09	0.00	0.00	0.00	122.22	122.22
H_2	49,320.05	0.00	49,318.65	49,315.57	3.08	0.00	0.00	1.40	49,315.57	49,315.57
H_2O	644,123.00	377,040.00	641,146.00	10,589.33	630,553.00	372,753.00	378,312.00	1,704.51	2,216.81	2,216.81
H_2S	4,824.68	0.00	4,819.66	4,796.36	23.21	0.44	0.00	5.03	4,796.16	4,796.16
HCl	48.32	0.00	0.00	0.00	0.00	48.32	48.32	0.00	0.00	0.00
HCN	0.10	0.00	0.10	0.07	0.02	0.00	0.00	0.00	0.07	0.07
N_2	4,862.01	0.00	4,861.93	4,861.73	0.21	0.00	0.00	0.08	4,861.73	4,861.73
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	29.34	0.00	3.02	0.00	0.00	25.74	22.99	3.34	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.01	0.00	0.01	0.01	0.00	0.00	0.00	0.00	0.01	0.01
Total	1,695,872.88	377,040.00	1,692,609.66	1,060,992.10	631,619.11	372,832.48	378,383.31	1,919.12	1,052,618.38	1,052,618.38
ALL PHASES										
Total Mole Flow, kmole/h	87,343.65	20,928.90	87,169.99	52,142.66	35,027.12	20,693.88	21,002.16	100.40	51,677.89	51,677.89
Total Mass Flow, kg/h	1,695,870.00	377,040.00	1,692,610.00	1,060,990.00	631,619.00	372,832.00	378,383.00	1,919.12	1,052,620.00	1,052,620.00
Temperature, C	160.00	90.00	135.77	39.91	54.67	127.02	125.55	111.53	29.44	27.53
Pressure, bar	8.97	1.20	8.00	6.83	6.83	2.50	2.36	1.60	35.22	34.88
Vapor Frac	1.00	0.00	1.00	1.00	0.00	0.00	0.00	1.00	1.00	1.00
Density, kg/m^3	4.90	965.14	4.63	5.35	982.54	937.09	938.34	0.97	29.30	28.84
Average MW	19.42	18.02	19.42	20.35	18.03	18.02	18.02	19.11	20.37	20.37

Exhibit 4-7 Mass Balance for Syngas Cooling and Scrubbing –Option OSU-2.

STREAM	13	14	15	16	17	18	19	20	21	22
Description	Syngas Cooler Exit to Water Scrubber	Scrubber Water	Scrubbed Syngas	Syngas Exiting Knockout Section	Recovered Knockout Liquid	Sour-Water Stripper Feed	Sour-Water Stripper Bottoms	Sour-Water Stripper Overhead	Compressed Syngas to Mercury Beds	To Acid Gas Removal
VAPOR / LIQUID										
Mole Flow, kmole/hr										
Ar	1.70	0.00	0.16	0.16	0.00	0.00	0.00	0.00	0.16	0.16
CH_4	0.07	0.00	748.34	748.31	0.03	0.00	0.00	0.01	748.31	748.31
C_2H_6	0.01	0.00	0.01	0.01	0.00	0.00	0.00	0.00	0.01	0.01
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.81	0.00	12,922.03	12,921.62	0.41	0.00	0.00	0.22	12,921.62	12,921.62
CO_2	169.59	0.00	7,693.93	7,686.63	6.15	0.88	0.00	2.43	7,686.62	7,686.62
COS	1.18	0.00	1.18	1.18	0.00	0.00	0.00	0.00	1.18	1.18
H_2	0.90	0.00	24,516.35	24,515.55	0.81	0.01	0.00	0.54	24,515.55	24,515.55
H_2O	78.67	13,069.34	17,089.36	523.14	16,565.07	11,980.81	11,994.90	0.20	58.51	58.51
H_2S	0.39	0.00	72.68	72.48	0.19	0.06	0.00	0.09	72.48	72.48
HCl	0.00	0.00	0.00	0.00	0.00	0.68	0.68	0.00	0.00	0.00
HCN	0.00	0.00	0.01	0.01	0.00	0.00	0.00	0.00	0.01	0.01
N_2	138.35	0.00	205.78	205.77	0.00	0.00	0.00	0.00	205.77	205.77
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	1.16	0.00	0.00	1.47	1.45	0.02	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.25	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	0.57	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.19	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	392.69	13,069.34	63,251.00	46,674.87	16,574.99	11,983.93	11,997.03	3.51	46,210.23	46,210.23
Mass Flow, kg/hr										
Ar	67.91	0.00	6.51	6.51	0.00	0.00	0.00	0.00	6.51	6.51
CH_4	1.13	0.00	12,005.46	12,004.95	0.50	0.01	0.00	0.24	12,004.95	12,004.95
C_2H_6	0.19	0.00	0.35	0.32	0.03	0.01	0.00	0.01	0.32	0.32
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	22.70	0.00	361,951.00	361,940.00	11.45	0.12	0.00	6.15	361,940.00	361,940.00
CO_2	7,463.68	0.00	338,608.00	338,287.00	270.62	38.92	0.00	106.76	338,287.00	338,287.00
COS	71.15	0.00	70.91	70.88	0.03	0.00	0.00	0.00	70.88	70.88
H_2	1.81	0.00	49,422.03	49,420.40	1.63	0.03	0.00	1.08	49,420.40	49,420.40
H_2O	1,417.31	235,448.00	307,870.00	9,424.57	298,424.00	215,838.00	216,091.00	3.68	1,054.14	1,054.14
H_2S	13.25	0.00	2,477.20	2,470.34	6.38	2.00	0.00	2.92	2,470.20	2,470.20
HCl	0.00	0.00	0.00	0.00	0.00	24.88	24.88	0.00	0.00	0.00
HCN	0.00	0.00	0.37	0.32	0.05	0.01	0.00	0.01	0.32	0.32
N_2	3,875.56	0.00	5,764.51	5,764.38	0.13	0.00	0.00	0.08	5,764.38	5,764.38
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.08	0.00	19.79	0.00	0.02	25.12	24.67	0.36	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	48.27	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	146.73	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	12.39	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	13,142.16	235,448.00	1,078,196.13	779,389.67	298,806.26	215,929.10	216,140.56	121.29	771,019.10	771,019.10
ALL PHASES										
Total Mole Flow, kmole/h	392.69	13,069.34	63,251.00	46,674.87	16,574.99	11,983.93	11,997.03	3.51	46,210.22	46,210.22
Total Mass Flow, kg/h	13,142.16	235,448.00	1,078,200.00	779,390.00	298,807.00	215,929.00	216,141.00	121.29	771,018.00	771,018.00
Temperature, C	237.78	90.00	121.98	39.91	54.42	122.45	125.53	46.32	29.44	26.90
Pressure, bar	0.71	1.20	8.00	6.83	6.83	2.50	2.36	1.60	35.22	34.88
Vapor Frac	1.00	0.00	1.00	1.00	0.00	0.00	0.00	1.00	1.00	1.00
Density, kg/m^3	0.56	965.14	4.18	4.38	983.76	940.80	938.26	2.09	23.60	23.34
Average MW	33.47	18.02	17.05	16.70	18.03	18.02	18.02	34.54	16.69	16.69

Exhibit 4-8 Process Flow Diagram for Acid Gas Removal, Sulfur Recovery and CO_2 Compression (OSU-1 and OSU-2)

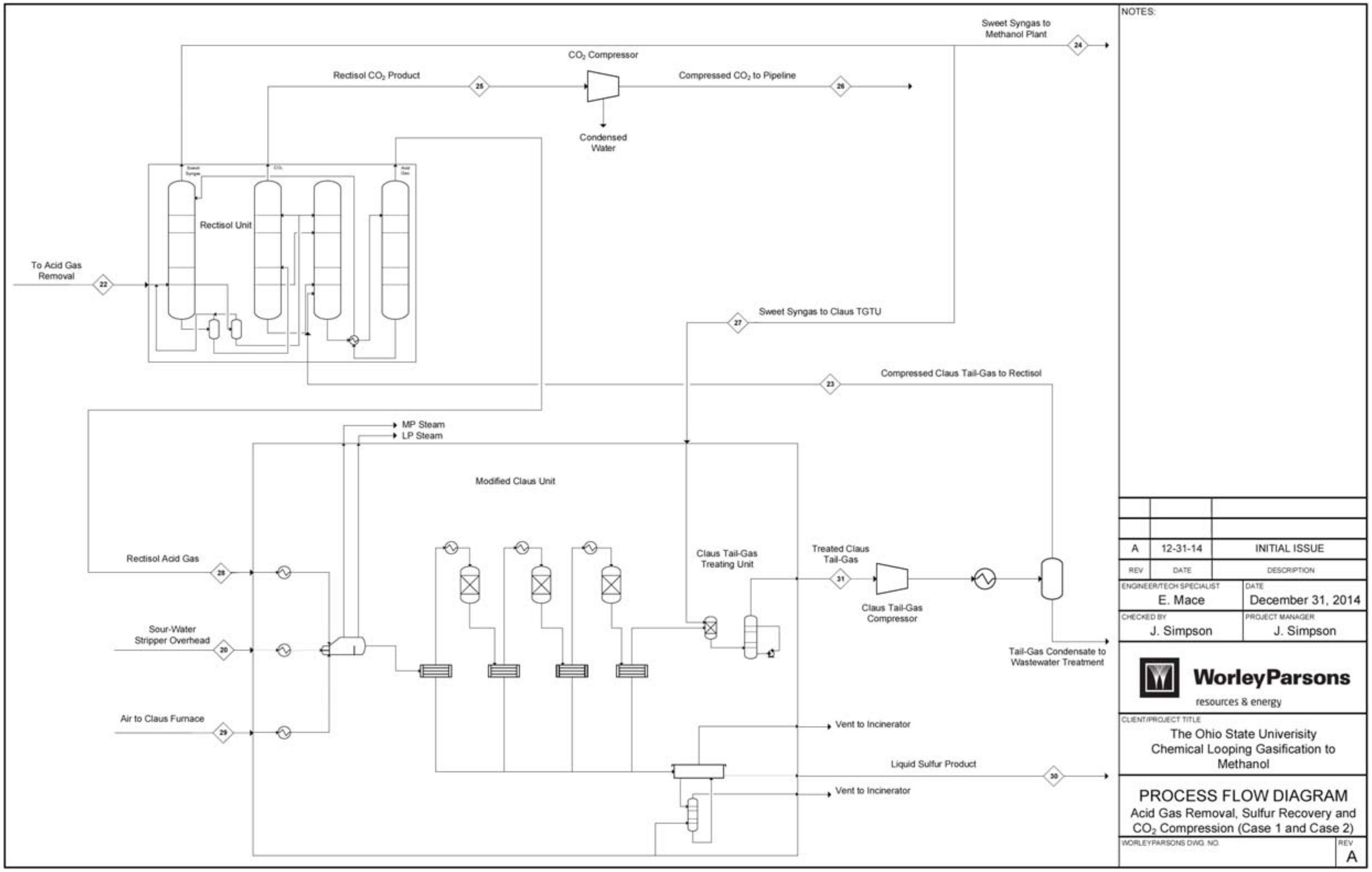


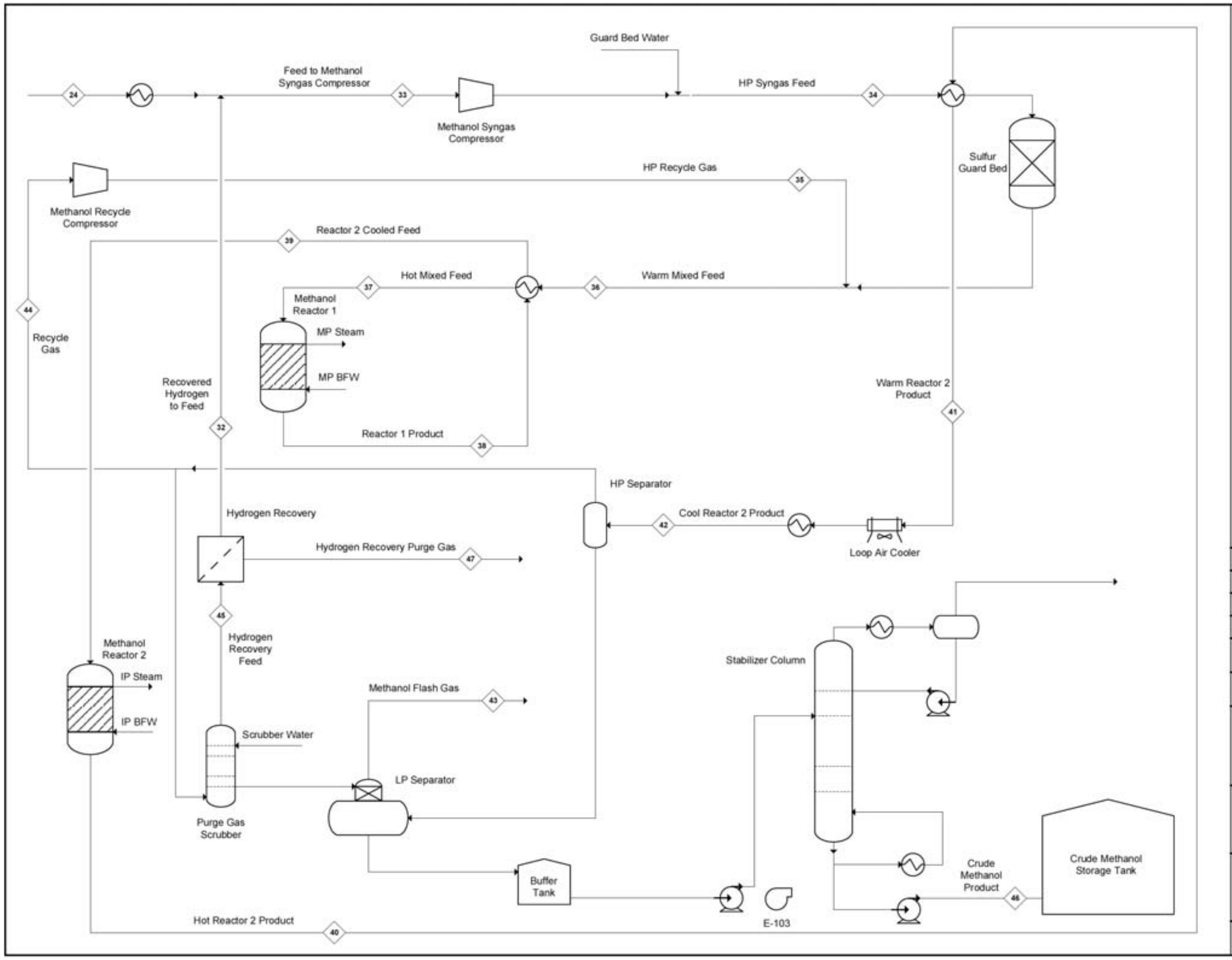
Exhibit 4-9 Mass Balance for Acid Gas Removal, Sulfur Recovery and CO_2 Compression – Option OSU-1

STREAM	23	24	25	26	27	28	29	30	31
Description	Compressed Claus Tail-Gas to Rectisol	Sweet Syngas to MeOH Plant	Rectisol CO_2 Product	Compressed CO_2 to Pipeline	Sweet Syngas to Claus TGTU	Rectisol Acid Gas	Air to Claus Furnace	Liquid Sulfur Product	Treated Claus Tail-Gas
VAPOR / LIQUID									
Mole Flow, kmole/hr									
Ar	3.30	3.64	0.00	0.00	0.00	0.00	3.30	0.00	3.30
CH_4	0.01	143.84	5.70	5.70	0.00	0.01	0.00	0.00	0.01
C_2H_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.07	11,334.15	95.12	95.12	0.24	0.52	0.00	0.00	0.07
CO_2	337.88	1,845.70	13,356.84	13,356.84	0.04	330.44	0.00	0.00	337.93
COS	0.01	0.00	0.00	0.00	0.00	2.04	0.00	0.00	0.01
H_2	1.77	24,435.75	28.70	28.70	0.51	0.35	0.00	0.00	1.77
H_2O	4.91	121.55	0.01	0.01	0.00	6.40	3.50	0.01	312.07
H_2S	3.07	0.00	0.00	0.00	0.00	143.80	0.00	0.00	3.08
HCl	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	268.38	441.92	0.00	0.00	0.01	0.00	268.27	0.00	268.38
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	71.97	0.00	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	6.69	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	12.30	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	619.41	38,326.56	13,486.37	13,486.37	0.81	483.56	347.04	19.01	926.63
Mass Flow, kg/hr									
Ar	131.71	145.34	0.00	0.00	0.00	0.00	131.70	0.01	131.71
CH_4	0.20	2,307.63	91.50	91.50	0.05	0.18	0.00	0.00	0.20
C_2H_6	0.00	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	2.08	317,474.00	2,664.28	2,664.28	6.68	14.64	0.00	0.00	2.08
CO_2	14,870.23	81,228.97	587,832.00	587,832.00	1.71	14,542.41	0.00	0.00	14,872.41
COS	0.31	0.00	0.00	0.00	0.00	122.53	0.00	0.00	0.31
H_2	3.57	49,259.54	57.86	57.86	1.04	0.71	0.00	0.00	3.57
H_2O	88.50	2,189.77	0.23	0.23	0.05	115.27	63.08	0.13	5,622.10
H_2S	104.70	0.00	0.00	0.00	0.00	4,900.86	0.00	0.01	104.88
HCl	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.07	0.00	0.00	0.00
N_2	7,518.35	12,379.81	0.00	0.00	0.26	0.00	7,515.09	0.08	7,518.16
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.01	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.04
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	2,303.02	0.12	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1,287.13	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	3,156.45	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.06	0.00
Total	22,719.65	464,985.09	590,645.86	590,645.86	9.79	19,696.67	10,012.89	4,443.99	28,255.46
ALL PHASES									
Total Mole Flow, kmole/h	619.41	38,326.56	13,486.38	13,486.38	0.81	483.56	347.04	23.33	926.63
Total Mass Flow, kg/h	22,719.65	464,985.00	590,646.00	590,646.00	9.79	19,696.67	10,012.89	4,582.46	28,255.46
Temperature, C	34.95	23.38	14.00	35.00	23.38	30.00	15.00	85.03	75.81
Pressure, bar	6.48	34.19	1.00	152.56	34.19	2.00	1.01	8.61	1.01
Vapor Frac	1.00	1.00	1.00	1.00	1.00	1.00	1.00	0.00	1.00
Density, kg/m^3	9.43	16.81	1.85	779.27	16.81	3.27	1.22	2,152.91	1.07
Average MW	36.68	12.13	43.80	43.80	12.13	40.73	28.85	196.44	30.49

Exhibit 4-10 Mass Balance for Acid Gas Removal, Sulfur Recovery and CO_2 Compression – Option OSU-2

STREAM	23	24	25	26	27	28	29	30	31
Description	Compressed Claus Tail-Gas to Rectisol	Sweetened Syngas to MeOH Plant	Rectisol CO_2 Product	Compressed CO_2 to Pipeline	Sweet Syngas to Claus TGTU	Rectisol Acid Gas	Air to Claus Furnace	Liquid Sulfur Product	Treated Claus Tail-Gas
VAPOR / LIQUID									
Mole Flow, kmole/hr									
Ar	1.70	1.86	0.00	0.00	0.00	0.00	1.70	0.00	1.70
CH_4	0.11	719.82	28.54	28.54	0.01	0.06	0.00	0.00	0.11
C_2H_6	0.00	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.10	12,813.47	107.53	107.53	0.13	0.59	0.00	0.00	0.10
CO_2	171.57	933.75	6,757.25	6,757.25	0.01	167.17	0.00	0.00	171.59
COS	0.00	0.00	0.00	0.00	0.00	1.18	0.00	0.00	0.00
H_2	1.41	24,487.60	28.76	28.76	0.24	0.35	0.00	0.00	1.41
H_2O	2.52	57.97	0.01	0.01	0.00	3.05	1.81	0.00	109.34
H_2S	1.66	0.00	0.00	0.00	0.00	74.14	0.00	0.00	1.66
HCl	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.00	0.00
N_2	138.45	344.21	0.00	0.00	0.00	0.00	138.33	0.00	138.35
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	37.11	0.00	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	3.37	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	6.41	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	317.51	39,358.71	6,922.09	6,922.09	0.39	246.55	178.95	9.79	424.27
Mass Flow, kg/hr									
Ar	67.96	74.47	0.00	0.00	0.00	0.00	67.91	0.00	67.91
CH_4	1.79	11,547.87	457.86	457.86	0.11	0.90	0.00	0.00	1.79
C_2H_6	0.00	0.32	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	2.70	358,910.00	3,011.98	3,011.98	3.56	16.55	0.00	0.00	2.71
CO_2	7,550.71	41,094.35	297,385.00	297,385.00	0.41	7,357.03	0.00	0.00	7,551.85
COS	0.27	0.00	0.00	0.00	0.00	71.15	0.00	0.00	0.27
H_2	2.84	49,364.06	57.98	57.98	0.49	0.71	0.00	0.00	2.84
H_2O	45.34	1,044.38	0.11	0.11	0.01	54.97	32.53	0.04	1,969.73
H_2S	56.55	0.00	0.00	0.00	0.00	2,526.75	0.00	0.00	56.66
HCl	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.32	0.00	0.00	0.00
N_2	3,878.36	9,642.65	0.00	0.00	0.10	0.00	3,875.08	0.04	3,875.67
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.02	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.05
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	1,187.53	0.06	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	649.07	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1,643.96	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.03	0.00
Total	11,606.54	471,678.12	300,912.93	300,912.93	4.68	10,028.38	5,163.05	2,293.23	13,529.47
ALL PHASES									
Total Mole Flow, kmole/h	317.51	39,358.70	6,922.09	6,922.09	0.39	246.55	178.95	12.02	424.27
Total Mass Flow, kg/h	11,606.54	471,679.00	300,913.00	300,913.00	4.68	10,028.38	5,163.05	2,364.63	13,529.47
Temperature, C	34.94	21.07	14.00	35.00	21.07	30.00	15.00	85.03	71.63
Pressure, bar	6.48	34.19	1.00	152.56	34.19	2.00	1.01	8.61	1.01
Vapor Frac	1.00	1.00	1.00	1.00	1.00	1.00	1.00	0.00	1.00
Density, kg/m^3	9.39	16.69	1.83	756.39	16.69	3.26	1.22	2,155.72	1.13
Average MW	36.55	11.98	43.47	43.47	11.98	40.67	28.85	196.80	31.89

Exhibit 4-11 Process Flow Diagram for Methanol Plant (OSU-1 and OSU-2)



NOTES:

REV	DATE	DESCRIPTION
A	12-31-14	INITIAL ISSUE

ENGINEER/TECH SPECIALIST: E. Mace
DATE: December 31, 2014
CHECKED BY: J. Simpson
PROJECT MANAGER: J. Simpson

WorleyParsons
resources & energy

CLIENT/PROJECT TITLE
The Ohio State Univerisity
Chemical Looping Gasification to
Methanol

PROCESS FLOW DIAGRAM
Methanol Plant
(Case 1 and Case 2)

WORLEYPARSONS DWG. NO.
REV A

Exhibit 4-12 Mass Balance for Methanol Plant – Option OSU-1.

