AOI3: HIGH HYDROGEN, LOW METHANE SYNGAS FROM LOW-RANK COALS FOR COAL-TO-LIQUIDS PRODUCTION

FINAL TECHNICAL REPORT

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ABSTRACT

An experimental program was undertaken to develop and demonstrate novel steam reforming catalysts for converting tars, C2+ hydrocarbons, and methane under high temperature and sulfur environments at lab scale. Several catalysts were developed and synthesized along with some catalysts based on recipes found in the literature. Of these, two had good resistance at 90 ppm H_2S with one almost not affected at all. Higher concentrations of H_2S did affect methane conversion across the catalyst, but performance was fairly stable for up to 200 hours. Based on the results of the experimental program, a techno-economic analysis was developed for IGCC and CTL applications and compared to DOE reference cases to examine the effects of the new technology. In the IGCC cases, the reformer/POX system produces nearly the same amount of electricity for nearly the same cost, however, the reformers/POX case sequesters a higher percentage of the carbon when compared to IGCC alone. For the CTL case the economics of the new process were nearly identical to the CTL case, but due to improved yields, the greenhouse gas emissions for a given production of fuels was approximately 50% less than the baseline case.

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EXECUTIVE SUMMARY

This report summarizes the experimental program designed to develop and demonstrate novel steam reforming catalysts for converting tars, C2+ hydrocarbons, and methane under high temperature and sulfur environments at lab scale. The goal of the experiments was to increase the yield and H_2 to CO ratios of surrogate low-rank coal syngas simulating the Transport Reactor Integrated Gasifier (TRIG) and Lurgi Fixed Bed Dry Bottom (FBDB) Gasifiers with H_2S concentrations up to 500 ppm. The ultimate goal was to demonstrate the potential commercial viability of catalytic steam reforming under severe contaminant environments as an alternative to the water-/sour- gas shift reactions to produce cost-effective, high H_2 syngas from low-rank coals. Several catalysts were developed and synthesized along with some catalysts based on recipes found in the literature. Of these, two had good resistance at 90 ppm with one almost not affected at all. Higher concentrations did affect conversion across the catalyst, but performance was fairly level for up to 200 hours. Based on the results of the laboratory scale experimental program, a techno-economic analysis was developed for IGCC and CTL applications and compared to DOE reference cases to examine the effects of the new technology. In the TRIG IGCC case, the reformer/POX system produces nearly the same amount of electricity for nearly the same cost, however, the reformers/POX case sequesters a higher percentage of the carbon when compared to a TRIG IGCC alone. The additional carbon that is captured in the reformer/POX case must be compressed for sequestration increasing the parasitic load lowering the overall net electricity output compared to the base case. For the CTL case the economics of the new process were nearly identical to the CTL case, but due to improved yields the greenhouse gas emissions for a given production of fuels was approximately 50% less than the baseline CTL case.

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1 INTRODUCTION

Gasification of coal and coal-biomass mixtures has one of the greatest potentials for increasing the energy efficiency of electricity generation with IGCC and carbon capture, minimizing greenhouse gas emissions relative to traditional coal combustion. A higher efficiency process will lower the overall emissions which are necessary to meet current and future regulation guidelines as well as lower costs to produce energy. The goal of this project is to develop and demonstrate potential commercial viability of a high temperature catalytic steam reforming process that can withstand the severe contaminant conditions of a near-raw syngas derived from low-rank coal gasification with the objective to increase the yield and H_2 to CO ratios while eliminating methane, ammonia, and tars from the process. Research was performed at the lab-scale using simulated low-rank coal syngas that contains methane, ammonia, as well as a tar surrogate to test the conversion efficiency and catalyst longevity. The catalysts developed and synthesis techniques used were limited to ones that have the potential for scalability and commercial viability in the near term (i.e., by 2020).

2 BACKGROUND

2.1 COAL AND COAL/BIOMASS GASIFICATION

Coal gasification is a complex, multi-step process that uses steam with air or oxygen to convert solid coal into its gaseous derivative components: H_2, CO, H_2O, CO_2, CH_4, and contaminants. There have been tremendous advances in gasification and post-gasification processing technologies and recent breakthroughs in different related technological areas (i.e., membranes / monoliths, high temperature sorbents, catalysts) are paving the way for transitioning gasification into a new standard for obtaining clean liquid transportation fuels, producing hydrogen, and electricity. Recent advances in carbon capture have made it possible for a coal-based feedstock to have comparable CO_2 emissions with its petroleum counterpart; with the potential to be even or below those levels when mixed with small amounts of biomass (Tarka, Wimer et al. 2009). This project focuses on advancing gas clean up technologies even further by developing a novel high-temperature (>900°C) steam reforming catalyst that will not deactivate in the presence of sulfur.

There are three main classifications of coal gasification technologies that are commercialized today: fluidized-bed, fixed-/moving- bed, and entrained-bed (Breault 2010). Fluidized-bed gasifiers use relatively narrow distribution of small particle sizes of coal that is fed into an inert bed which maintains a uniform gasification temperature. The fluidized bed operates at lower temperatures and generally requires more reactive, low rank coals to achieve high conversions (Ke Liu 2010). Fixed-bed gasifiers use preheated lumped coal that is fed from the top of the gasifier where air or oxygen and steam are fed from the bottom. Fixed-bed gasifiers consume lower amounts of O_2 but produce more tars and methane (Ke Liu 2010). Entrained-bed gasification uses pulverized coal or coal water slurries under high temperatures (1500-1900°C) which requires higher O_2 feed rates to maintain the reaction (Ke Liu 2010). Transport reactor gasification (e.g., TRIG) is a recently developed gasification technology that was designed specifically for converting low-rank coals with low gas-solid transfer resistance and high gas-solid contact (Breault 2010). The TRIG system uses a limestone sorbent that is co-fed with the coal into the gasifier to remove the majority of the sulfur *in situ* down to levels as low as 100 ppmv which is substantially lower for the raw gas relative to other gasification technologies (Ke Liu 2010). The temperature of operation is moderate and fairly high carbon conversion efficiencies of 97% to syngas are possible.

A major challenge in any coal gasification application is gas clean-up which represents 36-41% of the total plant cost when including the secondary application such as: SNG (synthetic natural gas), coal-to-methanol, or IGCC application (Magee 1987; Ke Liu 2010). Additionally, the syngas derived from coal is typically lean in H_2 relative to distillate fuels which have H_2:CO ratios of approximately 2:1.

2.2 STEAM REFORMING

In 2010 there were 12 trillion standard cubic feet of H_2 produced annually, primarily from steam reforming of natural gas. Commercial applications typically convert all sulfur species in the natural gas into H_2S via a H_2 reduction, remove the H_2S with ZnO scrubbing, and then steam reforming occurs over a Ni-based catalyst where the support is typically Al_2O_3 (Subramani, Sharma et al. 2010). This reaction is typically carried out at temperatures over 700°C, is thermodynamically more favorable with higher ratios of steam to carbon and at lower pressures (Subramani, Sharma et al. 2010).

Several studies have looked at novel methods for steam cracking of tar using Ni supported on dolomite (Wang, Chang et al. 2004; González, Román et al. 2011). At 800°C a nickel-ceria perovskite was shown to have favorable steam reforming activity of methane with high selectivities towards CO and H_2 production (Zhang, Muratsugu et al.). Other studies have shown Ni hexaaluminates as a high-temperature steam reforming or partial oxidation catalyst with relatively high carbon deposition resistance, especially in the presence of steam (Machida, Teshima et al. 1991; Gardner, Spivey et al. 2010). A partial oxidation of decalin study was performed at over 850°C using Sr-, La-, and Ba-substituted hexaaluminates and showed deactivation when adding 50 ppmw sulfur in the form of dibenzothiophene during the reaction (Gardner, Shekhawat et al. 2007).

2.3 PROPOSED COMMERCIAL PROCESS

Figure 1 shows a process that converts low-rank coals (i.e., sub-bituminous and lignite) into distillate, hydrocarbon fuels using four main processing steps: gasification, clean up, syngas upgrading, and

Fischer-Tropsch Synthesis. This process begins by feeding steam, low-rank coal, and oxygen into a gasifier [e.g., TRI (ThermoChem Recovery, Inc.), TRIG (Transport Reactor Integrated Gasifier) or Lurgi's FBDB (Fixed-Bed Dry-Bottom)] to produce a raw syngas which contains hydrogen, carbon monoxide, methane, carbon dioxide; trace contaminants: acid halides, mercury, phosphorous, antimony, cadmium, hydrogen selenide, and arsine (AsH_3); in addition: tar, particulates, ammonia, and sulfur (H_2S, COS) that are at or near percentage levels. Removing these contaminants is required to meet emission regulations for power generation as well as to prevent catalyst deactivation in the FT reactor. If a TRIG gasifier is used it would also require feeding limestone sorbent into the gasifier and remove the majority of sulfur from the syngas.

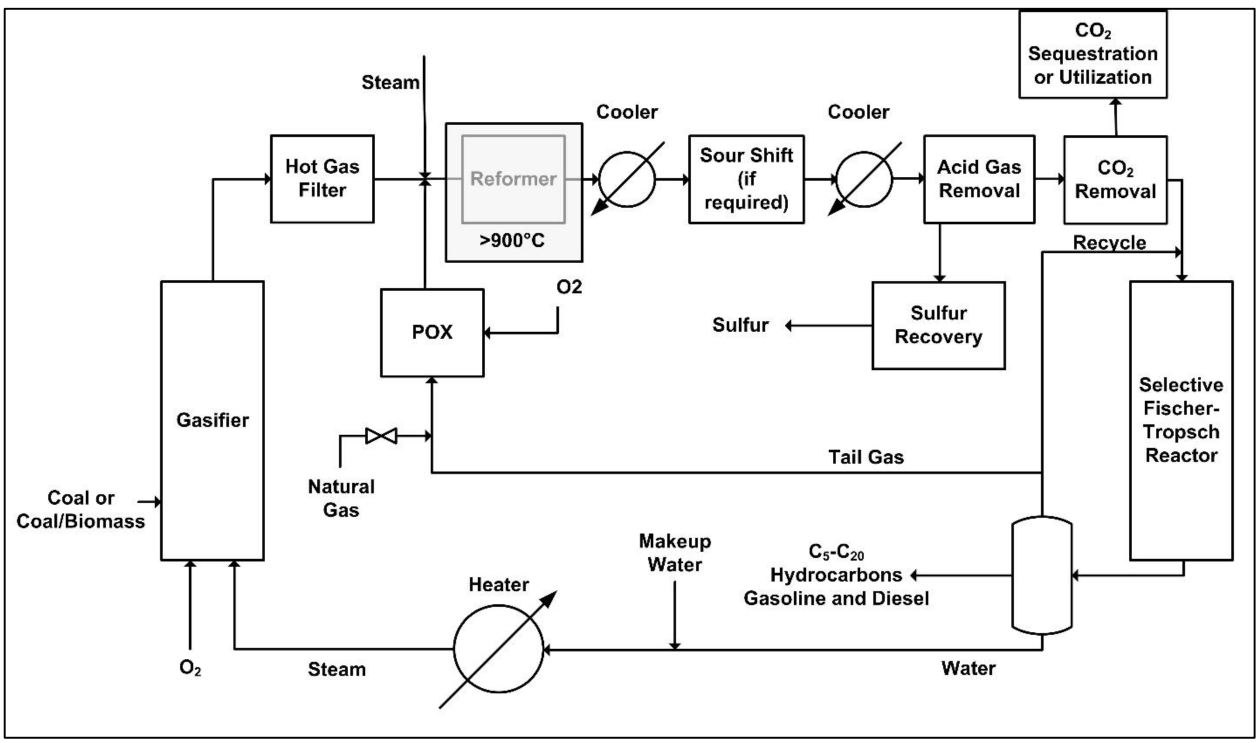

Figure 1. Simplified flow diagram for commercial embodiment of CTL process with proposed reformer.

An acid halide sorbent is fed to the raw syngas prior to the hot gas filter which removes the remaining particulates and spent sorbent. Following hot gas filtration is a high temperature >900°C) catalytic steam reforming step to convert the ammonia, methane, and tars into syngas which increases the H_2:CO ratio and eliminates these hydrocarbon and ammonia by-products from the syngas. Development of a catalyst that is stable under the severe deactivating conditions of a near-raw low rank coal (or coal-biomass) syngas containing tars, ammonia and methane is the focus of this proposed project. The high steam feed rates to the gasifier to produce higher H_2:CO ratios in the raw syngas also result in higher CH_4 formation which are produced on a 3:1 ratio in this reaction. Catalytic partial oxidation of light hydrocarbons produced upstream (i.e., Fischer-Tropsch derived CH_4 and LPG) will further increase the H_2:CO ratio of the syngas; additionally, the heat generated from the partial oxidation step will be sufficient to maintain the temperature required for steam reforming which is endothermic. After cooling, the syngas can be subjected to an optional sour shift if needed for H_2/CO ratio adjustment. It is then desulfurized and further cleaned using conventional technologies followed by amine-based CO_2 capture. Optionally the gas could be treated with warm gas cleanup technology such as that being developed by Research Triangle Institute under a cooperative agreement with the DOE to investigate the optimum sorbent-condition combinations. This technology has been successfully tested using a slip stream at the Eastman Chemical Plant's coal-to-chemical gasifier in Kingsport Tennessee (Gupta, Turk et al. 2009). The H_2:CO syngas is then fed to the selective FT reactor which maximizes the yield to liquid transportation fuels by making a wax-free product while maintaining relatively low yields to light hydrocarbons (C_{4-}<25%) without needing an additional hydrocracking step. This technology is being developed by Chevron Corporation and demonstrated by Southern Research under a cooperative agreements with the Department of Energy. These projects are testing the catalyst at the NCCC (National Carbon Capture Center) in Wilsonville, Alabama to demonstrate bench-scale syngas conversion to FT liquids using a Co-based catalyst with slip stream of coal- and coal/biomass-derived syngas post (gas) clean up. The FT liquids are passed to the distillation column and are separated into their respective fractions: light hydrocarbons (CH_4, LPG), naptha, jet, and diesel. The light hydrocarbons are recycled to the POX unit to provide high quality heat for the steam reforming and to further increase the yield to liquid transportation fuels.

3 PROJECT OBJECTIVES

The objective of the proposed project is to increase the yield and H_2:CO ratio of a surrogate low-rank coal syngas by developing, testing, and optimizing steam reforming catalysts for converting tars, C2+ hydrocarbons, NH_3 and methane under high temperature and sulfur environments. A lab-scale microreactor version of the catalytic steam reforming process was built with the capability of controlling the flow rate of the surrogate syngas (H_2, CO, CO_2, CH_4, H_2O, C_2H_6, C_3H_8, NH_3, tar surrogate, and H_2S) as well as the reaction temperature and pressure to perform a parametrically designed study of these effects on the productivity and deactivation rates of the catalyst under commercially relevant reaction environments and space velocities. A thorough mass balance was performed during these studies and catalyst characterizations including measuring carbon/sulfur deposition on the catalyst's surface and BET (Brunauer Emmett Teller) surface area after reaction. The results of this study were inputted into a Techno-economic model that will be used to evaluate the commercial viability of this process relative to the current state-of-the art and will help identify potential barriers through a sensitivity analysis.

The syngas composition of commercial gasifiers: TRIG (Transport Reactor Integrated Gasifier) and Lurgi's FBDB (Fixed-Bed Dry-Bottom) were used as the basis for the lab-scale experiments and Techno-economic modeling of the commercial embodiment. Using a surrogate syngas from commercially available gasification technologies significantly increases the potential impact beyond the focus of the scope of work proposed here which is a modified Fischer-Tropsch coal-to-liquids commercial embodiment with carbon capture. There are other areas where this project has commercial relevance for applications related to coal gasification which include: IGCC (Integrated Gasification Combined Cycle), fuel cells, H_2 production, chemical production, and/or synthetic oxygenated fuels. The ultimate goals of this project are to improve the state-of-the-art of coal or coal-biomass gasification technologies by:

- Creating a single step for tar, methane, light hydrocarbons, and NH_3 removal
- Replacing/eliminating a water-gas shift step
- Recycle the Fischer-Tropsch (FT) tail gas to provide heat for the steam reforming and maximize yield to FT liquids
- Sulfur tolerant catalyst allows use of system at any stage of clean up
- Adaptable to virtually any coal or coal-biomass gasification technology
- Increasing the energy efficiency of gasification using high-temperature clean up

4 EXPERIMENTAL METHODS AND EQUIPMENT

4.1 EXPERIMENTAL GOALS

The purpose of the experimental program was to develop a catalyst and demonstrate its performance under severe deactivating conditions expected from coal-derived syngas. For the experiments conducted on the lab scale reformer system, the primary objectives are to demonstrate the performance of each catalyst. Specifically the experiments need to:

- Quantify conversion of methane, tars, and ammonia;
- Demonstrate that the conversion is to desired products, CO, H_2, and in the case of ammonia decomposition, N_2
- Quantify the amounts of all products, including CO_2 and light hydrocarbons

Relevant reactions are steam reforming of methane

1) $CH_4 + H_2O \rightarrow CO + 3H_2$,

Steam reforming of toluene (representative of tar destruction)

2) $C_7H_8 + 7H_2O \rightarrow 7CO + 11H_2$,

Ammonia decomposition

3) $2NH_3 \rightarrow N_2 + 3H_2$

And water gas shift.

4) $CO + H_2O \leftrightarrow CO_2 + H_2$

Ammonia can also react with methane to form HCN

5) $CH_4 + NH_3 \leftrightarrow HCN + 3H_2$

At the temperatures of the experiments, no HCN should be formed, and all of the ammonia and toluene should be completely destroyed. Note that reactions 1 through 3 all result in increases in the number of moles and the flow rate exiting the reactor. Because of this issue, the experimental data needs to quantify inlet and outlet flows and concentrations to determine the performance of the catalysts.

4.2 LABORATORY SCALE EXPERIMENTAL PROCESS

Figure 2 shows a diagram of the laboratory scale apparatus. Three mass flow controllers (Brooks Instruments) are available to control the flow rate of hydrogen, nitrogen, or simulated syngas. Simulated syngas approximating either Lurgi or TRIG gasifiers was tested with varying H_2S concentrations. Table 1 shows the target compositions for syngas fed to the catalyst bed. Laboratory experiments focused on Nitrogen is also used to provide a setpoint for the Equilibar back pressure regulator that sets the reactor pressure. Toluene was supplied as a tar simulant using a Teledyne Model 100DM Syringe pump equipped with their ISCO controller. Water (for steam) was supplied with an Eldex Optos Model 1LM metering pump. The toluene and water were mixed with the syngas mixture and vaporized in a static mixer prior to being fed to the tubular catalytic reactor purchased from MTI

Corporation. Table 2 shows typical flow settings for gas and liquid flows for each experiment. Temperature was controlled by their OTF-1200X single zone programmable tube furnace. Figures 3 and 4 show a diagram of the reactor tube and a loading diagram, respectively. A Julabo FP 35 chiller is then used to cool and condense water out of the syngas exiting the reactor prior to online analysis of the remaining gas using an online gas chromatograph purchased from SRI Instruments, a modified MG3 configuration of their 8610 GC. Data from SRI's PeakSimple software was periodically downloaded intro a spreadsheet to calculate methane, tar, and ammonia conversion as well as the relative changes in CO, H_2, and CO_2. Pressure was initially measured using gauges, but an Omega DPG409-500G digital pressure gauge was installed for later experiments. A sulfur specific sorbent was used to reduce the H_2S concentration leaving the process and entering the hood. The entire system was installed in a walk-in fume hood for all experiments conducted with H_2S concentrations greater than 90 ppm (Figure 5). An Agilent Model ADM 1000 flowmeter is available to check flows leaving the system.

Table 1. Syngas composition of coal gasification (Probstein and Hicks 2006; Driscoll 2008)

	H_2	CO	CO_2	CH_4	H_2O	H_2S	NH_3	Tar1
TRIG - lignite	29.2%	34.3%	13.6%	2.5%	18.9%	100-250 ppmw	.28%	.10%
Lurgi's FBDB - lignite	18.7%	7.5%	15.7%	5.2%	51.6%	.29%	.58%	.41%

Table 2. Typical flow rates for TRIG and Lurgi simulated syngas experiments

TRIG	Space Velocity	Custom Gas Mix Flow	Water/ Ammonia	NH4OH	Tar
	cc/gcatalyst/ hr	SCCM	ml/min	g/l	ml/min
	12000	161.2	0.029	28.81	0.00088
	24000	322.3	0.058	28.81	0.00176
Lurgi	cc/gcatalyst/ hr	SCCM	ml/min	g/l	ml/min
	12000	94.9	0.078	21.86	0.00361
	24000	189.8	0.156	21.86	0.00722

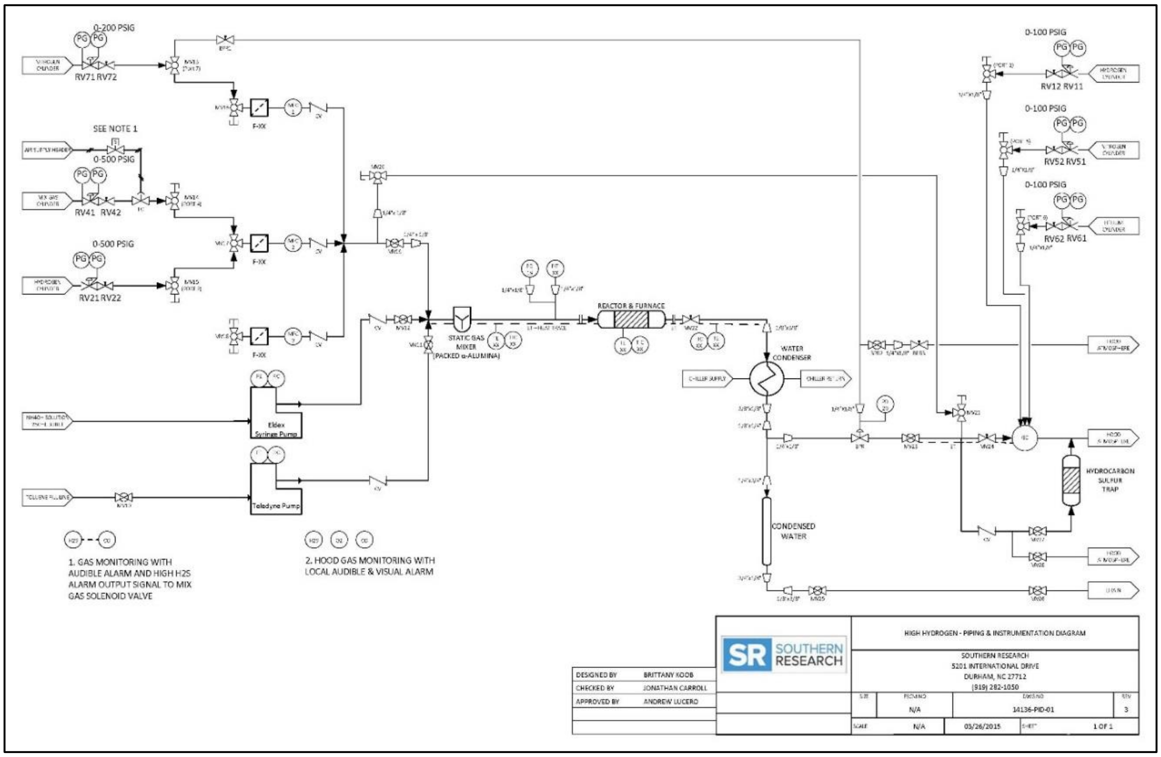

Figure 2. Process flow diagram for laboratory steam reformer system.

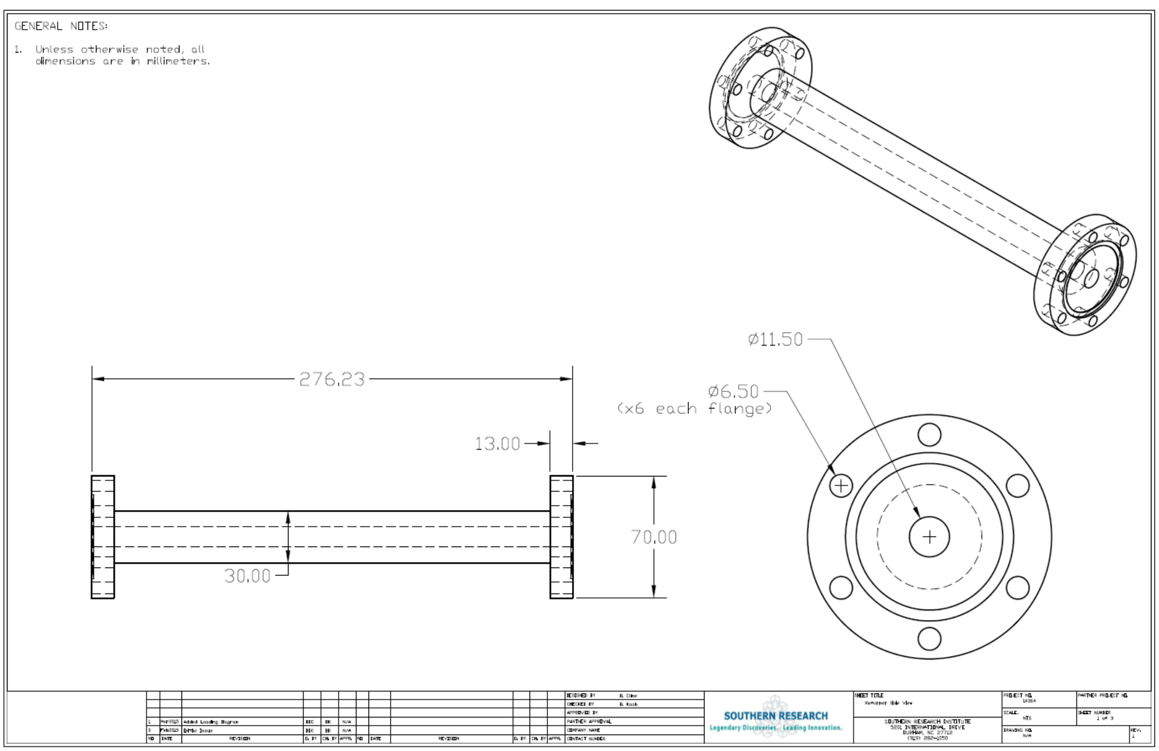

Figure 3. Diagram of high temperature reforming reactor.

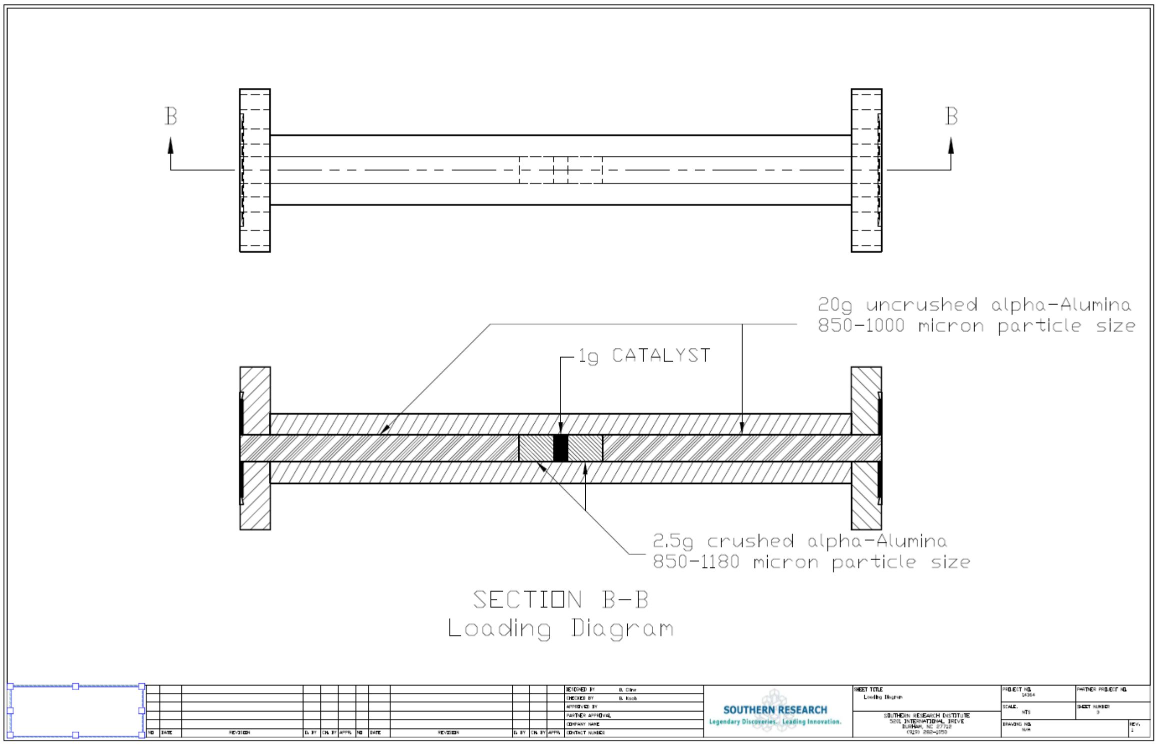

Figure 4. Catalyst loading diagram for laboratory steam reformer.