STREAM	32	33	34	35	36	37	38	39	40	41	42	43	44	45	46
Description	Recovered Hydrogen to Feed	Feed to Methanol Syngas Compressor	HP Syngas Feed	HP Recycle Gas	Warm Mixed Feed	Hot Mixed Feed	Reactor 1 Product	Reactor 2 Cooled Feed	Hot Reactor 2 Product	Warm Reactor 2 Product	Cool Reactor 2 Product	Methanol Flash Gas	Recycle Gas	Hydrogen Recovery Feed	Crude Methanol Product
VAPOR / LIQUID															
Mole Flow, kmole/hr															
Ar	0.00	3.64	3.64	78.59	82.23	82.23	82.23	82.23	82.23	82.23	82.23	0.34	78.59	3.27	0.02
CH_4	0.14	143.99	143.99	2,959.33	3,103.32	3,103.27	3,103.27	3,103.27	3,103.27	3,103.27	3,103.27	18.90	2,959.33	123.30	1.73
C_2H_6	0.00	0.00	0.00	0.00	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.00	0.00	0.00	0.00
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	501.30	501.30	501.26	9,220.83	9,220.83	12,228.39	12,228.39	12,228.39	7.13	501.30	0.00	11,719.96
1-Propanol	0.00	0.00	0.00	0.03	0.03	0.03	1.25	1.25	2.30	2.30	2.30	0.00	0.03	0.00	2.28
Formaldehyde	0.00	0.00	0.00	0.22	0.22	0.22	0.72	0.72	0.38	0.38	0.38	0.01	0.22	0.01	0.14
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.00	0.00	0.00	21.15	21.15	21.15	26.06	26.06	30.26	30.26	30.26	1.15	21.15	0.88	7.08
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.00	11,334.15	11,334.15	3,044.42	14,378.57	14,378.19	6,186.50	6,186.50	3,174.48	3,174.48	3,174.48	3.18	3,044.42	126.85	0.04
CO_2	0.00	1,845.70	1,845.70	10,623.58	12,469.29	12,469.15	11,927.24	11,927.24	11,920.50	11,920.50	11,920.50	387.91	10,623.58	442.11	466.90
COS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	676.24	25,111.99	25,111.99	24,969.09	50,081.08	50,081.03	32,072.43	32,072.43	26,027.82	26,027.82	26,027.82	18.20	24,969.09	1,040.38	0.16
H_2O	0.00	121.55	121.55	9.88	131.43	131.43	680.72	680.72	693.76	693.76	693.76	9.60	9.88	3.63	1,114.71
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCl	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	0.75	442.67	442.67	10,291.36	10,734.03	10,733.79	10,733.79	10,733.79	10,733.79	10,733.79	10,733.79	13.41	10,291.36	428.81	0.21
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	677.13	0.00	0.00	52,498.95	91,502.65	91,501.75	74,035.06	74,035.06	67,997.20	67,997.20	67,997.20	459.84	52,498.95	2,169.24	13,313.23
Mass Flow, kg/hr															
Ar	0.00	145.34	145.34	3,139.60	3,284.94	3,284.86	3,284.86	3,284.86	3,284.86	3,284.86	3,284.86	13.64	3,139.60	130.81	0.81
CH_4	2.32	2,309.95	2,309.95	47,475.87	49,785.82	49,785.02	49,785.02	49,785.02	49,785.02	49,785.02	49,785.02	303.27	47,475.87	1,978.12	27.76
C_2H_6	0.00	0.01	0.01	0.14	0.15	0.15	0.15	0.15	0.15	0.15	0.15	0.00	0.14	0.01	0.00
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	16,062.63	16,062.63	16,061.31	295,455.00	295,455.00	391,824.00	391,824.00	391,824.00	228.51	16,062.63	0.00	375,533.00
1-Propanol	0.00	0.00	0.00	1.52	1.52	1.52	75.32	75.32	138.31	138.31	138.31	0.02	1.52	0.00	136.77
Formaldehyde	0.00	0.00	0.00	6.46	6.46	6.46	21.73	21.73	11.40	11.40	11.40	0.36	6.46	0.27	4.31
Acetaldehyde	0.00	0.00	0.00	0.01	0.01	0.01	0.05	0.05	0.09	0.09	0.09	0.00	0.01	0.00	0.08
Dimethy-Ether	0.00	0.00	0.00	974.58	974.58	974.52	1,200.72	1,200.72	1,394.21	1,394.21	1,394.21	53.02	974.58	40.59	326.01
Acetone	0.00	0.00	0.00	0.03	0.03	0.03	0.15	0.15	0.25	0.25	0.25	0.00	0.03	0.00	0.22
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.01	317,474.00	317,474.00	85,275.30	402,749.00	402,739.00	173,287.00	173,287.00	88,918.55	88,918.55	88,918.55	89.00	85,275.30	3,553.13	1.13
CO_2	0.00	81,228.97	81,228.97	467,542.00	548,771.00	548,765.00	524,916.00	524,916.00	524,619.00	524,619.00	524,619.00	17,071.91	467,542.00	19,457.21	20,547.96
COS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	1,363.23	50,622.77	50,622.77	50,334.69	100,958.00	100,957.00	64,654.17	64,654.17	52,468.97	52,468.97	52,468.97	36.70	50,334.69	2,097.27	0.32
H_2O	0.00	2,189.77	2,189.77	178.04	2,367.82	2,367.83	12,263.27	12,263.27	12,498.23	12,498.23	12,498.23	172.90	178.04	65.41	20,081.87
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCl	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	20.90	12,400.71	12,400.71	288,297.00	300,698.00	300,691.00	300,691.00	300,691.00	300,691.00	300,691.00	300,691.00	375.56	288,297.00	12,012.34	6.02
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.01	0.01	0.02	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.00	0.02	0.00	0.01
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	1,386.46	466,371.54	466,371.54	959,287.89	1,425,659.99	1,425,633.75	1,425,634.48	1,425,634.48	1,425,634.08	1,425,634.08	1,425,634.08	18,344.90	959,287.89	39,335.17	416,666.27
ALL PHASES															
Total Mole Flow, kmole/h	677.13	39,003.69	39,003.69	52,498.96	91,502.65	91,501.75	74,035.07	74,035.07	67,997.20	67,997.20	67,997.20	459.84	52,498.96	2,169.24	13,313.23
Total Mass Flow, kg/h	1,386.46	466,372.00	466,372.00	959,288.00	1,425,660.00	1,425,630.00	1,425,630.00	1,425,630.00	1,425,630.00	1,425,630.00	1,425,630.00	18,344.90	959,288.00	39,335.17	416,666.00
Temperature, C	40.93	51.48	99.90	46.04	166.00	205.11	246.11	204.44	221.11	122.64	40.00	48.83	39.95	35.93	38.68
Pressure, bar	47.31	34.05	52.06	52.06	51.71	51.50	50.81	50.47	50.12	49.57	49.30	5.51	48.95	48.00	26.20
Vapor Frac	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	0.97	0.80	1.00	1.00	1.00	0.00
Density, kg/m^3	3.61	14.82	19.52	35.70	21.57	19.73	22.36	24.24	25.49	33.06	48.12	8.38	34.30	33.77	606.73
Average MW	2.05	11.96	11.96	18.27	15.58	15.58	19.26	19.26	20.97	20.97	20.97	39.89	18.27	18.13	31.30

Exhibit 4-13 Mass Balance for Methanol Plant – Option OSU-2.

STREAM	32	33	34	35	36	37	38	39	40	41	42	43	44	45	46
Description	Recovered Hydrogen to Feed	Feed to Methanol Syngas Compressor	HP Syngas Feed	HP Recycle Gas	Warm Mixed Feed	Hot Mixed Feed	Reactor 1 Product	Reactor 2 Cooled Feed	Hot Reactor 2 Product	Warm Reactor 2 Product	Cool Reactor 2 Product	Methanol Flash Gas	Recycle Gas	Hydrogen Recovery Feed	Crude Methanol Product
VAPOR / LIQUID															
Mole Flow, kmole/hr															
Ar	0.00	1.86	1.86	40.53	42.39	42.39	42.39	42.39	42.39	42.39	42.39	0.16	40.53	1.69	0.01
CH_4	0.73	720.55	720.55	14,925.97	15,646.52	15,646.49	15,646.49	15,646.49	15,646.49	15,646.49	15,646.49	88.10	14,925.97	621.90	10.53
C_2H_6	0.00	0.01	0.01	0.13	0.14	0.14	0.14	0.14	0.14	0.14	0.14	0.00	0.13	0.01	0.00
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	655.87	655.87	655.86	9,263.63	9,263.63	12,786.17	12,786.17	12,786.17	5.53	655.87	0.00	12,124.76
1-Propanol	0.00	0.00	0.00	0.03	0.03	0.03	1.26	1.26	2.31	2.31	2.31	0.00	0.03	0.00	2.28
Formaldehyde	0.00	0.00	0.00	0.51	0.51	0.51	1.28	1.28	0.86	0.86	0.86	0.02	0.51	0.02	0.31
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.00	0.00	0.00	22.90	22.90	22.90	27.81	27.81	32.01	32.01	32.01	0.91	22.90	0.95	7.24
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.00	12,813.47	12,813.47	14,985.47	27,798.95	27,798.01	19,216.48	19,216.48	15,625.40	15,625.40	15,625.40	15.27	14,985.47	624.39	0.26
CO_2	0.00	933.75	933.75	8,334.59	9,268.34	9,268.18	9,227.66	9,227.66	9,285.07	9,285.07	9,285.07	242.05	8,334.59	346.79	361.64
COS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	342.71	24,830.31	24,830.31	12,654.07	37,484.38	37,484.54	20,200.68	20,200.68	13,190.34	13,190.34	13,190.34	8.92	12,654.07	527.25	0.10
H_2O	0.00	57.97	57.97	0.99	58.96	58.96	106.86	106.86	55.75	55.75	55.75	8.53	0.99	4.06	643.01
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCl	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	0.58	344.80	344.80	8,026.52	8,371.32	8,371.31	8,371.31	8,371.31	8,371.31	8,371.31	8,371.31	10.14	8,026.52	334.44	0.22
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	344.03	39,702.73	39,702.73	59,647.59	99,350.33	99,349.34	82,106.00	82,106.00	75,038.25	75,038.25	75,038.25	379.63	59,647.59	2,461.50	13,150.37
Mass Flow, kg/hr															
Ar	0.00	74.47	74.47	1,619.00	1,693.47	1,693.47	1,693.47	1,693.47	1,693.47	1,693.47	1,693.47	6.51	1,619.00	67.46	0.50
CH_4	11.68	11,559.55	11,559.55	239,454.00	251,013.00	251,013.00	251,013.00	251,013.00	251,013.00	251,013.00	251,013.00	1,413.35	239,454.00	9,976.98	168.86
C_2H_6	0.00	0.32	0.32	3.85	4.17	4.17	4.17	4.17	4.17	4.17	4.17	0.09	3.85	0.16	0.07
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	21,015.63	21,015.63	21,015.10	296,827.00	296,827.00	409,696.00	409,696.00	409,696.00	177.16	21,015.63	0.00	388,504.00
1-Propanol	0.00	0.00	0.00	1.96	1.96	1.96	75.75	75.75	138.75	138.75	138.75	0.02	1.96	0.00	136.77
Formaldehyde	0.00	0.00	0.00	15.37	15.37	15.37	38.41	38.41	25.90	25.90	25.90	0.62	15.37	0.64	9.27
Acetaldehyde	0.00	0.00	0.00	0.01	0.01	0.01	0.05	0.05	0.09	0.09	0.09	0.00	0.01	0.00	0.08
Dimethy-Ether	0.00	0.00	0.00	1,054.91	1,054.91	1,054.88	1,281.08	1,281.08	1,474.57	1,474.57	1,474.57	42.14	1,054.91	43.94	333.59
Acetone	0.00	0.00	0.00	0.04	0.04	0.04	0.16	0.16	0.26	0.26	0.26	0.00	0.04	0.00	0.22
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	0.07	358,911.00	358,911.00	419,749.00	778,660.00	778,633.00	538,261.00	538,261.00	437,674.00	437,674.00	437,674.00	427.81	419,749.00	17,489.52	7.26
CO_2	0.00	41,094.35	41,094.35	366,804.00	407,898.00	407,891.00	406,107.00	406,107.00	408,634.00	408,634.00	408,634.00	10,652.40	366,804.00	15,262.19	15,915.70
COS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	690.87	50,054.93	50,054.93	25,509.08	75,564.01	75,564.34	40,722.14	40,722.14	26,590.14	26,590.14	26,590.14	17.98	25,509.08	1,062.87	0.21
H_2O	0.00	1,044.38	1,044.38	17.85	1,062.23	1,062.23	1,925.16	1,925.16	1,004.38	1,004.38	1,004.38	153.65	17.85	73.18	11,584.08
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCl	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	16.30	9,658.95	9,658.95	224,851.00	234,510.00	234,510.00	234,510.00	234,510.00	234,510.00	234,510.00	234,510.00	284.04	224,851.00	9,368.76	6.07
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.02	0.02	0.03	0.05	0.05	0.05	0.05	0.05	0.05	0.05	0.00	0.03	0.00	0.02
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	718.92	472,397.97	472,397.97	1,300,095.72	1,772,492.85	1,772,458.62	1,772,458.45	1,772,458.45	1,772,458.78	1,772,458.78	1,772,458.78	13,175.76	1,300,095.72	53,345.69	416,666.70
ALL PHASES															
Total Mole Flow, kmole/h	344.03	39,702.73	39,702.73	59,647.59	99,350.32	99,349.34	82,106.00	82,106.00	75,038.25	75,038.25	75,038.25	379.63	59,647.59	2,461.50	13,150.37
Total Mass Flow, kg/h	718.92	472,397.00	472,397.00	1,300,100.00	1,772,490.00	1,772,460.00	1,772,460.00	1,772,460.00	1,772,460.00	1,772,460.00	1,772,460.00	13,175.77	1,300,100.00	53,345.69	416,666.00
Temperature, C	39.64	51.57	100.20	45.92	166.00	205.88	246.11	204.44	221.11	116.79	40.00	50.18	39.94	34.64	40.17
Pressure, bar	47.31	34.05	52.06	52.06	51.71	51.50	50.81	50.47	50.12	49.57	49.30	5.51	48.95	48.00	26.20
Vapor Frac	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	0.98	0.83	1.00	1.00	1.00	0.00
Density, kg/m^3	3.69	14.73	19.39	43.16	24.70	22.54	25.03	27.13	28.69	37.75	53.50	7.23	41.47	41.08	599.40
Average MW	2.09	11.90	11.90	21.80	17.84	17.84	21.59	21.59	23.62	23.62	23.62	34.71	21.80	21.67	31.68

Exhibit 4-14 Process Flow Diagram Fired Boiler (OSU-1)

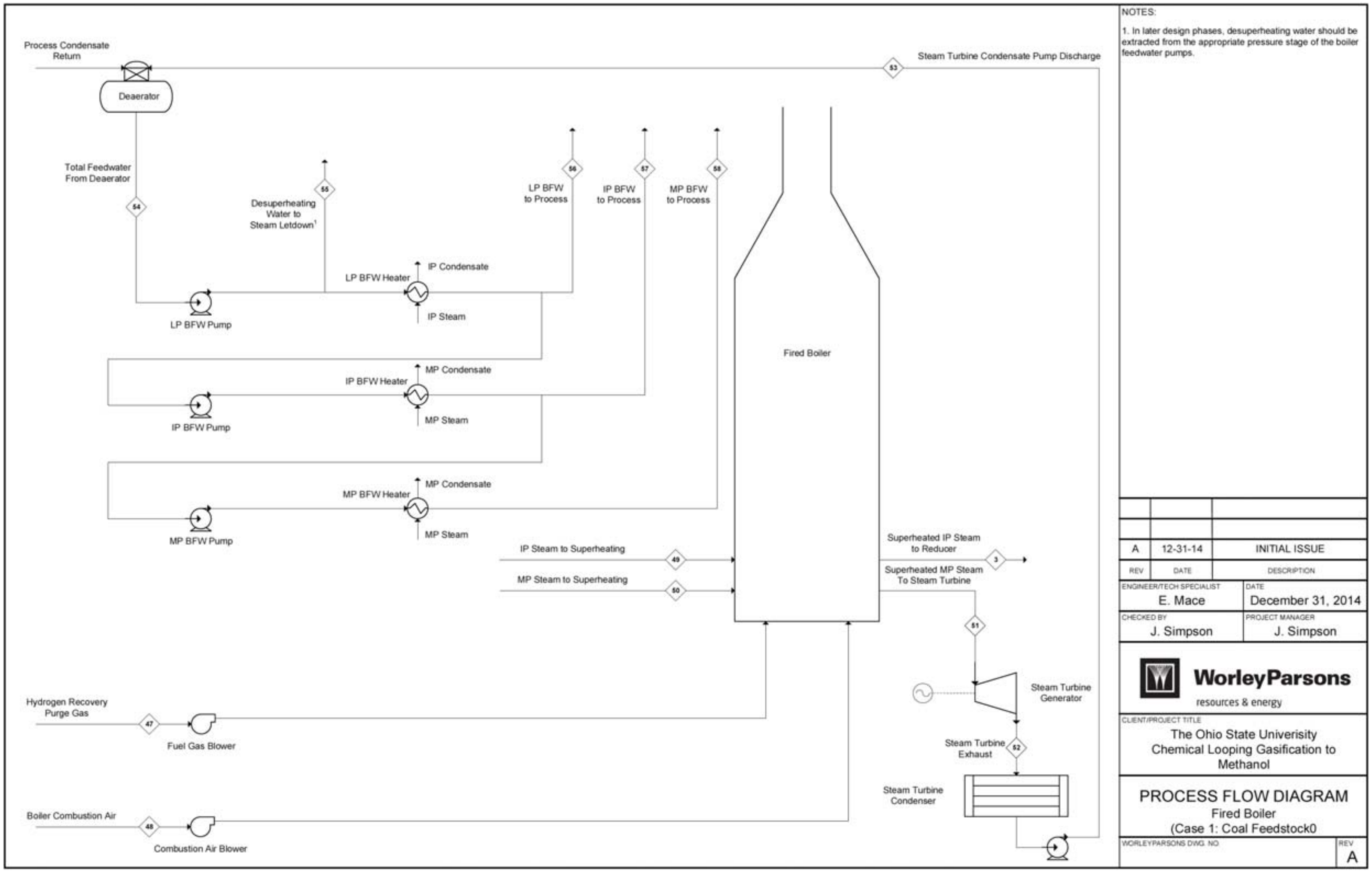


Exhibit 4-15 Mass Balance Fired Boiler – Option OSU-1

STREAM	47	48	49	3	50	51	52	53	54	55	56	57	58
Description	Hydrogen Recovery Purge Gas	Boiler Combustion Air	IP Steam to Superheating	Superheated IP Steam to Reducer	MP Steam to Superheating	Superheated MP Steam to Steam Turbine	Steam Turbine Exhaust	Steam Turbine Condensate Pump Discharge	Total Feedwater from Dearator	Desuperheating Water to Steam Letdown	LP BFW to Process	IP BFW to Process	MP BFW to Process
VAPOR / LIQUID													
Mole Flow, kmole/hr													
Ar	3.27	27.03	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	123.16	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_2H_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.88	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	126.85	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	442.11	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	364.13	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	3.63	28.71	40,897.84	40,897.84	7,923.44	7,923.44	7,923.44	7,923.44	93,213.36	545.15	8,290.26	49,536.65	34,841.30
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCl	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	428.06	2,199.78	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	590.17	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	1,492.11	2,845.70	40,897.84	40,897.84	7,923.44	7,923.44	7,923.44	7,923.44	93,213.36	545.15	8,290.26	49,536.65	34,841.30
Mass Flow, kg/hr													
Ar	130.81	1,079.96	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	1,975.80	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_2H_6	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.27	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	40.59	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	3,553.12	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	19,457.21	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	734.04	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	65.41	517.27	736,786.00	736,786.00	142,743.00	142,743.00	142,743.00	142,743.00	1,679,270.00	9,821.08	149,351.00	892,417.00	627,676.00
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCl	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	11,991.44	61,623.49	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	18,884.70	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	37,948.71	82,105.42	736,786.00	736,786.00	142,743.00	142,743.00	142,743.00	142,743.00	1,679,270.00	9,821.08	149,351.00	892,417.00	627,676.00
ALL PHASES													
Total Mole Flow, kmole/h	1,492.11	2,845.70	40,897.84	40,897.84	7,923.44	7,923.44	7,923.44	7,923.44	93,213.36	545.15	8,290.26	49,536.65	34,841.30
Total Mass Flow, kg/h	37,948.71	82,105.42	736,786.00	736,786.00	142,743.00	142,743.00	142,743.00	142,743.00	1,679,270.00	9,821.08	149,351.00	892,417.00	627,676.00
Temperature, C	30.93	15.00	191.60	250.00	235.72	296.31	94.41	94.50	138.34	138.41	147.60	186.05	194.00
Pressure, bar	1.22	1.01	13.00	10.00	31.00	30.00	0.78	10.00	3.45	7.55	6.55	15.00	32.00
Vapor Frac	1.00	1.00	1.00	1.00	1.00	1.00	0.90	0.00	0.00	0.00	0.00	0.00	0.00
Density, kg/m^3	1.23	1.22	6.47	4.29	15.04	12.43	0.51	804.81	770.09	770.14	762.01	724.48	716.59
Average MW	25.43	28.85	18.02	18.02	18.02	18.02	18.02	18.02	18.02	18.02	18.02	18.02	18.02