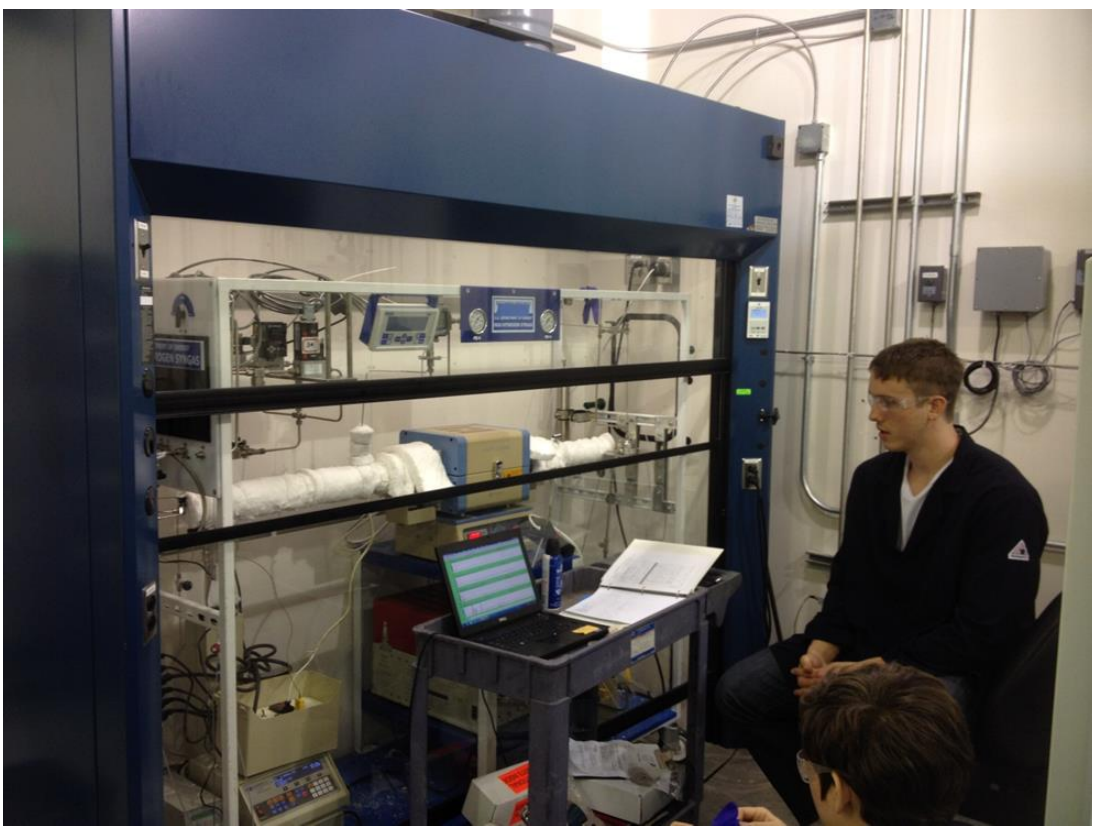

Figure 5. Reforming reactor skid installed in walk-in hood.

4.3 CATALYST CHARACTERIZATION

A series of catalysts was prepared using standard techniques such as coprecipitation and impregnation. Active metals were included with promoters and supports. A Micromeritics ASAP2020 was used to test the pore size distribution, mesoporous pore volume, and BET surface area. Samples were also characterized by X-ray diffraction to determine crystalline phases and dimensions. Several catalyst samples were characterized before and after reforming experiments.

5 RESULTS AND DISCUSSION

5.1 STEAM REFORMING CATALYST TESTING

5.1.1 CATALYST SCREENING

Initial experiments were conducted using a simulated TRIG gasifier syngas first with the reactor empty, second with the reactor loaded with alumina, and finally with a catalyst synthesized at Southern. Almost all tests were conducted at space velocities of 12000 and 24000. Although initial plans were to screen catalyst performance with simulated syngas containing no sulfur then check the best catalysts in the presence of H_2S, the project team decided to conduct all catalyst screening tests with a 35 ppm of H_2S since tolerance is a primary goal of the study. Subsequent deactivation tests increased the concentration to 90 ppm then 250, and 500. Experiments were typically conducted for at least 24 hours, but many were extended for 48 hours.

Figure 6 shows methane conversion data for tests conducted with the empty reactor and with the reactor filled with alumina. Early catalyst tests at low and high pressure are shown for comparison. Virtually all the ammonia and toluene (used as a simulant for tar) were destroyed in the empty reactor, but methane conversion was less than 10% with no additional hydrogen formed in the reaction. Results with the reactor filled with alumina were identical to the empty reactor. Neither toluene nor ammonia were detected after the reformer in any experiment conducted. Figure 7 shows a summary of the average performance of several catalysts synthesized for this project.

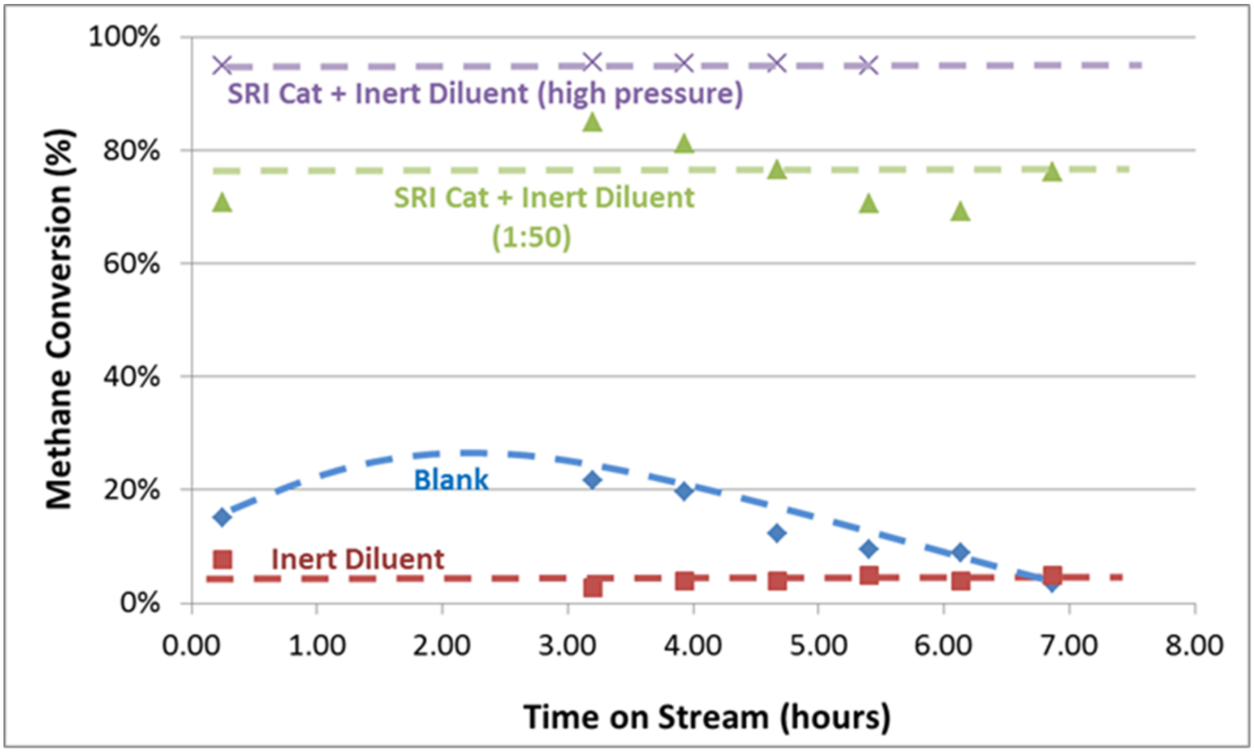

Figure 6. Methane conversion for empty and inert filled reactor compared to active catalyst.

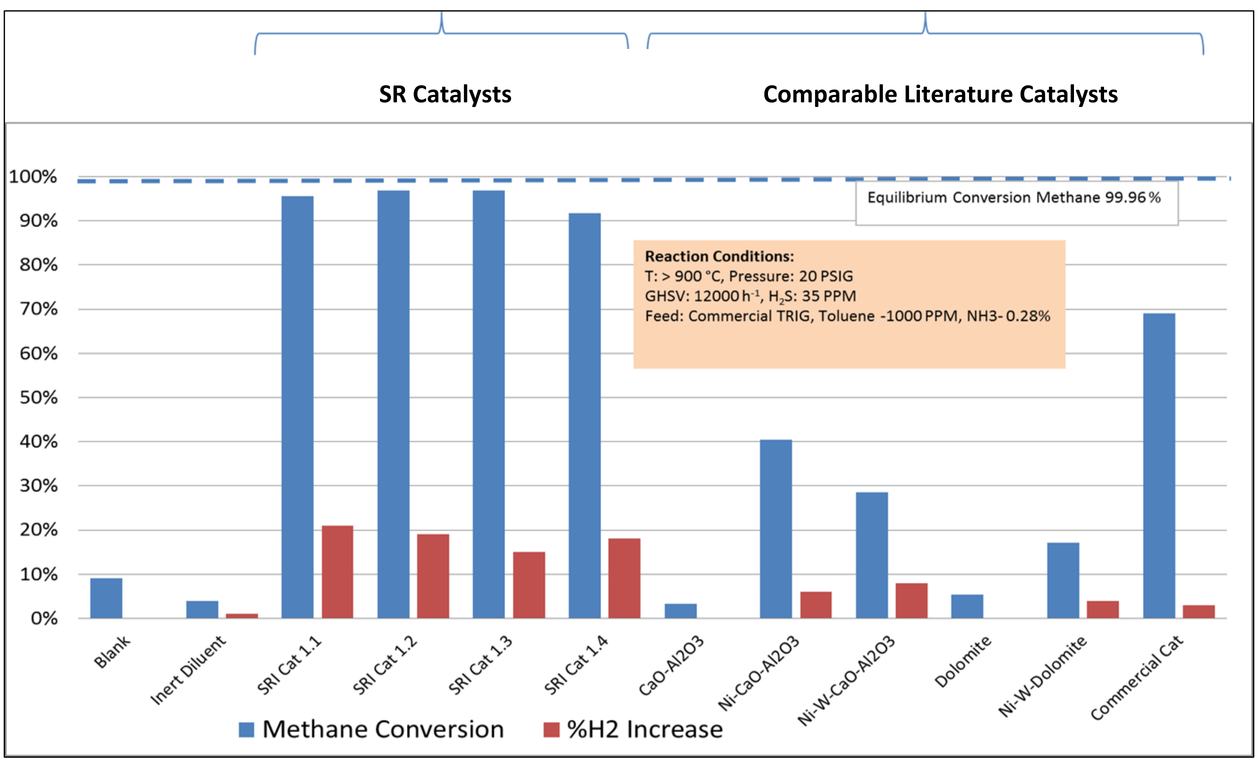

Figure 7. Comparison of the performance of several catalysts.

SR catalyst formulas designated as 1.1 and 1.2 had the highest overall performance and were chosen for further tests. Catalyst 4.7 (not shown on the figure) is closely related to catalyst 1.1 and was also subjected to extensive testing. Experiments were generally conducted for 24 to 48 hours, but individual catalyst samples were used for multiple experiments for up to about 200 hours of time for catalysts to be subjected to syngas containing sulfur. As mentioned previously, the offgas composition was measured throughout the experiments using and online GC purchased from SRI Instruments. This allowed for calculations of conversion and the hydrogen to CO ratio approximately every 30 minutes. Figures 8 and 9 show graphs of typical calculations over time for tests with catalyst 1.1 using simulated TRIG and Lurgi gasifier syngas feed, respectively. Both have high methane conversions through the test. With only steam that would come from the syngas, hydrogen increased slightly for both tests, about 15% for the TRIG syngas, and about 20% for the Lurgi syngas (Lurgi syngas has more methane and tar to be converted to CO and H_2). With the TRIG syngas we observed a slight increase we observed a slight increase in the H_2 to CO ratio, but with the Lurgi syngas the H_2 to CO ratio decreased. Note that the ratio the TRIG syngas starts at 0.85 and the ratio in Lurgi syngas is at 2.49. In the TRIG syngas the CO_2 concentration is 40% of the CO concentration where in the Lurgi syngas, the CO_2 concentration is over twice that of the CO. The reverse of the water-gas shift reaction is driving the already too high for FT H_2:CO ratio down for experiments in the reformer with the Lurgi gasifier syngas. Although the H_2 to CO ratio did not increase significantly in experiments using simulated TRIG syngas, the yield of syngas (CO and H_2) increased by about 15% on the simulated TRIG feed with the small amount of methane and tar present before recycle of lights from the FT system. Figure 10 shows these effects during a typical experiment. Figure 11 shows several experiments conducted with the same catalyst totaling over 200 hours of exposure to 35 ppm H_2S. Modifying experimental conditions such as feed gas, pressure, and space velocity all had significant effects on the methane conversion, but the catalyst performance was stable for about 200 hours.

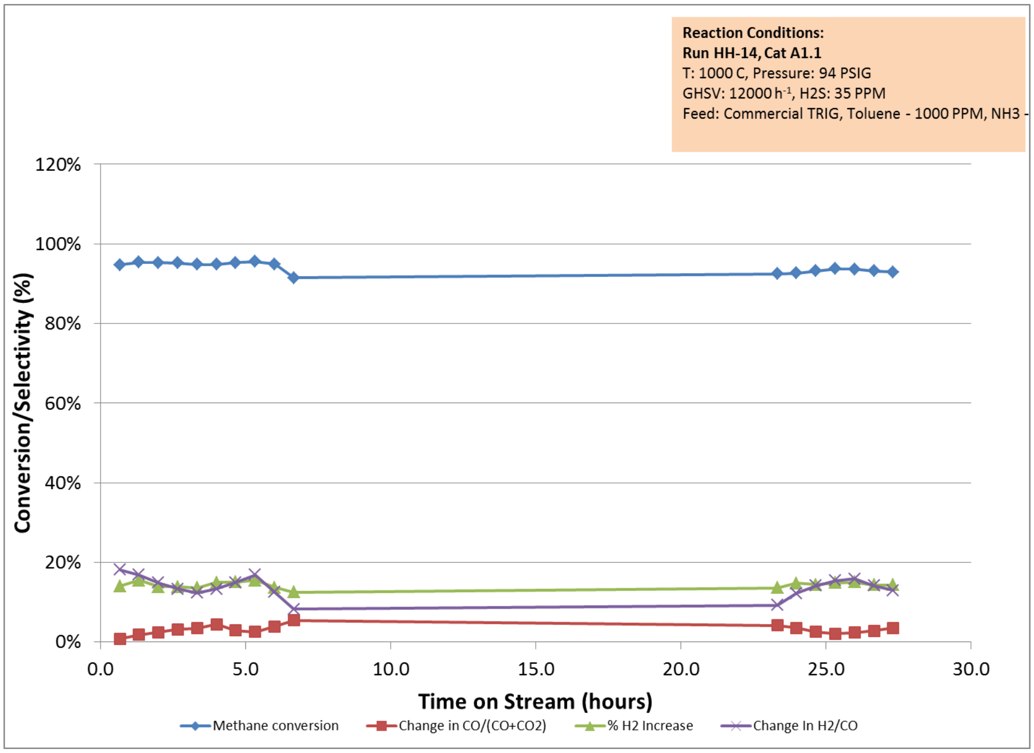

Figure 8. Performance of catalyst 1.1 over time using simulated TRIG gasifier feed except 35 ppm H_2S.

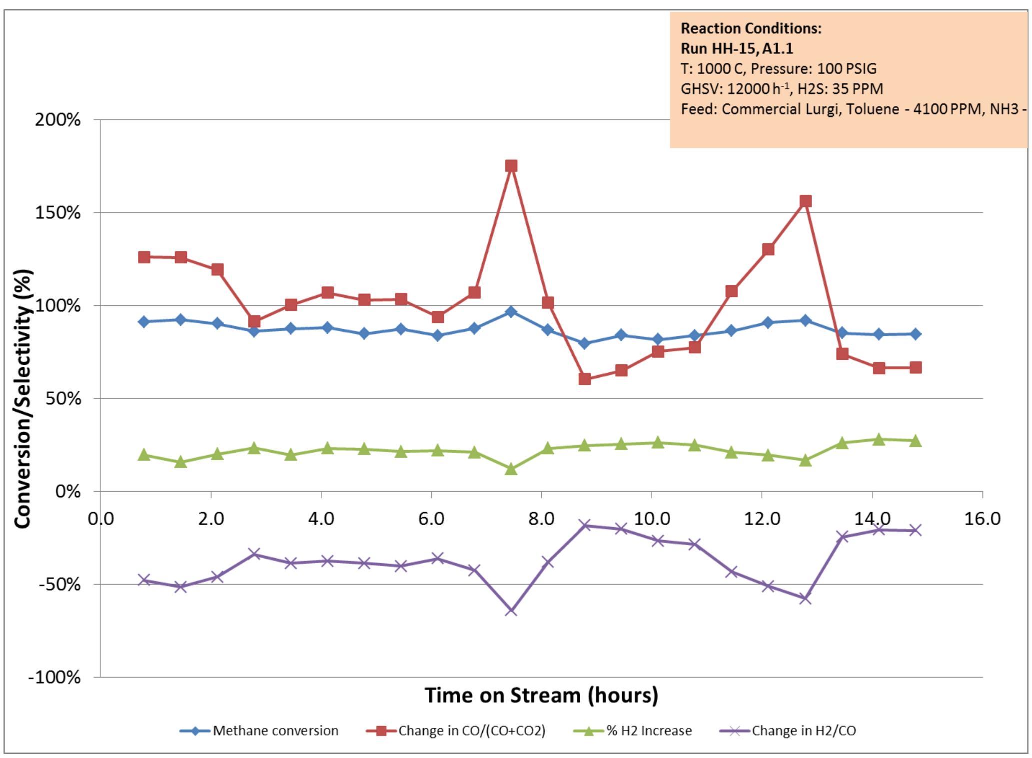

Figure 9. Performance of catalyst 1.1 using simulated Lurgi syngas feed except for 35 ppm H_2S.

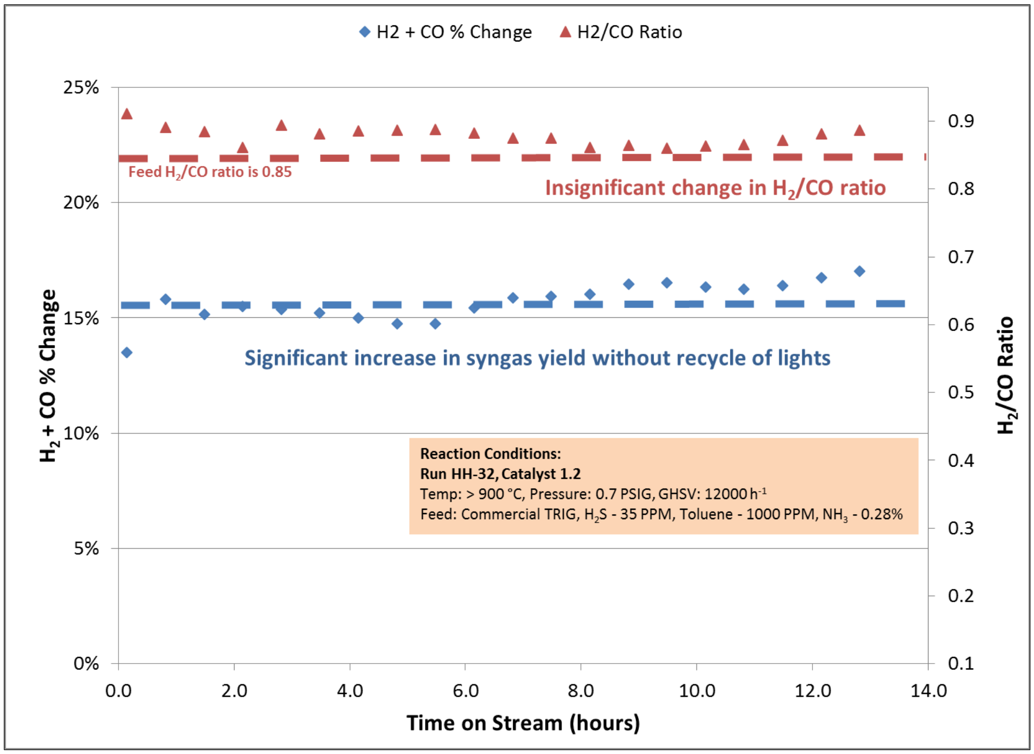

Figure 10. Effect of reformer on syngas yield and H_2:CO Ratio

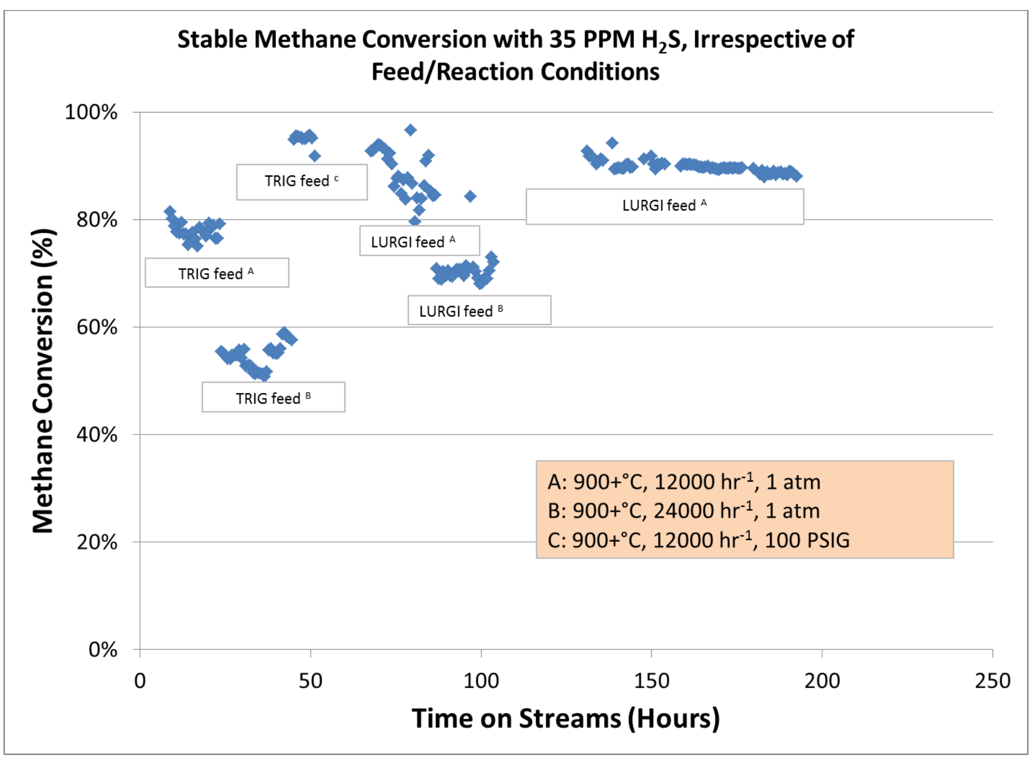

Figure 11. Stable performance of catalyst 1.1 over several experiments with H_2S

A total of 45 screening experiments were conducted on 15 catalyst formulations including one commercial sample. Nearly all tests were conducted at space velocities of 12000 and 24000. Virtually all the tests were conducted under conditions that would be expected from the gasifier – CO, H_2, CO_2, CH_4, H_2O, NH_3, and tar concentrations as expected from raw gasifier feed. As mentioned previously, Screening tests are conducted with 35 ppm H_2S. Toluene is used to simulate tars for this program.

Figures 12 and 13 show a comparison of the performance of catalyst 1.2 and a commercial reforming catalyst sample at identical conditions. The commercial catalyst appeared to be slowly losing activity while the methane conversion was trending slightly up over time with Southern's 1.2 catalyst. Southern's catalyst also had significantly higher methane conversion and slightly higher syngas yield accompanying the higher methane conversion.

Figure 14 shows the effect of increasing space velocity on catalyst 1.2. As expected, conversion and resulting syngas yield drop as space velocity is increased. Figure 15 shows the effect of pressure on catalyst 1.2. The results indicate that moderate pressure increases appear to increase conversion. The H_2 to CO ratio appeared to drop with pressure increases. Figure 16 shows the conversion and H_2 to CO ratio change on catalyst 1.1 with both TRIG and Lurgi feed. The conversion is higher with the Lurgi feed, likely due to the high concentration of water in the stream, but the H_2 to CO ratio decreased significantly due to reverse water-gas shift from the high CO_2 concentration in the feed.

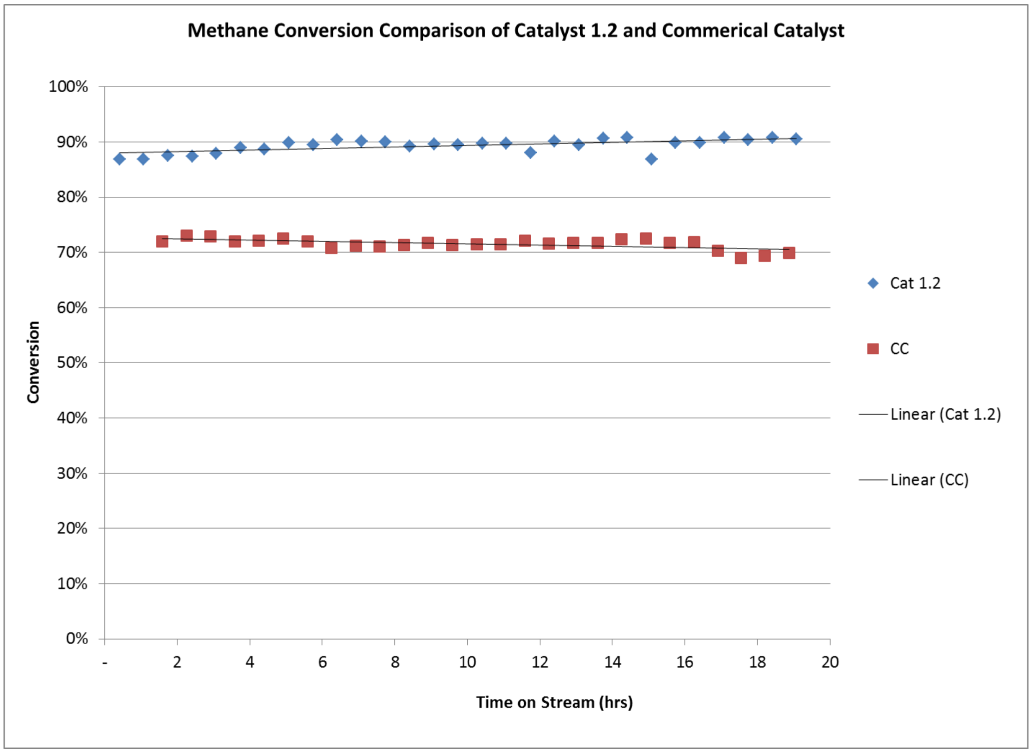

Figure 12. Comparison of methane conversion versus time for Southern catalyst 1.2 and commercial reforming catalyst

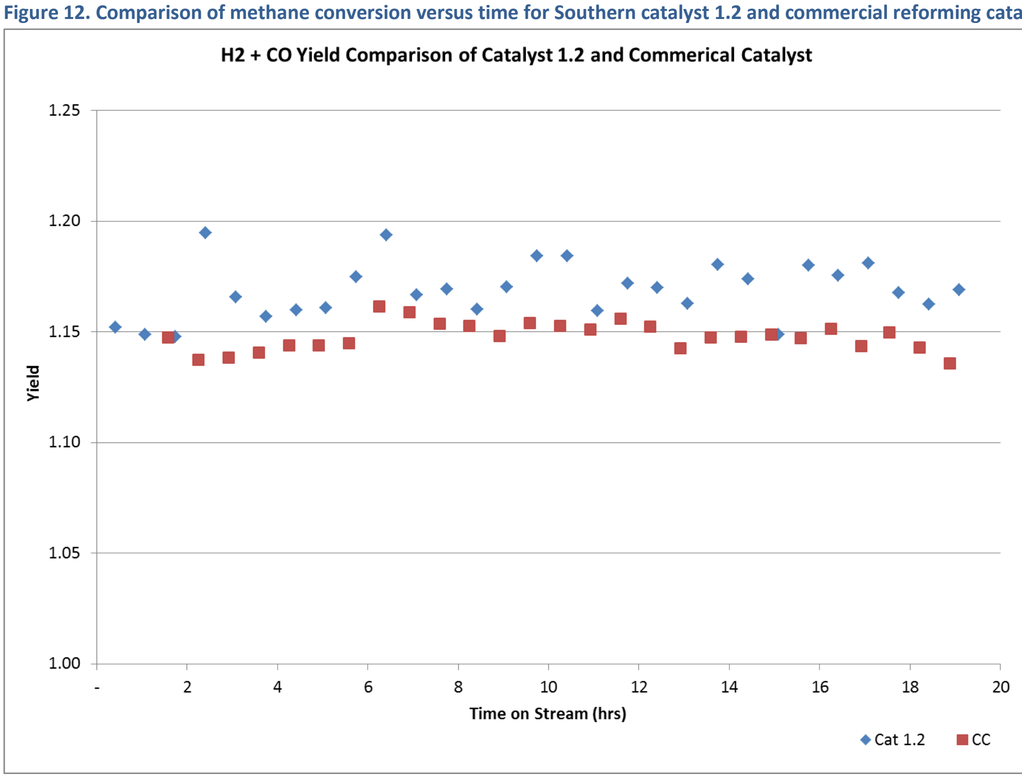

Figure 13. Comparison of additional syngas yields for Southern catalyst and the commercial catalyst.