Exhibit 4-16 Process Flow Diagram Fired Boiler (OSU-2)

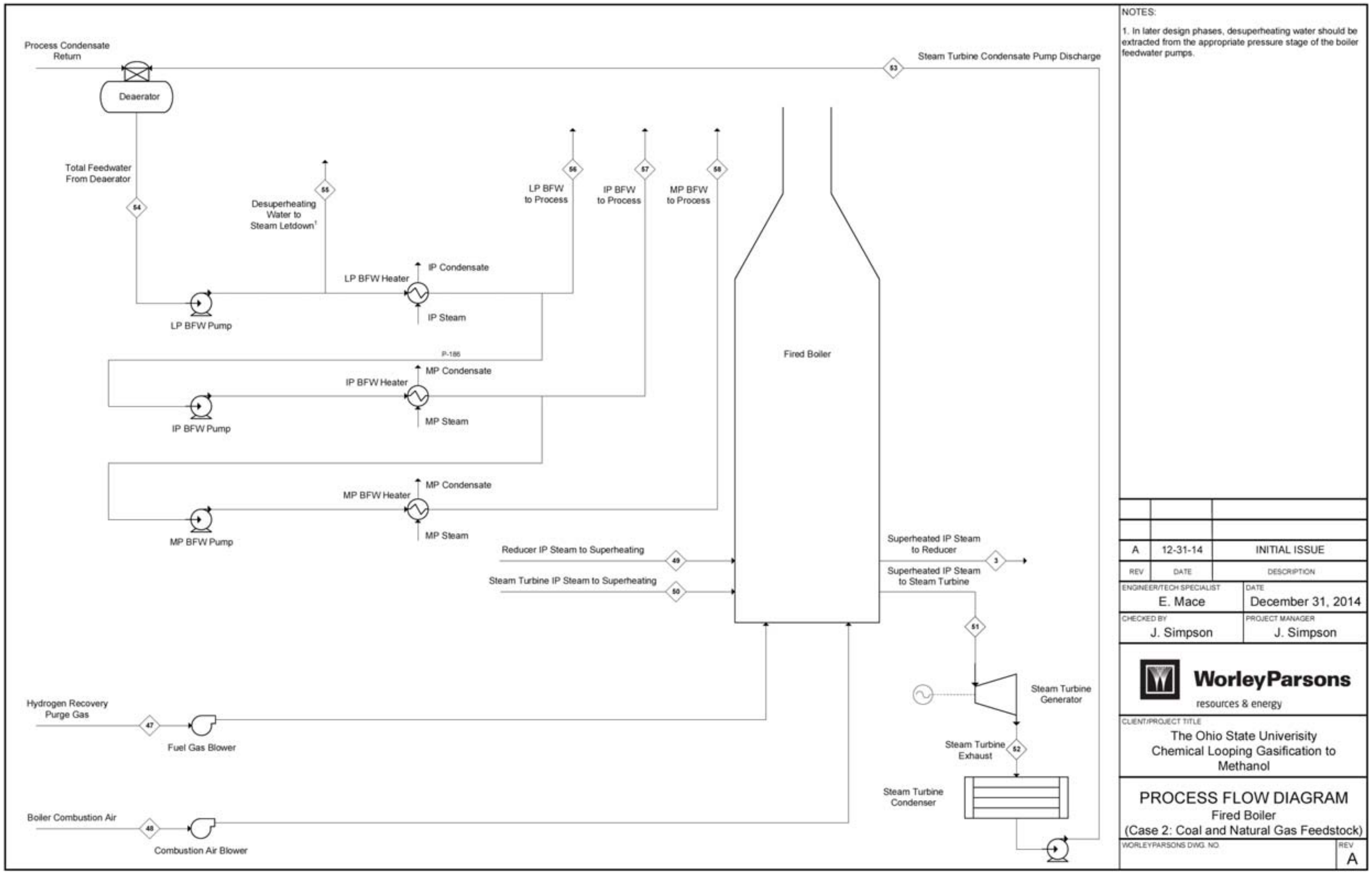


Exhibit 4-17 Mass Balance Fired Boiler – Option OSU-2

STREAM	47	48	49	3	50	51	52	53	54	55	56	57	58
Description	Hydrogen Recovery Purge Gas	Boiler Combustion Air	Reducer IP Steam to Superheating	Superheated IP Steam to Reducer	Steam Turbine IP Steam to Superheating	Superheated IP Steam to Steam Turbine	Steam Turbine Exhaust	Steam Turbine Condensate Pump Discharge	Total Feedwater from Dearator	Desuperheating Water to Steam Letdown	LP BFW to Process	IP BFW to Process	MP BFW to Process
VAPOR / LIQUID													
Mole Flow, kmole/hr													
Ar	1.69	90.52	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	621.17	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_2H_6	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	0.95	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	624.39	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	346.79	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	184.54	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	4.06	96.15	17,463.45	17,463.45	7,842.42	7,842.42	7,842.42	7,842.42	73,831.12	762.77	5,924.53	37,030.26	30,113.56
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCl	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	333.86	7,365.92	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	1,976.17	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	2,117.48	9,528.75	17,463.45	17,463.45	7,842.42	7,842.42	7,842.42	7,842.42	73,831.12	762.77	5,924.53	37,030.26	30,113.56
Mass Flow, kg/hr													
Ar	67.46	3,616.22	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH_4	9,965.30	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_2H_6	0.16	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C_3H_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-C_4H_{10}	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
1-Propanol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Formaldehyde	0.64	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetaldehyde	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dimethy-Ether	43.94	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Acetone	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Cl_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	17,489.45	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO_2	15,262.19	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2	372.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	73.18	1,732.08	314,609.00	314,609.00	141,283.00	141,283.00	141,283.00	141,283.00	1,330,090.00	13,741.53	106,732.00	667,111.00	542,504.00
H_2S	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCl	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	9,352.46	206,345.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NaOH	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
O_2	0.00	63,234.99	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
S_8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
SO_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	52,626.77	274,928.29	314,609.00	314,609.00	141,283.00	141,283.00	141,283.00	141,283.00	1,330,090.00	13,741.53	106,732.00	667,111.00	542,504.00
ALL PHASES													
Total Mole Flow, kmole/h	2,117.48	9,528.75	17,463.45	17,463.45	7,842.42	7,842.42	7,842.42	7,842.42	73,831.12	762.77	5,924.53	37,030.26	30,113.56
Total Mass Flow, kg/h	52,626.77	274,928.00	314,609.00	314,609.00	141,283.00	141,283.00	141,283.00	141,283.00	1,330,090.00	13,741.53	106,732.00	667,111.00	542,504.00
Temperature, C	29.64	15.00	191.60	250.00	191.60	471.39	47.58	47.49	138.34	138.41	147.60	186.05	194.00
Pressure, bar	1.22	1.01	13.00	10.00	13.00	12.00	0.10	10.00	3.45	7.55	6.55	15.00	32.00
Vapor Frac	1.00	1.00	1.00	1.00	1.00	1.00	0.99	0.00	0.00	0.00	0.00	0.00	0.00
Density, kg/m^3	1.21	1.22	6.47	4.29	6.47	3.54	0.07	835.75	770.09	770.14	762.01	724.48	716.59
Average MW	24.85	28.85	18.02	18.02	18.02	18.02	18.02	18.02	18.02	18.02	18.02	18.02	18.02

The performance results for OSU CLG cases are summarized and compared to the DOE/NETL Baseline cases in Exhibit 4-18. The electrical and steam loads of the major power consumers are listed in Exhibit 4-19.

Exhibit 4-18: Comparison of System Performance

Stream	Mass Flow lb/hr			
	DOE/NETL MBL-1, MBL-2	DOE/NETL, MBL-3	OSU-1	OSU-2
As Received Coal	1,618,190	NA	1,395,457	718,631
Dried Coal	1,277,850	NA	NA	NA
Natural Gas to Reformer, OSU CLG	NA	583,677	NA	272,290
Oxygen from ASU containing 95% O_2	1,010,968	682,554	NA	NA
Steam to gasifier, reformer, OSU CLG	158,326	130,751	1,624,318	693,587
Air for coal drying	434,322	NA	NA	NA
N_2 to coal drying	3,100,000	NA	NA	NA
Air to direct-fired boiler	121,518	NA	181,009	606,106
Total makeup water	5,596,974	4,550,133	3,618,118	2,499,200
Raw syngas from gasifier to scrubber	6,543,540	NA	3,738,715	2,381,570
Clean syngas for methanol production	1,183,080	1,183,090	1,025,106	1,039,864
Fuel gas (for coal drying)	27,080	NA	NA	NA
Flash gas (for coal drying or recycled to OSU CLG)	90,917	NA	40,443	29,048
Tail gas from Claus unit (for coal drying or recycled to Rectisol in OSU cases)	61,476	NA	50,089	25,589
Purge Gas (for coal drying and power Generation)	148,223	NA	83,662	116,021
Crude Methanol	941,823	940,989	918,582	918,582
Sulfur (S_8)	11,576	NA	9,797	5,214
Captured CO_2 (MLB-2, MLB-3, OSU-1 and OSU-2 only)	1,569,410	235,808	1,302,138	663,393
Water discharge	2,248,798	1,398,202	1,832,047	988,985

Exhibit 4-19 Comparison of Auxiliary Loads

Electrical Loads	DOE/NETL MBL-1	DOE/NETL MBL-2	DOE/NETL MBL-3	OSU-1	OSU-2
Coal handling and milling	9,090	9,090	N.A.	7,840	4,040
Ash handling	1,940	1,940	N.A.	1,990	1,030
ASU	179,940	179,940	122,104	N.A.	N.A.
Oxidizer air compressor	N.A.	N.A.	N.A.	133,665	114,651
Oxidizer spent air expander	N.A.	N.A.	N.A.	(38,152)	(32,725)
Reducer & oxidizer	N.A.	N.A.	N.A.	0	0
Syngas cooling	N.A.	N.A.	N.A.	6,809	3,282
Methanol syngas recycle compressor	6,600	6,600	3,370	N.A.	N.A.
Syngas compressor	N.A.	N.A.	20,760	83,760	75,658
Incinerator air blower	2,680	2,680	N.A.	N.A.	N.A.
Fired boiler air blower	310	310	310	2,329	7,799
Fired boiler fuel gas blower	N.A.	N.A.	N.A.	972	1,373
Flash bottoms pumps	720	720	N.A.	N.A.	N.A.
Scrubber pumps	1,070	1,070	N.A.	464	285
Sour-water stripper pumps	N.A.	N.A.	N.A.	4	2
Acid gas removal (Rectisol)	51,270	51,270	N.A.	40,672	29,791
Sulfur recovery plant (Modified-Claus)	250	250	N.A.	4,155	1,906
CO$_2$ compressor	9,000	68,820	N.A.	58,751	30,310
Methanol syngas compressor	20,760	20,760	N.A.	17,218	17,551
Methanol recycle compressor	3,370	3,370	N.A.	3,011	3,376
Water treatment	3,530	3,530	3,530	3,039	1,736
Methanol plant/air cooler fans	1,800	1,800	1,800	2,977	2,977
Methanol flash gas recycle compressor	N.A.	N.A.	N.A.	313	261
Circulating water pumps	9,110	9,430	9,110	5,060	4,637
Boiler feedwater pumps	1,500	1,500	1,500	1,756	1,439
Cooling tower fans	360	510	360	2,602	2,384
Steam turbine auxiliaries	100	100	100	100	100
Miscellaneous BOP	5,000	5,000	5,000	5,000	5,000
Transformer losses	N.A.	N.A.	N.A.	295	321
Total auxiliary power	**308,400**	**368,690**	**167,944**	**344,334**	**277,183**
Power Generation					
Steam turbine	N.A.	N.A.	N.A.	20,830	31,491
NGCC Power Generation	320,680	390,170	194,070	N.A.	N.A.
Power Generation Excess**	**12,280**	**21,480**	**26,126**	**(323,504)**	**(245,692)**
Notes					
*N.A. = Not applicable					
** Negative value indicates power purchase requirement					

4.2 Environmental Performance

4.2.1 NOx Emissions

The proposed CLG process is capable of low thermal NOx formation by avoiding excessive temperatures in the oxidizer and strong reducing conditions in the reducer reactor. The exhaust from the boiler is a potential source of NOx emissions. The boiler design incorporates low NOx burners to limit the NOx formation and meet the emissions requirements.

4.2.2 Mercury

To achieve 90 percent mercury capture target set for this study, a sulfur impregnated activated carbon bed is used. This technology has been shown to have a removal efficiency of 95 percent based on Eastman Chemical's operating experience at its coal-to-methanol plant in Kingsport, Tennessee[9]. Similar to the Methanol Baseline study, a removal efficiency of 95 percent was used as the assumed performance level for this study.

4.2.3 Particular Matter

The primary potential source of particulate emissions results from the attrition of the oxygen carrier (OC) particles and fly ash entrained in the oxygen carrier. To mitigate these emissions, two steps are implemented. The first is the continuous separation of the fines from the OC with a particulate classifier located after the reducing reactor. Additionally particulate control devices are located after the oxidizer to separate the particulate from the air stream prior to expansion through the expander and subsequent emission to the atmosphere.

4.2.4 Solid Waste-ash/Spent Oxygen Carrier

Fly ash from coal and attrition products from the oxygen carrier are the primary solid wastes discharged from the OSU CLG. These two streams will represent the fine material in the circulating oxygen carrier and will be separated from coarser OC carrier by a cyclone separator. The fine material can be further separated into fly ash and attritted oxygen carrier based on the fly ash particle size being smaller that the attritted oxygen carrier. The fly ash will be sent to a solid land fill while the attritted oxygen carrier will be recycled into new oxygen carrier. Based on the coal flow to the CLG reactor, the 450,373 and 231,921 ton/yr of fly ash will be produced from the OSU-1 and OSU-2 options. 20,696 and 17,739 ton/yr of attritted oxygen carrier will be recycled for options OSU-1 and OSU-2.

4.3 Carbon, Sulfur, Waste Water and Makeup Water Balances

Carbon Balance

The OSU-1 and OSU-2 carbon balances for the overall plant are shown in Exhibit 4-20 and Exhibit 4-21.

Exhibit 4-20 Carbon Balance Table for OSU CLG-Case 1

Carbon Balance			
	Carbon In, kg/h		**Carbon Out, kg/h**
Coal	316,919	Ash	0
Natural Gas	0	Oxidizer Flue Gas	0
		CO_2 to Pipeline	161,636
		Claus Vent	0
		Fired Boiler	8,334
		Crude Methanol	146,648
		Miscellaneous Losses	301
Total	316,919	**Total**	316,919

Exhibit 4-21 Carbon Balance Table for OSU CLG-Case 2

Carbon Balance			
	Carbon In, kg/h		**Carbon Out, kg/h**
Coal	163,207	Ash	0
Natural Gas	89,207	Oxidizer Flue Gas	0
		CO_2 to Pipeline	82,794
		Claus Vent	0
		Fired Boiler	19,149
		Crude Methanol	150,360
		Miscellaneous Losses	112
Total	252,414	**Total**	252,414

Sulfur Balance

The OSU-1 and OSU-2 sulfur balances for the overall plant are shown in Exhibit 4-22 and Exhibit 4-23.

Exhibit 4-22 Sulfur Balance Table for OSU CLG-Case 1

Sulfur Balance			
	Sulfur In, kg/h		**Sulfur Out, kg/h**
Coal	4,605	Sulfur Product	4,582
Natural Gas	0	Miscellaneous Losses	23
Total	4,605	**Total**	4,605

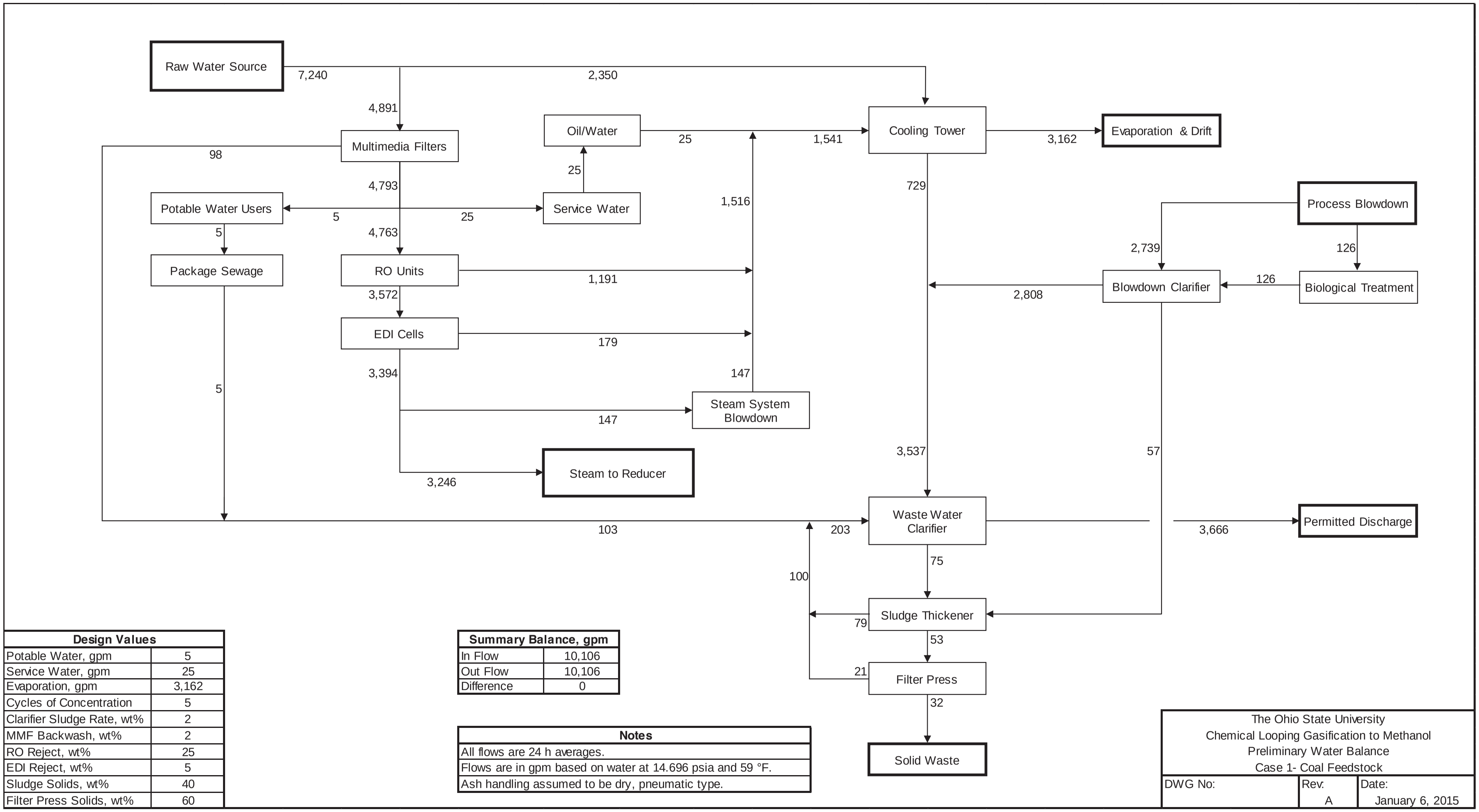
Exhibit 4-23 Sulfur Balance Table for OSU CLG-Case 2

Sulfur Balance			
	Sulfur In, kg/h		**Sulfur Out, kg/h**
Coal	2,371	Sulfur Product	2,364
Natural Gas	0	Miscellaneous Losses	7
Total	2,371	**Total**	2,371

Waste Water and Make-up Water

The OSU CLG – Case 1 and OSU CLG – Case 2 water balances for the overall plant are shown in Exhibit 4-24 and Exhibit 4-25.