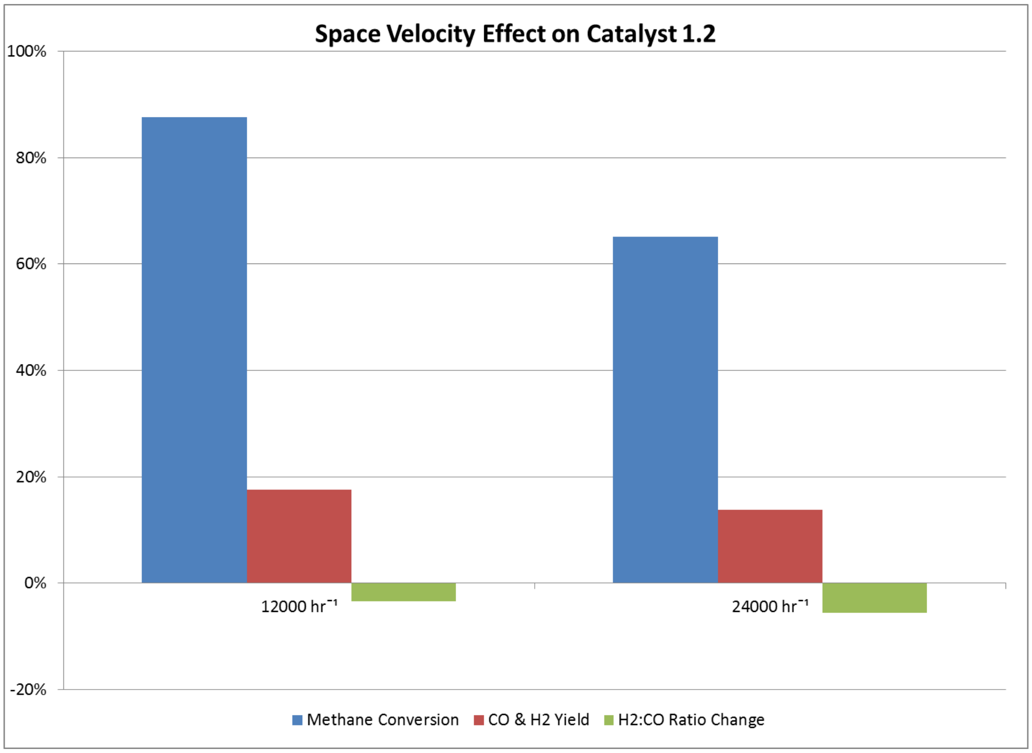

Figure 14. Effect of space velocity on catalyst performance

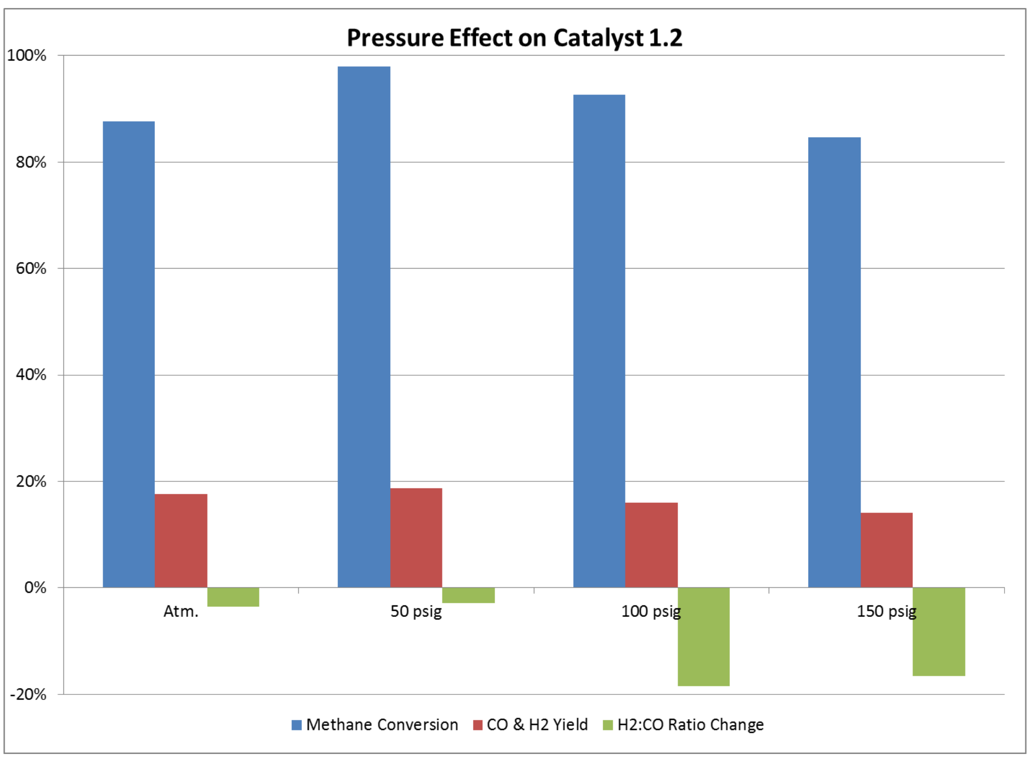

Figure 15. Effect of pressure on catalyst 1.2

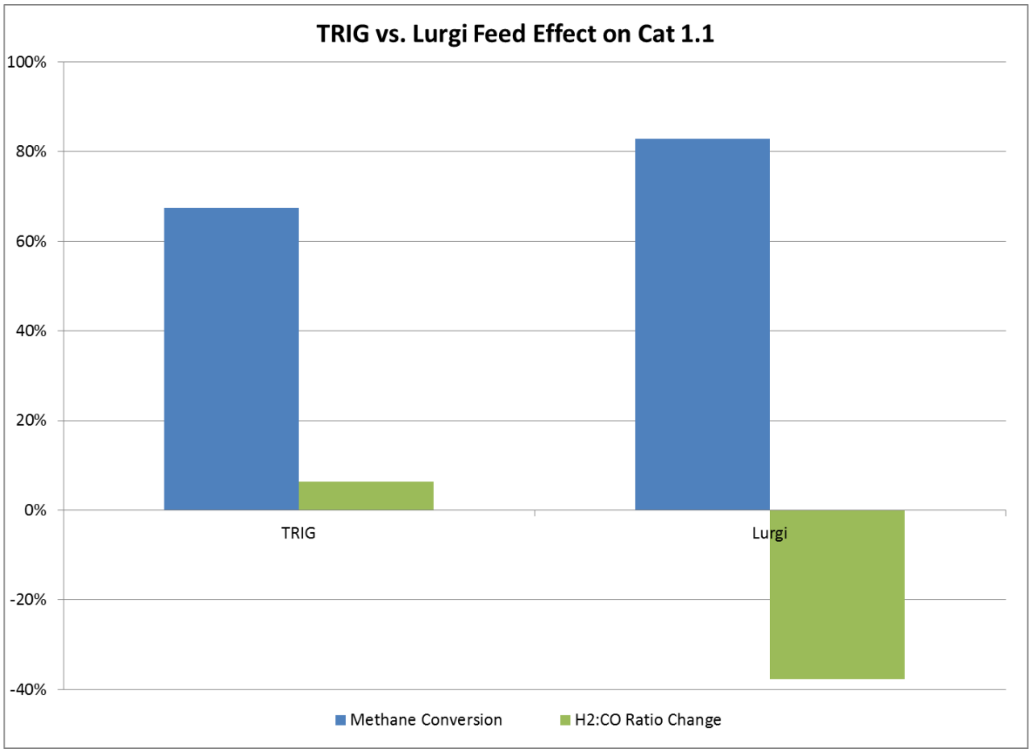

Figure 16. Effect of feed composition on methane conversion and H_2 to CO ratio

5.1.2 CATALYST DEACTIVATION

The primary objectives for deactivation were to increase the concentration of H_2S in simulated experiments. Figure 17 shows a comparison of the results for catalyst 1.1 at 90 ppm which similar operating conditions using 35 ppm gas. Within the scatter in the data little or no deactivation was observed in the performance of the catalyst at 90 ppm. Based on results from previous experiments, one final catalyst formulation was prepared and characterized. Table 3 summarizes operating conditions for 12 deactivation experiments conducted. Two catalyst samples designated as 1.2 and 4.7 were utilized in experiments containing simulated Lurgi or TRIG gasifier syngas containing H_2S concentrations from 90 ppm up to 500 ppm. Catalyst 4.7 is similar to catalyst 1.1 with an additional metal added. Most of the experiments were conducted at slightly over atmospheric pressure and space velocity of 12000.

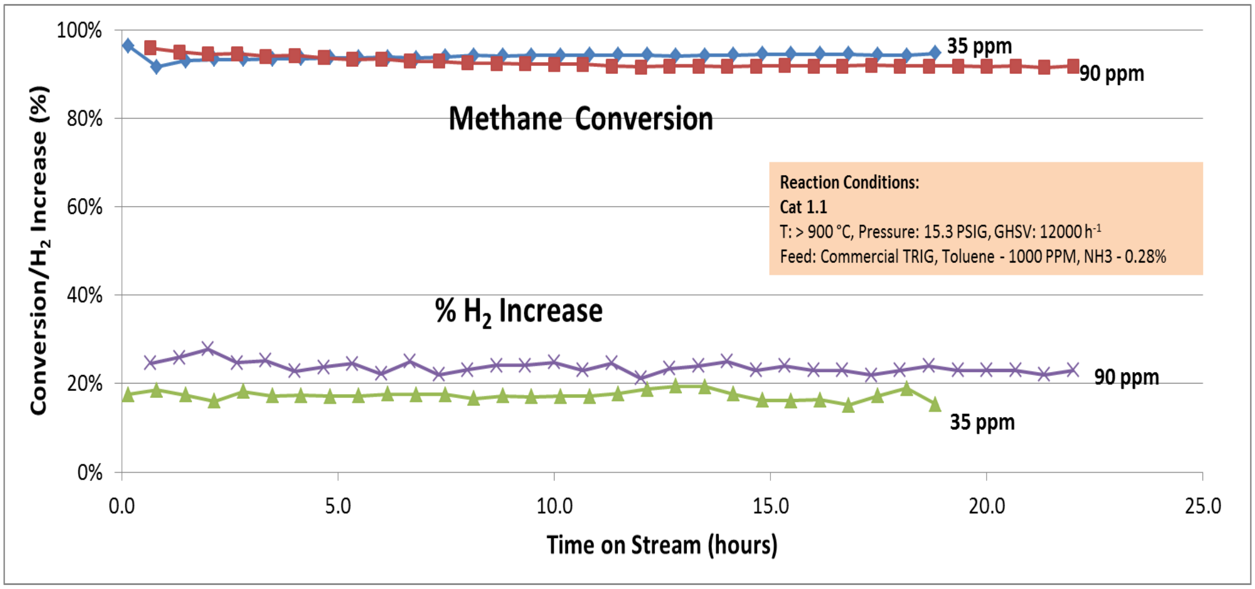

Figure 17. Comparison of catalyst performance with H_2S concentrations at 35 ppm and 90 ppm.

Table 3. Summary of catalyst deactivation experiments conducted.

Date	Run No.	Catalyst	Feed Mixture	H_2S PPM	Total time on stream for catalyst sample.
2/20/15	51	4.7	TRIG	90	48
2/23/15	52	4.7	Lurgi	500	96
2/27/15	53	4.7	Lurgi	500	144
3/5/15	54	4.7	Lurgi	500	192
3/9/15	55	1.2	TRIG	90	48
3/11/15	56 (no regen)	1.2	TRIG	250	96
3/17/15	57 regen	1.2	TRIG	250	144
3/19/15	58	1.2	TRIG	500	192
3/25/15	60	4.7	Lurgi	90	48
3/31/15	61	4.7	Lurgi	500	96
4/8/15	62	4.7	Lurgi after Regen	500	120
4/21/15	63	1.2	TRIG	250	48

Each experiment was conducted for 48 hours, although individual catalyst samples were exposed to syngas containing H_2S for up to 192 hours. Most 48 hour experiments were followed by a high temperature regeneration treatment for one hour in 5% hydrogen. Figures 18 through 29 show methane conversion and change in hydrogen concentration for each experiment. The steam reforming experiments were conducted using the steam available in the syngas. Additional steam was not added for these experiments, though it is recognized that the resulting hydrogen concentration can be manipulated with additional steam.

Figures 18 and 22 show that both catalysts cause high conversion of methane in TRIG syngas with catalyst 4.7 slightly higher with less deactivation over time. Catalyst 1.2 deactivated more rapidly in 250 ppm H_2S as shown in Figures 23, 24, and 29. Figures 23 and 24 show experiments where the catalyst had previously been exposed to H_2S and Figure 29 shows an experiment with a fresh catalyst sample exposed to 250 ppm H_2S. Figure 25 shows the same catalyst in a 500 ppm H_2S run. In every case with both catalysts, increasing the H_2S concentration immediately resulted in a decrease of methane conversion. In some cases the conversion remained steady at the lower conversion rate, but it was apparent that increasing the H_2S concentration immediately affected the catalyst performance. In the case of catalyst 1.2 the conversion was about 85% in the presence of 90ppm H_2S (figure 22). The conversion immediately decreased to about 70% in the presence of 250 ppm of H_2S (even for the fresh catalyst sample) and decreased to about 50% (Figure 23). The following run at 500 ppm (Figure 25) immediately decreased the conversion to 40% and continued deactivation from there. Similar effects were observed with catalyst 4.7 although it was not affected by 90ppm H_2S (Figure 18). With Lurgi syngas in 90ppm H_2S the conversion was about 90% (Figure 26). When H_2S was increased to 500 ppm (Figure 27), the methane conversion decreased about 70% (Figure 27).

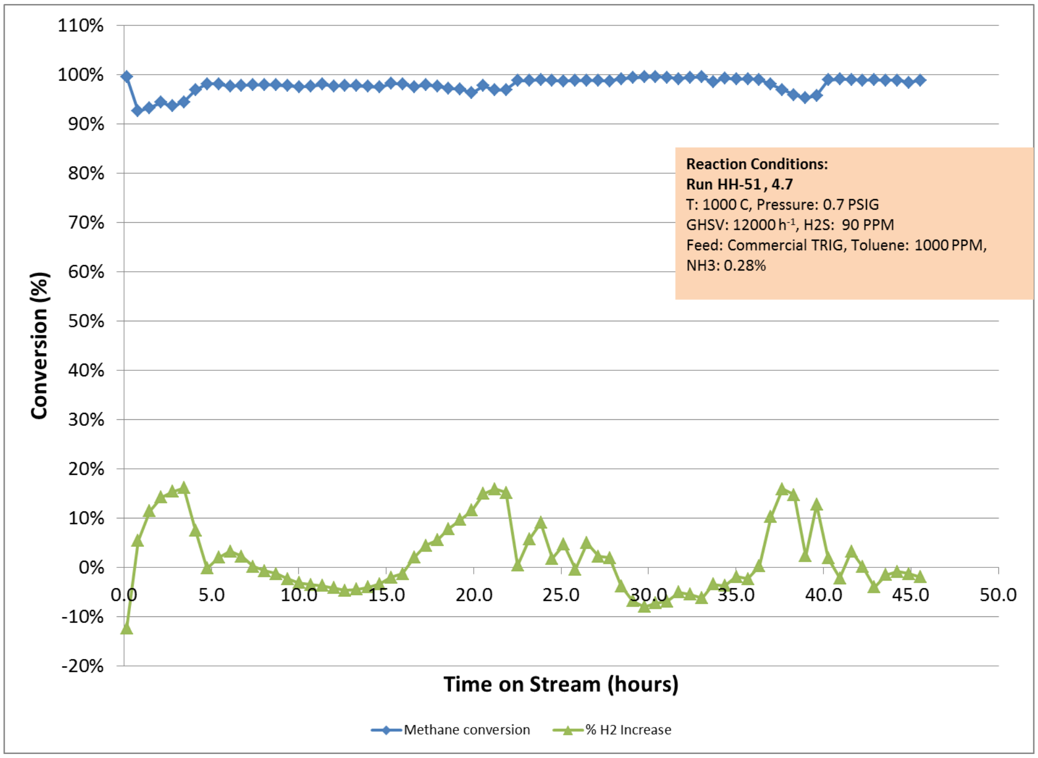

Figure 18. Deactivation experiment with catalyst 4.7 using simulated TRIG syngas containing 90 ppm H_2S.

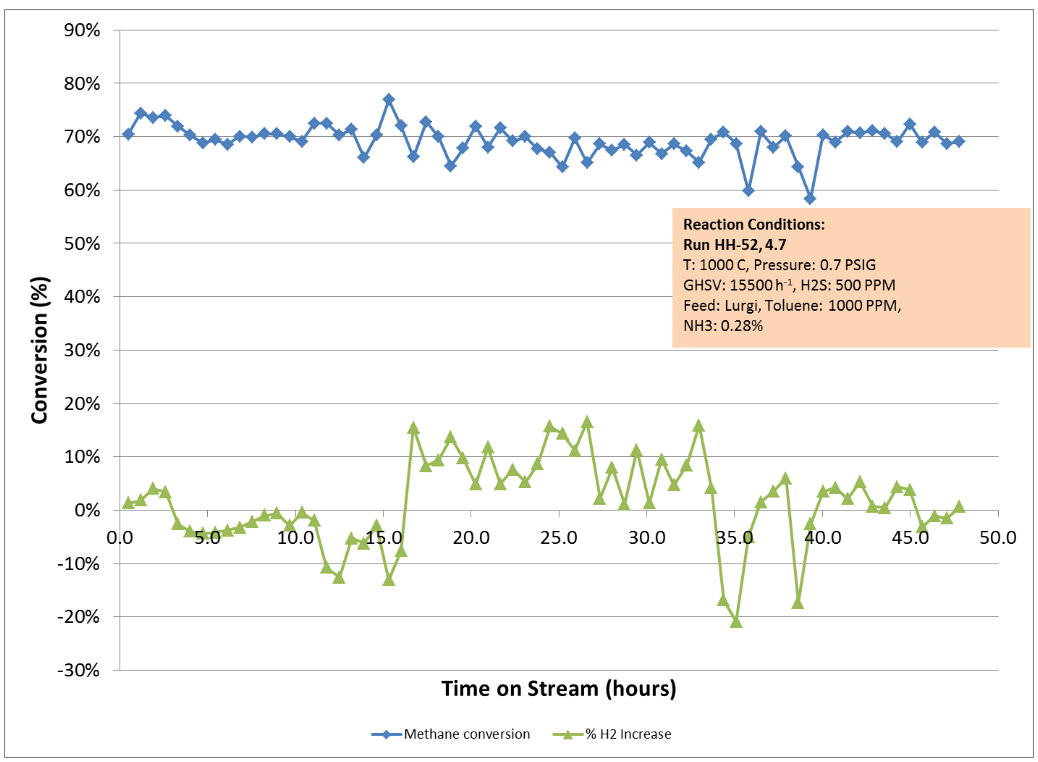

Figure 19. Deactivation experiment with the same catalyst 4.7 sample using a modified simulated Lurgi syngas containing 500 ppm H_2S after regeneration.

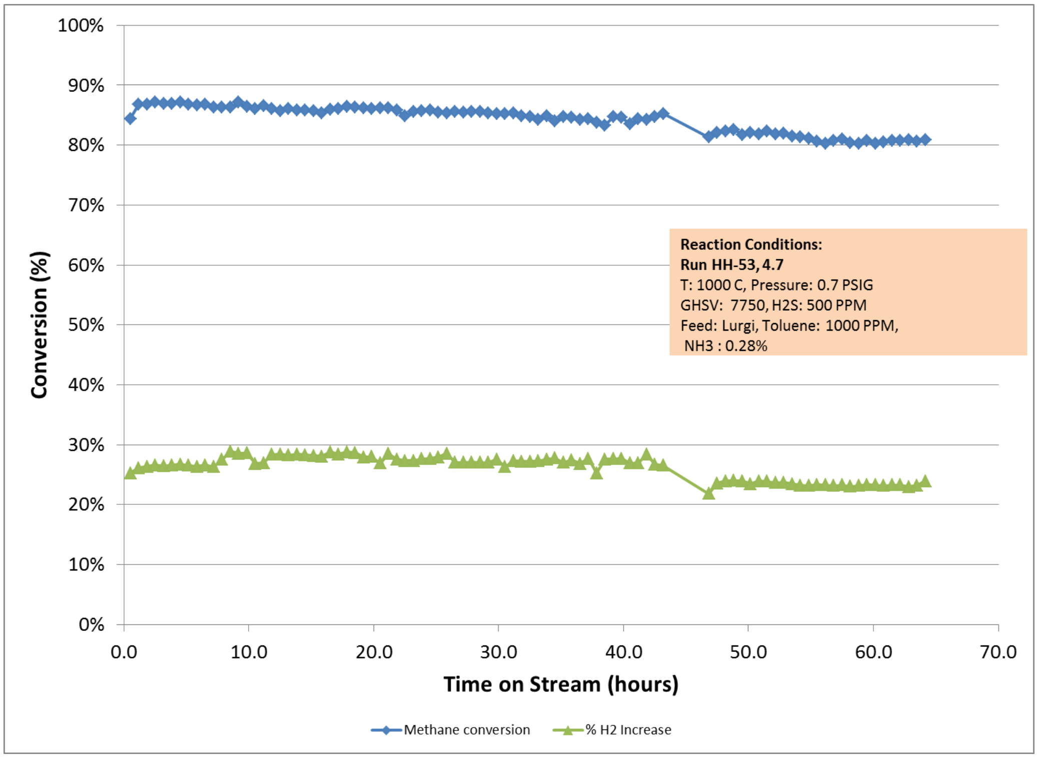

Figure 20. Deactivation experiment with the same catalyst 4.7 sample after regeneration using a modified simulated Lurgi syngas containing 500 ppm H_2S at lower space velocity.

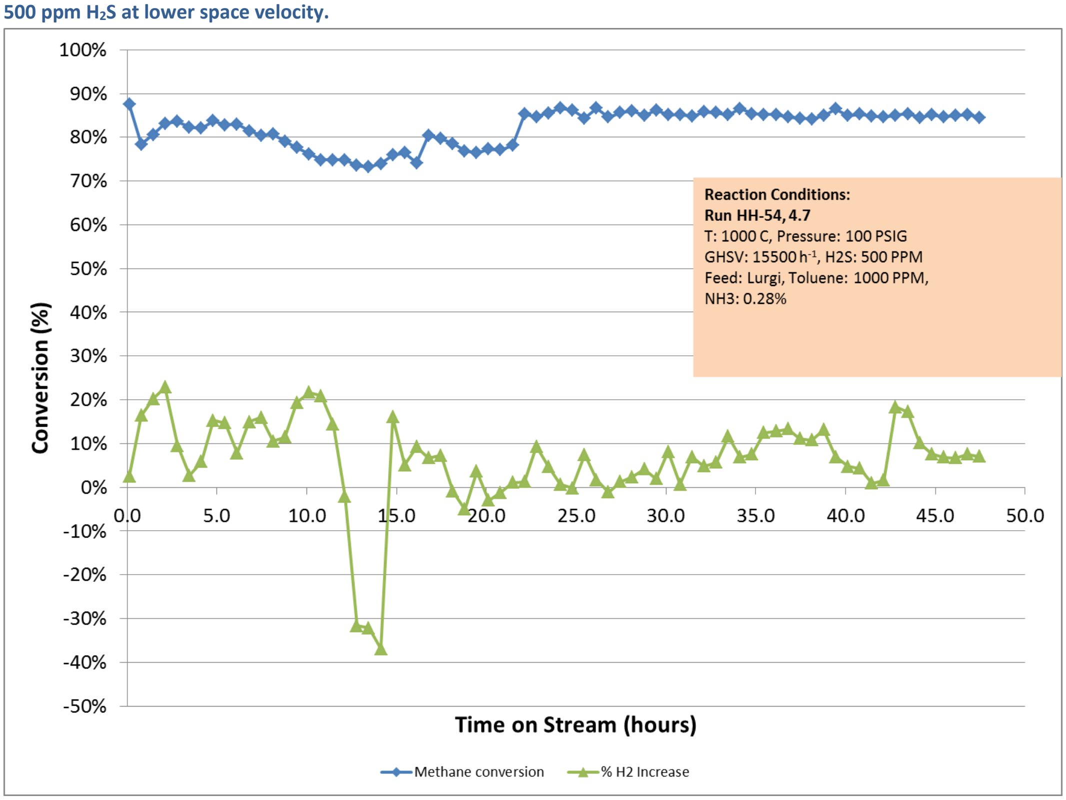

Figure 21. Deactivation experiment at 100 PSIG with the same catalyst 4.7 sample using a modified simulated Lurgi syngas containing 500 ppm H_2S after regeneration.

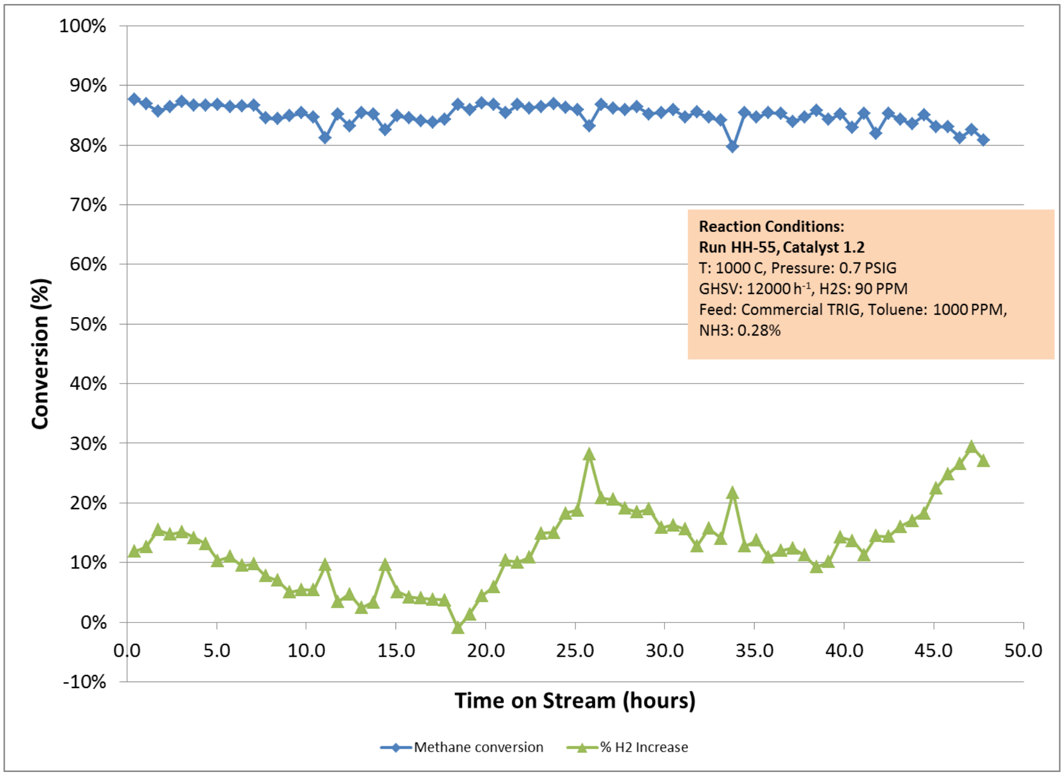

Figure 22. Deactivation test with catalyst 1.2 sample using simulated TRIG syngas containing 90ppm H_2S

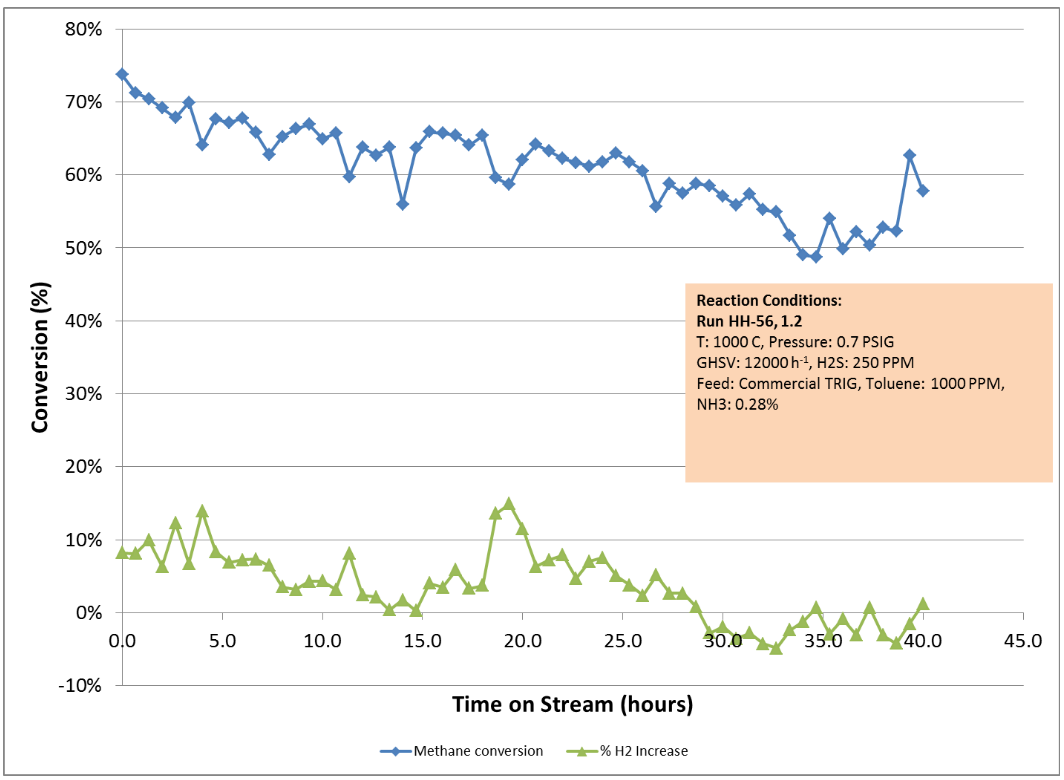

Figure 23. Deactivation test with same catalyst 1.2 sample using simulated TRIG syngas containing 250ppm H_2S without regeneration

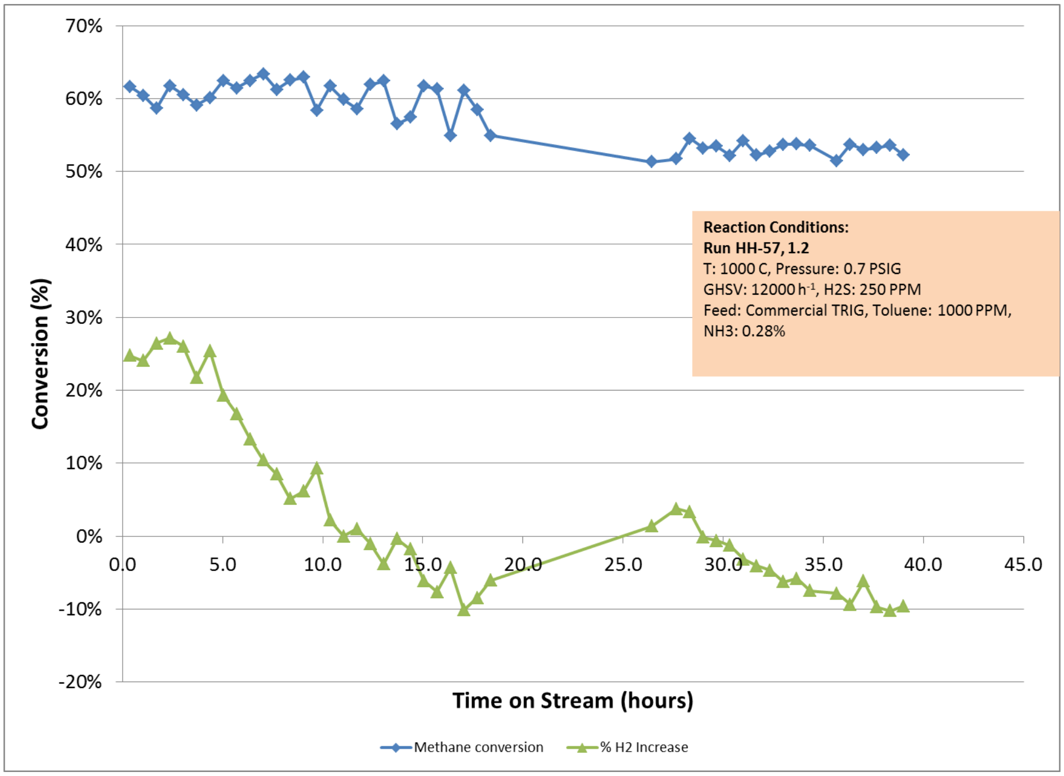

Figure 24. Repeat of test in Figure 23 with same catalyst sample after catalyst regeneration procedure

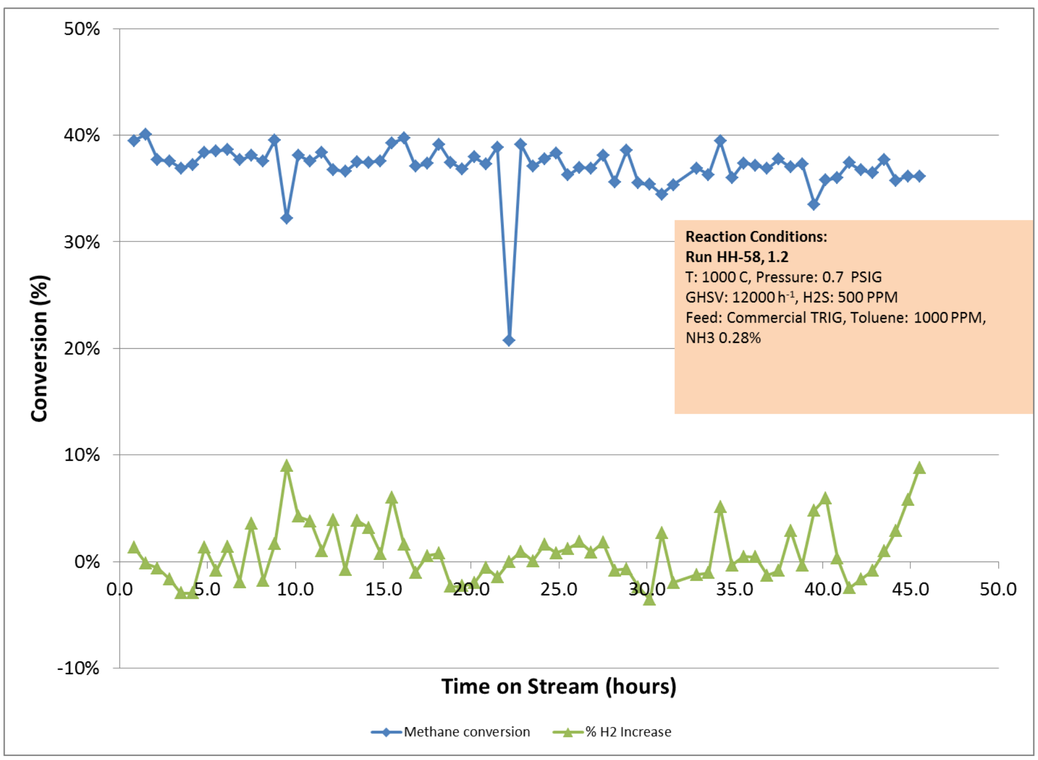

Figure 25. Catalyst deactivation test with same catalyst 1.2 sample using TRIG syngas containing 500 PPM H_2S after regeneration procedure.

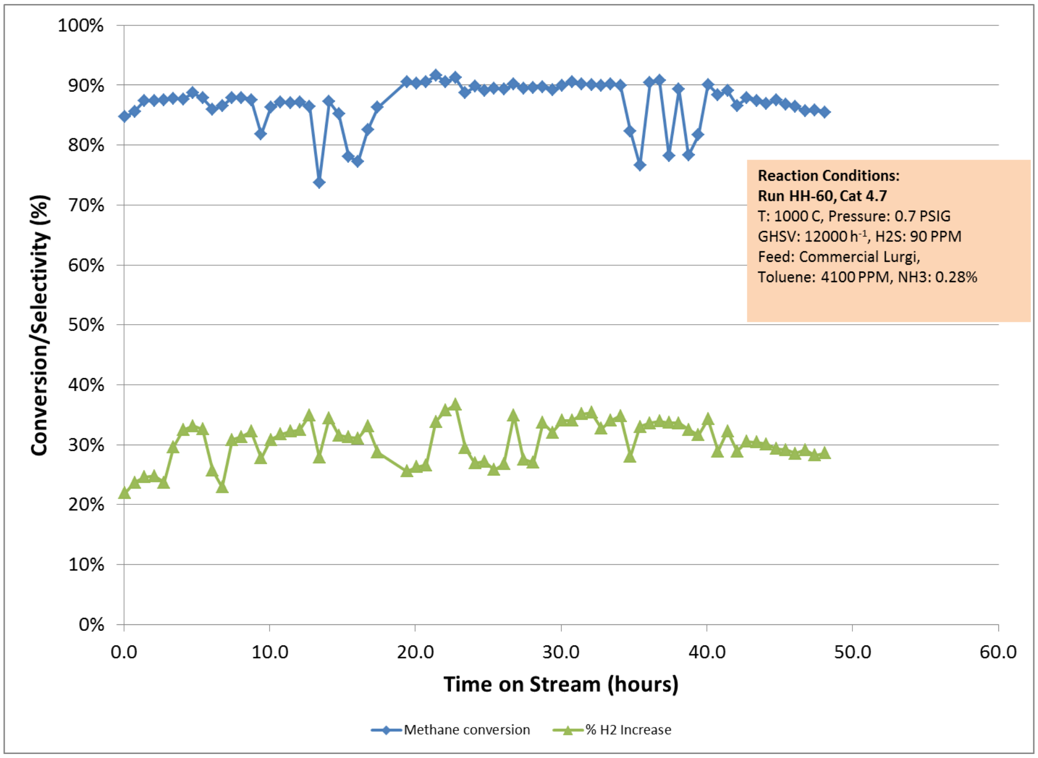

Figure 26. Catalyst deactivation test with new catalyst 4.7 sample using simulated Lurgi syngas containing 90ppm H_2S

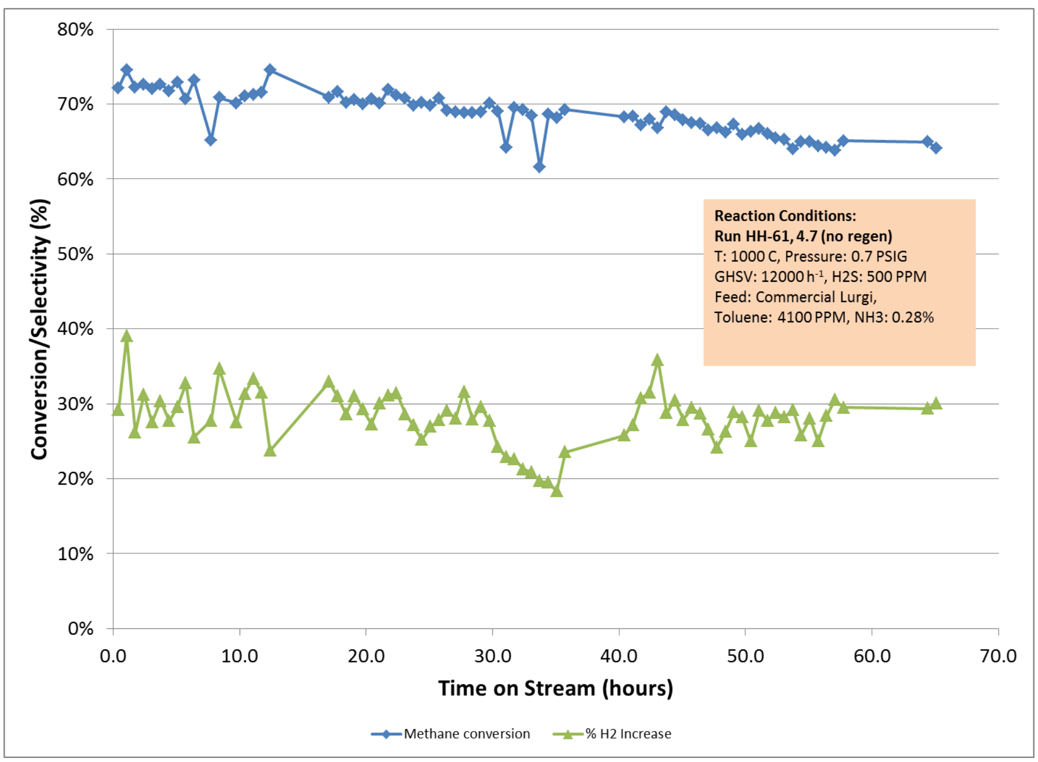

Figure 27. Catalyst deactivation test with same catalyst 4.7 sample using Lurgi syngas containing 500 ppm H_2S without regeneration.

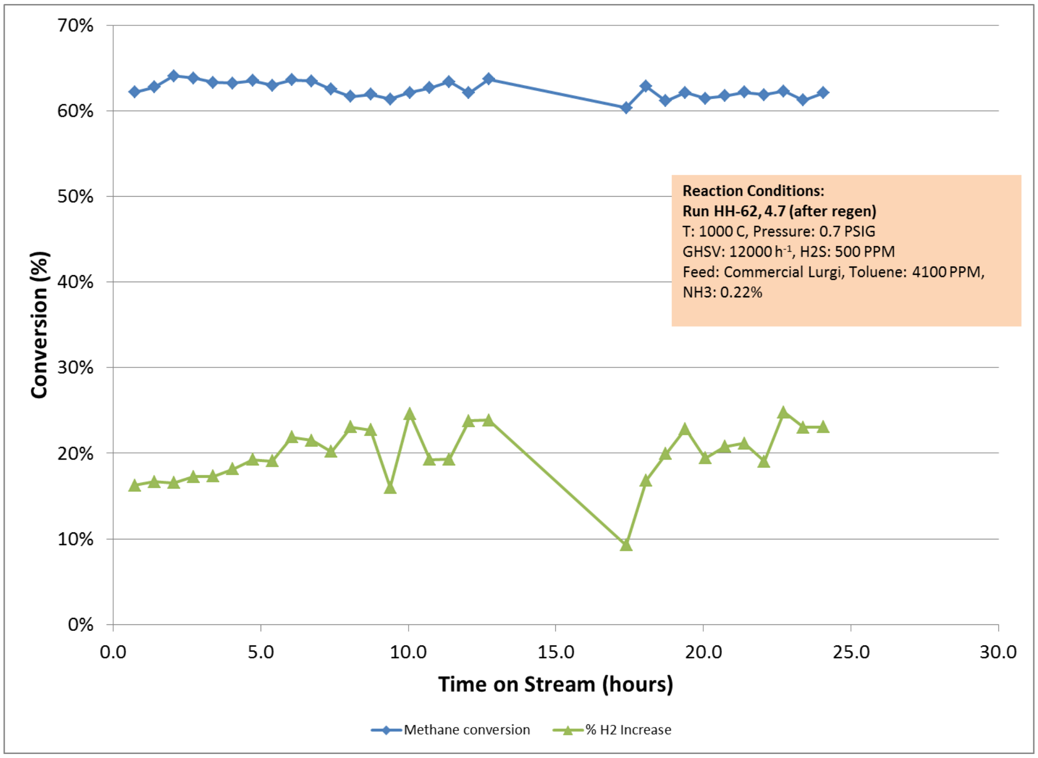

Figure 28. Catalyst deactivation test with same catalyst 4.7 sample using Lurgi syngas containing 500 ppm H_2S after regeneration.

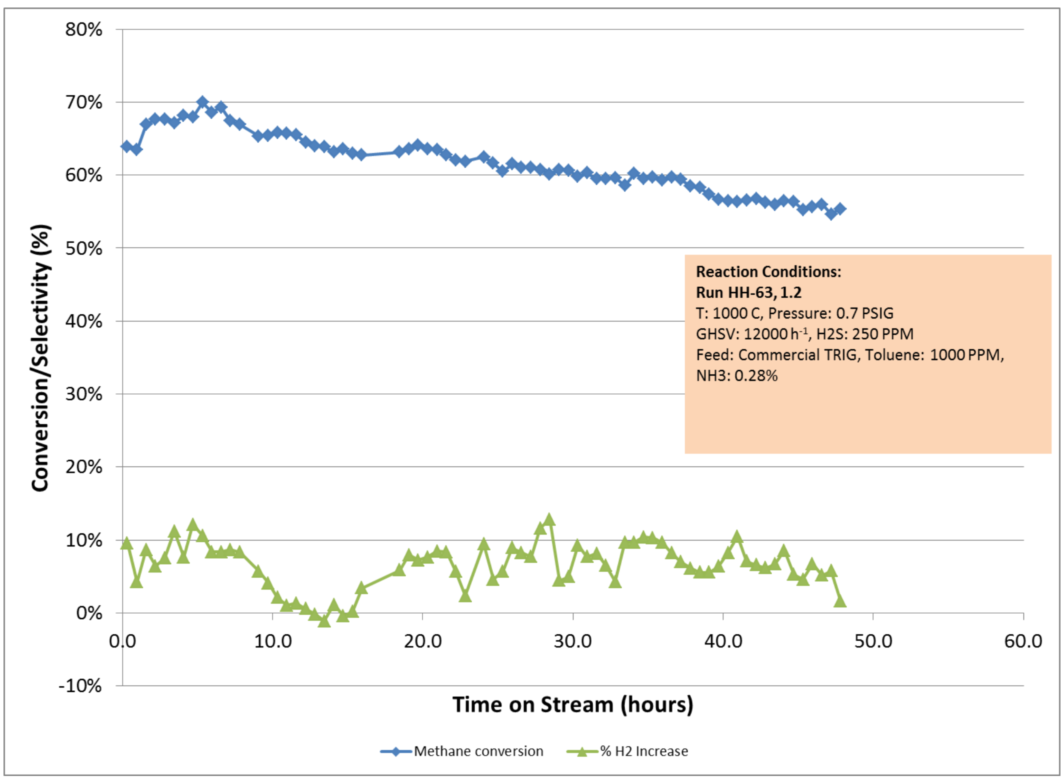

Figure 29. Catalyst deactivation test with fresh sample of catalyst 1.2 using simulated TRIG syngas containing 250 ppm H_2S.

5.2 TECHNOECONOMIC ANALYSES

To assess the steam reforming technology, several techno-economic and life-cycle analyses were conducted comparing a plant featuring the sulfur tolerant steam reformer and a base case without the technology. An in depth analysis was conducted by NEXANT Inc. on a CTL plant and IGCC using the TRIG gasifier based on ASPEN plus simulations provided by Southern. Additional limited scope cases utilizing more assumptions for IGCC and CTL plants were conducted by Southern using Lurgi and ThermoChem Recovery Inc. (TRI) gasifiers.

The primary comparison cases for the steam reformer are based on the Department of Energy's National Energy Technology Laboratory's cost and performance baseline reports for both IGCC and CTL. For each case an ASPEN Plus model simulation was generated to replicate the mass balance found in the NETL report. For the IGCC, the transport gasifier (TRIG) case was used as the base case. For the CTL case, the single Shell Gasifier was replaced with five TRIG gasifiers to provide the same liquid output. Both Southern Research (SR) comparison cases we built using these ASPEN simulations simply adding in the steam reforming block and associated utilities.

Once the TRIG IGCC and TRIG CTL simulations were completed for the base case and the modified SR cases the mass balances were provided to NEXANT Inc. Utilizing the mass balances NEXANT used the NETL reports as a guide for their TEA and LCA analysis. The resulting information was provided to Southern in the form of two reports which can be found in the Appendix.

Summary of Nexant IGCC Report

The methodology and results of the report can be seen in the Appendix, however, an excerpt of the findings of the Nexant IGCC study is shown below detailing the comparison between the basecase (P1) and the modified case (P2) with the proposed new technology:

- *Despite having a higher efficiency and lower COE, the reference Case P1 TRIG IGCC is unable to meet the 90% CO_2 capture criteria set out by DOE/NETL due to the high methane concentration in the TRIG syngas. Also, the fate of the tars and higher hydrocarbons in the raw syngas leaving the TRIG is unknown. There is likelihood that these heavy hydrocarbon compounds may condense and plug coolers downstream of the TRIG, rendering the whole process infeasible.*

- *Case P2 does not face the same issues as Case P1. With the addition of the SRI catalytic reformer, practically all the tars are destroyed and most of the methane is reformed, in the presence of oxygen to form CO and H2. The CO is converted to CO_2 downstream in the shift reactors, thus enabling a CO_2 capture rate of 90% or greater. Despite the higher capital cost and slightly reduced efficiency, this case meets DOE/NETL's CO_2 capture guidelines.*

- *Case P2 consumes more auxiliary power mainly due to the following:*
    - *More oxygen has to be supplied by the ASU as the catalytic reformer consumes oxygen to reform methane and other higher hydrocarbons to CO and H2*
    - *The CO_2 capture and compression process consumes more power since more CO2 is captured and subsequently compressed in Case P2.*

So while the modified case (P2) has a slightly higher cost of electricity (COE) than the base case, it was found that the addition of the POX/reformer allowed the system to operate more realistically by destroying the tars while simultaneously improving the carbon sequestration capability of the plant overall. The downside to the higher sequestration rate in the POX/reformer plant is the increased parasitic load required to compress the additional gas. However, the sensitivity analyses conducted for the IGCC report showed the modified case closing the gap on the base case, the base case always maintained a cost edge over the range of analysis.

Summary of Nexant CTL report

In the report for the CTL cases, each plant was sized to produce 50,000 barrels of liquid fuel. All other equipment, variable costs, etc were all scaled for this plant size. The report found that the SRI modified CTL case used approximately 26% less coal than the base case, however, the lower volume and energy content of the resultant tail gas from the modified case produced significantly less electricity. The base case was a net exporter of electricity due to the high methane content of the tail gas, while the modified case was a slight net importer of electricity. Despite these differences, each plant produced a very similar liquid fuel cost of production (COP), with the base case coming to $115/barrel while the modified case was $119/barrel.

Sensitivity analyses conducted for this report showed that several variables would produce cases where the modified case would outperform the base case in COP. The variables studied in this report were; POX/reformer TPC, plant capacity factor, coal price, CO_2 sale price, and cost of electricity. While variables such as plant capacity factor and CO_2 sale saw a greater decline in the COP for the base case than for the modified case, the remaining variables showed that the modified case can improve the COP if certain targets are met. When we see lower TPC costs, higher feedstock costs, or lower electricity prices, then the modified case becomes more competitive. Table 4 shows both the value used in the Nexant reports to form the basis of comparison, but it also shows the point where the modified case surpasses the basecase.

Table 4 Breakeven point for sensitivity analysis variables for modified POX/reformer case vs base case

	Value used for Report	Break even value
POX/Reformer TPC as % of Total TPC	13%	8.67%
Plant capacity factor (availability)	90%	NA
Coal price	$19.63/ton	$40/ton
CO_2 sale price	$0/tonne	NA
Price of electricity	$60/MWhr	$40/MWhr

For both the IGCC and CTL cases, the SR POX/reformer is slightly more expensive than the base case, however, the modified case does provide some additional non-monetary advantages. In the case of the IGCC, the destruction of tars and/or the increased CO_2 capture and sequestration may be more appealing. While for the CTL plants, the modified case has identified three variables which would move the COP lower than in the base case.

Lurgi and ThermoChem cases

In the cases studied by Nexant, the TRIG gasifier was used because it is considered to be the most commercially ready gasifier available for these applications. However, the Lurgi and ThermoChem (TRI) gasifiers could offer an alternative to the TRIG gasifier if it can be shown to produce lower cost electricity or liquid fuels. The following cases are basic estimates to determine if a more detailed analysis should be done in the future which could be directly compared to the rigorous cases listed above.

Both the Lurgi and ThermoChem gasifiers produce a syngas that has a different molar composition than the TRIG gasifier. In the two comparison cases (Lurgi & TRI), the syngas contains much more methane and tars than the TRIG gasifier. The syngas compositions used for the following analyses can be seen below in Table 5. The IGCC models for these alternative cases were simulated using an ASPEN plus simulation of the POX/reformer only. It was assumed that the remainder of the plant would remain constant, however, the BTU content of the gas was different and the combustion and heat recovery steam generators were sized for the energy content.

Table 5 Molar syngas composition for comparison cases

	TRIG	Lurgi	TRI
H2	24.93%	18.70%	33.02%
CO	34.07%	7.50%	29.24%
CO2	7.63%	15.70%	16.76%
N2	0.54%	0.00%	0.36%
CH4	0.02%	5.20%	4.04%
H2O	31.39%	51.60%	14.88%
C2H4	0.00%	0.00%	0.19%
AR	0.63%	0.00%	0.54%
H2S	0.54%	0.29%	0.23%
HCL	0.02%	0.00%	0.00%
COS	0.04%	0.00%	0.02%
NH3	0.10%	0.59%	0.26%
Tars	0.10%	0.42%	0.44%
HCN	0.00%	0.00%	0.01%
Total	100.00%	100.00%	100.00%

A comparison of the inlet syngas listed above in Table 5 with the syngas composition found exiting the POX/reformer in the IGCC cases, listed below in Table 6, shows that the Lurgi gasifier sees a dramatic increase in hydrogen, while in the TRI case a slight increase is seen at the cost of the CO. The increase in hydrogen seen in the Lurgi syngas is primarily due to the reforming activity coupled with a water-gas shift (WGS) reaction. Given the high concentration of water in the Lurgi syngas the equilibrium lies heavily in the forward direction, consuming the water and generating H_2 and CO_2. However, in the other two cases, the CO produced through reforming is insufficient to feed the WGS reaction therefore, the feed CO is consumed to generate some additional hydrogen.

In the case of the IGCC the relative concentrations of the CO does not matter much as the total energy available in the gas stream. In the all the cases TRIG, Lurgi and TRI case, the exit stream from the reformer has slightly less energy than the input. In the TRIG and TRI case the partial oxidation consumes just enough methane and tar to get the gas to the reforming temperature, however, does not contain enough hydrocarbons to boost the overall energy content. In the Lurgi case, it appears that there is a penalty for heating the excess steam inherent in the gas to the reforming temperatures.

Table 6 Molar composition of syngas leaving reformer (IGCC cases)

	TRIG	Lurgi	TRI
H2	33.30%	28.59%	36.83%
CO	17.77%	7.72%	28.80%
CO2	21.85%	19.17%	18.37%
N2	0.61%	0.37%	0.55%
CH4	0.04%	0.01%	0.14%
H2O	25.18%	43.69%	14.41%
C2H4	0.00%	0.00%	0.00%
AR	0.68%	0.21%	0.67%
H2S	0.53%	0.24%	0.22%
HCL	0.02%	0.00%	0.00%
COS	0.02%	0.01%	0.01%
NH3	0.00%	0.00%	0.00%
Tars	0.00%	0.00%	0.00%
HCN	0.00%	0.00%	0.00%
Total	100.00%	100.00%	100.00%

In order to calculate the capital cost for each case, only the cost of the gasifier and the cost of the electricity generation components were changed. The coal handling, gas cleanup, ASU, O/M, variable costs, etc were kept constant. It is possible that some of these costs will vary for each case, however, that is beyond the scope of this study.

The Lurgi gasifier was sized based on the mass balances found in the literature (Beychok, 1974; Blazek et Al., 1979) as was the cost (National Research Council, 1973). This gasifier did not compare well because it produces very little syngas compared to the feed rate of the coal. The oil collected from the bottoms would need to be reformed to improve the economics.

The TRI gasifier mass balance was provided by ThermoChem Recovery International, however, the capital cost was generated from an old NETL project report (TRI, 2001). The costs were a preliminary engineering estimation for a 40 ton/hr plant, which itself used scaling factors. The numbers provided here scaled from these to the 6935 tons/day size and should be considered very rough.

Comparing the economics in of the three gasifiers shown in **Error! Reference source not found.** shows that the base case for all three gasifiers performs slightly better than the POX/reformer case whose performance is shown in **Error! Reference source not found.**. The difference is small considering the increase in capital as well as reduced production. However, what we do see is that the TRI gasifier performs better than the TRIG gasifier, while the overall capital cost appears to be low, the gas quality produces more electricity per ton of coal. The COE was calculated using the simple COE equation found in QGESS Cost Estimation Methodology for NETL Assessments of Power Plant Performance.