Exhibit 4-24 Water Balance for OSU-1 Option



Design Values	
Potable Water, gpm	5
Service Water, gpm	25
Evaporation, gpm	3,162
Cycles of Concentration	5
Clarifier Sludge Rate, wt%	2
MMF Backwash, wt%	2
RO Reject, wt%	25
EDI Reject, wt%	5
Sludge Solids, wt%	40
Filter Press Solids, wt%	60

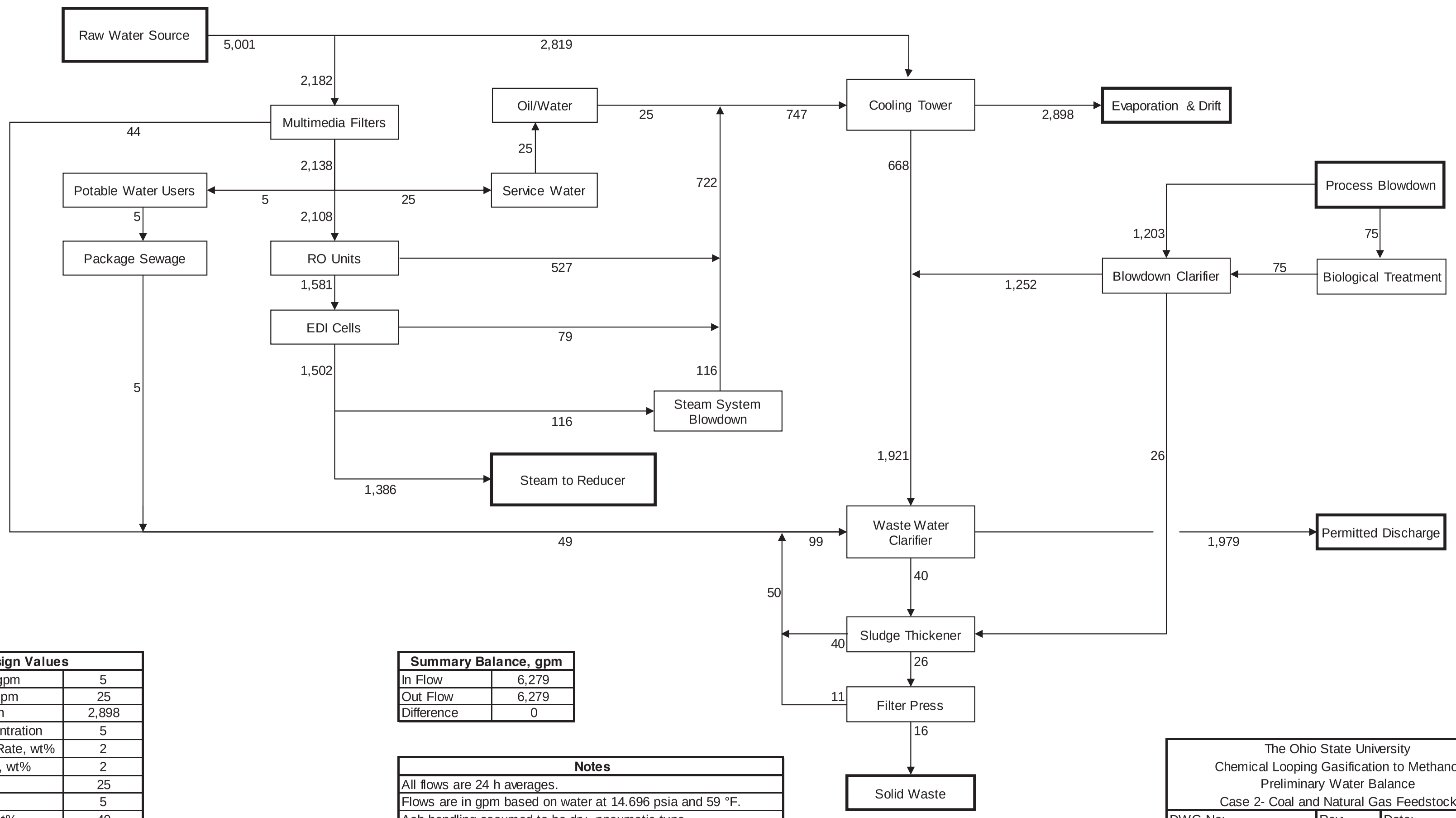
Summary Balance, gpm	
In Flow	10,106
Out Flow	10,106
Difference	0

Notes
All flows are 24 h averages.
Flows are in gpm based on water at 14.696 psia and 59 °F.
Ash handling assumed to be dry, pneumatic type.

The Ohio State University
Chemical Looping Gasification to Methanol
Preliminary Water Balance
Case 1- Coal Feedstock

DWG No:	Rev.	Date:
	A	January 6, 2015

Exhibit 4-25 Water Balance for OSU-2 Option



Design Values	
Potable Water, gpm	5
Service Water, gpm	25
Evaporation, gpm	2,898
Cycles of Concentration	5
Clarifier Sludge Rate, wt%	2
MMF Backwash, wt%	2
RO Reject, wt%	25
EDI Reject, wt%	5
Sludge Solids, wt%	40
Filter Press Solids, wt%	60

Summary Balance, gpm	
In Flow	6,279
Out Flow	6,279
Difference	0

Notes
All flows are 24 h averages.
Flows are in gpm based on water at 14.696 psia and 59 °F.
Ash handling assumed to be dry, pneumatic type.

The Ohio State University
Chemical Looping Gasification to Methanol
Preliminary Water Balance
Case 2- Coal and Natural Gas Feedstock

DWG No:	Rev.	Date:
	A	January 6, 2015

5. COST ESTIMATING RESULTS

The cost estimating methodology is described in Section 3.3 of this report.

Exhibit 5-1 shows a high level cost summary of the design cases utilizing the OSU CLG technology in comparison to DOE/NETL Methanol Baseline Cases. Total plant capital costs for the OSU CLG cases, organized by cost account, are presented in Exhibit 5-2 and Exhibit 5-3. Initial and annual O&M costs are provided in Exhibit 5-4 and Exhibit 5-5.

Exhibit 5-1: Comparative Summary of Capital and Operating Costs

	Case				
	MBL-1	**MBL-2**	**MBL-3**	**OSU-1**	**OSU-2**
Total Plant Costs (2011 MM$)	4,586	4,775	2,028	3,497	2,996
Total Overnight Cost (2011 MM$)	5,572	5,802	2,485	4,236	3,634
Bare Erected Cost	3,444	3,584	1,605	2,564	2,187
Home Office Expenses	320	333	153	239	205
Project Contingency	162	167	15	138	124
Process contingency	660	690	255	556	480
Owners Costs	986	1,027	457	739	638
Total As Spent Capital (2011 MM$)	6,580	6,852	2,935	5,003	4,291
Annual Fixed Operating Costs (x1000 $/yr)	156,650	162,051	70,644	108,291	94,034
Variable Operating Costs ($/gal)	0.085	0.088	0.032	0.056	0.049

For similar sized plants, there is a significant capital cost reduction for the OSU technology compared to the gasification based coal to methanol process, 27% when comparing the MBL-2 and OSU-1 cases with CO_2 capture. While there is some cost savings that can be attributed to having the power generated off site for the OSU options, the majority of the cost savings is attributed to replacing the ASU and gasifiers in the MBL cases with the OSU CLG technology. When comparing the OSU options with the natural gas fed MBL-3, the MBL-3 has the lower capital costs. The lower capital cost for the MBL-3 is driven by the elimination of coal handling equipment and the lower equipment costs for the reformers.

Exhibit 5-2: Total Plant Cost Summary OSU CLG Case 1

Client: Ohio State University **Report Date:** 2014-Dec-30
Project: Coal to Methanol via Chemical Looping Gasification

TOTAL PLANT COST SUMMARY

Case: Case 1 - 100% Coal Feedstock
Plant Size: -325.0 MW,net **Estimate Type:** Conceptual **Cost Base (Jun)** 2011 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		TOTAL PLANT COST $
				Direct	Indirect				Process	Project	
1	COAL & SORBENT HANDLING	$33,295	$5,838	$25,466	$0	$0	**$64,599**	$5,725	$0	$14,065	**$84,389**
2	COAL & SORBENT PREP & FEED	$28,731	$129,544	$92,149	$0	$0	**$250,424**	$21,446	$0	$54,374	**$326,244**
3	FEEDWATER & MISC. BOP SYSTEMS	$13,182	$4,803	$10,477	$0	$0	**$28,462**	$2,589	$0	$7,272	**$38,323**
4	GASIFIER & ACCESSORIES										
4.1	Reducer Reactor, Oxidizer, Xfer & Ash Separator	$94,830	$47,415	$94,830	$0	$0	**$237,076**	$23,708	$39,118	$44,985	**$344,886**
4.2	Syngas Cooler	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**
4.3	Air Separation Unit	$0	$0	w/equip.	$0	$0	**$0**	$0	$0	$0	**$0**
4.4-4.9	Other Gasification Equipment	$268,864	$52,016	$113,161	$0	$0	**$434,041**	$41,589	$0	$92,583	**$568,212**
	SUBTOTAL 4	**$363,694**	**$99,431**	**$207,991**	**$0**	**$0**	**$671,117**	**$65,297**	**$39,118**	**$137,568**	**$913,099**
5A	GAS CLEANUP & PIPING	$688,430	$23,249	$61,696	$0	$0	**$773,375**	$72,171	$97,156	$189,091	**$1,131,792**
5B	CO2 COMPRESSION	$72,976	$0	$27,323	$0	$0	**$100,299**	$9,363	$0	$21,933	**$131,595**
5C	METHANOL PRODUCTION	$111,584	$70,884	$129,178	$0	$0	**$311,646**	$31,165	$0	$68,562	**$411,373**
6	COMBUSTION TURBINE/ACCESSORIES										
6.1	Combustion Turbine Generator	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**
6.2-6.9	Combustion Turbine Other	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**
	SUBTOTAL 6	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
7	HRSG, DUCTING & STACK										
7.1	Heat Recovery Steam Generator	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**
7.2-7.9	HRSG Accessories, Ductwork, & Stack	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**
	SUBTOTAL 7	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	$5,787	$0	$553	$0	$0	**$6,340**	$555	$0	$689	**$7,584**
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	$4,451	$169	$2,048	$0	$0	**$6,669**	$569	$0	$1,412	**$8,650**
	SUBTOTAL 8	**$10,238**	**$169**	**$2,601**	**$0**	**$0**	**$13,008**	**$1,124**	**$0**	**$2,102**	**$16,235**
9	COOLING WATER SYSTEM	$10,721	$11,721	$8,731	$0	$0	**$31,173**	$2,781	$0	$6,897	**$40,852**
10	ASH/SPENT SORBENT HANDLING SYS	$75,457	$2,853	$86,813	$0	$0	**$165,123**	$14,141	$0	$18,537	**$197,802**
11	ACCESSORY ELECTRIC PLANT	$32,826	$21,307	$37,153	$0	$0	**$91,286**	$7,757	$0	$19,899	**$118,942**
12	INSTRUMENTATION & CONTROL	$12,582	$3,803	$10,553	$0	$0	**$26,938**	$2,368	$1,347	$5,491	**$36,143**
13	IMPROVEMENTS TO SITE	$5,333	$3,143	$13,997	$0	$0	**$22,473**	$2,225	$0	$7,409	**$32,108**
14	BUILDINGS & STRUCTURES	$0	$6,890	$7,322	$0	$0	**$14,212**	$1,262	$0	$2,600	**$18,073**
	TOTAL COST	**$1,459,049**	**$383,636**	**$721,451**	**$0**	**$0**	**$2,564,137**	**$239,414**	**$137,620**	**$555,800**	**$3,496,970**

Exhibit 5-2: Total Plant Cost Summary OSU CLG Case 1 (Continued)

Client: Ohio State University
Project: Coal to Methanol via Chemical Looping Gasification
Report Date: 2014-Dec-30

TOTAL PLANT COST SUMMARY

Case: Case 1 - 100% Coal Feedstock
Plant Size: -325.0 MW,net
Estimate Type: Conceptual
Cost Base (Jun) 2011 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		TOTAL PLANT COST $
				Direct	Indirect				Process	Project	
1	COAL & SORBENT HANDLING										
1.1	Coal Receive & Unload	$8,743	$0	$4,214	$0	$0	$12,958	$1,126	$0	$2,817	$16,901
1.2	Coal Stackout & Reclaim	$11,299	$0	$2,702	$0	$0	$14,000	$1,188	$0	$3,038	$18,226
1.3	Coal Conveyors & Yd Crush	$10,505	$0	$2,673	$0	$0	$13,178	$1,120	$0	$2,860	$17,158
1.4	Other Coal Handling	$2,748	$0	$619	$0	$0	$3,367	$285	$0	$730	$4,382
1.5	Switchgrass Receive & Unload	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.6	Switchgrass Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.7	Switchgrass Conveyors	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.8	Other Switchgrass Handling	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.9	Coal & Switchgrass Hnd.Foundations	$0	$5,838	$15,259	$0	$0	$21,097	$2,005	$0	$4,620	$27,723
	SUBTOTAL 1.	**$33,295**	**$5,838**	**$25,466**	**$0**	**$0**	**$64,599**	**$5,725**	**$0**	**$14,065**	**$84,389**
2	COAL & SORBENT PREP & FEED										
2.1	Coal Crushing	$5,928	$106,183	$37,250	$0	$0	$149,361	$12,220	$0	$32,316	$193,897
2.2	Prepared Coal Storage & Feed	$8,336	$2,003	$1,289	$0	$0	$11,628	$961	$0	$2,518	$15,107
2.3	Coal Injection System	$9,883	$1,196	$29,317	$0	$0	$40,396	$3,772	$0	$8,834	$53,001
2.4	Misc.Coal Prep & Feed	$4,584	$3,349	$9,865	$0	$0	$17,797	$1,591	$0	$3,878	$23,266
2.5	Switchgrass Prep Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.6	Switchgrass Storage & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.7	Switchgrass Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.9	Coal & Switchgrass Feed Foundation	$0	$16,814	$14,428	$0	$0	$31,242	$2,902	$0	$6,829	$40,974
	SUBTOTAL 2.	**$28,731**	**$129,544**	**$92,149**	**$0**	**$0**	**$250,424**	**$21,446**	**$0**	**$54,374**	**$326,244**
3	FEEDWATER & MISC. BOP SYSTEMS										
3.1	Feedwater System	$3,313	$1,547	$1,864	$0	$0	$6,724	$594	$0	$1,463	$8,781
3.2	Water Makeup & Pretreating	$675	$70	$371	$0	$0	$1,115	$103	$0	$365	$1,583
3.3	Other Feedwater Subsystems	$3,696	$1,093	$1,775	$0	$0	$6,564	$574	$0	$1,427	$8,565
3.4	Service Water Systems	$394	$787	$2,712	$0	$0	$3,893	$370	$0	$1,279	$5,541
3.5	Other Boiler Plant Systems	$2,126	$794	$1,955	$0	$0	$4,876	$444	$0	$1,064	$6,384
3.6	FO Supply Sys & Nat Gas	$123	$253	$225	$0	$0	$601	$56	$0	$132	$789
3.7	Waste Treatment Equipment	$913	$0	$565	$0	$0	$1,478	$142	$0	$486	$2,106
3.8	Misc. Power Plant Equipment	$1,943	$259	$1,010	$0	$0	$3,212	$306	$0	$1,055	$4,574
	SUBTOTAL 3.	**$13,182**	**$4,803**	**$10,477**	**$0**	**$0**	**$28,462**	**$2,589**	**$0**	**$7,272**	**$38,323**
4	GASIFIER & ACCESSORIES										
4.1	Reducer Reactor, Oxidizer, Xfer & Ash Separator	$94,830	$47,415	$94,830	$0	$0	$237,076	$23,708	$39,118	$44,985	$344,886
4.2	Syngas Cooler	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.3	Air Separation Unit	$0	$0	w/equip.	$0	$0	$0	$0	$0	$0	$0
4.4	Low Temperature Cooling	$48,003	$0	$10,017	$0	$0	$58,020	$5,529	$0	$12,710	$76,258
4.5	Black Water & Sour Gas Section	$6,927	$2,717	$5,304	$0	$0	$14,949	$0	$0	$0	$14,949
4.6	Soot Recovery & SARU	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.7	Looping System Auxiliary Components	$213,934	$46,327	$92,654	$0	$0	$352,914	$35,291	$0	$77,641	$465,847
4.8	Major Component Rigging (w/4.1)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.9	Gasification Foundations	$0	$2,972	$5,186	$0	$0	$8,158	$769	$0	$2,232	$11,159
	SUBTOTAL 4.	**$363,694**	**$99,431**	**$207,991**	**$0**	**$0**	**$671,117**	**$65,297**	**$39,118**	**$137,568**	**$913,099**

Exhibit 5-2: Total Plant Cost Summary OSU CLG Case 1 (Continued)

Client: Ohio State University **Report Date:** 2014-Dec-30
Project: Coal to Methanol via Chemical Looping Gasification

TOTAL PLANT COST SUMMARY

Case: Case 1 - 100% Coal Feedstock
Plant Size: -325.0 MW,net **Estimate Type:** Conceptual **Cost Base (Jun)** 2011 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		TOTAL PLANT COST $
				Direct	Indirect				Process	Project	
5A	GAS CLEANUP & PIPING										
5A.1	Rectisol System	$483,425	$0	w/equip	$0	$0	$483,425	$44,419	$96,685	$124,906	$749,435
5A.2	Elemental Sulfur Plant	$10,437	$2,034	$13,374	$0	$0	$25,845	$2,455	$0	$5,660	$33,960
5A.3	Mercury Removal	$5,360	$0	$4,051	$0	$0	$9,412	$887	$471	$2,154	$12,923
5A.4a	COS Hydrolysis	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5A.4b	Shift Reactors	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5A.4c	Particulate Removal	$96,273	$0	$11,255	$0	$0	$107,528	$9,941	$0	$23,494	$140,962
5A.5	Syngas Compressor	$72,116	$14,423	$28,847	$0	$0	$115,386	$11,539	$0	$25,385	$152,310
5A.6	Blowback Gas Systems	$20,818	$3,504	$1,959	$0	$0	$26,280	$2,425	$0	$5,741	$34,447
5A.7	Fuel Gas Piping	$0	$277	$181	$0	$0	$458	$40	$0	$100	$598
5A.9	HGCU Foundations	$0	$3,010	$2,029	$0	$0	$5,039	$466	$0	$1,651	$7,156
	SUBTOTAL 5A.	**$688,430**	**$23,249**	**$61,696**	**$0**	**$0**	**$773,375**	**$72,171**	**$97,156**	**$189,091**	**$1,131,792**
5B	CO2 COMPRESSION										
5B.1	CO2 Removal System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5B.2	CO2 Compression & Drying	$72,976	$0	$27,323	$0	$0	$100,299	$9,363	$0	$21,933	$131,595
	SUBTOTAL 5B.	**$72,976**	**$0**	**$27,323**	**$0**	**$0**	**$100,299**	**$9,363**	**$0**	**$21,933**	**$131,595**
5C	METHANOL PRODUCTION										
5C.1	Methanol Synthesis	$108,631	$45,703	$91,406	$0	$0	$245,741	$24,574	$0	$54,063	$324,378
5C.2	Gasoline Synthesis	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5C.3	Gasoline Refining	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5C.4	Process Interconnects	$0	$24,000	$36,000	$0	$0	$60,000	$6,000	$0	$13,200	$79,200
5C.5	Direct Fired Boiler	$2,953	$1,181	$1,772	$0	$0	$5,905	$591	$0	$1,299	$7,795
	SUBTOTAL 5C.	**$111,584**	**$70,884**	**$129,178**	**$0**	**$0**	**$311,646**	**$31,165**	**$0**	**$68,562**	**$411,373**
6	COMBUSTION TURBINE/ACCESSORIES										
6.1	Combustion Turbine Generator	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 6.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
7	HRSG, DUCTING & STACK										
7.1	Heat Recovery Steam Generator	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.2	HRSG Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.3	Ductwork	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.4	Stack	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.9	HRSG,Duct & Stack Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 7.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	$5,787	$0	$553	$0	$0	$6,340	$555	$0	$689	$7,584
8.2	Turbine Plant Auxiliaries	$28	$0	$82	$0	$0	$110	$11	$0	$12	$133
8.3	Condenser & Auxiliaries	$2,206	$0	$262	$0	$0	$2,469	$228	$0	$270	$2,967
8.4	Steam Piping	$2,216	$0	$1,405	$0	$0	$3,621	$286	$0	$977	$4,885
8.9	TG Foundations	$0	$169	$299	$0	$0	$469	$44	$0	$154	$667
	SUBTOTAL 8.	**$10,238**	**$169**	**$2,601**	**$0**	**$0**	**$13,008**	**$1,124**	**$0**	**$2,102**	**$16,235**

Exhibit 5-2: Total Plant Cost Summary OSU CLG Case 1 (Continued)

Client: Ohio State University
Project: Coal to Methanol via Chemical Looping Gasification
Report Date: 2014-Dec-30

TOTAL PLANT COST SUMMARY

Case: Case 1 - 100% Coal Feedstock
Plant Size: -325.0 MW,net **Estimate Type:** Conceptual **Cost Base (Jun)** 2011 ($x1000)