Table 7 Summary of Economics for Alternative Gasifier IGCC Base Cases

	CTL Base TRIG	CTL Lurgi	CTL TRI
Total Overnight Cost $1000	$2,078,113	$1,468,658	$1,817,263
Total annual fixed O/M, $1000	$60,500	$60,500	$60,500
Total annual variable O/M (100% CF), $1000	$45,900	$45,900	$45,900
Total annual feedstock cost (100% CF) $1000	$49,700	$49,700	$49,700
COE, mills/kWh	99.2	354.2	74.1
Power Generated kWh	4,004,859	921,601	4,952,928

Table 8 Summary of Economics for Alternative Gasifier IGCC Case with POX/reformer

	CTL Base TRIG	CTL Lurgi	CTL TRI
Total Overnight Cost $1000	$2,153,870	$1,535,052	$1,883,060
Total annual fixed O/M, $1000	$62,400	$62,400	$62,400
Total annual variable O/M (100% CF), $1000	$47,500	$47,500	$47,500
Total annual feedstock cost (100% CF) $1000	$49,700	$49,700	$49,700
COE, mills/kWh	106.4	384.6	79.9
Power Generated kWh	3,848,240	877,867	4,734,022

The simplified case for the CTL alternative gasifier cases focused on the effects of the POX/reformer on the combined gasifier gas and recycled tail gas when coupled with a selective Fischer-Tropsch catalyst that produces only liquids and does not require any additional product upgrading.

Figure 30 shows the ASPEN process flow diagram used for each case. The feed syngas composition listed in Table 5 was used for each case and the recycle was also added to the POX/reformer.

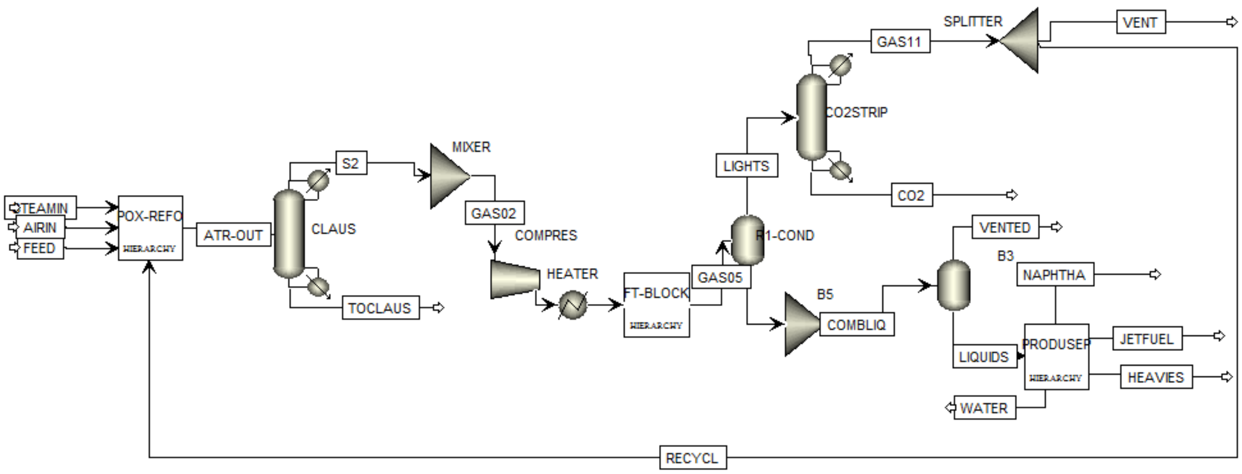

Figure 30 ASPEN Process Flow Diagram for Simplified CTL cases

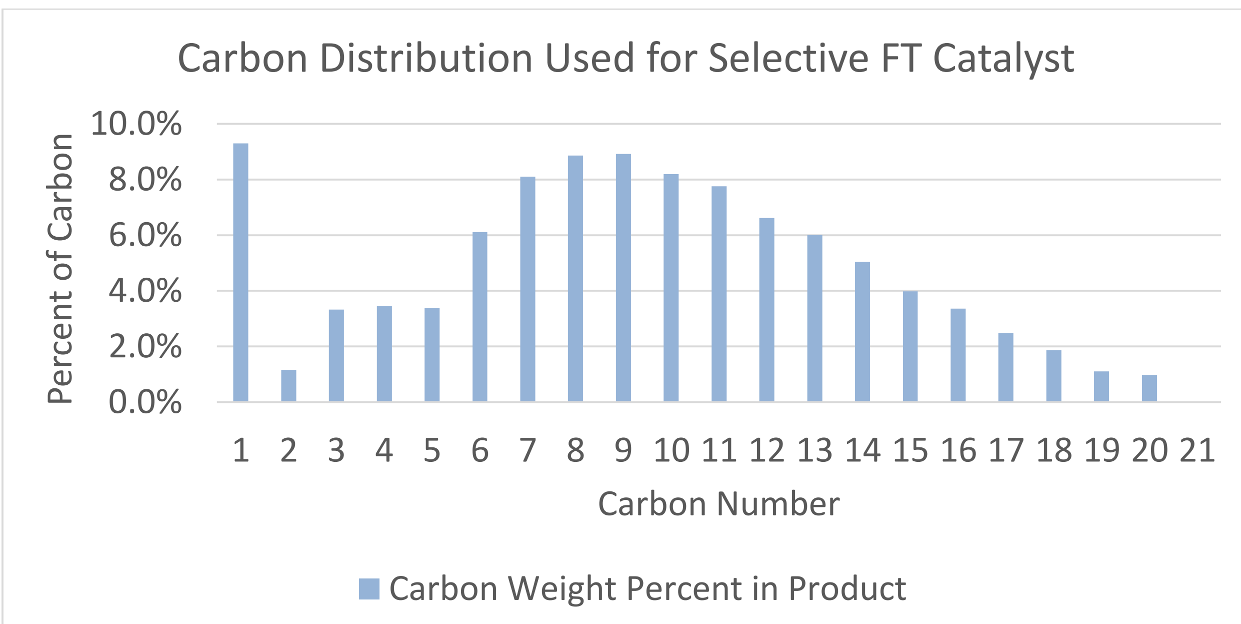

Figure 31 Carbon Weight Distribution Used in FT Catalyst Product

Figure 31 shows the carbon distribution of the catalyst that was modeled in the FT reactor. This selective product distribution is based on the selective cobalt zeolite catalyst developed by Chevron (Kibby, 2011; Dyer, 1987).

Table 9 shows the composition of the syngas leaving the POX/reformer for each case. The Lurgi case was not pursued any further because the H2:CO ratio of 3.7 was not useful in a Fischer-Tropsch plant. The large amount of water in the gas caused a pronounced WGS reaction causing the CO to nearly be depleted. In this case, dropping the temperature to condense and remove the water then reheating the syngas would not be useful for this technology.

However, both the TRIG and TRI have a hydrogen to carbon monoxide ratio of 1.79, which is slightly low for a cobalt based Fischer-Tropsch reactor (1.8-2.1 is preferred) but could easily be supplemented by using some of the shifted tail gas heading to the power block. The ratio would be useful in a Fischer-Tropsch reactor featuring an iron catalyst.

Table 9 Molar composition of syngas leaving reformer (CTL cases)

	TRIG	Lurgi	TRI
H2	36.56%	28.59%	38.18%
CO	20.40%	7.72%	21.43%
CO2	18.27%	19.17%	17.52%
N2	1.91%	0.37%	1.52%
CH4	0.08%	0.01%	0.10%
H2O	20.11%	43.69%	19.17%
C2H4	0.00%	0.00%	0.00%
AR	2.23%	0.21%	1.94%
H2S	0.40%	0.24%	0.14%
HCL	0.01%	0.00%	0.00%
COS	0.01%	0.01%	0.01%
NH3	0.00%	0.00%	0.00%
Tars	0.00%	0.00%	0.00%
HCN	0.00%	0.00%	0.00%
Total	100.00%	100.00%	100.00%

Comparing the economics of the CTL base cases for TRIG and the TRI gasifier in Table 10 with the economics for the modified cases containing the POX/reformer in Table 11 shows the similar results as the Nexant CTL cases. The modified case is slightly more expensive, however, the process is more chemically efficient. There is less excess gas making its way to the combustion turbine and adding extra revenue through electricity export. This means less feedstock is necessary and less carbon is making its way into the atmosphere. This is even the case with the selective FT catalyst producing more light hydrocarbons to reform in the POX/reformer.

Table 10 Summary of Economics for Alternative Gasifier CTL Base Cases

	CTL Base TRIG	CTL TRI
Total Overnight Cost $1000	$6,509,105	$6,918,139
Total annual fixed O/M, $1000	$190,000	$190,000
Total annual variable O/M (90% CF), $1000	$132,000	$132,000
Total annual feedstock cost (90% CF) $1000	$143,533	$188,439
Total annual power credit (90% CF) $1000	-$101,307	-$79,279
COP F-T Fuel	$137.48	$147.00
Barrels/day FT-liquid	49,996	49,996

Table 11 Summary of Economics for Alternative Gasifier CTL Cases with POX/reformer

	CTL Base TRIG	CTL TRI
Total Overnight Cost $1000	$6,826,618	$7,487,989
Total annual fixed O/M, $1000	$184,000	$190,000
Total annual variable O/M (90% CF), $1000	$120,000	$132,000
Total annual feedstock cost (90% CF) $1000	$142,947	$203,047
Total annual power credit (90% CF) $1000	-$30,655	$18,912
COP F-T Fuel with POX/reformer	$144.87	$160.65
Barrels/day FT-liquid	49,996	49,996

Life cycle analysis

The life-cycle analysis of the IGCC case is conducted using the emissions factors found in the Argonne National Laboratory's Greenhouse Gas, Regulated Emission and Energy Use in Transportation Model (GREET) and information gathered from the SR generated ASPEN mass balance. In both cases, the TRIG IGCC plant with and without the POX/reformer will be studied. The boundary of the model will encompass the coal mining, cleaning, and transportation (captured in GREET coal emission factor), the coal gasification/syngas combustion, and the mitigation due to CO_2 sequestration. The GREET model emission factor used for the upstream processes was the "Bituminous Coal for Central Hydrogen Plant, FTD Plant, Methanol Plant, and DME Plant Use" value of 139.5 kg CO_2e/ton coal. For the coal gasification/syngas combustion, it was assumed that 100% of the carbon in the feed coal would be converted to CO_2 in the IGCC, while any CO_2 that was captured and sequestered would be subtracted from this total. Table 12 shows that the SR modified case featuring the POX/reformer produced about a 24% reduction in CO_2e per kilowatt-hour when compared to the base case electricity generation via IGCC. The benefit is derived from converting the methane and tars into CO_2 before combustion in the turbine, which allows the carbon to be sequestered at a higher rate per kWh.

Table 12 Emissions for IGCC electricity production (kg CO2e/kWh)

	Base Case (kg CO_2e/kWh)	Modified Case w/ POX/Reformer (kg CO_2e/kWh)
Upstream PRB Coal	2.1	2.2
PRB Coal Gasification	27.8	29.0
Sequestered CO2	-23.3	-26.1
Total Emissions	6.7	5.1

The life-cycle analysis of the CTL plant is calculated in a similar manner to the IGCC cases. Both the cases under consideration will be the TRIG gasification based CTL plant. Emission factors for the various input streams as well as net electricity will use the GREET emission factors and the SR CTL mass balances. For the output streams such as vent gas, transportation fuel, etc. the carbon content of the stream will be converted with 100% efficiency to CO_2. For this analysis, the combustion of the fuel was included to make this a cradle-grave type of LCA. Again the upstream coal emission factor was 139.5 kg CO_2e/ton coal, while the electricity factor was the "Electricity: Coal-Fired (IGCC Turbine) Plant" value of

1.1 kg CO_2e/kWh, and finally the combustion of the fuel was based on the "Diesel Car" vehicle operation factor of 75.7 kg CO_2e/MJ fuel equivalent. The emissions were normalized per barrel of fuel produced.

The modified POX/reformer case despite being slightly more expensive produced a significant improvement in GHG emissions. This is primarily due to the more efficient utilization of the carbon in the incoming coal. By converting most of the hydrocarbons found in the syngas and the recycled gasses into either CO or CO_2, there is much less slippage of carbon to the gas turbine & steam generators. The carbon that does enter the plant featuring the POX/reformer is either directed into fuel production or it is sequestered rather than being burned in the gas turbine. This diversion of carbon in the modified case manifests itself in lowered electricity production (requiring some import) as well as lower feedstock requirements to produce the same amount of fuel. Table 13 shows the carbon reduction for the modified case with POX/reformer compared to the base case.

Table 13 Emissions for Fischer-Tropsch fuel (kg CO2e/barrel)

	Base Case (kg CO_2e/barrel)	Modified Case w/ POX/Reformer (kg CO_2e/barrel)
Upstream PRB Coal	100.0	73.6
PRB Coal Gasification	1316.6	969.0
Sequestered CO_2	-754.0	-598.0
Transportation of Fuel	2.6	2.6
Electricity	-4.4	0.6
Fuel Export	-436.3	-436.3
Combustion of Fuel	436.3	436.3
Total Emissions	660.8	447.8

6 CONCLUSIONS

A laboratory scale experimental program was undertaken to develop novel H_2S resistant steam reforming catalysts to support a process designed to increase the yield and H_2:CO ratio of a surrogate low-rank coal syngas by converting tars, C2+ hydrocarbons, NH_3 and methane under high temperature and sulfur environments. A novel steam reforming catalyst was developed that was not affected by up to 90 ppm H_2S, and was still useful at higher concentrations up to 500 ppm H_2S. A techno economic analysis developed for the process indicated near equal economics compared to baseline IGCC and CTL processes with large decreases in greenhouse gas emissions for equivalent power or FT liquids yields.

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10 ACRONYMS AND ABBREVIATIONS

Ar	Argon
Aspen	ASPEN Plus®
ASU	Air Separation Unit
ATR	Autothermal reformer
BTU	British Thermal Unit
CF	Capacity Factor
CH_4	Methane
CO	Carbon monoxide
CO_2	Carbon dioxide
CO_2e	CO_2 equivalent
COE	Cost of Electricity
COP	Cost of Production
COS	Carbonyl sulfide
C_1	Methane
C_3H_8	Propane
C_4	Butane
C_7	Heptane
CTL	Coal to Liquid
DOE	Department of Energy

FT	Fischer-Tropsch
GC	Gas chromatograph
GHG	Greenhouse gas
GREET	Greenhouse Gas, Regulated Emission and Energy Use in Transportation Model
HCl	Hydrogen chloride
hr	Hour
H_2	Hydrogen
H_2O	Water
H_2S	Hydrogen sulfide
HCN	Hydrogen Cyanide
IGCC	Integrated Gasification Combined Cycle
kg	kilogram
kg/hr	kilogram per hour
kW	Kilowatt
kWh	Kilowatt hour
LCA	Life Cycle Analysis
MJ	MegaJoule
mole%	Mole Percent
mol/hr	Moles per hour
NETL	National Energy Technology Laboratory
NH_3	Ammonia
N_2	Nitrogen
O_2	Oxygen
O/M	Operation & Maintenance
PCI	Precision Combustion Incorporated
PFD	Process flow diagram
POX	Partial Oxidation
PRB	Powder River Basin

Syngas	Synthesis gas
SLPM	Standard liters per minute
SR	Southern Research
TPC	Total Plant Cost
TRI	ThermoChem Recovery Incorporated
TRIG™	Transport Reactor Integrated Gasification
TEA	Techno-Economic Analysis
WGS	Water-gas shift
°C	Degrees Celcius
°F	Degrees Fahrenheit

11 APPENDIX

SRI POX TEA
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SRI POX TEA
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Southern Research Institute (SRI) High Temperature Catalytic Steam Reforming Process for Hydrogen-Rich Syngas Production - IGCC Application

Preliminary Cost Estimation and Techno-Economic Analysis, Based on SRI's Aspen Process Simulation and DOE's Reference TRIG IGCC Design

Submitted to

Southern Research Institute (SRI)

By

Nexant

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1.1 BACKGROUND

With funding from DOE “Advanced Gasification Technologies Development and Gasification Scoping Studies for Innovative Initiatives, Area of Interest 3, ‘High-Hydrogen Syngas Production” program, Southern Research Institute (SRI) is developing a high temperature catalytic steam reforming process that can withstand the severe contaminant conditions of a near-raw syngas derived from low-rank coal gasification with the objective to increase the H_2-to-CO ratios and eliminate methane, ammonia and tars from the process. SRI is developing the technology for both IGCC and coal-to-liquid applications. In addition to catalyst development, SRI is also carried out Aspen simulation of the overall process. Nexant was asked to assist with preliminary cost estimation and techno-economic analysis (TEA) of the process, based on SRI’s simulation results. The TEA is to be carried out based on DOE NETL’s methodology.

1.2 STUDY OBJECTIVES

The objective of this TEA study is to assess the performance and economic potential of integrating the catalytic steam reforming process offered by SRI with a TRIG gasifier. The catalytic reformer will treat raw syngas exiting the gasifier and generate a hydrogen-rich, tar-free syngas with minimal methane content. This will enable near-zero emissions from coal gasification for both power and coal-to-liquids productions with a carbon capture rate of more than 90%.

The current report presents the results of the techno-economic analysis (TEA) performed for the IGCC case, utilizing KBR’s Transport Gasifier (TRIG) coupled with SRI’s catalytic reformer. It was carried out, to the maximum extent possible, in accordance with:

- The guidelines as set forth in the Attachment 2 of the DE-FOA-0000784 document (“Design Basis for Techno-economic Analyses Deliverables”)
- The plant balance and cost data available in the TRIG IGCC Reference Case design, per “*Cost and Performance Baseline for Fossil Energy Plants, Volume 3a: Low Rank Coal to Electricity*, May 2011, DOE/NETL. 2010/1399” (NETL Report 1399), and
- Aspen simulation results provided by SRI for both the IGCC designs with and without its high-temperature, contaminant-resistant, catalytic reforming unit.

Section 2 IGCC Design Basis

2.1 DESIGN REFERENCES

The process design references used for this study follow the recommended reference studies set forth by Attachment 2 of the FOA. These are namely:

- "*Cost and Performance Baseline for Fossil Energy Plants, Volume 3a: Low Rank Coal to Electricity*, May 2011, DOE/NETL. 2010/1399" (NETL Report 1399)

NETL Report 1399 contains an IGCC design based on the KBR TRIG, which serves as the reference case for comparison with the case with SRI's catalytic reformer. This report also contains a comprehensive set of IGCC design bases and assumptions, as well as reference costs and economic evaluation guidelines, allowing it to serve as the design reference for the IGCC design utilizing SRI's catalytic reforming system.

- NETL's Series of Quality Guidelines for Energy Systems Studies (QGESS):
    - "*Specifications for Selected Feedstocks*, January 2012, DOE/NETL. 341/011812"
    - "*Process Modeling Design Parameters*, January 2012, DOE/NETL. 341/081911"
    - "*CO_2 Impurity Design Parameters*, January 2012, DOE/NETL. 341/011212"
    - "*Detailed Coal Specifications*, January 2012, DOE/NETL-401/01211"
    - "*Cost Estimation Methodology for NETL Assessments of Power Plant Performance*, April 2011, DOE/NETL. 2011/1455"
    - "*Capital Cost Scaling Methodology*, January 2013, DOE/NETL. 341/013113"
    - "*Fuel Prices for Selected Feedstocks in NETL Studies*, November 2012, DOE/NETL 341/11212"
- "*Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases*, August 2012, DOE/NETL-341/082312" (NETL Report 341/082312), was used as the reference to develop the updated capital and operating cost estimates, in June 2011 dollars.

2.2 CASE CONFIGURATIONS

An IGCC configuration utilizing KBR TRIG coupled with SRI's catalytic reformer is evaluated to assess the feasibility of integrating SRI's catalytic reformer into a TRIG-based IGCC plant with CO_2 capture. This configuration is compared against the reference TRIG-only design in the NETL Report 1399 (Case S2B) to identify any potential advantages.

The two IGCC configurations are identified in the IGCC case study matrix shown in Table 2-1.

Table 2-1
Case Study Matrix for IGCC with CO_2 Capture

	Case P1[1]	Case P2
Gasification Technology		
TRIG Gasifier	✓	✓
SRI Catalytic Reformer		✓
Gas Cleanup		
Two-Stage Selexol for CO_2 and Sulfur Removal[2]	✓	✓
Water Gas Shift		
Sour Shift	✓	✓
GE 7FB Advanced Gas Turbine	✓	✓
CO_2 Drying and Compression (to 2,200 psig)	✓	✓

1 Reference case based on SRI's benchmark simulation of the NETL 1399 Report S2B case
2 Selexol removes H_2S and CO_2. Additional trace contaminant cleanup technologies is included as defined by DOE/NETL baseline studies

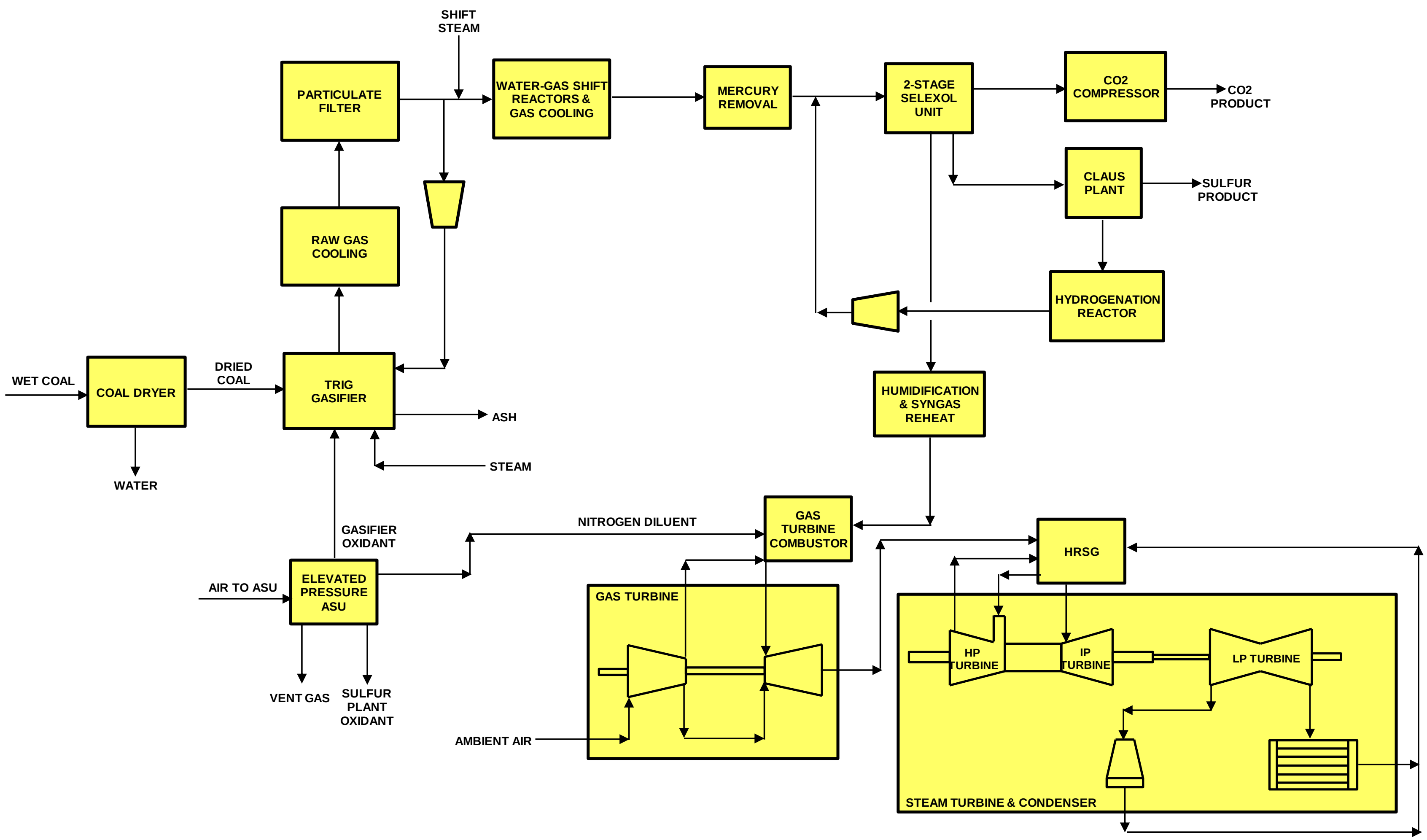
2.2.1 Case P1: Reference TRIG IGCC Power Plant with Selexol-Based AGR

Per DE-FOA-000784 AOI 3's requirements, it is required to establish one reference IGCC case using for comparison with the advanced IGCC power plant utilizing the SRI's catalytic reformer. DOE-NETL recommended that one of the power cases in its Baseline Studies is to be used for the reference case, of which SRI selected Case S2B, the TRIG-based IGCC case for the current analysis.

The reference TRIG-based IGCC case is a coal-fired IGCC plant generating enough fuel gas to fill two advanced GE 7F-turbines rated nominally at 215 MW each, for a total of 430 MWe at the Montana site's elevation. The power plant is equipped with a heat recovery steam generator (HRSG) and steam turbine to generate additional power from waste heat from the flue gas. Adding in the steam turbine power and subtracting auxiliary loads (including CO_2 capture and compression), the reference IGCC plant's nominal net export capacity is 450 MWe. The two-stage Selexol process is used for acid gas removal (AGR), whereby it captures both H_2S and CO_2. However, due to the elevated methane concentration in the syngas, the overall CO_2 capture rate does not exceed 85%.

The Case P1 Reference IGCC plant simplified Block Flow Diagram (BFD) is shown in Figure 2-1. It is assumed to operate as a base-loaded unit with an annual on-stream factor of 80 percent or 7,000 hrs/year at full capacity.

Figure 2-1
Case P1: Reference TRIG IGCC Power Plant with Selexol-Based AGR - Simplified BFD



2.2.2 Case P2: Coupled TRIG/SRI Catalytic Reformer with Selexol-Based AGR

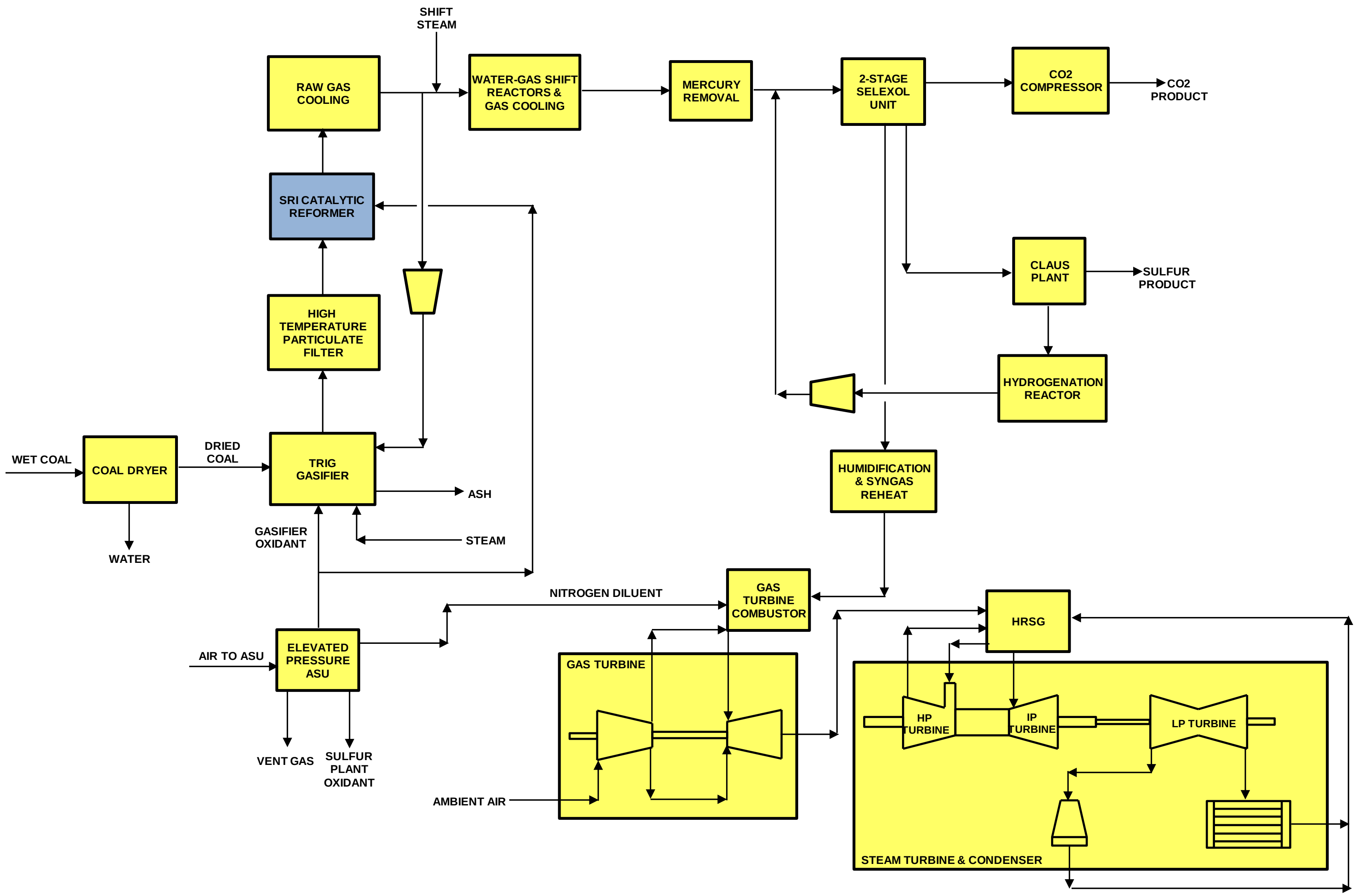
Case P2 is a preliminary conceptual design that couples the SRI catalytic reformer with the TRIG process. Like Case P1, the power plant is equipped with a HRSG and steam turbine to generate additional power from waste heat from the flue gas. Due to the different flue gas quantity and IGCC waste heat recovery scheme, the steam turbine output may differ substantially from that of Case P1.

Similar to Case P1, the two-stage Selexol process is used for AGR. However, as the catalytic reformer is able to convert practically all the tar and most of the methane in the raw syngas to CO and H_2, carbon slippage into the fuel gas is minimized since CO can be shifted to CO_2 via the shift reactors and subsequently captured by the Selexol units. With the coupling of the SRI catalytic reformer to the TRIG, the Case P2 IGCC is able to achieve a CO_2 capture rate that is greater than 90%.

The simplified BFD for the Case P2 IGCC plant is shown in Figure 2-2. It is assumed to operate as a base-loaded unit with an annual on-stream factor of 80 percent or 7,000 hrs/year at full capacity.

The blue SRI catalytic reformer block in Figure 2-2 is the differentiating technology of interest in this study. All other blocks (in yellow) are expected to have the same unit operations as those in Case P1, but with different throughputs due to differing flow quantities. Costs for these systems will be scaled based on system capacities from Case S2B using factors given in the QGESS *Capital Cost Scaling Methodology* document wherever possible.

Figure 2-2
Case P2: TRIG w/SRI Catalytic Reformer IGCC Plant - Simplified BFD



2.3 PROCESS DESIGN CRITERIA

SRI carried out a simulation of the P1 and P2 IGCC cases on ASPEN Plus to obtain the process heat and material balances (HMB). The results from SRI's simulations were then provided to Nexant. The HMB were used to estimate the overall plant utility balance and determine the overall IGCC plant performance. Based on the SRI-provided HMB, the IGCC power plant cost can be estimated as well, using the cost estimation methodology described in Section 2.4 below.

2.4 CAPITAL COST ESTIMATION METHODOLOGY

2.4.1 General

For IGCC plants with CO_2 capture, the NETL 1399 Baseline Study provided a code of accounts grouped into 14 major systems. Each of these major systems is broken down further into different subsystems. This type of code-of-accounts structure has the advantage of grouping all reasonably allocable components of a system or process into a specific system account.

For the IGCC cases evaluated in this study, capital cost scaling following the guidelines and parameters that are described in the NETL *Capital Cost Scaling Methodology* document was used to perform the cost estimation for systems that are not related to the SRI catalytic reformer. In general, this cost estimation methodology involves determining the scaling parameters, exponents and coefficients from the *Capital Cost Scaling Methodology*, as well as the reference cost and baseline capacity from the 1399 Baseline Study. Once these have been established, the capital cost can be estimated based on the revised capacity from the HMB developed by SRI's ASPEN models of the IGCC cases.

Nexant performed a bottoms-up, major-equipment factored cost estimation for the SRI catalytic reformer, with SRI providing guidance on the catalyst and reactor vessel.

2.4.2 IGCC Plant Capital Cost Estimate Criteria

The capital cost estimates for the IGCC systems that are unrelated to the STI catalytic reformer were developed based on the Case S2B TRIG IGCC plant with CO_2 capture case in the NETL 1399 Baseline Study. The costs were adjusted for differences in unit or plant capacity according to NETL's Guidelines as described in the NETL *Capital Cost Scaling Methodology* QGESS document.

Table 2-2 shows the code of accounts for the IGCC plant. These systems are further broken down to include the various subsystems. The scaling parameters for these BOP subsystems, as laid out by the NETL *Capital Cost Scaling Methodology* document, are also shown in this table.

Table 2-2
Code of Accounts for Report IGCC Plant

Acct No.	Item/Description	Scaling Parameter
1	COAL & SORBENT HANDLING	
1.1	Coal Receive & Unload	Coal Feed Rate
1.2	Coal Stackout & Reclaim	Coal Feed Rate
1.3	Coal Conveyors & Yard Crush	Coal Feed Rate
1.4	Other Coal Handling	Coal Feed Rate
1.9	Coal & Sorbent Handling Foundations	Coal Feed Rate
2	COAL & SORBENT PREP & FEED	
2.1	Coal Crushing & Drying	Coal Feed Rate
2.2	Prepared Coal Storage & Feed	Coal Feed Rate
2.3	Dry Coal Injection System	Coal Feed Rate
2.4	Misc Coal Prep & Feed	Coal Feed Rate
2.9	Coal & Sorbent Feed Foundation	Coal Feed Rate
3	FEEDWATER & MISC. BOP SYSTEMS	
3.1	Feedwater System	BFW (HP only)
3.2	Water Makeup & Pretreating	Raw Water Makeup
3.3	Other Feedwater Subsystems	BFW (HP only)
3.4	Service Water Systems	Raw Water Makeup
3.5	Other Boiler Plant Systems	Raw Water Makeup
3.6	FO Supply Sys and Nat Gas	Coal Feed Rate
3.7	Waste Treatment Equipment	Raw Water Makeup
3.8	Misc Power Plant Equipment	Coal Feed Rate
4	GASIFIER & ACCESSORIES	
4.1	Gasifier, Syngas Cooler & Auxiliaries	Syngas Throughput
4.3	ASU/Oxidant Compression	O_2 Production
4.4	LT Heat Recovery and Fuel Gas Saturation	Syngas Flow
4.6	Other Gasification Equipment	Syngas Flow
4.9	Gasification Foundations	Syngas Flow
5A	GAS CLEANUP & PIPING	
5A.1	Double Stage Selexol	Gas Flow to AGR
5A.2	Elemental Sulfur Plant	Sulfur Production
5A.3	Mercury Removal	Hg Bed Carbon Fill
5A.4	Shift Reactors	WGS/COS Catalyst
5A.5	Blowback Gas Systems	Candle Filter Flow
5A.6	Fuel Gas Piping	Fuel Gas Flow
5A.9	HGCU Foundations	Sulfur Production
5B	CO_2 REMOVAL & COMPRESSION	
5B.2	CO_2 Compression & Drying	CO_2 Flow
6	COMBUSTION TURBINE/ACCESSORIES	
6.1	Combustion Turbine Generator	Fuel Gas Flow
6.2	Combustion Turbine Foundations	Fuel Gas Flow
7	HRSG, DUCTING & STACK	
7.1	Heat Recovery Steam Generator	HRSG Duty
7.3	Ductwork	Vol Flow to Stack
7.4	Stack	Vol Flow to Stack
7.9	HRSG, Duct & Stack Foundations	Vol Flow to Stack

Acct No.	Item/Description	Scaling Parameter
8	STEAM TURBINE GENERATOR	
8.1	Steam TG & Accessories	Turbine Capacity
8.2	Turbine Plant Auxiliaries	Turbine Capacity
8.3a	Condenser & Auxiliaries	Condenser Duty
8.3b	Air Cooled Condenser	Condenser Duty
8.4	Steam Piping	BFW (HP Only)
8.9	TG Foundations	Turbine Capacity
9	COOLING WATER SYSTEM	
9.1	Cooling Towers	Cooling Tower Duty
9.2	Circulating Water Pumps	Circ H_2O Flow Rate
9.3	Circ. Water System Auxiliaries	Circ H_2O Flow Rate
9.4	Circ Water Piping	Circ H_2O Flow Rate
9.5	Makeup Water System	Raw Water Makeup
9.6	Component Cooling Water System	Circ H_2O Flow Rate
9.9	Circ. Water System Foundations	Circ H_2O Flow Rate
10	ASH/SPENT SORBENT HANDLING SYS	
10.1	Slag Dewatering & Cooling	Slag Production
10.6	Ash Storage Silos	Slag Production
10.7	Ash Transport & Feed Equipment	Slag Production
10.8	Misc. Ash Handling System	Slag Production
10.9	Ash/Spent Sorbent Foundation	Slag Production
11	ACCESSORY ELECTRIC PLANT	
11.1	Generator Equipment	Turbine Capacity
11.2	Station Service Equipment	Auxiliary Load
11.3	Switchgear & Motor Control	Auxiliary Load
11.4	Conduit & Cable Tray	Auxiliary Load
11.5	Wire & Cable	Auxiliary Load
11.6	Protective Equipment	Auxiliary Load
11.7	Standby Equipment	Total Gross Output
11.8	Main Power Transformers	Total Gross Output
11.9	Electrical Foundations	Total Gross Output
12	INSTRUMENTATION & CONTROL	
12.4	Other Major Component Control	Auxiliary Load
12.6	Control Boards, Panels & Racks	Auxiliary Load
12.7	Computer & Accessories	Auxiliary Load
12.8	Instrument Wiring & Tubing	Auxiliary Load
12.9	Other I & C Equipment	Auxiliary Load
13	IMPROVEMENT TO SITE	
13.1	Site Preparation	Accounts 1-12
13.2	Site Improvements	Accounts 1-12
13.3	Site Facilities	Accounts 1-12
14	BUILDING & STRUCTURES	
14.1	Combustion Turbine Area	Gas Turbine Power
14.2	Steam Turbine Building	Accounts 1-12
14.3	Administration Building	Accounts 1-12
14.4	Circulation Water Pumphouse	Circ H_2O Flow Rate
14.5	Water Treatment Buildings	Raw Water Makeup
14.6	Machine Shop	Accounts 1-12
14.7	Warehouse	Accounts 1-12
14.8	Other Buildings & Structures	Accounts 1-12
14.9	Waste Treating Building & Structures	Raw Water Makeup

2.4.3 Home Office, Engineering Fees and Project/Process Contingencies

Engineering and Construction Management Fees and Home Office cost, project and process contingencies were factored from the each subsystem's TFC. These were then added to the TFC to come up with the total project cost (TPC) of the system. Factors from Case S2B in the NETL 1399 Baseline Report were used.

2.4.4 Owner's Cost

Owner's cost was then added to TPC to come up with the total overnight cost (TOC) for the system. Owner's costs as defined in the NETL 1399 Baseline Study include the following:

- Preproduction Costs –
    - 6 months of all labor cost
    - 1 month of maintenance materials
    - 1 month of non-fuel consumables
    - 1 month of waste disposal
    - 25% of 1 month fuel cost at 100% capacity factor
    - 2% TPC

- Inventory Capital -
    - 60 day supply of fuel and consumable at 100% CF
    - 0.5% TPC

- Initial Cost for Catalyst and Chemicals per design
- Land Cost = $900,000 at 300 acres x $3,000/acre
- Other Owner's Costs at 15% TPC
- Financing Costs at 2.7% TPC

2.5 OPERATION & MAINTENANCE COSTS

The operation and maintenance (O&M) costs pertain to those charges associated with operating and maintaining the power plants over their expected life. These costs include:

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

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 power generation. Variable O&M costs were estimated based on 80% capacity factor.

2.5.1 Fixed Costs

Operating labor cost was determined based on the number of operators required to work in the plant. Other assumptions used in calculating the total fixed cost include:

- 2011 Base hourly labor rate, $/hr — $39.7
- Length of work-week, hrs — 50
- Labor burden, % — 30
- Administrative/Support labor, % O&M Labor — 25
- Maintenance material + labor, % TPC — 2.8
- Maintenance labor only, % maintenance material + labor — 35
- Property Taxes and insurances, % TPC — 2

2.5.2 Variable Costs

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

The unit costs for major consumables and waste disposal was selected from NETL 1399 Baseline Report, QGESS *Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases* and from the QGESS *Fuel Prices for Selected Feedstocks in NETL Studies* document.

The 2011 coal price as delivered to the Montana IGCC plant is $19.63/ton, per the QGESS *Fuel Prices for Selected Feedstocks in NETL Studies* document.

2.5.3 CO_2 Transport and Storage Costs

As specified in DE-FOA-0000784 Attachment 2, CO_2 Transport and Storage (T&S) costs used for the Montana IGCC plant location is $22/tonne. Per the TEA reporting requirements, the COEs are reported both with and without the cost of CO_2 T&S.

2.6 FINANCIAL MODELING BASIS

2.6.1 Cost of Electricity

The metrics used to evaluate overall financial performance are the cost of electricity (COE) for the IGCC plant. All costs were expressed in the “first-year-of-construction” year dollars, and the resulting COE was also expressed in “first-year-of-construction” year dollars.

The same financial modeling methodology was used for this study as per the NETL 1399 Baseline Study, and guidelines in the QGESS *Cost Estimation Methodology for NETL Assessments of Power Plant Performance* document. This is a simplified method that is a function of the plant TPC, capital charge factor, fixed and variable operating costs, capacity factor and net power generation, as shown in the equation below:

$$COE = \frac{\begin{matrix} first\ year \\ capital\ charge \end{matrix} + \begin{matrix} first\ year \\ fixed\ operating \\ costs \end{matrix} + \begin{matrix} first\ year \\ variable\ operating \\ costs \end{matrix}}{\begin{matrix} annual\ net\ megawatt\ hours \\ of\ power\ generated \end{matrix}}$$

$$COE = \frac{(CCF)(TOC) + OC_{FIX} + (CF)(OC_{VAR})}{(CF)(MWH)}$$

The capital charge factor (CCF) used in evaluating the COE was pre-calculated using the NETL Power Systems Financial Model (PSFM). This factor is valid for global economic assumptions used for a pre-determined finance structure and capital expenditure period. For the IGCC with CO_2 capture cases, the financial performance evaluations are in accordance with the high-risk, Investor Owned Utility (IOU) finance structure with a 5 year capital expenditure period. The resulting CCF is 0.1243.

2.6.2 CO_2 Sales Price

As outlined in the TEA's reporting requirements, sensitivity analysis is to be done to determine the impact of CO_2 sales on IGCC COE. The varying parameter is the CO_2 sales price at the IGCC plant gate and is to range between $0/tonne (baseline case assuming no value to the product CO2) and $60/tonne.

The formula used to calculate the revised COE after taking into account CO_2 sales is shown below:

$$COE = Baseline\ COE\ - \frac{(CO_2\ Sales\ Price)\ \times annual\ tonnes\ of\ CO_2\ product}{annual\ net\ megawatt\ hours\ of\ power\ generated}$$

2.6.3 Cost of CO_2 Emissions

The TEA also requires sensitivity analysis on cost of CO2 emissions to be performed. The varying parameter is the CO_2 emissions cost. The range of the emissions cost is between $0/tonne (baseline case assuming no CO_2 emissions cost) and $60/tonne.

The formula used to calculate the revised COE after taking into cost of CO_2 emissions is shown below:

$$COE = Baseline\ COE + \frac{(Cost\ of\ CO_2\ Emissions)\ \times annual\ tonnes\ of\ CO_2\ emitted}{annual\ net\ megawatt\ hours\ of\ power\ generated}$$

Section 3 Case P1: TRIG IGCC with Selexol-Based CO_2 Capture

3.1 PROCESS OVERVIEW

The reference Case P1 TRIG IGCC power plant, which is based on the NETL 1399 S2B IGCC case, is a Montana PRB coal-fired IGCC plant designed to generate enough hydrogen-rich fuel gas to fill two advanced GE 7F-turbines rated nominally at 215 MW each for a total of 430 MW at the Montana site's elevation. The power plant is equipped with a heat recovery steam generator (HRSG) and steam turbines to maximize power recovery.

In order to maximize CO_2 removal and maintain the same syngas heat content (Btu/SCF) to the GT, the raw syngas must be converted to hydrogen-rich syngas by the water-gas shift (WGS) reaction. Steam for the WGS reaction is provided partly by the steam generated from quench cooling of the syngas and partly by the water content in scrubber overhead gas that is also feeding to the unit. The balance of the WGS steam requirement is provided by the steam header from offsite.

The two-stage Selexol process is used to capture CO_2 and H_2S in the syngas. Due to the carbon slippage from the unconverted tar and methane in the syngas leaving the TRIG, the overall carbon capture rate is 83.5% of the raw syngas' carbon content. H_2S recovered from the Selexol unit is converted into elemental sulfur in the Claus plant.

The nominal net IGCC power export capacity after accounting for the auxiliary loads which include CO_2 capture and compression is 460 MWe.

The IGCC plant is assumed to operate as a base-loaded unit with annual on-stream capacity factor of 80 percent or 7,000 hrs/year at full capacity.

3.2 PERFORMANCE RESULTS

The SRI-modeled Case P1 IGCC plant with CO_2 capture consumes 6,935 tpd PRB coal at the Montana site and produces a net output of 457 MWe with a net plant efficiency of 31.5 percent on a HHV basis. Overall performance for the Case P1 IGCC plant is summarized in Table 3-1, which includes auxiliary power requirements.

Table 3-1
Case P1 Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	Case P1
Gas Turbine Power	426,372
Steam Turbine Power	192,447
TOTAL POWER, kWe	**618,819**
Auxiliary Load Summary, kWe	
Coal Handling	510
Coal Milling	730
Slag Handling	631
Coal Dryer Circulation Blower	2,563
Air Separation Unit Auxiliaries	1,002
Air Separation Unit Main Air Compressor	53,796
Oxygen Compressor	6,600
Nitrogen Compressors	30,060
CO_2 Compressor	28,180
Boiler Feedwater Pumps	5,074
Condensate Pump	230
Syngas Recycle Compressor	1,544
Circulating Water Pump	2,035
Ground Water Pumps	248
Cooling Tower Fans	1,330
Air Cooled Condenser Fans	2,272
Acid Gas Removal	16,446
Gas Turbine Auxiliaries	1,000
Steam Turbine Auxiliaries	99
Claus Plant/TGTU Auxiliaries	247
Claus Plant TG Recycle Compressor	1,726
Miscellaneous Balance of Plant	3,000
Transformer Losses	2,321
TOTAL AUXILIARIES, kWe	**161,643**
NET POWER, kWe	**457,176**
Net Plant Efficiency, % (HHV)	31.5%
Net Plant Heat Rate, Btu/kWh	10,826
CONDENSER COOLING DUTY, MMBtu/hr	**958**
CONSUMABLES	
As-Received Coal Feed, lb/hr	577,940
Thermal Input, kWt	1,450,549
Raw Water Withdrawal, gpm	2,898
Raw Water Consumption, gpm	2,423

3.3 EQUIPMENT LIST

As the Case P1 TRIG IGCC is based on the NETL 1399 report's S2B case, the reader should refer to the S2B equipment list in the NETL 1399 report.

3.4 CAPITAL COST

Table 3-2 shows the cost breakdown of the Case P1 TRIG IGCC with Selexol-based AGR, consistent with the Code of Accounts format as expressed in the NETL 1399 report.

Table 3-3 shows the calculation and addition of owner's costs to determine the TOC, used to calculate COE.

The estimated TOC of the Case P1 TRIG IGCC with Selexol-based CO_2 capture using PRB coal in 2011 dollars is $4,546/kW.

Table 3-2
Case P1 Total Plant Cost Summary

Case P1: TRIG IGCC with Selexol-Based AGR | **Coal Feed, lb/hr** 577,940 | **Plant Size** 457.2 MW, net
Low Rank Western Coal Baseline Study PRB | **Coal HHV, Btu/lb** 8,564 | **Net Efficiency** 31.5%

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	$	$/kW
1	COAL & SORBENT HANDLING	**$19,277**	**$3,380**	**$14,744**	**$0**	**$0**	**$37,401**	**$3,393**	**$0**	**$8,158**	**$48,953**	**$107**
2	COAL & SORBENT PREP & FEED	**$93,928**	**$7,624**	**$15,906**	**$0**	**$0**	**$117,458**	**$10,188**	**$0**	**$25,530**	**$153,176**	**$335**
3	FEEDWATER & MISC BOP SYSTEMS	**$8,921**	**$7,926**	**$8,285**	**$0**	**$0**	**$25,132**	**$2,362**	**$0**	**$6,278**	**$33,772**	**$74**
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries (TRIG)	$155,267	$0	$66,713	$0	$0	$221,980	$19,820	$51,180	$44,892	$337,872	$739
4.2	Syngas Cooling (w/4.1)	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.3	ASU/Oxidant Compression	$155,889	$0	$0	$0	$0	$155,889	$15,111	$0	$17,100	$188,100	$411
4.4	LT Heat Recovery & FG Saturation	$11,084	$0	$4,185	$0	$0	$15,269	$1,491	$0	$3,352	$20,111	$44
4.x	Other Gasification Equipment	$0	$11,857	$6,836	$0	$0	$18,693	$1,719	$0	$5,009	$25,421	$56
	SUBTOTAL 4.	**$322,241**	**$11,857**	**$77,733**	**$0**	**$0**	**$411,831**	**$38,140**	**$51,180**	**$70,353**	**$571,504**	**$1,250**
5A	GAS CLEANUP & PIPING	**$92,047**	**$3,189**	**$76,367**	**$0**	**$0**	**$171,603**	**$16,576**	**$27,772**	**$43,334**	**$259,284**	**$567**
5B	CO2 REMOVAL & COMPRESSION	**$34,702**	**$0**	**$11,592**	**$0**	**$0**	**$46,294**	**$4,457**	**$0**	**$10,150**	**$60,901**	**$133**
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	$111,211	$0	$7,881	$0	$0	$119,092	$11,290	$11,909	$14,229	$156,520	$342
6.x	Combustion Turbine Other	$0	$923	$1,067	$0	$0	$1,991	$186	$0	$653	$2,830	$6
	SUBTOTAL 6.	**$111,211**	**$923**	**$8,948**	**$0**	**$0**	**$121,083**	**$11,476**	**$11,909**	**$14,882**	**$159,350**	**$349**
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	$27,667	$0	$5,356	$0	$0	$33,023	$3,140	$0	$3,617	$39,780	$87
7.x	Ductwork and Stack	$3,969	$2,833	$3,696	$0	$0	$10,498	$972	$0	$1,864	$13,334	$29
	SUBTOTAL 7.	**$31,635**	**$2,833**	**$9,053**	**$0**	**$0**	**$43,521**	**$4,113**	**$0**	**$5,481**	**$53,114**	**$116**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$29,377	$0	$4,482	$0	$0	$33,859	$3,249	$0	$3,711	$40,818	$89
8.x	Turbine Plant Auxiliaries and Steam Piping	$39,697	$881	$13,725	$0	$0	$54,303	$5,123	$0	$12,912	$72,338	$158
	SUBTOTAL 8.	**$69,074**	**$881**	**$18,206**	**$0**	**$0**	**$88,162**	**$8,372**	**$0**	**$16,624**	**$113,157**	**$248**
9	COOLING WATER SYSTEM	**$5,698**	**$7,026**	**$5,905**	**$0**	**$0**	**$18,629**	**$1,715**	**$0**	**$4,265**	**$24,609**	**$54**
10	ASH/SPENT SORBENT HANDLING SYS	**$24,387**	**$1,857**	**$12,005**	**$0**	**$0**	**$38,249**	**$3,670**	**$0**	**$4,576**	**$46,496**	**$102**
11	ACCESSORY ELECTRIC PLANT	**$33,678**	**$14,787**	**$26,980**	**$0**	**$0**	**$75,445**	**$6,491**	**$0**	**$15,677**	**$97,613**	**$214**
12	INSTRUMENTATION & CONTROL	**$12,615**	**$2,556**	**$8,293**	**$0**	**$0**	**$23,464**	**$2,125**	**$1,174**	**$4,488**	**$31,252**	**$68**
13	IMPROVEMENTS TO SITE	**$3,619**	**$2,133**	**$9,497**	**$0**	**$0**	**$15,250**	**$1,505**	**$0**	**$5,026**	**$21,780**	**$48**
14	BUILDINGS & STRUCTURES	**$0**	**$7,251**	**$8,182**	**$0**	**$0**	**$15,433**	**$1,406**	**$0**	**$2,776**	**$19,614**	**$43**
	CALCULATED TOTAL COST	**$863,033**	**$74,224**	**$311,696**	**$0**	**$0**	**$1,248,954**	**$115,988**	**$92,035**	**$237,598**	**$1,694,575**	**$3,707**

Table 3-3
Case P1 Total Plant Cost Summary

Owner's Costs	$ x $1,000	$/kW
Preproduction Costs		
6 months All Labor	$13,301	$29
1 Month Maintenance Materials	$2,941	$6
1 Month Non-Fuel Consumables	$390	$1
1 Month Waste Disposal	$490	$1
25% of 1 Months Fuel Cost at 100% CF	$1,035	$2
2% of TPC	$33,891	$74
Total	**$52,049**	**$114**
Inventory Capital		
60 day supply of fuel at 100% CF	$8,168	$18
60 day supply of non-fuel consumables at 100% CF	$588	$1
0.5% of TPC (spare parts)	$8,473	$19
Total	**$17,229**	**$38**
Initial Cost for Catalyst and Chemicals	**$13,421**	**$29**
Land	**$900**	**$2**
Other Owner's Cost	**$254,186**	**$556**
Financing Costs	**$45,754**	**$100**
Total Owner's Costs	**$383,539**	**$839**
Total Overnight Costs (TOC)	**$2,078,114**	**$4,546**

3.5 OPERATING COSTS

Table 3-9 shows the operating cost breakdown for the Case P1 Reference TRIG IGCC.

Table 3-4
Case P1 Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES			
Case:	Case P1: TRIG IGCC with Selexol-Based AGR		
Plant Size (MWe)	**457**	**Heat Rate (Btu/kWh):**	10,826
Primary/Secondary Fuel:	**PRB**	**Fuel Cost ($/MMBtu):**	
Design/Construction	5 years	**Book Life (yrs):**	20
TPC (Plant Cost) Year	June 2011	**TPI Year:**	2016
Capacity Factor (%)	80	**CO2 Captured (TPD)**	10306

OPERATING & MAINTENANCE LABOR

Operating Labor

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

Operating Labor Requirements per Shift	**units/mod**	**Total Plant**
Skilled Operator	2.0	2.0
Operator	10.0	10.0
Foreman	1.0	1.0
Lab Tech's etc	3.0	3.0
TOTAL Operating Jobs	16.0	16.0

	Annual Cost $	Annual Unit Cost $/kW-net
Annual Operating Labor Cost	$7,233,658	
Maintenance Labor Cost	$14,048,361	
Administration & Support Labor	$5,320,505	
Property Taxes and Insurance	$33,891,490	
TOTAL FIXED OPERATING COSTS	**$60,494,014**	

VARIABLE OPERATING COSTS

	Consumption Initial	Consumption /Day	Unit Cost	Initial Fill Cost		$/kWh-net
Maintenance Material Cost					**$28,233,392**	
Water(/1000 gallons)	0	2,087	1.67	**$0**	**$1,020,007**	
Chemicals						
MU & WT Chem (lb)	0	12433	0.27	$0	$972,411	
Carbon (Hg Removal) (lb)	89517	123	1.63	$145,913	$58,543	
COS Catalyst (m3)	0	0	3751.70	$0	$0	
Water Gas Shift Catalyst (ft3)	5031	3.45	771.99	$3,884,169	$776,618	
Selexol Solution (gal)	255267	80.83	36.79	$9,391,269	$868,371	
SCR Catalyst (m3)	0	0	0.00	$0	$0	
Ammonia (19% NH3) (ton)	0	0	0.00	$0	$0	
Claus Catalyst (ft3)	w/equip	0.75	203.15	$0	$44,620	
Subtotal Chemicals				**$13,421,350**	**$2,720,563**	
Other						
Supplemental Fuel (MMBtu)	0	0	0.00	$0	$0	
Gases, N2 etc.(/100scf)	0	0	0.00	$0	$0	
LP Steam (/1000 lbs)	0	0	0.00	$0	$0	
Subtotal Other				**$0**	**$0**	
Waste Disposal:						
Spent Mercury Catalyst (lb)	0	123	0.65	$0	$23,345	
Flyash (ton)	0	0	0.00	$0	$0	
Slag (ton)	0	639	25.11	$0	$4,683,624	
Subtotal Waste Disposal				**$0**	**$4,706,969**	
By-products & Emissions						
Sulfur (tons)	0	50	0.00	$0	$0	
Subtotal By-Products				**$0**	**$0**	
TOTAL VARIABLE OPERATING COSTS				**$13,421,350**	**$36,680,931**	
Fuel (tons)	**0**	**6935**	**19.63**	**$0**	**$39,752,768**	

3.6 COST OF ELECTRICITY

Table 3-5 shows a summary of the power output, CAPEX, OPEX, COE and cost of CO_2 capture for the Case P1 TRIG IGCC with Selexol-based AGR and CO_2 capture. The Case P1 IGCC COE is estimated to be 123.4 mills/kWh

Table 3-5
Plant Performance and Economic Summary

Case	Case P1
CAPEX, $MM	
Total Installed Cost (TIC)	$1,249
Total Plant Cost (TPC)	$1,695
Total Overnight Cost (TOC)	$2,078
OPEX, $MM/yr (100% Capacity Factor Basis)	
Fixed Operating Cost (OC_{fix})	$60.5
Variable Operating Cost, less Fuel (OC_{var})	$45.9
Fuel (OC_{fuel})	$49.7
Power Production, MWe	
Gas Turbine	426.4
Steam Turbine	192.4
Auxiliary Power Consumption	161.6
Net Power Output	457.2
Power Generated, MWh/yr (MWH)	4,004,859
COE, excl CO_2 TS&M, mills/kWh	**123.4**
COE, incl CO_2 TS&M, mills/kWh	**142.1**
Cost of CO_2 Avoided excl CO2 TS&M, $/ton CO_2	**58.4**
Cost of CO_2 Avoided incl CO2 TS&M, $/ton CO_2	**83.3**

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Section 4 Case P2: TRIG with SRI Catalytic Reformer IGCC

4.1 PROCESS OVERVIEW

The Case P2 IGCC power plant, like the Case P1 plant, is a Montana PRB coal-fired TRIG-based IGCC plant designed to generate enough hydrogen-rich fuel gas to fill two advanced GE 7F-turbines to generate nominally a total of 430 MW at the Montana site's elevation. To maximize power recovery, the power plant is equipped with a heat recovery steam generator (HRSG) and steam turbines.