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		TOTAL PLANT COST
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$
9	COOLING WATER SYSTEM										
9.1	Cooling Towers	$6,997	$0	$1,264	$0	$0	$8,262	$766	$0	$1,354	$10,382
9.2	Circulating Water Pumps	$2,198	$0	$67	$0	$0	$2,265	$192	$0	$369	$2,826
9.3	Circ.Water System Auxiliaries	$187	$0	$27	$0	$0	$214	$20	$0	$35	$269
9.4	Circ.Water Piping	$0	$8,308	$2,012	$0	$0	$10,320	$865	$0	$2,237	$13,422
9.5	Make-up Water System	$388	$0	$534	$0	$0	$922	$85	$0	$201	$1,209
9.6	Component Cooling Water Sys	$950	$1,136	$780	$0	$0	$2,866	$256	$0	$624	$3,746
9.9	Circ.Water System Foundations	$0	$2,278	$4,047	$0	$0	$6,325	$597	$0	$2,077	$8,998
	SUBTOTAL 9.	**$10,721**	**$11,721**	**$8,731**	**$0**	**$0**	**$31,173**	**$2,781**	**$0**	**$6,897**	**$40,852**
10	ASH/SPENT SORBENT HANDLING SYS										
10.1	Slag Dewatering & Cooling	$70,397	$0	$68,958	$0	$0	$139,355	$13,177	$0	$15,253	$167,785
10.2	Gasifier Ash Depressurization	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.3	Cleanup Ash Depressurization	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.4	High Temperature Ash Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.5	Slag Handling System	$0	$0	$15,326	$0	$0	$15,326	$0	$0	$1,533	$16,859
10.6	Ash Storage Silos	$1,082	$0	$1,169	$0	$0	$2,251	$213	$0	$370	$2,834
10.7	Ash Transport & Feed Equipment	$1,737	$0	$406	$0	$0	$2,143	$191	$0	$350	$2,684
10.8	Misc. Ash Handling Equipment	$2,241	$2,747	$815	$0	$0	$5,802	$538	$0	$951	$7,291
10.9	Ash/Spent Sorbent Foundation	$0	$106	$139	$0	$0	$245	$23	$0	$80	$349
	SUBTOTAL 10.	**$75,457**	**$2,853**	**$86,813**	**$0**	**$0**	**$165,123**	**$14,141**	**$0**	**$18,537**	**$197,802**
11	ACCESSORY ELECTRIC PLANT										
11.1	Generator Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
11.2	Station Service Equipment	$7,335	$0	$674	$0	$0	$8,010	$737	$0	$875	$9,621
11.3	Switchgear & Motor Control	$13,539	$0	$2,517	$0	$0	$16,056	$1,485	$0	$2,631	$20,172
11.4	Conduit & Cable Tray	$0	$6,879	$21,205	$0	$0	$28,084	$2,617	$0	$7,675	$38,377
11.5	Wire & Cable	$0	$13,263	$8,069	$0	$0	$21,333	$1,448	$0	$5,695	$28,476
11.6	Protective Equipment	$0	$949	$3,524	$0	$0	$4,473	$430	$0	$735	$5,638
11.7	Standby Equipment	$372	$0	$371	$0	$0	$744	$70	$0	$122	$936
11.8	Main Power Transformers	$11,579	$0	$206	$0	$0	$11,785	$894	$0	$1,902	$14,581
11.9	Electrical Foundations	$0	$216	$587	$0	$0	$802	$76	$0	$264	$1,142
	SUBTOTAL 11.	**$32,826**	**$21,307**	**$37,153**	**$0**	**$0**	**$91,286**	**$7,757**	**$0**	**$19,899**	**$118,942**
12	INSTRUMENTATION & CONTROL										
12.1	IGCC Control Equipment	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0
12.2	Combustion Turbine Control	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	$0	$0	$0
12.4	Other Major Component Control	$622	$0	$424	$0	$0	$1,046	$98	$52	$180	$1,376
12.5	Signal Processing Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards,Panels & Racks	$413	$0	$270	$0	$0	$684	$64	$34	$156	$938
12.7	Computer & Accessories	$6,615	$0	$216	$0	$0	$6,831	$626	$342	$780	$8,578
12.8	Instrument Wiring & Tubing	$0	$3,803	$7,199	$0	$0	$11,002	$890	$550	$3,110	$15,552
12.9	Other I & C Equipment	$4,932	$0	$2,444	$0	$0	$7,375	$689	$369	$1,265	$9,698
	SUBTOTAL 12.	**$12,582**	**$3,803**	**$10,553**	**$0**	**$0**	**$26,938**	**$2,368**	**$1,347**	**$5,491**	**$36,143**

Exhibit 5-2: Total Plant Cost Summary OSU CLG Case 1 (Continued)

Client: Ohio State University
Project: Coal to Methanol via Chemical Looping Gasification
Report Date: 2014-Dec-30

TOTAL PLANT COST SUMMARY

Case: Case 1 - 100% Coal Feedstock
Plant Size: -325.0 MW,net **Estimate Type:** Conceptual **Cost Base (Jun)** 2011 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $
13	IMPROVEMENTS TO SITE										
13.1	Site Preparation	$0	$167	$3,804	$0	$0	$3,972	$387	$0	$1,308	$5,666
13.2	Site Improvements	$0	$2,976	$4,207	$0	$0	$7,183	$712	$0	$2,369	$10,264
13.3	Site Facilities	$5,333	$0	$5,986	$0	$0	$11,319	$1,126	$0	$3,733	$16,178
	SUBTOTAL 13.	**$5,333**	**$3,143**	**$13,997**	**$0**	**$0**	**$22,473**	**$2,225**	**$0**	**$7,409**	**$32,108**
14	BUILDINGS & STRUCTURES										
14.1	Combustion Turbine Area	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.2	Steam Turbine Building	$0	$701	$998	$0	$0	$1,699	$153	$0	$278	$2,130
14.3	Administration Building	$0	$1,043	$754	$0	$0	$1,797	$157	$0	$293	$2,247
14.4	Circulation Water Pumphouse	$0	$349	$196	$0	$0	$544	$47	$0	$89	$680
14.5	Water Treatment Buildings	$0	$1,321	$1,288	$0	$0	$2,609	$231	$0	$426	$3,265
14.6	Machine Shop	$0	$569	$390	$0	$0	$958	$83	$0	$156	$1,198
14.7	Warehouse	$0	$913	$584	$0	$0	$1,497	$130	$0	$244	$1,870
14.8	Other Buildings & Structures	$0	$617	$480	$0	$0	$1,097	$96	$0	$238	$1,431
14.9	Waste Treating Building & Str.	$0	$1,379	$2,633	$0	$0	$4,011	$366	$0	$875	$5,253
	SUBTOTAL 14.	**$0**	**$6,890**	**$7,322**	**$0**	**$0**	**$14,212**	**$1,262**	**$0**	**$2,600**	**$18,073**
	TOTAL COST	**$1,459,049**	**$383,636**	**$721,451**	**$0**	**$0**	**$2,564,137**	**$239,414**	**$137,620**	**$555,800**	**$3,496,970**

Exhibit 5-3: Total Plant Cost Summary OSU CLG Case 2

Client: Ohio State University
Project: Coal to Methanol via Chemical Looping Gasification
Report Date: 2014-Dec-30

TOTAL PLANT COST SUMMARY

Case: Case 2 - 50% Coal / 50% Natural Gas Feedstock
Plant Size: -246.0 MW,net **Estimate Type:** Conceptual **Cost Base (Jun)** 2011 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $
1	COAL & SORBENT HANDLING	$22,064	$3,869	$16,876	$0	$0	**$42,808**	$3,794	$0	$9,320	**$55,923**
2	COAL & SORBENT PREP & FEED	$18,540	$83,596	$59,465	$0	$0	**$161,602**	$13,839	$0	$35,088	**$210,530**
3	FEEDWATER & MISC. BOP SYSTEMS	$12,606	$5,212	$9,730	$0	$0	**$27,548**	$2,510	$0	$7,035	**$37,094**
4	GASIFIER & ACCESSORIES										
4.1	Reducer Reactor, Oxidizer, Xfer & Ash Separator	$79,857	$39,929	$79,857	$0	$0	**$199,643**	$19,964	$32,941	$37,882	**$290,431**
4.2	Syngas Cooler	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**
4.3	Air Separation Unit	$0	$0	w/equip.	$0	$0	**$0**	$0	$0	$0	**$0**
4.4-4.9	Other Gasification Equipment	$235,285	$46,666	$100,310	$0	$0	**$382,261**	$36,920	$0	$82,190	**$501,370**
	SUBTOTAL 4	**$315,142**	**$86,595**	**$180,167**	**$0**	**$0**	**$581,904**	**$56,884**	**$32,941**	**$120,072**	**$791,801**
5A	GAS CLEANUP & PIPING	$623,193	$21,404	$52,511	$0	$0	**$697,107**	$65,063	$89,907	$170,966	**$1,023,043**
5B	CO2 COMPRESSION	$44,005	$0	$16,476	$0	$0	**$60,481**	$5,646	$0	$13,226	**$79,353**
5C	METHANOL PRODUCTION	$112,111	$71,116	$129,737	$0	$0	**$312,963**	$31,296	$0	$68,852	**$413,112**
6	COMBUSTION TURBINE/ACCESSORIES										
6.1	Combustion Turbine Generator	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**
6.2-6.9	Combustion Turbine Other	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**
	SUBTOTAL 6	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
7	HRSG, DUCTING & STACK										
7.1	Heat Recovery Steam Generator	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**
7.2-7.9	HRSG Accessories, Ductwork, & Stack	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**
	SUBTOTAL 7	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	$7,925	$0	$660	$0	$0	**$8,585**	$751	$0	$934	**$10,269**
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	$4,254	$235	$2,291	$0	$0	**$6,781**	$581	$0	$1,462	**$8,823**
	SUBTOTAL 8	**$12,180**	**$235**	**$2,951**	**$0**	**$0**	**$15,365**	**$1,332**	**$0**	**$2,395**	**$19,092**
9	COOLING WATER SYSTEM	$10,097	$11,123	$8,277	$0	$0	**$29,497**	$2,631	$0	$6,531	**$38,659**
10	ASH/SPENT SORBENT HANDLING SYS	$49,536	$1,974	$56,950	$0	$0	**$108,460**	$9,284	$0	$12,197	**$129,942**
11	ACCESSORY ELECTRIC PLANT	$30,765	$19,104	$33,659	$0	$0	**$83,527**	$7,092	$0	$18,095	**$108,714**
12	INSTRUMENTATION & CONTROL	$12,857	$3,886	$10,783	$0	$0	**$27,526**	$2,419	$1,376	$5,611	**$36,933**
13	IMPROVEMENTS TO SITE	$5,408	$3,188	$14,195	$0	$0	**$22,791**	$2,257	$0	$7,514	**$32,562**
14	BUILDINGS & STRUCTURES	$0	$7,214	$7,786	$0	$0	**$15,001**	$1,333	$0	$2,731	**$19,065**
	TOTAL COST	**$1,268,503**	**$318,516**	**$599,563**	**$0**	**$0**	**$2,186,582**	**$205,381**	**$124,224**	**$479,634**	**$2,995,821**

Exhibit 5-3: Total Plant Cost Summary OSU CLG Case 2 (Continued)

Client: Ohio State University
Project: Coal to Methanol via Chemical Looping Gasification
Report Date: 2014-Dec-30

TOTAL PLANT COST SUMMARY

Case: Case 2 - 50% Coal / 50% Natural Gas Feedstock
Plant Size: -246.0 MW,net **Estimate Type:** Conceptual Cost Base (Jun) 2011 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		TOTAL PLANT COST $
				Direct	Indirect				Process	Project	
1	COAL & SORBENT HANDLING										
1.1	Coal Receive & Unload	$5,794	$0	$2,793	$0	$0	$8,587	$746	$0	$1,867	$11,200
1.2	Coal Stackout & Reclaim	$7,487	$0	$1,790	$0	$0	$9,277	$787	$0	$2,013	$12,078
1.3	Coal Conveyors & Yd Crush	$6,961	$0	$1,771	$0	$0	$8,732	$742	$0	$1,895	$11,370
1.4	Other Coal Handling	$1,821	$0	$410	$0	$0	$2,231	$189	$0	$484	$2,904
1.5	Switchgrass Receive & Unload	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.6	Switchgrass Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.7	Switchgrass Conveyors	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.8	Other Switchgrass Handling	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.9	Coal & Switchgrass Hnd.Foundations	$0	$3,869	$10,112	$0	$0	$13,981	$1,329	$0	$3,062	$18,371
	SUBTOTAL 1.	**$22,064**	**$3,869**	**$16,876**	**$0**	**$0**	**$42,808**	**$3,794**	**$0**	**$9,320**	**$55,923**
2	COAL & SORBENT PREP & FEED										
2.1	Coal Crushing	$3,825	$68,521	$24,038	$0	$0	$96,385	$7,885	$0	$20,854	$125,124
2.2	Prepared Coal Storage & Feed	$5,379	$1,292	$832	$0	$0	$7,504	$620	$0	$1,625	$9,749
2.3	Coal Injection System	$6,378	$772	$18,919	$0	$0	$26,068	$2,434	$0	$5,700	$34,202
2.4	Misc.Coal Prep & Feed	$2,958	$2,161	$6,366	$0	$0	$11,485	$1,027	$0	$2,502	$15,014
2.5	Switchgrass Prep Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.6	Switchgrass Storage & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.7	Switchgrass Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.9	Coal & Switchgrass Feed Foundation	$0	$10,850	$9,311	$0	$0	$20,161	$1,873	$0	$4,407	$26,441
	SUBTOTAL 2.	**$18,540**	**$83,596**	**$59,465**	**$0**	**$0**	**$161,602**	**$13,839**	**$0**	**$35,088**	**$210,530**
3	FEEDWATER & MISC. BOP SYSTEMS										
3.1	Feedwater System	$2,941	$1,373	$1,654	$0	$0	$5,968	$527	$0	$1,299	$7,794
3.2	Water Makeup & Pretreating	$632	$65	$347	$0	$0	$1,044	$96	$0	$342	$1,483
3.3	Other Feedwater Subsystems	$3,280	$970	$1,576	$0	$0	$5,826	$509	$0	$1,267	$7,603
3.4	Service Water Systems	$369	$737	$2,540	$0	$0	$3,646	$346	$0	$1,198	$5,190
3.5	Other Boiler Plant Systems	$1,991	$744	$1,831	$0	$0	$4,567	$416	$0	$996	$5,979
3.6	FO Supply Sys & Nat Gas	$560	$1,059	$225	$0	$0	$1,844	$171	$0	$403	$2,418
3.7	Waste Treatment Equipment	$855	$0	$529	$0	$0	$1,384	$133	$0	$455	$1,973
3.8	Misc. Power Plant Equipment	$1,977	$264	$1,027	$0	$0	$3,268	$312	$0	$1,074	$4,654
	SUBTOTAL 3.	**$12,606**	**$5,212**	**$9,730**	**$0**	**$0**	**$27,548**	**$2,510**	**$0**	**$7,035**	**$37,094**
4	GASIFIER & ACCESSORIES										
4.1	Reducer Reactor, Oxidizer, Xfer & Ash Separator	$79,857	$39,929	$79,857	$0	$0	$199,643	$19,964	$32,941	$37,882	$290,431
4.2	Syngas Cooler	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.3	Air Separation Unit	$0	$0	w/equip.	$0	$0	$0	$0	$0	$0	$0
4.4	Low Temperature Cooling	$37,509	$0	$7,827	$0	$0	$45,337	$4,320	$0	$9,931	$59,588
4.5	Black Water & Sour Gas Section	$4,849	$1,902	$3,713	$0	$0	$10,464	$0	$0	$0	$10,464
4.6	Soot Recovery & SARU	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.7	Looping System Auxiliary Components	$192,926	$41,792	$83,584	$0	$0	$318,302	$31,830	$0	$70,027	$420,159
4.8	Major Component Rigging (w/4.1)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.9	Gasification Foundations	$0	$2,972	$5,186	$0	$0	$8,158	$769	$0	$2,232	$11,159
	SUBTOTAL 4.	**$315,142**	**$86,595**	**$180,167**	**$0**	**$0**	**$581,904**	**$56,884**	**$32,941**	**$120,072**	**$791,801**

OHIO STATE UNIVERSITY TECHNO-ECONOMIC ANALYSIS, PRELIMINARY REPORT

Exhibit 5-3: Total Plant Cost Summary OSU CLG Case 2 (Continued)

Client: Ohio State University
Project: Coal to Methanol via Chemical Looping Gasification
Report Date: 2014-Dec-30

TOTAL PLANT COST SUMMARY

Case: Case 2 - 50% Coal / 50% Natural Gas Feedstock
Plant Size: -246.0 MW,net **Estimate Type:** Conceptual **Cost Base (Jun)** 2011 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		TOTAL PLANT COST $
				Direct	Indirect				Process	Project	
5A	GAS CLEANUP & PIPING										
5A.1	Rectisol System	$447,332	$0	w/equip	$0	$0	$447,332	$41,102	$89,466	$115,580	$693,482
5A.2	Elemental Sulfur Plant	$6,700	$1,306	$8,585	$0	$0	$16,590	$1,576	$0	$3,633	$21,799
5A.3	Mercury Removal	$5,015	$0	$3,790	$0	$0	$8,805	$830	$440	$2,015	$12,090
5A.4a	COS Hydrolysis	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5A.4b	Shift Reactors	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5A.4c	Particulate Removal	$77,018	$0	$9,004	$0	$0	$86,022	$7,953	$0	$18,795	$112,770
5A.5	Syngas Compressor	$67,744	$13,549	$27,098	$0	$0	$108,390	$10,839	$0	$23,846	$143,075
5A.6	Blowback Gas Systems	$19,383	$3,262	$1,824	$0	$0	$24,469	$2,258	$0	$5,346	$32,073
5A.7	Fuel Gas Piping	$0	$277	$181	$0	$0	$458	$40	$0	$100	$598
5A.9	HGCU Foundations	$0	$3,010	$2,029	$0	$0	$5,039	$466	$0	$1,651	$7,156
	SUBTOTAL 5A.	**$623,193**	**$21,404**	**$52,511**	**$0**	**$0**	**$697,107**	**$65,063**	**$89,907**	**$170,966**	**$1,023,043**
5B	CO2 COMPRESSION										
5B.1	CO2 Removal System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5B.2	CO2 Compression & Drying	$44,005	$0	$16,476	$0	$0	$60,481	$5,646	$0	$13,226	$79,353
	SUBTOTAL 5B.	**$44,005**	**$0**	**$16,476**	**$0**	**$0**	**$60,481**	**$5,646**	**$0**	**$13,226**	**$79,353**
5C	METHANOL PRODUCTION										
5C.1	Methanol Synthesis	$109,637	$46,126	$92,252	$0	$0	$248,015	$24,802	$0	$54,563	$327,380
5C.2	Gasoline Synthesis	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5C.3	Gasoline Refining	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5C.4	Process Interconnects	$0	$24,000	$36,000	$0	$0	$60,000	$6,000	$0	$13,200	$79,200
5C.5	Direct Fired Boiler	$2,474	$990	$1,484	$0	$0	$4,948	$495	$0	$1,089	$6,532
	SUBTOTAL 5C.	**$112,111**	**$71,116**	**$129,737**	**$0**	**$0**	**$312,963**	**$31,296**	**$0**	**$68,852**	**$413,112**
6	COMBUSTION TURBINE/ACCESSORIES										
6.1	Combustion Turbine Generator	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 6.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
7	HRSG, DUCTING & STACK										
7.1	Heat Recovery Steam Generator	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.2	HRSG Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.3	Ductwork	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.4	Stack	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.9	HRSG,Duct & Stack Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 7.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	$7,925	$0	$660	$0	$0	$8,585	$751	$0	$934	$10,269
8.2	Turbine Plant Auxiliaries	$39	$0	$114	$0	$0	$153	$15	$0	$17	$184
8.3	Condenser & Auxiliaries	$2,009	$0	$364	$0	$0	$2,373	$220	$0	$259	$2,852
8.4	Steam Piping	$2,206	$0	$1,398	$0	$0	$3,604	$285	$0	$972	$4,862
8.9	TG Foundations	$0	$235	$415	$0	$0	$650	$61	$0	$214	$925
	SUBTOTAL 8.	**$12,180**	**$235**	**$2,951**	**$0**	**$0**	**$15,365**	**$1,332**	**$0**	**$2,395**	**$19,092**

Exhibit 5-3: Total Plant Cost Summary OSU CLG Case 2 (Continued)

Client: Ohio State University
Project: Coal to Methanol via Chemical Looping Gasification
Report Date: 2014-Dec-30

TOTAL PLANT COST SUMMARY

Case: Case 2 - 50% Coal / 50% Natural Gas Feedstock
Plant Size: -246.0 MW,net **Estimate Type:** Conceptual **Cost Base (Jun)** 2011 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		TOTAL PLANT COST
				Direct	Indirect				Process	Project	$
9	COOLING WATER SYSTEM										
9.1	Cooling Towers	$6,582	$0	$1,189	$0	$0	$7,772	$721	$0	$1,274	$9,766
9.2	Circulating Water Pumps	$2,068	$0	$67	$0	$0	$2,135	$181	$0	$347	$2,664
9.3	Circ.Water System Auxiliaries	$178	$0	$25	$0	$0	$203	$19	$0	$33	$255
9.4	Circ.Water Piping	$0	$7,884	$1,910	$0	$0	$9,793	$821	$0	$2,123	$12,737
9.5	Make-up Water System	$367	$0	$505	$0	$0	$872	$81	$0	$191	$1,143
9.6	Component Cooling Water Sys	$901	$1,078	$740	$0	$0	$2,719	$243	$0	$592	$3,555
9.9	Circ.Water System Foundations	$0	$2,161	$3,841	$0	$0	$6,002	$566	$0	$1,970	$8,539
	SUBTOTAL 9.	**$10,097**	**$11,123**	**$8,277**	**$0**	**$0**	**$29,497**	**$2,631**	**$0**	**$6,531**	**$38,659**
10	ASH/SPENT SORBENT HANDLING SYS										
10.1	Slag Dewatering & Cooling	$46,035	$0	$45,094	$0	$0	$91,129	$8,617	$0	$9,975	$109,721
10.2	Gasifier Ash Depressurization	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.3	Cleanup Ash Depressurization	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.4	High Temperature Ash Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.5	Slag Handling System	$0	$0	$10,106	$0	$0	$10,106	$0	$0	$1,011	$11,117
10.6	Ash Storage Silos	$749	$0	$809	$0	$0	$1,558	$147	$0	$256	$1,961
10.7	Ash Transport & Feed Equipment	$1,202	$0	$281	$0	$0	$1,483	$132	$0	$242	$1,857
10.8	Misc. Ash Handling Equipment	$1,551	$1,900	$564	$0	$0	$4,015	$372	$0	$658	$5,045
10.9	Ash/Spent Sorbent Foundation	$0	$73	$96	$0	$0	$170	$16	$0	$56	$241
	SUBTOTAL 10.	**$49,536**	**$1,974**	**$56,950**	**$0**	**$0**	**$108,460**	**$9,284**	**$0**	**$12,197**	**$129,942**
11	ACCESSORY ELECTRIC PLANT										
11.1	Generator Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
11.2	Station Service Equipment	$6,524	$0	$600	$0	$0	$7,124	$655	$0	$778	$8,557
11.3	Switchgear & Motor Control	$12,042	$0	$2,238	$0	$0	$14,280	$1,321	$0	$2,340	$17,941
11.4	Conduit & Cable Tray	$0	$6,118	$18,860	$0	$0	$24,978	$2,328	$0	$6,827	$34,133
11.5	Wire & Cable	$0	$11,797	$7,177	$0	$0	$18,974	$1,288	$0	$5,065	$25,327
11.6	Protective Equipment	$0	$969	$3,598	$0	$0	$4,566	$439	$0	$751	$5,756
11.7	Standby Equipment	$378	$0	$376	$0	$0	$754	$71	$0	$124	$949
11.8	Main Power Transformers	$11,821	$0	$210	$0	$0	$12,031	$913	$0	$1,942	$14,886
11.9	Electrical Foundations	$0	$220	$599	$0	$0	$819	$78	$0	$269	$1,166
	SUBTOTAL 11.	**$30,765**	**$19,104**	**$33,659**	**$0**	**$0**	**$83,527**	**$7,092**	**$0**	**$18,095**	**$108,714**
12	INSTRUMENTATION & CONTROL										
12.1	IGCC Control Equipment	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0
12.2	Combustion Turbine Control	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	$0	$0	$0
12.4	Other Major Component Control	$636	$0	$433	$0	$0	$1,069	$100	$53	$183	$1,407
12.5	Signal Processing Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards,Panels & Racks	$422	$0	$276	$0	$0	$699	$65	$35	$160	$959
12.7	Computer & Accessories	$6,759	$0	$221	$0	$0	$6,980	$640	$349	$797	$8,766
12.8	Instrument Wiring & Tubing	$0	$3,886	$7,356	$0	$0	$11,242	$910	$562	$3,178	$15,892
12.9	Other I & C Equipment	$5,040	$0	$2,497	$0	$0	$7,536	$704	$377	$1,293	$9,910
	SUBTOTAL 12.	**$12,857**	**$3,886**	**$10,783**	**$0**	**$0**	**$27,526**	**$2,419**	**$1,376**	**$5,611**	**$36,933**