The IGCC plant operates as a base-loaded unit with an annual on-stream capacity factor of 80 percent.

In the Case P1 IGCC, the raw syngas leaving the TRIG contains a significant amount of methane and potentially some tar as well. The fate of the tar is unknown while the methane remains unconverted throughout the process until it is combusted in the gas turbine, producing CO_2 that is vented to the atmosphere. Due to the carbon slippage to the gas turbine in the form of methane, the Case P1 IGCC is unable to achieve the desired 90% carbon capture rate.

The Case P2 TRIG coupled with SRI Catalytic Reformer IGCC has the following characteristics that differentiate it from the Case P1 TRIG Reference IGCC:

- Tar and methane-containing syngas leaving the TRIG enters the SRI catalytic reformer, along with a stream of oxygen from the ASU. In the reformer, practically all the tars and most of the methane are destroyed and converted into CO and H_2. The hydrogen-rich syngas exits the reformer at about the same temperature as the raw syngas leaving the TRIG (1,800ºF) and is routed to the convective cooler downstream.

- The catalytic reformer requires an oxidant stream to reform the methane and tars into CO and H2. Additional oxygen has to be generated from the ASU and supplied to the reformer. The resulting ASU in the Case P2 IGCC is therefore larger and consumes more power than the Case P1 ASU.

- Most of the methane from the TRIG has been converted to CO and H_2 in the reformer upstream. The resulting syngas, now containing a larger quantity of CO and H2 is then cooled and sent to the shift reactor, where the CO is converted to CO_2 and H_2 in the presence of steam. More CO_2 is thus available for the two-stage Selexol unit to capture in order to meet the 90% CO_2 capture rate. However, this is achieved at the expense of more reboiling steam to the stripper column and higher CO_2 compression horsepower, resulting in lower steam turbine power generation and larger auxiliary power consumption respectively.

4.2 PERFORMANCE RESULTS

The SRI-modeled Case P2 IGCC plant with CO_2 capture consumes 6,935 tpd of PRB coal at the Montana site to produce a net output of 439 MWe at a net plant efficiency of 30.3 percent on a HHV basis. Overall performance for the Case P2 IGCC plant is summarized in Table 4-1, which includes auxiliary power requirements.

Table 4-1
Case P2 Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	Case P2
Gas Turbine Power	425,361
Steam Turbine Power	189,148
TOTAL POWER, kWe	**614,509**
Auxiliary Load Summary, kWe	
Coal Handling	510
Coal Milling	730
Slag Handling	631
Coal Dryer Circulation Blower	2,563
Air Separation Unit Auxiliaries	1,147
Air Separation Unit Main Air Compressor	61,608
Oxygen Compressor	7,575
Nitrogen Compressors	30,060
CO_2 Compressor	31,077
Boiler Feedwater Pumps	5,172
Condensate Pump	230
Syngas Recycle Compressor	1,710
Circulating Water Pump	2,062
Ground Water Pumps	254
Cooling Tower Fans	1,347
Air Cooled Condenser Fans	2,241
Acid Gas Removal	18,122
Gas Turbine Auxiliaries	998
Steam Turbine Auxiliaries	97
Claus Plant/TGTU Auxiliaries	247
Claus Plant TG Recycle Compressor	1,528
Miscellaneous Balance of Plant	3,000
Transformer Losses	2,305
TOTAL AUXILIARIES, kWe	**175,212**
NET POWER, kWe	**439,297**
Net Plant Efficiency, % (HHV)	30.3%
Net Plant Heat Rate, Btu/kWh	11,267
CONDENSER COOLING DUTY, MMBtu/hr	**945**
CONSUMABLES	
As-Received Coal Feed, lb/hr	577,940
Thermal Input, kWt	1,450,549
Raw Water Withdrawal, gpm	2,977
Raw Water Consumption, gpm	2,498

4.3 CAPITAL COST

Table 4-2 shows the cost breakdown of the Case P2 IGCC utilizing TRIG coupled with the SRI catalytic reformer-based IGCC, expressed in a consistent format with the Code of Accounts in the NETL 1399 report.

Table 4-3 shows the calculation and addition of owner's costs to determine the TOC, used to calculate COE.

The estimated TOC of the Case P2 TRIG-based IGCC with SRI catalytic reformer using PRB coal in 2011 dollars is $4,531/kW.

Table 4-2
Case 2b Total Plant Cost Summary

Case P2: Coupled TRIG/SRI Catalytic Reformer IGCC w/Selexol-Based AGR		Coal Feed, lb/hr	577,940	Plant Size	439.3 MW, net
Low Rank Western Coal Baseline Study	PRB	Coal HHV, Btu/lb	8,564	Net Efficiency	30.3%

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 $	TOTAL PLANT COST $/kW
1	COAL & SORBENT HANDLING	**$19,277**	**$3,380**	**$14,744**	**$0**	**$0**	**$37,401**	**$3,393**	**$0**	**$8,158**	**$48,953**	**$111**
2	COAL & SORBENT PREP & FEED	**$93,928**	**$7,624**	**$15,906**	**$0**	**$0**	**$117,458**	**$10,188**	**$0**	**$25,530**	**$153,176**	**$349**
3	FEEDWATER & MISC BOP SYSTEMS	**$9,689**	**$8,192**	**$9,022**	**$0**	**$0**	**$26,903**	**$2,530**	**$0**	**$6,678**	**$36,111**	**$82**
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries (TRIG)	$156,376	$0	$67,189	$0	$0	$223,565	$19,962	$51,545	$45,213	$340,284	$775
4.2	SRI Catalytic Reformer	$6,341	$0	$2,122	$0	$0	$8,463	$1,577	$2,146	$2,008	$14,194	$32
4.3	ASU/Oxidant Compression	$163,688	$0	$0	$0	$0	$163,688	$15,867	$0	$17,956	$197,510	$450
4.4	LT Heat Recovery & FG Saturation	$11,285	$0	$4,261	$0	$0	$15,546	$1,518	$0	$3,413	$20,476	$47
4.x	Other Gasification Equipment	$0	$12,072	$6,960	$0	$0	$19,032	$1,750	$0	$5,100	$25,882	$59
	SUBTOTAL 4.	**$337,690**	**$12,072**	**$80,532**	**$0**	**$0**	**$430,293**	**$40,673**	**$53,691**	**$73,689**	**$598,346**	**$1,362**
5A	GAS CLEANUP & PIPING	**$101,133**	**$3,218**	**$83,936**	**$0**	**$0**	**$188,286**	**$18,189**	**$31,057**	**$47,650**	**$285,182**	**$649**
5B	CO2 REMOVAL & COMPRESSION	**$37,053**	**$0**	**$12,377**	**$0**	**$0**	**$49,431**	**$4,759**	**$0**	**$10,838**	**$65,028**	**$148**
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	$111,211	$0	$7,881	$0	$0	$119,092	$11,290	$11,909	$14,229	$156,520	$356
6.x	Combustion Turbine Other	$0	$923	$1,067	$0	$0	$1,991	$186	$0	$653	$2,830	$6
	SUBTOTAL 6.	**$111,211**	**$923**	**$8,948**	**$0**	**$0**	**$121,083**	**$11,476**	**$11,909**	**$14,882**	**$159,350**	**$363**
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	$27,412	$0	$5,307	$0	$0	$32,719	$3,111	$0	$3,583	$39,414	$90
7.x	Ductwork and Stack	$3,974	$2,836	$3,701	$0	$0	$10,511	$974	$0	$1,866	$13,351	$30
	SUBTOTAL 7.	**$31,386**	**$2,836**	**$9,008**	**$0**	**$0**	**$43,230**	**$4,085**	**$0**	**$5,450**	**$52,765**	**$120**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$29,023	$0	$4,428	$0	$0	$33,451	$3,209	$0	$3,667	$40,327	$92
8.x	Turbine Plant Auxiliaries and Steam Piping	$39,423	$870	$13,624	$0	$0	$53,917	$5,086	$0	$12,822	$71,826	$164
	SUBTOTAL 8.	**$68,446**	**$870**	**$18,052**	**$0**	**$0**	**$87,368**	**$8,296**	**$0**	**$16,489**	**$112,153**	**$255**
9	COOLING WATER SYSTEM	**$5,753**	**$7,082**	**$5,956**	**$0**	**$0**	**$18,791**	**$1,730**	**$0**	**$4,301**	**$24,822**	**$57**
10	ASH/SPENT SORBENT HANDLING SYS	**$24,387**	**$1,857**	**$12,005**	**$0**	**$0**	**$38,249**	**$3,670**	**$0**	**$4,576**	**$46,496**	**$106**
11	ACCESSORY ELECTRIC PLANT	**$34,206**	**$15,340**	**$27,950**	**$0**	**$0**	**$77,496**	**$6,674**	**$0**	**$16,138**	**$100,307**	**$228**
12	INSTRUMENTATION & CONTROL	**$12,861**	**$2,606**	**$8,455**	**$0**	**$0**	**$23,923**	**$2,166**	**$1,197**	**$4,576**	**$31,862**	**$73**
13	IMPROVEMENTS TO SITE	**$3,661**	**$2,157**	**$9,605**	**$0**	**$0**	**$15,423**	**$1,522**	**$0**	**$5,083**	**$22,028**	**$50**
14	BUILDINGS & STRUCTURES	**$0**	**$7,302**	**$8,250**	**$0**	**$0**	**$15,552**	**$1,417**	**$0**	**$2,796**	**$19,764**	**$45**
	CALCULATED TOTAL COST	**$890,680**	**$75,461**	**$324,746**	**$0**	**$0**	**$1,290,887**	**$120,768**	**$97,854**	**$246,834**	**$1,756,344**	**$3,998**

Table 4-3
Case P2 Total Plant Cost Summary

Owner's Costs	$ x $1,000	$/kW
Preproduction Costs		
6 months All Labor	$13,621	$31
1 Month Maintenance Materials	$3,048	$7
1 Month Non-Fuel Consumables	$420	$1
1 Month Waste Disposal	$490	$1
25% of 1 Months Fuel Cost at 100% CF	$1,035	$2
2% of TPC	$35,127	$80
Total	**$53,742**	**$122**
Inventory Capital		
60 day supply of fuel at 100% CF	$8,168	$19
60 day supply of non-fuel consumables at 100% CF	$617	$1
0.5% of TPC (spare parts)	$8,782	$20
Total	**$17,567**	**$40**
Initial Cost for Catalyst and Chemicals	**$14,446**	**$33**
Land	**$900**	**$2**
Other Owner's Cost	**$263,452**	**$600**
Financing Costs	**$47,421**	**$108**
Total Owner's Costs	**$397,528**	**$905**
Total Overnight Costs (TOC)	**$2,153,872**	**$4,903**

4.4 OPERATING COSTS

Table 4-4 shows the operating cost breakdown for the Case P2 Coupled TRIG/SRI Catalytic Reformer-based IGCC.

Table 4-4
Case P2 Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES

Case:	Case P2- Coupled TRIG/SRI Catalytic Reformer IGCC w/ Selexol-based AGR		
Plant Size (MWe)	**439**	**Heat Rate (Btu/kWh):**	11,267
Primary/Secondary Fuel:	**PRB**	**Fuel Cost ($/MMBtu):**	
Design/Construction	5 years	**Book Life (yrs):**	20
TPC (Plant Cost) Year	June 2011	**TPI Year:**	2016
Capacity Factor (%)	80	**CO2 Captured (TPD)**	11373

OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate (base):		$39.70 $/hr				
Operating Labor Burden:		30.00 % of base				
Labor Overhead Charge		25.00 % of labor				
Operating Labor Requirements per Shift		**units/mod**		**Total Plant**		
Skilled Operator		2.0		2.0		
Operator		10.0		10.0		
Foreman		1.0		1.0		
Lab Tech's etc		3.0		3.0		
TOTAL Operating Jobs		16.0		16.0		
					Annual Cost $	Annual Unit Cost $/kW-net
Annual Operating Labor Cost					$7,233,658	
Maintenance Labor Cost					$14,560,440	
Administration & Support Labor					$5,448,524	
Property Taxes and Insurance					$35,126,872	
TOTAL FIXED OPERATING COSTS					**$62,369,494**	
VARIABLE OPERATING COSTS						
						$/kWh-net
Maintenance Material Cost					**$29,262,530**	
Consumables	Consumption Initial	Consumption /Day	Unit Cost	Initial Fill Cost		
Water(/1000 gallons)	0	2,144	1.67	**$0**	**$1,047,679**	
Chemicals						
MU & WT Chem (lb)	0	12770	0.27	$0	$998,792	
Carbon (Hg Removal) (lb)	89517	123	1.63	$145,913	$58,543	
SRI Reforming Catalyst (ft3)	w/equip	0.63	750	$0	$138,733	
Water Gas Shift Catalyst (ft3)	5119	3.51	771.99	$3,952,151	$790,211	
Selexol Solution (gal)	281275	89.07	36.79	$10,348,092	$956,844	
SCR Catalyst (m3)	0	0	0.00	$0	$0	
Ammonia (19% NH3) (ton)	0	0	0.00	$0	$0	
Claus Catalyst (ft3)	w/equip	0.75	203.15	$0	$44,611	
Subtotal Chemicals				**$14,446,156**	**$2,987,734**	
Other						
Supplemental Fuel (MMBtu)	0	0	0.00	$0	$0	
Gases, N2 etc.(/100scf)	0	0	0.00	$0	$0	
LP Steam (/1000 lbs)	0	0	0.00	$0	$0	
Subtotal Other				**$0**	**$0**	
Waste Disposal:						
Spent Mercury Catalyst (lb)	0	123	0.65	$0	$23,345	
Flyash (ton)	0	0	0.00	$0	$0	
Slag (ton)	0	639	25.11	$0	$4,683,622	
Subtotal Waste Disposal				**$0**	**$4,706,967**	
By-products & Emissions						
Sulfur (tons)	0	50	0.00	$0	$0	
Subtotal By-Products				**$0**	**$0**	
TOTAL VARIABLE OPERATING COSTS				**$14,446,156**	**$38,004,910**	
Fuel (tons)	**0**	**6935**	**19.63**	**$0**	**$39,752,768**	

4.5 COST OF ELECTRICITY

Table 4-5 shows a summary of the power output, CAPEX, OPEX, COE and cost of CO_2 capture for the Case P2 Coupled TRIG/SRI Catalytic Reformer-based IGCC. The COE for the Case P2 IGCC is 132.5 mills/kWh

Table 4-5
Plant Performance and Economic Summary

Case	Case P2
CAPEX, $MM	
Total Installed Cost (TIC)	$1,291
Total Plant Cost (TPC)	$1,756
Total Overnight Cost (TOC)	$2,154
OPEX, $MM/yr (100% Capacity Factor Basis)	
Fixed Operating Cost (OC_{fix})	$62.4
Variable Operating Cost, less Fuel (OC_{var})	$47.5
Fuel (OC_{fuel})	$49.7
Power Production, MWe	
Gas Turbine	425.4
Steam Turbine	189.1
Auxiliary Power Consumption less Hydraulic Turbine	175.2
Net Power Output	439.3
Power Generated, MWh/yr (MWH)	3,848,240
COE, excl CO_2 TS&M, mills/kWh	**132.5**
COE, incl CO_2 TS&M, mills/kWh	**154.0**
Cost of CO_2 Avoided excl CO2 TS&M, $/ton CO_2	**64.7**
Cost of CO_2 Avoided incl CO2 TS&M, $/ton CO_2	**90.9**

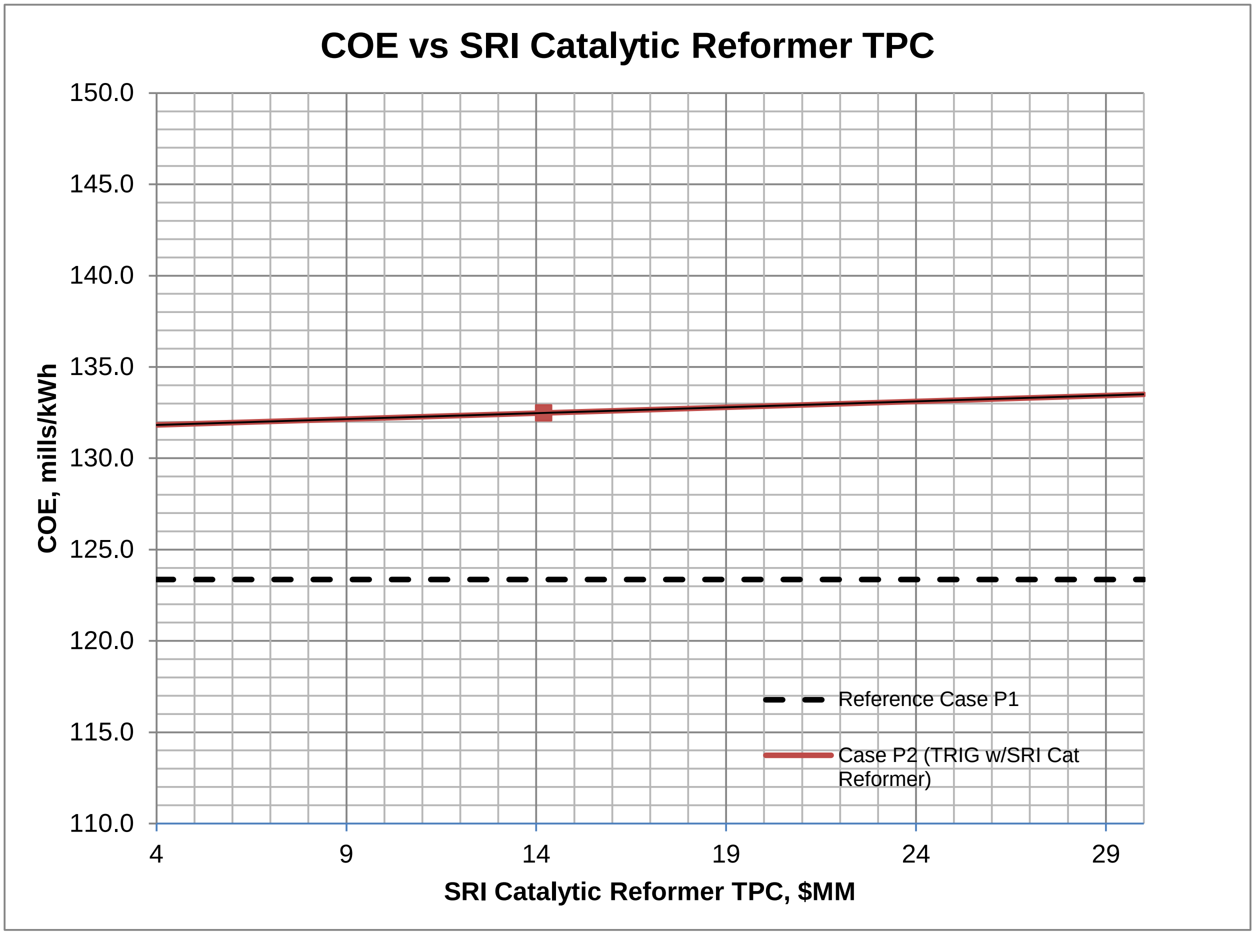
Section 5 Sensitivity Analysis

Sensitivity analysis was carried out to determine the effects of various parameters of the SRI catalytic reformer on the overall IGCC COE. The parameters investigated include: system capital cost, feedstock cost, IGCC plant capacity factor, CO_2 sales price, and cost of CO_2 emissions.

5.1 SRI CATALYTIC REFORMER COST

Figure 5-1 shows how the P2 IGCC COE changes as the SRI catalytic reformer TPC varies from $4MM to $30MM, compared with the baseline estimate of about $14.2MM. Also shown in figure is the reference Case P1 TRIG IGCC COE at 123.4 mills/kWh.

Figure 5-1
Sensitivity Analysis – COE vs Feed System TPC



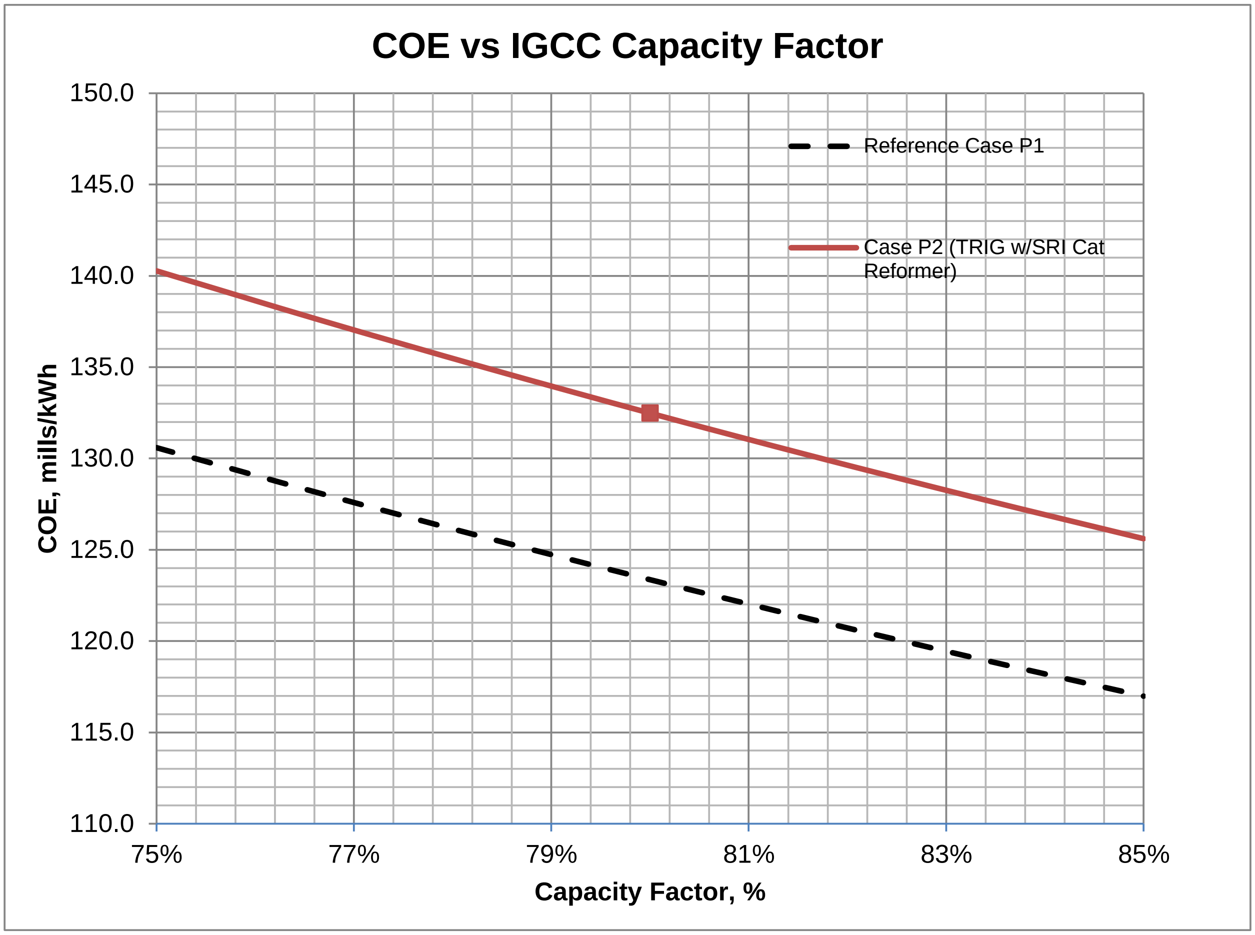
From Figure 6-1, it can be seen that the COE does not vary significantly with respect to the SRI-catalytic reformer TPC. This is because the SRI catalytic reformer is merely a very small contributor to the overall IGCC TPC ($32/kW out of $3,998/kW).

For the Case P2 coupled TRIG/SRI catalytic reformer-based IGCC cases, it takes a $15.6MM increase catalytic reformer TPC to increase the IGCC COE by 1 mill/kWh.

5.2 CAPACITY FACTOR

The baseline IGCC plant capacity factor used in this study is 80%. Figure 5-2 shows how the IGCC COE varies with plant capacity factor as it varies from 75% to 85%.

Figure 5-2
Sensitivity Analysis – COE vs IGCC Plant Capacity Factor

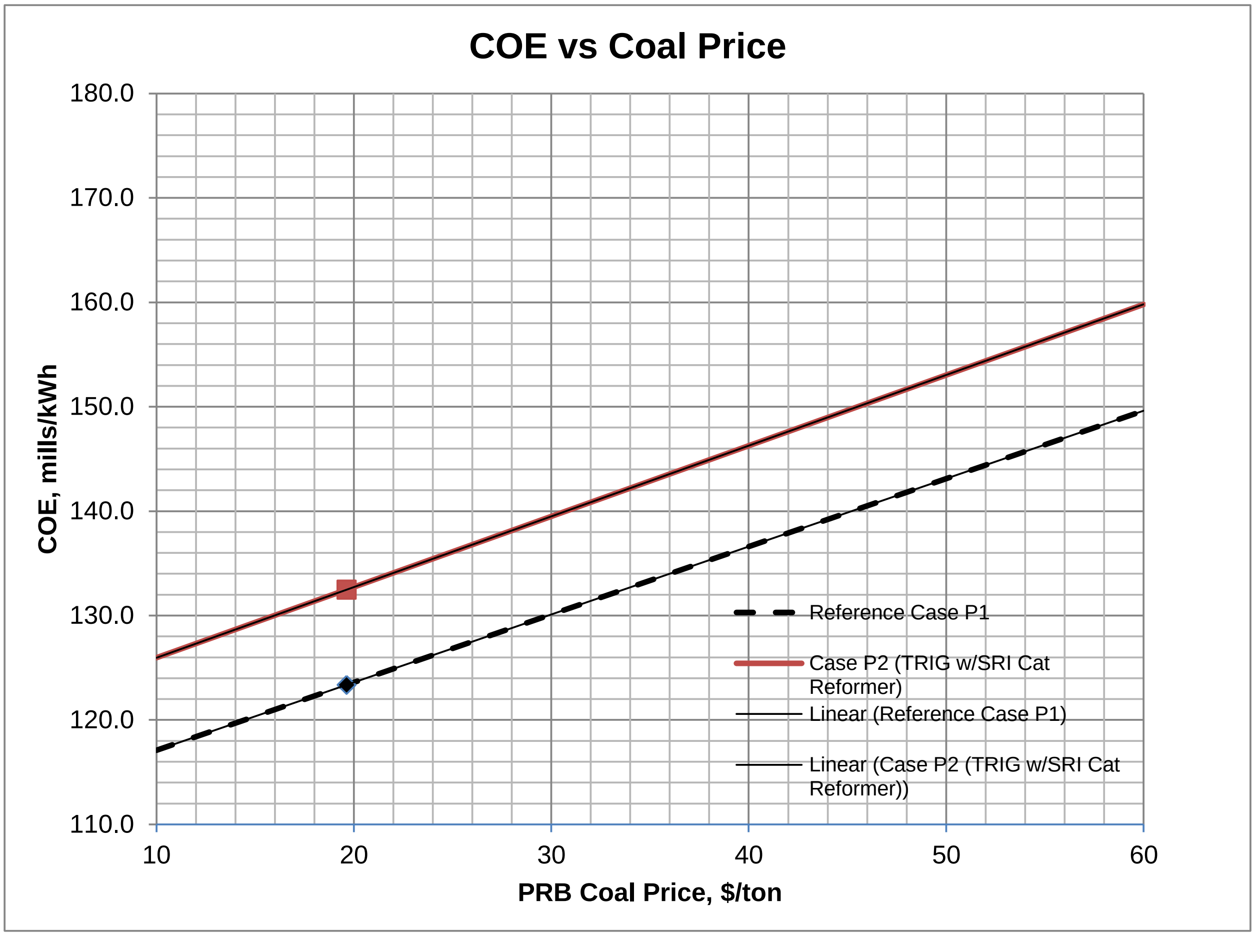


5.3 FEEDSTOCK PRICE

The baseline IGCC plant PRB coal feedstock price used in this study is $19.63/ton. Figure 5-3 shows how the IGCC COE varies with coal price as it varies from $10/ton to $60/ton.