Exhibit 5-3: Total Plant Cost Summary OSU CLG Case 2 (Continued)

Client: Ohio State University
Project: Coal to Methanol via Chemical Looping Gasification
Report Date: 2014-Dec-30

TOTAL PLANT COST SUMMARY

Case: Case 2 - 50% Coal / 50% Natural Gas Feedstock
Plant Size: -246.0 MW,net **Estimate Type:** Conceptual **Cost Base (Jun)** 2011 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		TOTAL PLANT COST
				Direct	Indirect				Process	Project	$
13	IMPROVEMENTS TO SITE										
13.1	Site Preparation	$0	$170	$3,858	$0	$0	$4,028	$393	$0	$1,326	$5,746
13.2	Site Improvements	$0	$3,018	$4,266	$0	$0	$7,284	$722	$0	$2,402	$10,409
13.3	Site Facilities	$5,408	$0	$6,071	$0	$0	$11,479	$1,141	$0	$3,786	$16,406
	SUBTOTAL 13.	**$5,408**	**$3,188**	**$14,195**	**$0**	**$0**	**$22,791**	**$2,257**	**$0**	**$7,514**	**$32,562**
14	BUILDINGS & STRUCTURES										
14.1	Combustion Turbine Area	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.2	Steam Turbine Building	$0	$1,006	$1,432	$0	$0	$2,438	$220	$0	$399	$3,056
14.3	Administration Building	$0	$1,043	$754	$0	$0	$1,797	$157	$0	$293	$2,247
14.4	Circulation Water Pumphouse	$0	$349	$196	$0	$0	$544	$47	$0	$89	$680
14.5	Water Treatment Buildings	$0	$1,321	$1,288	$0	$0	$2,609	$231	$0	$426	$3,265
14.6	Machine Shop	$0	$569	$390	$0	$0	$958	$83	$0	$156	$1,198
14.7	Warehouse	$0	$913	$584	$0	$0	$1,497	$130	$0	$244	$1,870
14.8	Other Buildings & Structures	$0	$623	$485	$0	$0	$1,107	$97	$0	$241	$1,445
14.9	Waste Treating Building & Str.	$0	$1,392	$2,659	$0	$0	$4,051	$370	$0	$884	$5,304
	SUBTOTAL 14.	**$0**	**$7,214**	**$7,786**	**$0**	**$0**	**$15,001**	**$1,333**	**$0**	**$2,731**	**$19,065**
	TOTAL COST	**$1,268,503**	**$318,516**	**$599,563**	**$0**	**$0**	**$2,186,582**	**$205,381**	**$124,224**	**$479,634**	**$2,995,821**

Exhibit 5-4: Initial and Annual Operating and Maintenance Costs OSU CLG Case 1

INITIAL & ANNUAL O&M EXPENSES

Case 1 - 100% Coal Feedstock		Cost Base (Jun):	2011
LPG (bbl/hr):	0	MWe-net:	-325
Methanol (bbl/hr):	3,300	Capacity Factor (%):	90

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate(base):	39.70	$/hour
Operating Labor Burden:	30.00	% of base
Labor O-H Charge Rate:	25.00	% of labor

		Total
Skilled Operator	2.0	2.0
Operator	10.0	10.0
Foreman	1.0	1.0
Lab Tech's, etc.	3.0	3.0
TOTAL-O.J.'s	16.0	16.0

	Annual Cost $	Annual Unit Cost $/bbl
Annual Operating Labor Cost	$7,233,658	$0.278
Maintenance Labor Cost	$23,447,898	$0.901
Administrative & Support Labor	$7,670,389	$0.295
Property Taxes and Insurance	$69,939,401	$2.689
TOTAL FIXED OPERATING COSTS	**$108,291,347**	**$4.163**

VARIABLE OPERATING COSTS

		$/bbl
Maintenance Material Cost	**$35,171,848**	**$1.35204**

Consumables	Consumption Initial Fill	Consumption /Day	Unit Cost	Initial Fill Cost		
Water(/1000 gallons)	0	2,907	1.67	**$0**	**$1,594,963**	**$0.06131**
Chemicals						
MU & WT Chem. (lb)	0	17,321	0.27	$0	$1,536,305	$0.05906
Carbon (Mercury Removal) (lb)	422,862	644	1.63	$689,265	$344,633	$0.01325
Oxygen Carrier (tonne)	10,871	63	600.00	$6,522,870	$12,417,300	$0.47733
COS Catalyst (lb)	0	0	3,751.70	$0	$0	$0.00000
Water Gas Shift Catalyst (ft3)	0	0.00	771.99	$0	$0	$0.00000
SCR Catalyst (m3)	0	0	0.00	$0	$0	$0.00000
Aqueous Ammonia (ton)	0	0	0.00	$0	$0	$0.00000
Claus Catalyst (ft3)	w/equip	1.78	131.27	$0	$76,878	$0.00296
Methanol Synthesis Catalyst (ft3)	4,207	3.84	500.00	$2,103,500	$631,050	$0.02426
DME Synthesis Catalyst (ft3)	0	0.00	750.00	$0	$0	$0.00000
Gasoline Synthesis Catalyst (ft3)	0	0.00	1,200.00	$0	$0	$0.00000
Aromatics Hydrotreater Catalyst (ft3)	138	0.13	475.00	$65,550	$19,665	$0.00076
Subtotal Chemicals				**$9,381,185**	**$15,025,830**	**$0.57761**
Other						
Supplemental Fuel (MBtu)	0	0	0.00	$0	$0	$0.00000
Supplemental Electricity (for consumption) (MWh)	0	7,800	60.00	$0	$153,738,000	$5.90984
Gases, N2 etc. (/100scf)	0	0	0.00	$0	$0	$0.00000
L.P. Steam (/1000 pounds)	0	0	0.00	$0	$0	$0.00000
Subtotal Other				**$0**	**$153,738,000**	**$5.90984**
Waste Disposal						
Spent Mercury Catalyst (lb.)	0	644	0.65	$0	$137,430	$0.00528
Slag (ton)	0	0	0.00	$0	$0	$0.00000
Flyash (ton)	0	1,371	25.11	$0	$11,309,571	$0.43475
Subtotal-Waste Disposal				**$0**	**$11,447,002**	**$0.44003**
By-products & Emissions						
Sulfur (tons)	0	121	0.00	$0	$0	$0.00000
Hydrogen (mmscf)	0	14.0	0.00	$0	$0	$0.00000
Supplemental Electricity (for sale) (MWh)	0	0	58.00	$0	$0	$0.00000
Subtotal By-Products				**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS				**$9,381,185**	**$216,977,643**	**$8.34083**
Fuel						
Coal (ton)	0	16,746	36.57	**$0**	**$201,173,801**	**$7.73332**

Exhibit 5-5: Initial and Annual Operating and Maintenance Costs OSU CLG Case 2

INITIAL & ANNUAL O&M EXPENSES		Cost Base (Jun):	2011
Case 2 - 50% Coal / 50% Natural Gas Feedstock			
LPG (bbl/hr):	0	MWe-net:	-246
Methanol (bbl/hr):	3,300	Capacity Factor (%):	90

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate(base):	39.70	$/hour
Operating Labor Burden:	30.00	% of base
Labor O-H Charge Rate:	25.00	% of labor

		Total
Skilled Operator	2.0	2.0
Operator	10.0	10.0
Foreman	1.0	1.0
Lab Tech's, etc.	3.0	3.0
TOTAL-O.J.'s	16.0	16.0

	Annual Cost $	Annual Unit Cost $/bbl
Annual Operating Labor Cost	$7,233,658	$0.278
Maintenance Labor Cost	$20,060,645	$0.771
Administrative & Support Labor	$6,823,576	$0.262
Property Taxes and Insurance	$59,916,418	$2.303
TOTAL FIXED OPERATING COSTS	**$94,034,297**	**$3.615**

VARIABLE OPERATING COSTS

		$/bbl
Maintenance Material Cost	**$30,090,968**	**$1.15673**

Consumables	Consumption Initial Fill	Consumption /Day	Unit Cost	Initial Fill Cost		
Water(/1000 gallons)	0	2,651	1.67	**$0**	**$1,454,347**	**$0.05591**
Chemicals						
MU & WT Chem. (lb)	0	15,794	0.27	$0	$1,400,860	$0.05385
Carbon (Mercury Removal) (lb)	384,464	585	1.63	$626,676	$313,338	$0.01205
Oxygen Carrier (tonne)	9,327	54	600.00	$5,596,200	$10,643,400	$0.40914
COS Catalyst (lb)	0	0	3,751.70	$0	$0	$0.00000
Water Gas Shift Catalyst (ft3)	0	0.00	771.99	$0	$0	$0.00000
SCR Catalyst (m3)	0	0	0.00	$0	$0	$0.00000
Aqueous Ammonia (ton)	0	0	0.00	$0	$0	$0.00000
Claus Catalyst (ft3)	w/equip	0.92	131.27	$0	$39,668	$0.00152
Methanol Synthesis Catalyst (ft3)	4,207	3.84	500.00	$2,103,500	$631,050	$0.02426
DME Synthesis Catalyst (ft3)	0	0.00	750.00	$0	$0	$0.00000
Gasoline Synthesis Catalyst (ft3)	0	0.00	1,200.00	$0	$0	$0.00000
Aromatics Hydrotreater Catalyst (ft3)	138	0.13	475.00	$65,550	$19,665	$0.00076
Subtotal Chemicals				**$8,391,926**	**$13,047,981**	**$0.50158**
Other						
Supplemental Fuel (MBtu)	0	0	0.00	$0	$0	$0.00000
Supplemental Electricity (for consumption) (MWh)	0	5,904	60.00	$0	$116,367,840	$4.47329
Gases, N2 etc. (/100scf)	0	0	0.00	$0	$0	$0.00000
L.P. Steam (/1000 pounds)	0	0	0.00	$0	$0	$0.00000
Subtotal Other				**$0**	**$116,367,840**	**$4.47329**
Waste Disposal						
Spent Mercury Catalyst (lb.)	0	585	0.65	$0	$124,951	$0.00480
Slag (ton)	0	0	0.00	$0	$0	$0.00000
Flyash (ton)	0	706	25.11	$0	$5,823,998	$0.22388
Subtotal-Waste Disposal				**$0**	**$5,948,949**	**$0.22868**
By-products & Emissions						
Sulfur (tons)	0	63	0.00	$0	$0	$0.00000
Hydrogen (mmscf)	0	7.1	0.00	$0	$0	$0.00000
Supplemental Electricity (for sale) (MWh)	0	0	58.00	$0	$0	$0.00000
Subtotal By-Products				**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS				**$8,391,926**	**$166,910,085**	**$6.41618**
Fuel						
Coal (ton)	0	8,624	36.57	**$0**	**$103,596,507**	**$3.98235**
Natural Gas (MMBtu)	0	147,727	6.13	**$0**	**$297,478,872**	**$11.43537**
TOTAI FUEL COSTS				-	**$401,075,378**	**$15.41772**

6. ECONOMIC ANALYSIS - COST OF METHANOL PRODUCTION ANALYSIS

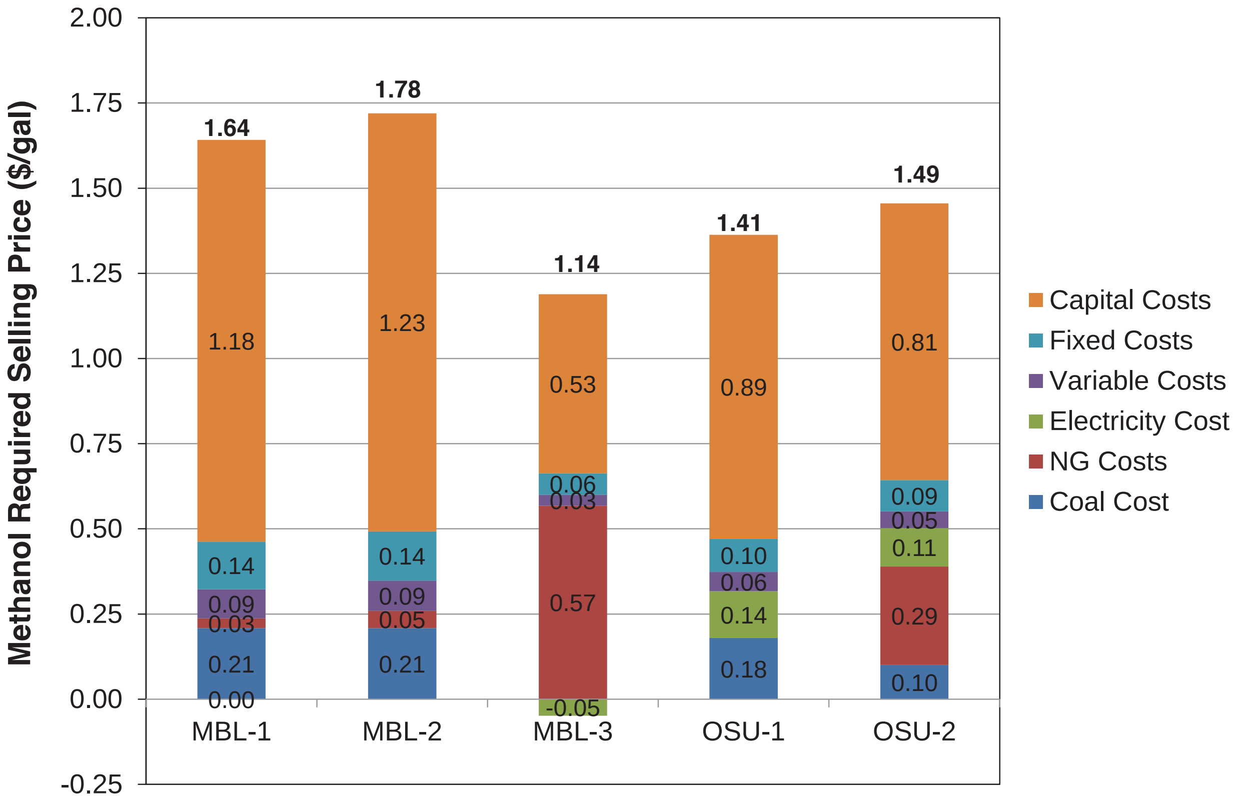
The methodology to determine the *Methanol Required Selling Price* is Based on the *DOE/NETL Quality Guidelines for Energy System Studies* and outlined in Section 3.4 of this report. The assumptions used in this analysis are summarized in Exhibit 3-6. All prices provided here are first year costs and are in 2011 dollars. The primary purpose of this analysis is comparative between the options developed in this study. Great caution should be used when comparing these costs to commodity price from various markets which are not only dependent on the cost components investigated here (capital, feedstock and O&M) but also on market forces including supply and demand.

The results of the reference conditions for the three MBL reference cases and the two OSU are presented in Exhibit 6-1 and Exhibit 6-2. Sensitivity studies for these options under varying economic conditions are provided following these exhibits. In this study, the methanol required selling price for the OSU options was less than that of the reference case, MBL-1. This is related to the higher efficiency of the process and lower capital costs associated with the OSU technology. Using the equation present in section 3.4.1 for the carbon capture costs, results in a negative carbon capture cost. Implications of a negative cost in this case are not clear.

Exhibit 6-1: Cost of Methanol Production for DOE/NETL Cases and OSU CLG Cases

	Case				
	MBL-1	**MBL-2**	**MBL-3**	**OSU-1**	**OSU-2**
Required Selling Price ($/ton, 2011$)	495.27	535.58	346.56	425.11	446.69
Required Selling Price ($/Gal, 2011$)	**1.64**	**1.78**	**1.15**	**1.41**	**1.48**
CO_2 TS&M Costs	0.00	0.06	0.01	0.05	0.03
CO_2 Emissions Costs	0.00	0.00	0.00	0.00	0.00
CO_2 Sales Revenue	0.00	0.00	0.00	0.00	0.00
Coal Cost	0.21	0.21	0.00	0.18	0.10
NG Costs	0.03	0.05	0.57	0.00	0.29
Electricity Cost	0.00	0.00	-0.05	0.14	0.11
Variable Costs	0.09	0.09	0.03	0.06	0.05
Fixed Costs	0.14	0.14	0.06	0.10	0.09
Capital Costs	1.18	1.23	0.53	0.89	0.81
Cost of CO_2 Captured	**NA**	**15.66**	**NA**	**NA**	**NA**

Exhibit 6-2: Cost Breakdown of Methanol Production for DOE/NETL Cases and OSU CLG Cases



The objectives of the sensitivity studies are to identify:

1. Parameters that have a significant impact on the production costs and thus identify opportunities to focus on for future cost reductions.
2. Parameters that could change and result in a change in the most favorable option.

In all of these cases, it is important to bound the parameters to a reasonable range. The parameters and cost investigated in this current OSU CLG and their ranges are summarized in Exhibit 6-3.

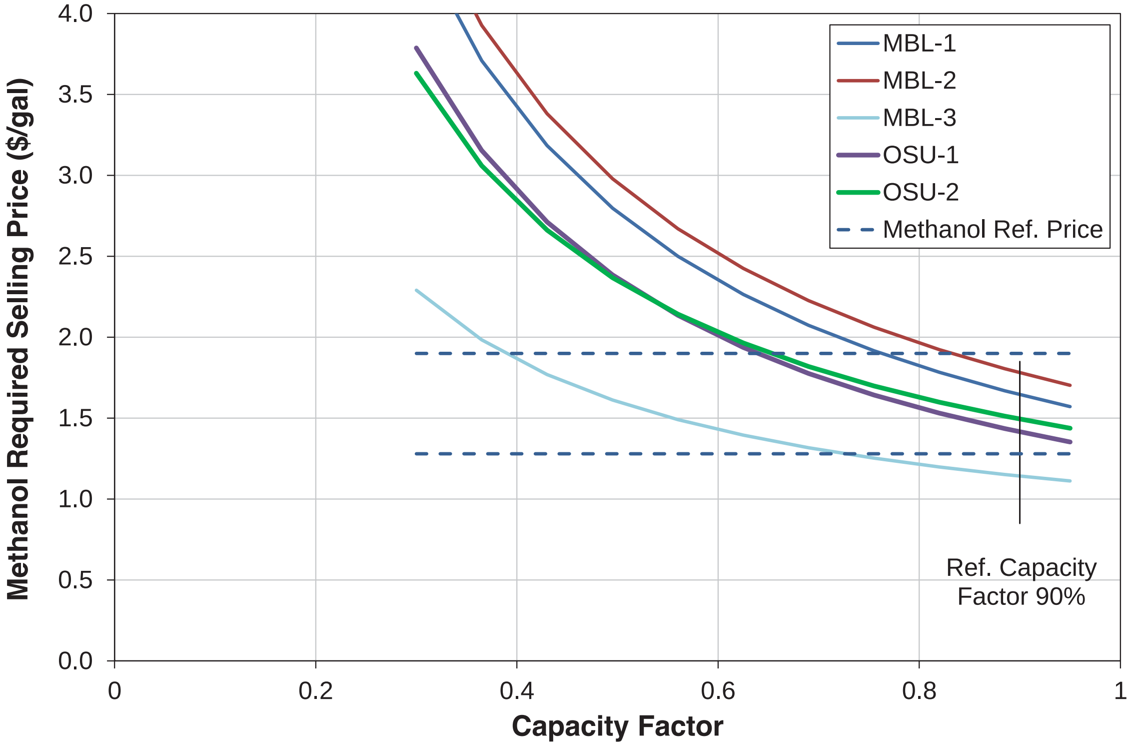
Exhibit 6-3: Summary of Parameters and Costs and Their Ranges used in Sensitivity Study

Parameter/Cost	Nominal	Min	Max	Range Basis
Capacity Factor	90%	50%	90%	
Coal Price, $/ton	36.57	20	60	Regional variation in coal price
Natural Gas Price ($/MMBtu)	6.13	2.00	12	Low price-observed during periods and in regions with excess supply. High price-typical world market price or periods with limited supply.
Electricity Price ($/MWh)	60	50	135	Maximum price equivalent to price with CO_2 capture
Capital Charge Factor	0.237	0.12	0.25	RFP Attachment 2
Oxygen Carrier Cost, $/tonne	600	500	900	Estimated range for producing a fired ceramic material
Oxygen Carrier Attrition Rate, /cycle	0.02%	0.01%	0.04%	OSU experimental work
Reactor Residence Time, min	50	40	90	OSU experimental work
CLG Equipment Cost	100%	50%	200%	RFP Attachment 2
CO_2 Emission, $/tonne CO_2	0	0	30	Low price set by current policy High price is typical highest emission cost
Captured CO_2 Sale Price	0	0	50	Low price set by sequestration. High price set under high demand for EOR

As part of the sensitivity study, the RSP is compared to the methanol market price as supplied by Methanex Corporation. From January 2011 to December 2014 the reference price of methanol has fluctuated from $1.28 to $1.90 and closed at $1.40/gal in December of 2014.[10} For comparative purposes, this methanol reference price range is included in the following sensitivity study exhibits. The prices from Methanex are for chemical grade methanol which is produced by distilling crude methanol to remove water and other impurities.[11]

The RSP sensitivity to capacity factor is illustrated in Exhibit 6-4. From the reference capacity factor of 90%, the RSP increases with decreasing capacity factor related to the capital contribution to the RSP being distributed over less annual production. This increase is less for the natural gas feedstock based facilities due to the capital costs providing a smaller contribution to the RSP. This smaller increase, for the natural gas cases, results in a switch in the economic ordering of the OSU-1 and OSU-2 options at a 50% capacity factor.

Exhibit 6-4: Sensitivity of Methanol Required Selling Price to Capacity Factor



The RSP sensitivity to coal price is illustrated in Exhibit 6-5 and for natural gas price in Exhibit 6-6. The MBL-1, MBL-2 and OSU-1 options which use coal as the only feedstock show similar sensitivities with respect to coal costs, approximately a $0.20/gal increase with a doubling in the reference coal cost. The OSU-2 option shows a lower sensitivity to coal cost due natural gas comprising 50% of the feed stock by thermal input. The RSP, for the options utilizing natural gas as a feedstock, has a much greater sensitivity to natural gas cost as illustrated in Exhibit 6-6, increasing by $0.30/gal for the OSU-2 option and $0.60/gal for the MBL-3 option. These increases in sensitivity are related to the feedstock cost contributing a greater portion to the RSP for the options utilizing natural gas. This sensitivity results in the OSU-2 option becoming more favorable than the OSU-1 option for natural gas prices less than $4.25/MMBtu and the OSU option becoming more favorable that the MBL-3 option for natural gas prices greater than $9.00/MMBtu. Note, the changes in coal and natural gas costs are assumed to have no impact on the reference methanol prices from Methanex.

Exhibit 6-5: Sensitivity of Methanol Required Selling Price to Coal Price

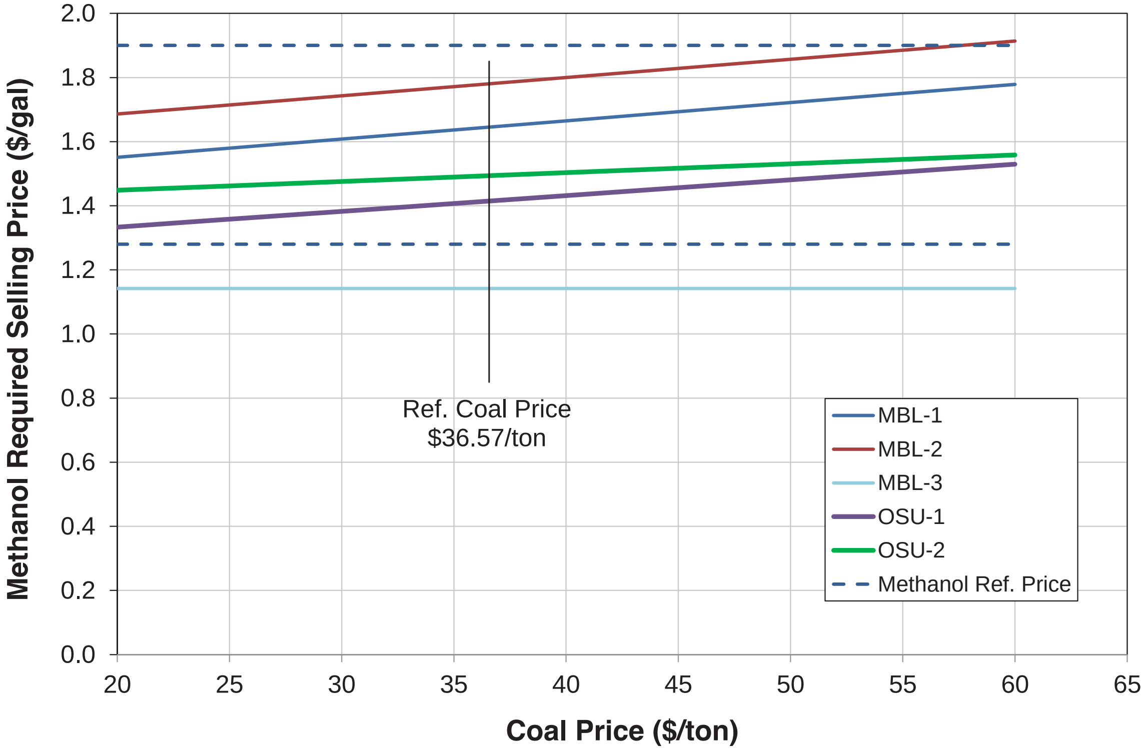
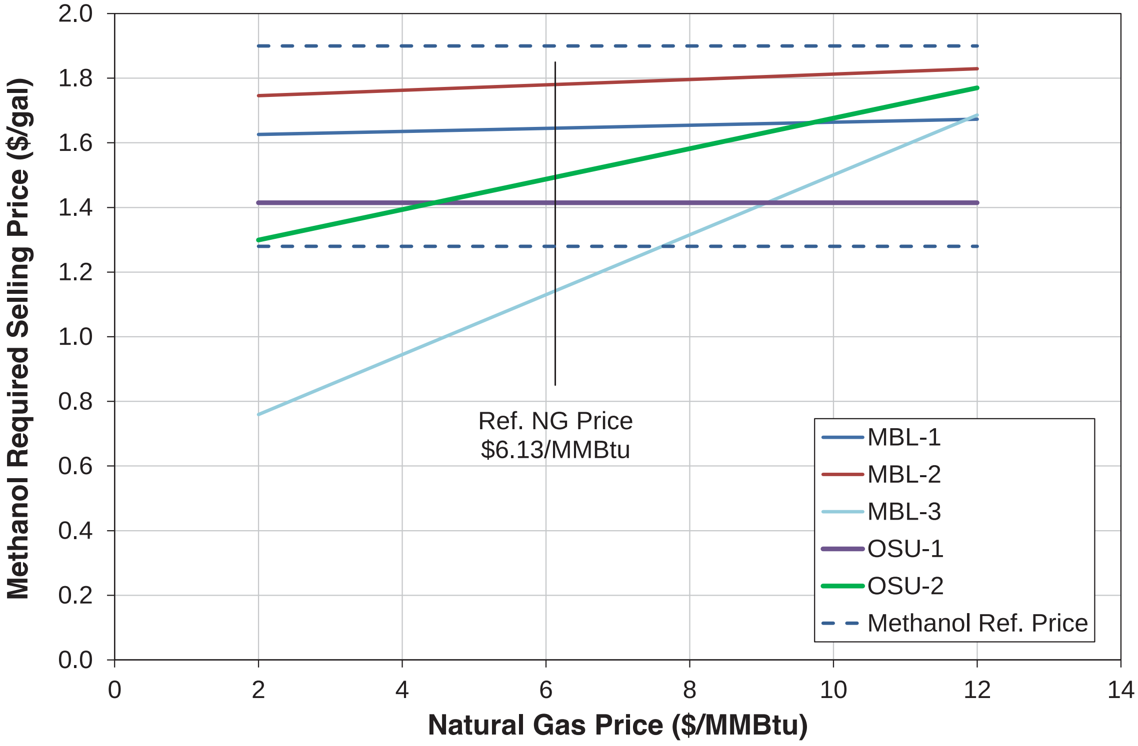
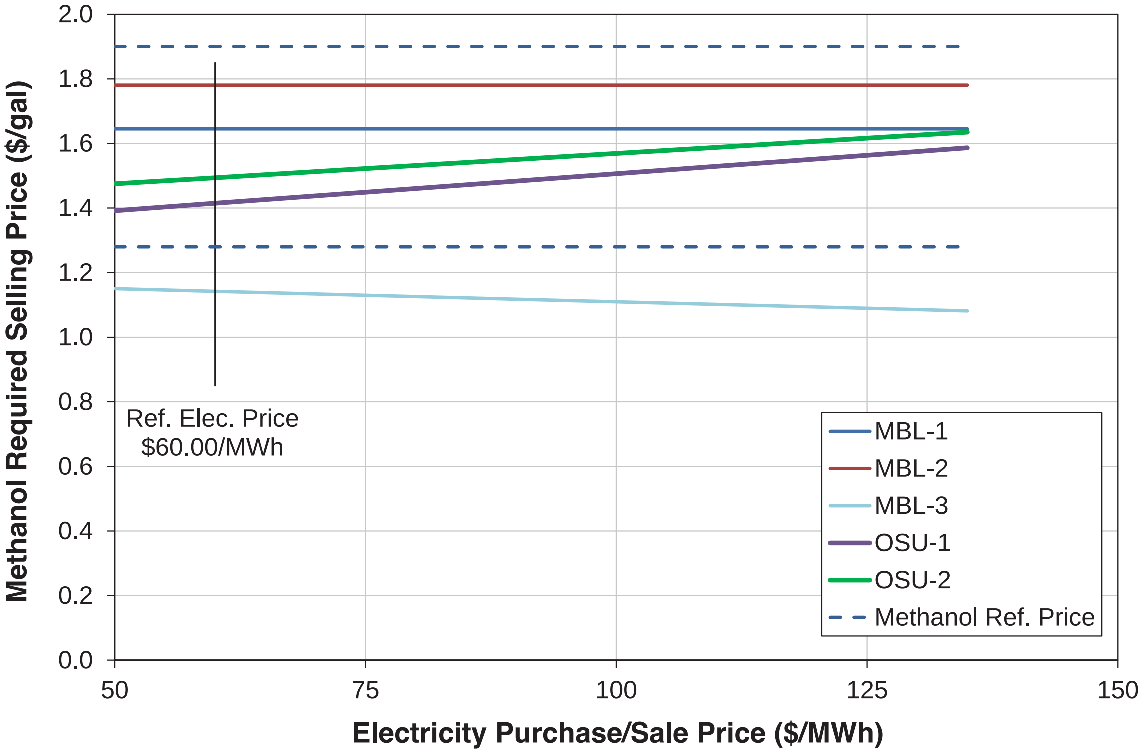


Exhibit 6-6: Sensitivity of Methanol Required Selling Price to Natural Gas Price



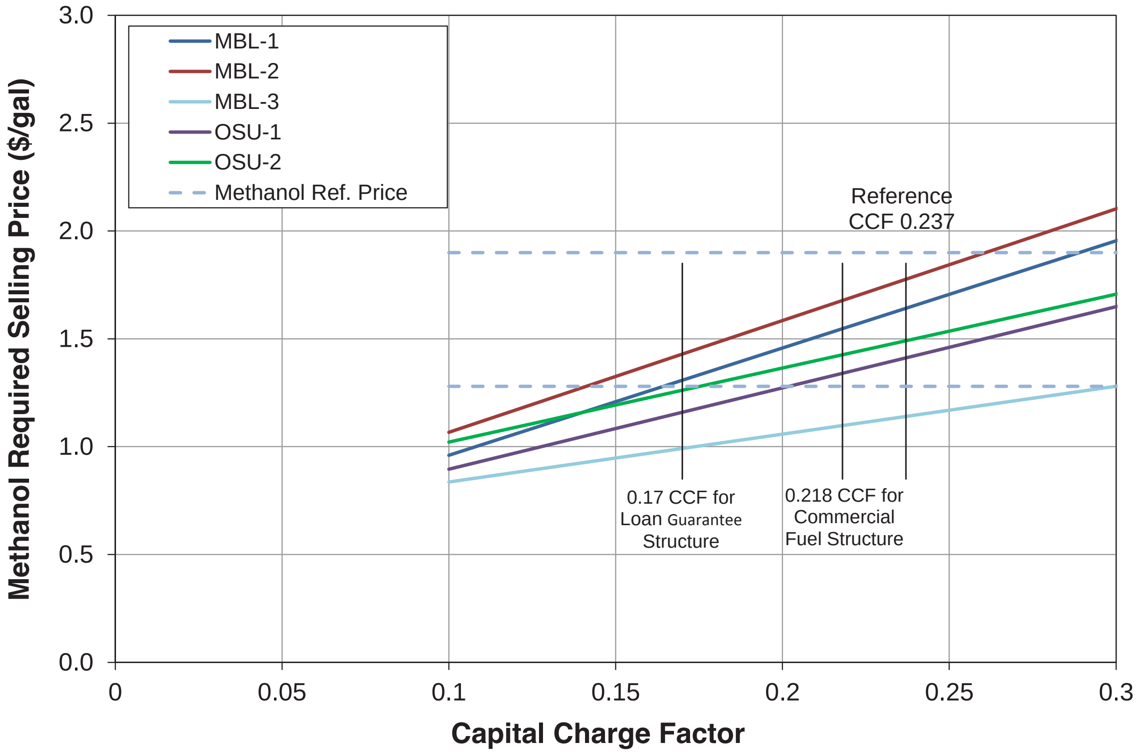
As designed, the OSU options rely on external power generation to drive the auxiliary loads. The RSP sensitivity to electricity price is illustrated in Exhibit 6-7. For an increase in power costs from $60/MWh to $135/MWh, an increase representative of switching from power generation without carbon capture to power generation with carbon capture, the RSP for the OSU options increase by $0.20/gal.

Exhibit 6-7: Sensitivity of Methanol Required Selling Price to Electricity Price



The capital charge factor (CCF) is the fraction of the total overnight costs that are allotted to the annual costs. As discussed in section 3.4, these CCF is dependent of the financial structure. The RSP sensitivity on the CCF is illustrated in Exhibit 6-8. Included in this exhibit are the CFF for the financial structures in the Methanol Baseline Study; CFF of 0.170 corresponding to the loan guarantee structure and CFF of 0.218 corresponding to the commercial fuel structure. The options that utilize natural gas a feedstock, MBL-3 and OSU-2, have lower capital costs and therefore a lower sensitivity to the CFF. Over the CFF range considered, the economic favorability of the MBL-1 option is greater than that of OSU-2 options for CFF less than 0.14.

Exhibit 6-8: Sensitivity of Methanol Required Selling Price to Capital Charge Factor



The performance parameter identified as critical to the methanol RSP are the oxygen carrier costs, the oxygen carrier attrition rate, and the reactor residence time. The RSP sensitivities to these parameters are provided in Exhibit 6-9, Exhibit 6-10, and Exhibit 6-11.

As described by OSU, there is no expectation of a significant increase in the manufacturing costs of the oxygen carrier. The required raw materials are common; primarily Al_2O_3 and Fe_2O_3, and the final product can be manufactured using existing methods in the ceramics industry. A cost increase of 50% is considered in Exhibit 6-9. This increase shows no significant impact on the methanol RSP, therefore the oxygen carrier costs are not a significant risk to the overall economics. Additionally, this sensitivity study shows that reducing the oxygen carrier costs is not a potential route to reducing the RSP so that costs become more competitive.

The attrition rate of the oxygen carrier and the replacement costs are another concern considering the large volume of oxygen carrier in the system. The RSP sensitivity to attrition rate was considered up to a 100% increase in the 0.02% per cycle rate estimated by OSU. Exhibit 6-10 illustrates that there is not a significant increase in the methanol RSP with this increase in the oxygen carrier attrition rate. Therefore, the oxygen carrier attrition rate does not pose significant risk to the project economics.

Exhibit 6-9: Sensitivity of Methanol Required Selling Price to Oxygen Carrier Cost

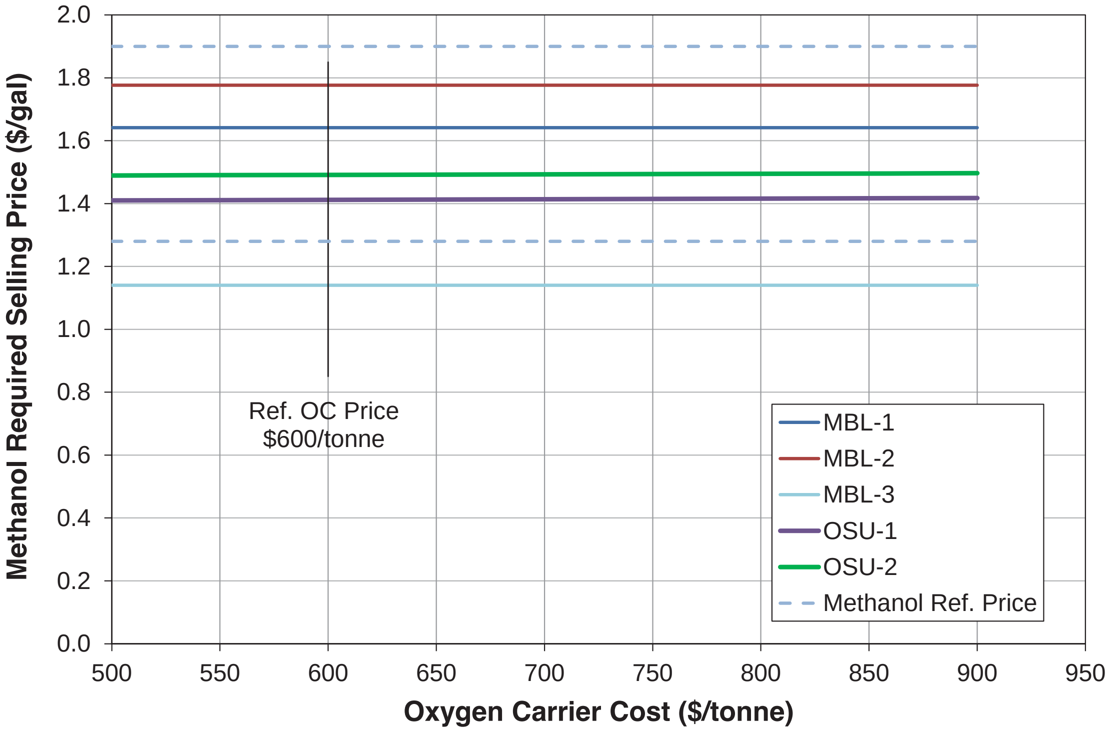
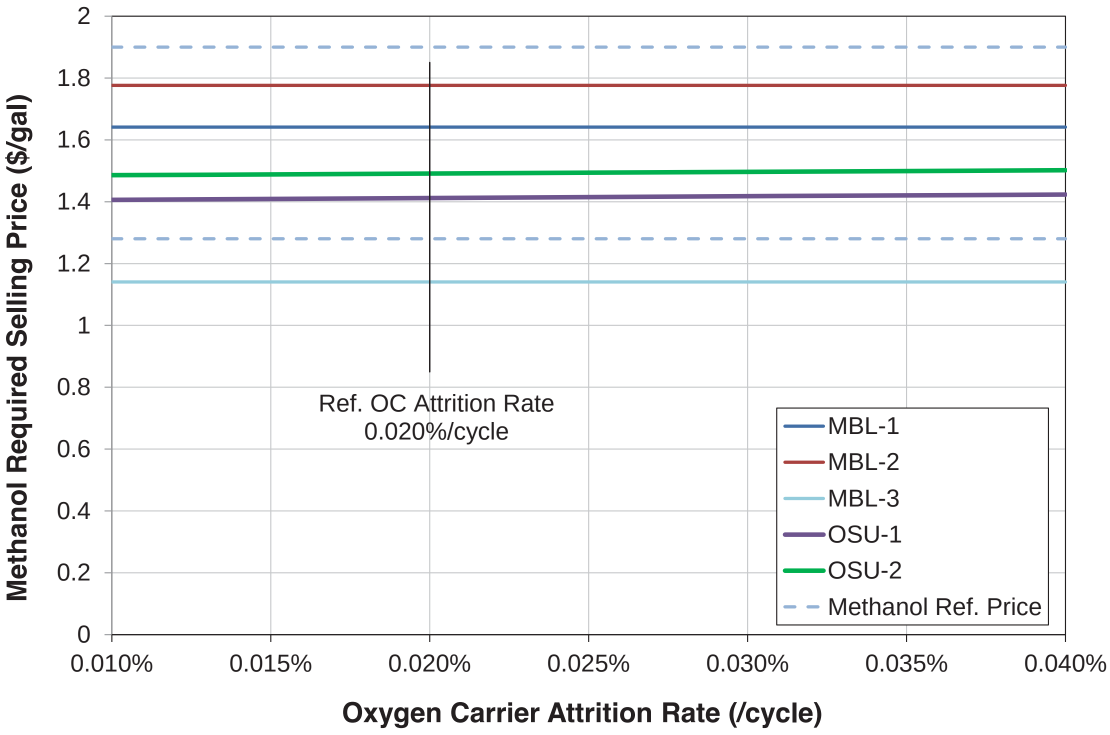


Exhibit 6-10: Sensitivity of Methanol Required Selling Price to Oxygen Carrier Attrition Rate



The reaction time between the solid oxygen carrier and the coal at the commercial scale pose a risk to the sizing of the CLG equipment. To understand this risk, a methanol RSP sensitivity to the reactor residence time was performed considering a required residency time up to 90 minutes compared to the 50 minutes reference value. The cost implications of changing residence time are:

1. The size of reactors,
2. The initial amount of oxygen carrier required, and
3. The amount of oxygen carrier that needs to be replaced due top attrition.

Exhibit 6-11 illustrates the impacts of changing the residency time considering these factors. With an increase from 50 to 90 minutes, the methanol RSP increases approximately $0.08/gal. This change does not impact the economic favorability ordering of the options. Additionally, this sensitivity illustrates that reducing the reactor residence time does not provide a potential route for significantly increasing the economic competitiveness of the OSU technology compared to the natural gas based MBL-3.

Exhibit 6-11: Sensitivity of Methanol Required Selling Price to Reactor Residence Time.

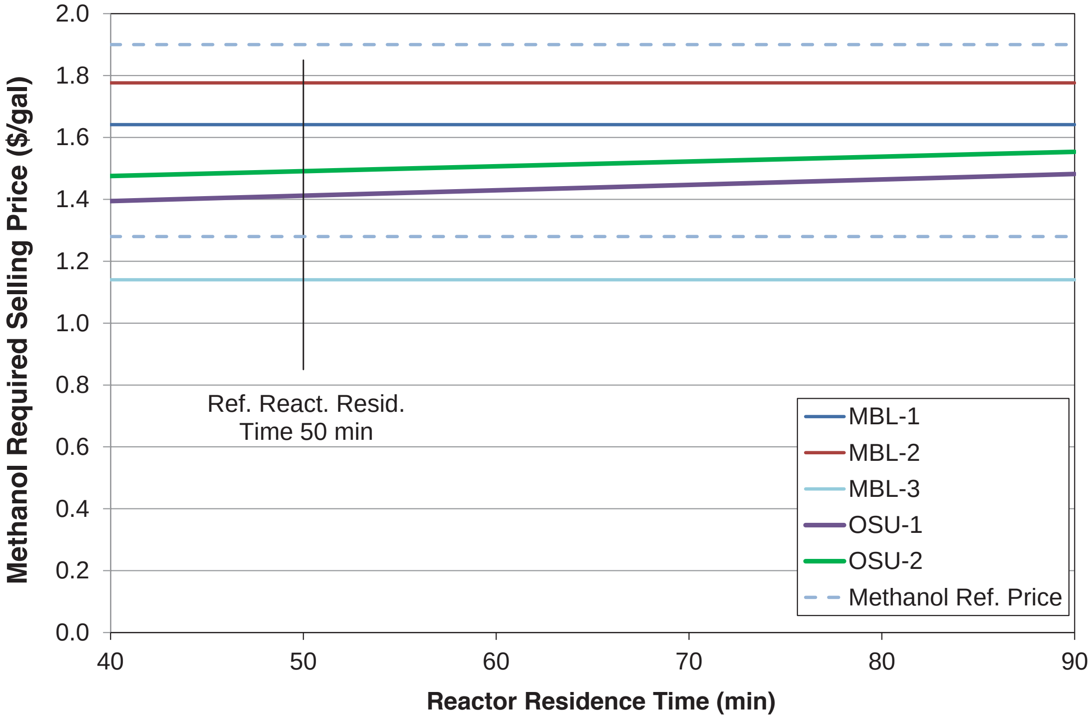
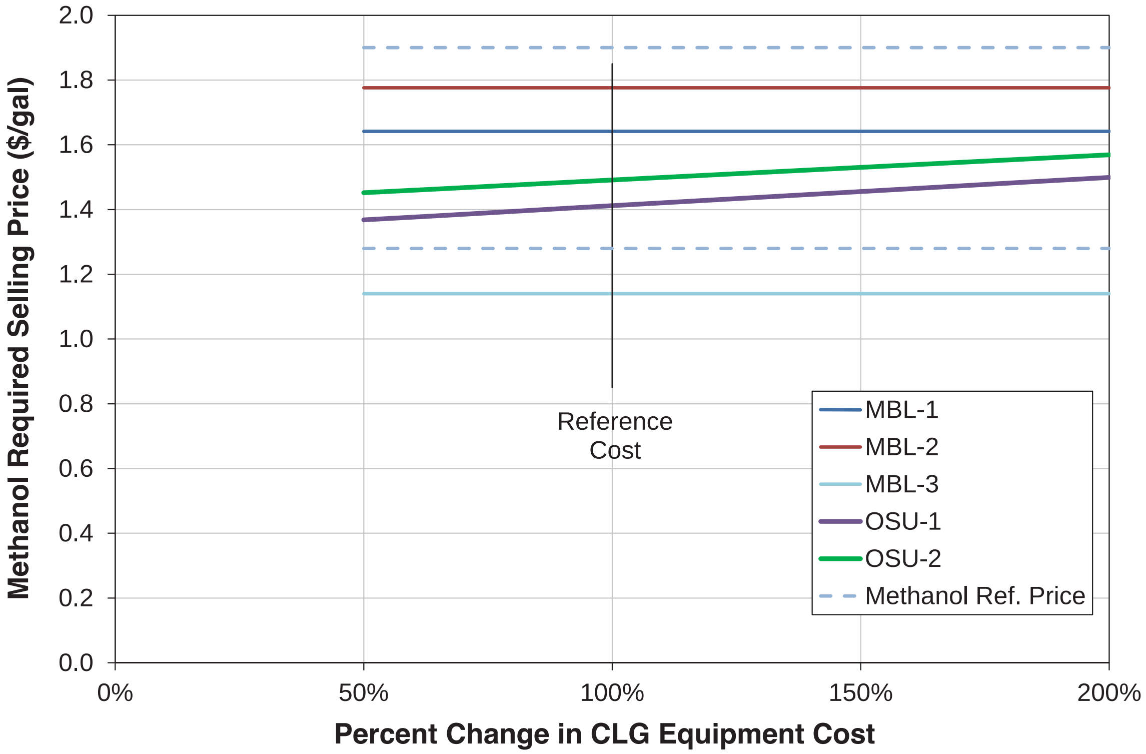


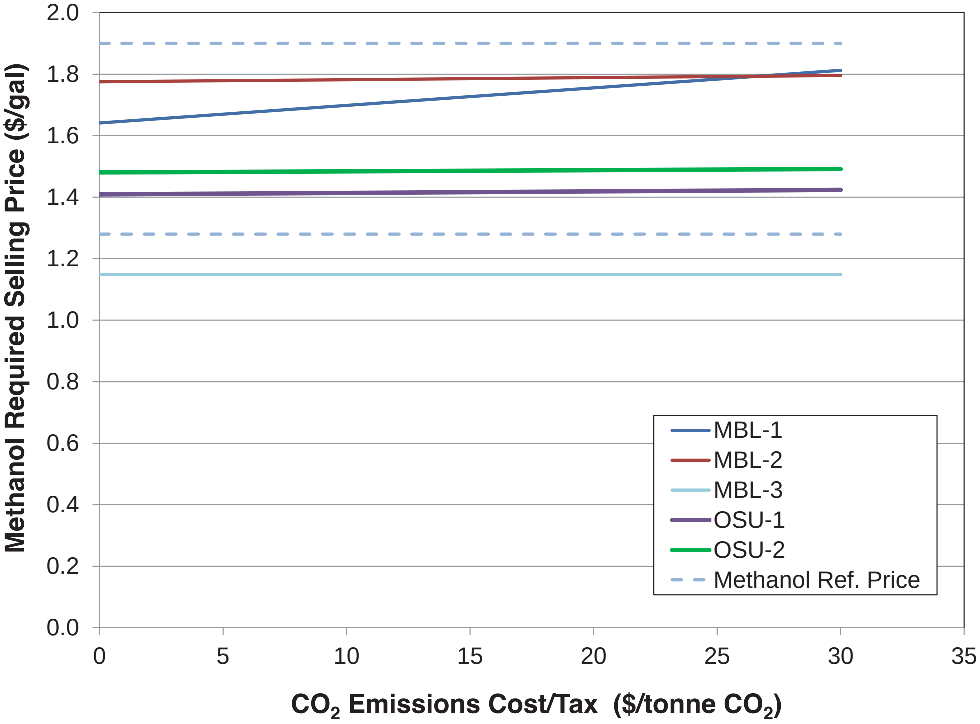
Exhibit 6-12 illustrates the RSP sensitivity to varying the OSU CLG equipment costs from -50% to +100%. For a 100% increase in the cost of the OSU CLG equipment there is a $0.10/gal increase in the methanol RSP. This change in RSP is not sufficient to change the economic favorability of the options.

Exhibit 6-12: Sensitivity Methanol Required Selling Price to CLG Equipment Cost



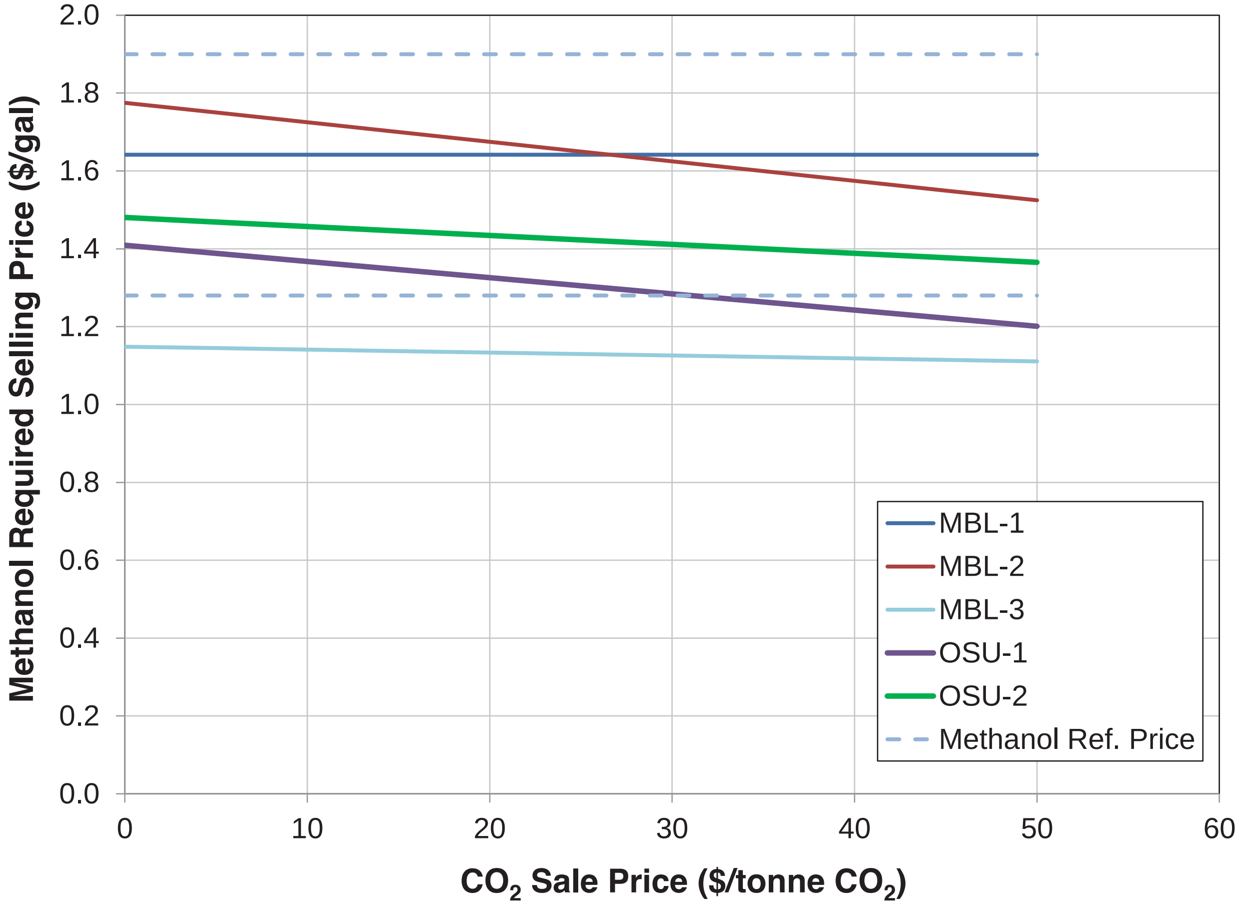
With future restrictions on the emissions of greenhouse gases, there is the potential for a tax to be applied to CO_2 emissions from industrial facilities. The CO_2 capture incorporated in these options provides a route to decreasing these taxes and offers an economic advantage over options that not include CO_2 capture such as the DOE/NETL MBL-1 case. Exhibit 6-13 illustrates the RSP sensitivity to CO_2 emissions costs/tax. As expected, the options that incorporate CO_2 capture show a very small sensitivity to this tax/cost. For the DOE/NETL baseline cases, the option with CO_2 capture, MLB-2, becomes economically favorable over MLB-1 with a CO_2 tax of ~$27/tonne. It is important to note that the implications of these taxes on the market price of CO_2 are not considered and the market values are assumed to be independent of this tax.

Exhibit 6-13: Sensitivity of Methanol Required Selling Price to CO_2 Emission Cost



With the incorporation of CO_2 capture into the methanol production facility, there is the potential to sell the CO_2 to other industries such as oil recovery. The value of the CO_2 to oil recovery is strongly dependent on the petroleum market and the availability of CO_2. Exhibit 6-14 illustrates the RSP sensitivity to CO_2 sale prices. The CO_2 capture options which use coal only as a feedstock, MBL-1 and OSU-1, show the greatest beneficial dependence on the sales of CO_2. For the baseline cases, the MLB-2 with CO_2 capture becomes economically favorable over MLB-1 with a CO_2 sale price of $27/tonne CO_2.

Exhibit 6-14: Sensitivity of Methanol Required Selling Price to Captured CO_2 Sale Price



7. STUDY FINDINGS

The current TEA investigated incorporating the OSU chemical looping gasification technology into coal fired IGCC power generation and coal based methanol production facilities at a conceptual level. One of the primary goals was to develop approaches the OSU CLG technology to produce high hydrogen content syngas required for IGCC power generation with pre-combustion CO_2 capture and control over the syngas composition for methanol production. While sufficient control over the syngas composition was achieved for syngas production, a syngas with a sufficiently high hydrogen content to reach 90% CO_2 capture was not. To achieve this level of hydrogen, reducing the iron in the oxygen carrier to the metallic state without the formation of elemental carbon is required. In the process models developed, conditions to achieve this objective could not be identified.

Two methanol production options incorporating the OSU CLG technology were developed for this, one with 100% coal as the feedstock to the process (OSU-1) and the other with 50% coal and 50% natural gas as the feedstock (OSU-2). A summary of the performance, capital and operating costs, and the economics of coal to methanol process developed under this project are provided in Exhibit 7-1. In this exhibit the results are compared to the results from the draft version of the DOE/NETL *Baseline Analysis of Crude Methanol Production from Coal and Natural Gas*. From the DOE/NETL study, case MBL-1 is based on a coal feed and gasifier without CO_2 capture, case MBL-2 is based on a coal feed and gasifier with CO_2 capture, and case MBL-3 is based on natural gas feed and a reformer with CO_2 capture

The OSU CLG technology provides several savings compared to the state-of-the-art coal base methanol production related to lower capital costs and higher efficiencies. In comparing the MBL-2 and OSU-1 options, which are both based on a coal only feed and have CO_2 capture, the following benefits are realized:

- A lower methanol RSP by $0.37/gal, a 21% decrease.
- Lower capital costs by 28%
- Higher efficiency based 14% in coal consumption
- A methanol RSP lower than the reference non-capture case, which results in CO_2 capture cost less than 0.

Under the reference economic conditions, the methanol RSP of the OSU-CLG based systems were not found to economically favorable compared to the natural gas fed MBL-3 option. The low capital costs of the reformer based facility were a significant economic advantage all of the coal fired cases that could not be overcome for the reference natural gas cost of $6.13/MMBtu.

Exhibit 7-1 Performance and Cost results for Methanol Production Options.

	Case				
	MBL-1	**MBL-2**	**MBL-3**	**OSU-1**	**OSU-2**
Performance Summary					
As Received Coal (lb/hr)	1,618,190	1,618,190	NA	1,395,457	718,631
Natural Gas to Reformer, OSU CLG (lb/hr)	NA	NA	583,677	NA	272,290
Crude Methanol (lb/hr)	941,823	941,823	940,989	918,582	918,582
Captured CO_2 (lb/hr)		1,569,410	235,808	1,302,138	663,393
Capital and Operating Cost Summary					
Total Plant Costs (2011 MM$)	4,586	4,775	2,028	3,497	2,996
Total Overnight Cost (2011 MM$)	5,572	5,802	2,485	4,236	3,634
Total As Spent Capital (2011 MM$)	6,580	6,852	2,935	5,003	4,291
Annual Fixed Operating Costs (x1000 $/yr)	156,650	162,051	70,644	108,291	94,034
Variable Operating Costs ($/gal)	0.085	0.088	0.032	0.056	0.049
Economic Comparison					
Methanol RSP ($/ton, 2011$)	495.27	535.58	346.56	425.106	446.69
Methanol RSP ($/Gal, 2011$)	**1.64**	**1.78**	**1.15**	**1.41**	**1.48**
CO2 TS&M Costs	0.00	0.06	0.01	0.05	0.03
Coal Cost	0.21	0.21	0.00	0.18	0.10
NG Costs	0.03	0.05	0.57	0.00	0.29
Electricity Cost	0.00	0.00	-0.05	0.14	0.11
Variable Costs	0.09	0.09	0.03	0.06	0.05
Fixed Costs	0.14	0.14	0.06	0.10	0.09
Capital Costs	1.18	1.23	0.53	0.89	0.81
Cost of CO_2 Captured	**NA**	**15.66**	**NA**	**NA***	**NA***

Notes:

*The methanol RSP for the OSU CLG based systems is less than the reference non CO_2 capture case, MBL-1. This results in a negative cost of CO_2 capture which does not have a clear interpretation.

7.1 Sensitivity Studies

Sensitivity studies were completed to identify parameters that have a significant impact on the production costs and thus identify opportunities to focus on for future cost reductions and parameters that could change and result in a change in the most favorable option. Parameters of greatest concern are related to the feedstock costs and the performance and cost of the OSU SLG technology.

7.1.1 Feedstock Costs

For the options that rely on coal as the feedstock (MLB-1, MLB-2, OSU-1 and OSU-2), the RSP sensitivity to feedstock cost is significantly less that the options that use natural gas as a feedstock (MLB-3); approximately 10% with a doubling of coal cost and 30% with a doubling of the natural gas cost. This is a result of the capital cost being the primary contributor to the coal based option RSPs and feedstock (natural gas) costs the primary contributor to the natural gas based option RSPs. Considering variations in natural gas costs, for natural gas costs greater than $9.00/MMBtu, the OSU-1 option becomes most favorable.

The options incorporating the OSU CLG technology in this study rely on purchase electricity for a portion of the electricity for auxiliary loads. With electricity cost increases from the reference price of $60/MWh to $130/MWh, costs indicative of current electricity costs and future electricity costs with carbon capture, the methanol RSP increases by 12%. With the high electricity costs $130/MWh, the methanol RSPs of the OSU CLG based options are less than the RSP for the MBL-1 and MBL-2 options.

7.1.2 OSU CLG Reactor Performance and Costs

The impact of the OSU CLG technology performance and capital and operating costs uncertainties on the methanol RSP were considered.

Slower reaction kinetics in the CLG will result in longer residence times and larger reactor size. With a doubling of the reactor residence time, the methanol RSP increased by 8%. This increase was not sufficiently significant to increase to RSP of the OSU CLG options above that of the coal based MLB options.

Uncertainties in the OSU CLG equipment may result in higher capital costs. A doubling of OSU CLG technology equipment costs results in a 10% increase in the methanol RSP. This increase was not sufficiently significant to increase to RSP of the OSU CLG options above that of the coal based MLB options.

The oxygen carrier (OC) cost is a significant concern as an initial cost and as a consumable related to the replacement of lost material. With a 50% increase in the OC cost, there is less than 1% increase in the methanol RSP. Doubling the OC attrition rate, thus doubling the OC consumption rate during operations, increases the methanol RSP by less than 1%.

7.1.3 CO_2 Emissions and Selling Price

The OSU CLG based options incorporate CO_2 capture and therefore are relatively independent of CO_2 emission taxes. The methanol RSP for the OSU CLG options were lower than that of the MBL coal based option without CO_2 capture, so no economic advantage was gained by the OSU CLG technologies.

The sale of captured CO_2 provides can provide a revenue source to reduce the methanol RSP. With CO2 revenue up to $50/tonne CO_2, the methanol RSP decreases by 15% and 8% for the OSU-1 and OSU-2 options respectively.

7.2 Recommendations for Future Evaluations

The production of a very high hydrogen syngas from the OSU CLG technology would be very beneficial. While OSU experimental work indicates that this should be achievable, process modeling to identify conditions where this occurs proved difficult. Future work, both experimental and modeling, should be performed to understand these differences.

Based on the findings of this study, the following specific recommendations are suggested:

1. While there is not a strong sensitivity to OSU CLG capital costs, as illustrated in this study, further work to substantiate these costs should be performed. This work should include the more detail design and sizing of the reactors and equipment requirements to incorporate that OSU CLG technology into the process.
2. Explore a OSU CLG configuration using natural gas as the only feedstock to the OSU CLG process
3. Explore the configurations to investigate the possibility of power generation as part of the OSU CLG system design
4. Incorporate sensitivity study findings in experimental work to direct future research to verify the OC costs and the reactor residence time assumptions at a larger integrated system scale
5. Explore chemistries beyond the Fe_2O_3-$FeAl_2O_4$ cycle for high hydrogen syngas compositions for IGCC power generation applications

8. REFERENCES

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