As the baseline case has a higher efficiency than Case P2, it is slightly less sensitive to coal price. For the Case P1 TRIG IGCC, the COE increases by 1 mill/kWh for every $1.54/ton increase in coal price, while in Case P2 Coupled TRIG/SRI Catalytic Reformer IGCC, the COE increases by 1 mill/kWh for every $1.48/ton increase in coal price

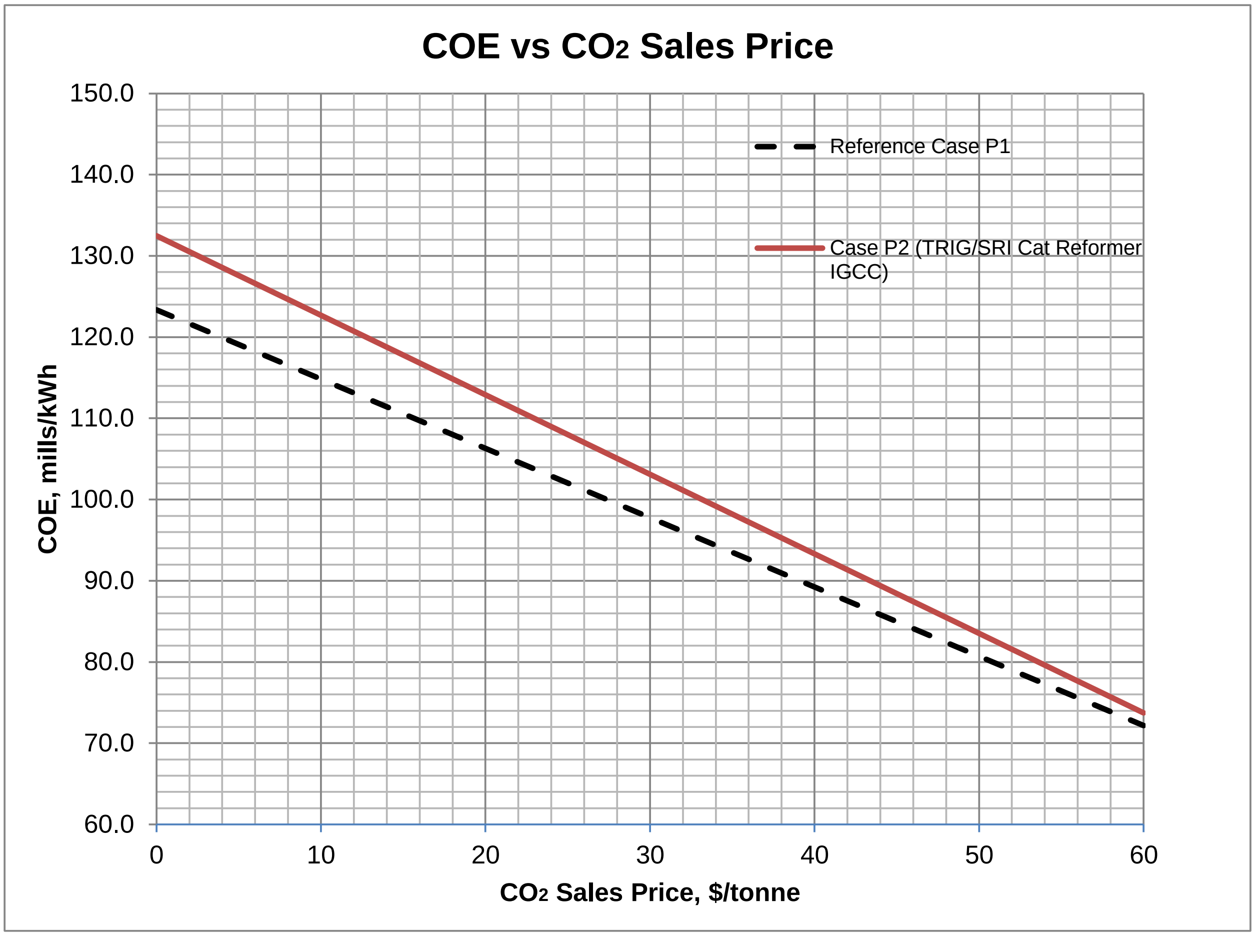
Figure 5-3
Sensitivity Analysis – COE vs Feedstock Price



5.4 CO_2 SALES PRICE

Sensitivity to CO_2 sales at plant gate prices is shown in Figure 5-4. The baseline case assumes that the CO_2 product carries no value ($0/tonne). The sales price is subsequently varied to a maximum of $60/tonne to determine its effect on the IGCC plant's COE.

Figure 5-4
Sensitivity Analysis – COE vs CO_2 Sales Price

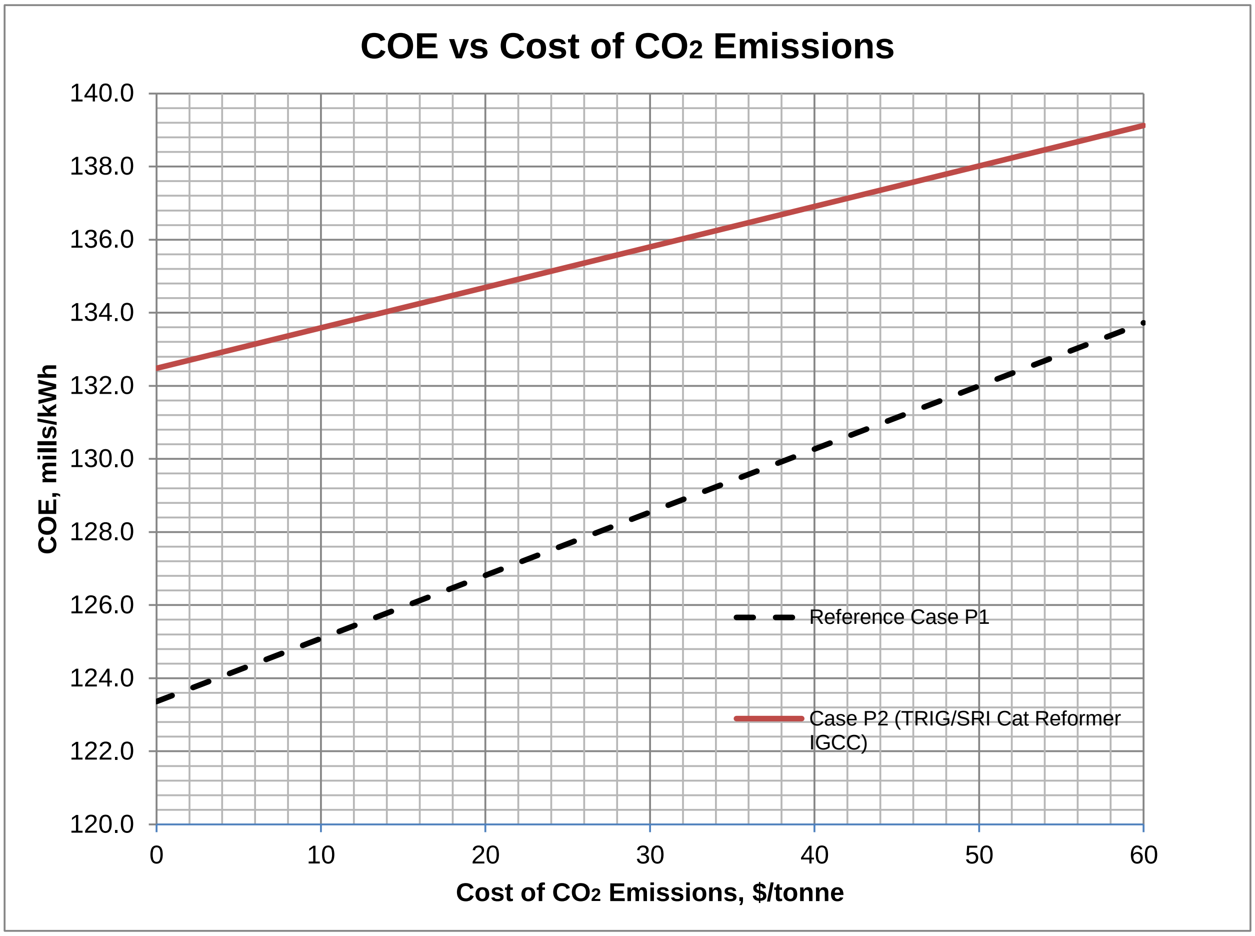


Due to the different rate of CO_2 capture between Case P1 (83.5% CO_2 capture) and Case P2 (90% CO_2 capture), the slopes are different for the two cases. Case P2 has a greater decrease in COE as CO_2 sales price increases since it captures more CO_2 for that is then subsequently sold. Based on extrapolation of the lines in Figure 5-4, Case P2's COE starts falling below the reference P1 IGCC's COE when the sales price of CO_2 exceeds $72.5/tonne.

5.5 COST OF CO_2 EMISSIONS

The sensitivity to CO_2 emissions costs is shown in Figure 8-6. The baseline case assumes that there are no costs associated with venting CO_2 to the atmosphere ($0/tonne). The cost of CO_2 emissions is subsequently varied to a maximum of $60/tonne to determine its effect on the IGCC plant's COE.

Figure 5-5
Sensitivity Analysis – COE vs Cost of CO_2 Emissions



For the same reason as explained in Section 5.4 the slopes are different between Cases P1 and P2 due to the different rate of CO_2 capture, hence different CO_2 emissions rate. Case P2, which has a higher CO_2 capture rate, vents less CO_2 to the atmosphere, so it has a smaller increase in COE as cost of CO_2 emissions increases.

The COEs are less sensitive to emissions cost than to CO_2 sales price because much more CO_2 is captured by the IGCC plant than is vented (90% vs 10%), hence the COE is about 9 times more sensitive to CO_2 sales price than to cost of CO_2 emissions.

Section 6 Conclusions

6.1 STUDY OBJECTIVE

The objective of this TEA study is to assess the performance and economic potential of integrating the catalytic steam reforming process offered by SRI with a TRIG gasifier. The catalytic reformer will treat raw syngas exiting the gasifier and generate a hydrogen-rich, tar-free syngas with minimal methane content. This will enable near-zero emissions from coal gasification for both power and coal-to-liquids productions with a carbon capture rate of more than 90%.

The current report presents the results of the techno-economic analysis (TEA) performed for the IGCC case, utilizing KBR's Transport Gasifier (TRIG) coupled with SRI's catalytic reformer.

6.2 CASE CONFIGURATIONS

The two IGCC configurations studied in this report are identified in the IGCC case study matrix shown in Table 6-1.

Table 6-1
Case Study Matrix for IGCC with CO_2 Capture

	Case P1[1]	Case P2[2]
Gasification Technology		
TRIG Gasifier	✓	✓
SRI Catalytic Reformer		✓
Gas Cleanup		
Two-Stage Selexol for CO_2 and Sulfur Removal[2]	✓	✓
Water Gas Shift		
Sour Shift	✓	✓
GE 7FB Advanced Gas Turbine	✓	✓
CO_2 Drying and Compression (to 2,200 psig)	✓	✓

[1] Case P1 modeled by SRI based on simulation of the DOE/NETL 2010/1399: Cost and Performance Baseline for Fossil Energy Plants, Volume 3a: Low Rank Coal to Electricity: IGCC Cases S2B case
[2] Case P2 modeled by SRI which added SRI Catalytic Reformer unit operation to the overall IGCC process

Both plant configurations were evaluated based on installation at a greenfield site (Montana, 3,400 ft elevation). To compare the plants on an equivalent basis, it was assumed that these plants would be dispatched any time they are available. The study capacity factor (CF) of 80% was chosen to reflect the maximum availability demonstrated by IGCC plants.

The gross and net output varies among the IGCC cases because of the gas turbine (GT) size constraint. Each case uses two combustion turbines for a combined potential gross output of 464 MW. Because these cases were operated at elevations higher than sea level, the output was reduced from the turbine's ISO condition potential. In the combined cycle, a heat recovery steam generator (HRSG) extracts heat from the CT exhaust to power a steam turbine.

6.3 RESULTS SUMMARY

Table 6-2 shows a summary comparison of the capital expenditure (CAPEX), operating expenditure (OPEX), power production, cost of electricity (COE) and cost of CO_2 avoided for the four cases.

Table 6-2
IGCC Results Summary

Case	Case P1	Case P2
IGCC Configuration		
Gasifier	KBR TRIG	KBR TRIG
Catalytic Reformer	None	SRI
Sulfur and CO_2 Removal	Selexol	Selexol
CAPEX, $MM		
Total Installed Cost (TIC)	$1,249	$1,291
Total Plant Cost (TPC)	$1,695	$1,756
Total Overnight Cost (TOC)	$2,078	$2,154
OPEX, $MM/yr (100% Capacity Factor Basis)		
Fixed Operating Cost (OC_{fix})	$60.5	$62.4
Variable Operating Cost, less Fuel (OC_{var})	$45.9	$47.5
Fuel (OC_{fuel})	$49.7	$49.7
Power Production, MWe		
Gas Turbine	426.4	425.4
Steam Turbine	192.4	189.1
Auxiliary Power Consumption	161.6	175.2
Net Power Output	457.2	439.3
Fuel Rate and Efficiency		
Coal Feed Rate, tpd AR Coal	6,935	6,935
CO_2 Capture Rate	83.5%	89.8%
Net Efficiency, %	31.5%	30.3%
Power Generated, MWh/yr (MWH)	4,004,859	3,848,240
COE, excl CO_2 TS&M, mills/kWh	**123.4**	**132.5**
COE, incl CO_2 TS&M, mills/kWh	**142.1**	**154.0**
Cost of CO_2 Avoided excl CO_2 TS&M, $/ton CO_2	**58.4**	**64.7**
Cost of CO_2 Avoided incl CO_2 TS&M, $/ton CO2	**83.3**	**90.9**

As shown, integrating the catalytic reformer does not add much value to an IGCC configuration – it incurs more costs (both capital and O&M), producing little more hydrogen but only burning it as fuel. It does allow the process to capture more CO_2, however.

It is also noted:

- Despite having a higher efficiency and lower COE, the reference Case P1 TRIG IGCC is ***unable to meet*** the 90% CO_2 capture criteria set out by DOE/NETL due to the high methane concentration in the TRIG syngas. Also, the fate of the tars and higher hydrocarbons in the raw syngas leaving the TRIG is unknown. There is likelihood that these heavy hydrocarbon compounds may condense and plug coolers downstream of the TRIG, rendering the whole process infeasible.

- Case P2 does not face the same issues as Case P1. With the addition of the SRI catalytic reformer, practically all the tars are destroyed and most of the methane is reformed, in the presence of oxygen to form CO and H_2. The CO is converted to CO_2 downstream in the shift reactors, thus enabling a CO_2 capture rate of 90% or greater. Despite the higher capital cost and slightly reduced efficiency, this case meets DOE/NETL's CO_2 capture guidelines.

- Case P2 consumes more auxiliary power mainly due to the following:
  - More oxygen has to be supplied by the ASU as the catalytic reformer consumes oxygen to reform methane and other higher hydrocarbons to CO and H_2
  - The CO_2 capture and compression process consumes more power since more CO_2 is captured and subsequently compressed in Case P2.

°F	Degree Fahrenheit
AGR	Acid Gas Removal
AOI	Area of Interest
ASU	Air Separation Unit
BEC	Bare Erected Cost
BFD	Block Flow Diagram
BFW	Boiler Feed Water
BOP	Balance of Plant
Btu	British Thermal Unit
CAPEX	Capital Expenditure
CCF	Capital Charge Factor
CF	Capacity Factor
CH_4	Methane
CO	Carbon Monoxide
CO_2	Carbon Dioxide
COE	Cost of Electricity
FOA	Funding Opportunity Announcement
ft	feet
GE	General Electric
H_2	Hydrogen
H_2O	Water
H_2S	Hydrogen Sulfide
HHV	Higher Heating Value
HMB	Heat and Material Balance
HP	High Pressure
hr	Hour
HRSG	Heat Recovery Steam Generator
IGCC	Integrated Gasification Combined Cycle
IOU	Investor Owned Utilitiy
kWe	Kilowatt electric
kWh	kilowatt hour
lb	Pound Mass
LHV	Lower Heating Value
LP	Low Pressure
max	Maximum
ME	Major Equipment
MEC	Major Equipment Cost

min	Minimum
Misc	Miscellaneous
MM	million
MU	Makeup
MWe	Megawatt electric
MWh	megawatt hour
N_2	Nitrogen
NETL	National Energy Technology Laboratory
O&M	Operating and Maintenance
O_2	Oxygen
PFD	Process Flow Diagram
ppmv	Parts per Million by Volume
ppmW, ppmw	Parts per Million by Weight
PRB	Powder River Basin
PSFM	Power Systems Financial Model
psi	Pounds Per Square Inch
psia	Pounds Per Square Inch, absolute
psig	Pounds Per Square Inch, gauge
QGESS	Quality Guidelines for Energy System Studies
SC	Supercritical
SO_2	Sulfur Dioxide
SOPO	Statement of Project Objectives
SRI	Southern Research Institute
STG	Steam Turbine Power Generation
T&S	Transportation and Storage
TDC	Total Direct Cost
TEA	Techno-Economic Analysis
TFC	Total Field Cost
TG	Turbine Generator
TGTU	Tail Gas Treatment Unit
TIC	Total Installed Cost
TOC	Total Overnight Cost
TPC	Total Plant Cost
tpd	tons per day
TRIG	Transport Gasifier
US, USA	United States of America
vol%	Percentage by Volume
WT	Waste Treatment

Southern Research Institute (SRI) High Temperature POX Steam Reforming Process for Hydrogen-Rich Syngas Production - FT CTL Application

Preliminary Cost Estimation and Techno-Economic Analysis, Based on SRI's Aspen Process Simulation and DOE's Reference FT CTL Design Cost Database

Submitted to

Southern Research Institute (SRI)

July 1, 2015

By

Nexant

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Section 1 Introduction

1.1 BACKGROUND

With funding from DOE "Advanced Gasification Technologies Development and Gasification Scoping Studies for Innovative Initiatives, Area of Interest 3, 'High-Hydrogen Syngas Production" program, Southern Research Institute (SRI) is developing a high temperature POX steam reforming process that can withstand the severe contaminant conditions of a near-raw syngas derived from low-rank coal gasification with the objective to increase the H_2-to-CO ratio and eliminate methane, ammonia and tars from the process. SRI is developing the technology for both IGCC and coal-to-liquid applications. In addition to catalyst development, SRI is also carried out Aspen simulation of the overall process. Nexant was asked to assist with preliminary cost estimation and techno-economic analysis (TEA) of the process, based on SRI's simulation results. The TEA is to be carried out based on DOE NETL's methodology.

1.2 STUDY OBJECTIVES

The objective of this TEA study is to assess the cost/performance of integrating the POX steam reforming process offered by SRI with a TRIG gasifier. The POX reformer will treat raw syngas exiting the gasifier and generate a hydrogen-rich, tar-free syngas with minimal methane content. This will enable near-zero emissions from coal gasification for both power and coal-to-liquids (CTL) productions with a carbon capture rate of more than 90%.

The current report presents the results of the cost/performance analysis for the Fischer Tropsch (FT) CTL plant, utilizing KBR's Transport Gasifier (TRIG) coupled with SRI's POX reformer. It was carried out, to the maximum extent possible, in accordance with:

- The guidelines as set forth in the Attachment 2 of the DE-FOA-0000784 document ("Design Basis for Techno-economic Analyses Deliverables")
- The FT plant balance and cost data available in the "*Cost and Performance Baseline for Fossil Energy Plants, Volume 4: Bituminous Coal to Liquid via Fischer-Tropsch Synthesis*, May 12, 2014, DOE/NETL 2011/1477" The TRIG gasifier plant balance and cost data available in the TRIG IGCC Reference Case design, per "*Cost and Performance Baseline for Fossil Energy Plants, Volume 3a: Low Rank Coal to Electricity*, May 2011, DOE/NETL. 2010/1399" (NETL Report 1399), and
- Aspen simulation results provided by SRI for both the FT CTL designs with and without its high-temperature, contaminant-resistant, POX reforming unit.

Section 2 FT CTL Design Basis

2.1 DESIGN REFERENCES

The process design references used for this study follow the recommended reference studies set forth by Attachment 2 of the FOA. These are namely:

- "*Cost and Performance Baseline for Fossil Energy Plants, Volume 3a: Low Rank Coal to Electricity*, May 2011, DOE/NETL. 2010/1399" (NETL Report 1399)
- "*Cost and Performance Baseline for Fossil Energy Plants, Volume 4: Bituminous Coal to Liquid via Fischer-Tropsch Synthesis*, May 12, 2014, DOE/NETL 2011/1477"
- NETL's Series of Quality Guidelines for Energy Systems Studies (QGESS):
  - "*Specifications for Selected Feedstocks*, January 2012, DOE/NETL. 341/011812"
  - "*Process Modeling Design Parameters*, January 2012, DOE/NETL. 341/081911"
  - "*CO_2 Impurity Design Parameters*, January 2012, DOE/NETL. 341/011212"
  - "*Detailed Coal Specifications*, January 2012, DOE/NETL-401/01211"
  - "*Cost Estimation Methodology for NETL Assessments of Power Plant Performance*, April 2011, DOE/NETL. 2011/1455"
  - "*Capital Cost Scaling Methodology*, January 2013, DOE/NETL. 341/013113"
  - "*Fuel Prices for Selected Feedstocks in NETL Studies*, November 2012, DOE/NETL 341/11212"
- "*Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases*, August 2012, DOE/NETL-341/082312" (NETL Report 341/082312), was used as the reference to develop the updated capital and operating cost estimates, in June 2011 dollars.

2.2 CASE CONFIGURATIONS

Two CTL plant configurations utilizing KBR TRIG gasifiers are evaluated to assess the feasibility and the cost/performance of integrating SRI's POX reformer into a TRIG-based FT CTL plant with CO_2 capture. The partial oxidation (POX) configuration which consists of the TRIG gasifier coupled with SRI's POX reformer is compared against the base configuration which does not include the POX unit.

The two FT CTL configurations are identified in the case study matrix shown in Table 2-1.

Table 2-1
Case Study Matrix for FT CTL with CO_2 Capture

	Base Case [1]	POX Case
Gasification Technology		
TRIG Gasifier	✓	✓
SRI POX Reformer		✓
Gas Cleanup		
Two-Stage Rectisol for CO_2 and Sulfur Removal[2]	✓	✓
Water Gas Shift	None	None
FT Synthesis, Product Upgrading and Amine CO_2 Removal	✓	✓
GE MS6001B Gas Turbine	✓	✓
CO_2 Drying and Compression (to 2,200 psig)	✓	✓

[1] Base case based on SRI's simulation of the KBR TRG gasifier FT CTL plant
[2] Rectisol removes H_2S and CO_2. Additional trace contaminant clean up technologies is included as defined by DOE/NETL baseline studies

2.2.1 Base Case: TRIG FT CTL Plant with Rectisol-Based AGR

The Base Case FT CTL plant is designed to produce a nominal 50,000 BPD of FT diesel and naphtha. It is consisted of a PRB coal fed TRIG gasification unit, a FT liquid fuels production unit and a CO_2 capture unit as shown in the simplified Block Flow Diagram (BFD) in Figure 2-1. It is assumed to operate with an annual on-stream factor of 90 percent.

The plant configuration is based on the DOE/NETL Report 1477 configuration with the exception of the Shell gasifier and the WGS units. The Illinois No. 6 coal fed Shell gasifier is replaced by the PRB coal fed TRIG gasifier. The WGS is eliminated. The configuration of the Fischer Tropsch synthesis, upgrading and fuel gas utilization units are identical to that of DOE Report 1477.

The Base Case FT CTL plant gasification section consists of the following units:

- Coal Handling
- Coal Prep, Drying & Feed
- Air Separation Unit (ASU)
- TRIG Gasifier System
- High Temperature Gas Cooling and Steam Generation
- Gas Cleaning (Syngas Scrubbing , Particulate Filters, and Hg Removal)
- Rectisol AGR for H_2S and CO_2 Removal
- CO_2 Compression and Purification Facilities
- Sour Water Stripping
- Sulfur Recovery and Tail Gas Treating

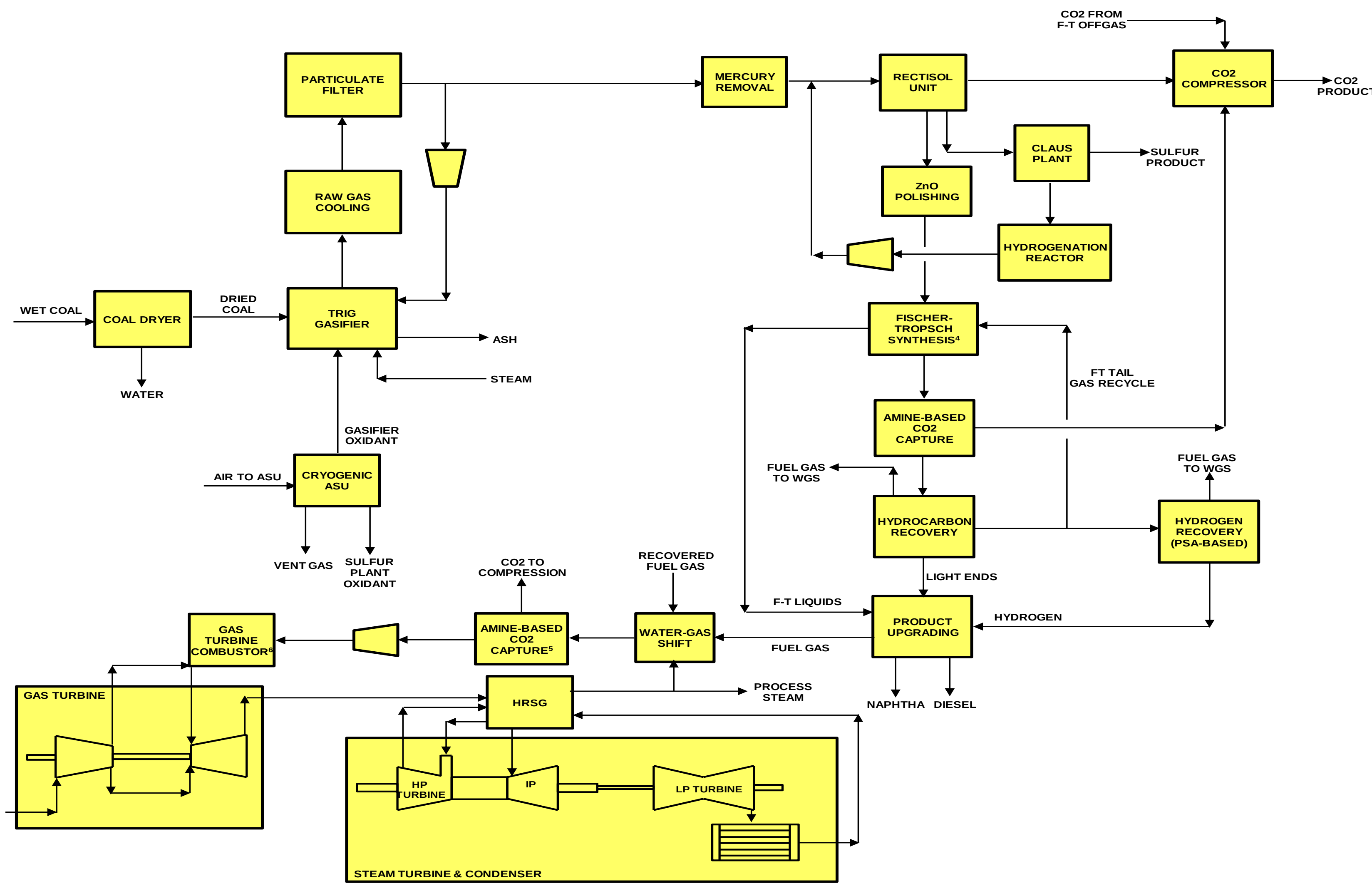
The Base Case FT CLT plant synthesis and product upgrading sections are consisted of the following major blocks:

- Fischer-Tropsch synthesis reactor
- Hydrocarbon recovery
- Product upgrading
- Amine CO_2 removal
- WGS
- Hydrogen recovery (PSA)

The steam and power generation sections are consisted of the following blocks:

- Steam system
- Combustion turbine power generation (CTG)
- HRSG, ducting and stack
- Steam turbine power generation (STG)

Figure 2-1
Base Case: TRIG Gasifier FT CTL Plant with Rectisol-Based AGR - Simplified BFD



2.2.2 POX Case: Coupled TRIG/SRI POX Reformer with Rectisol-Based AGR

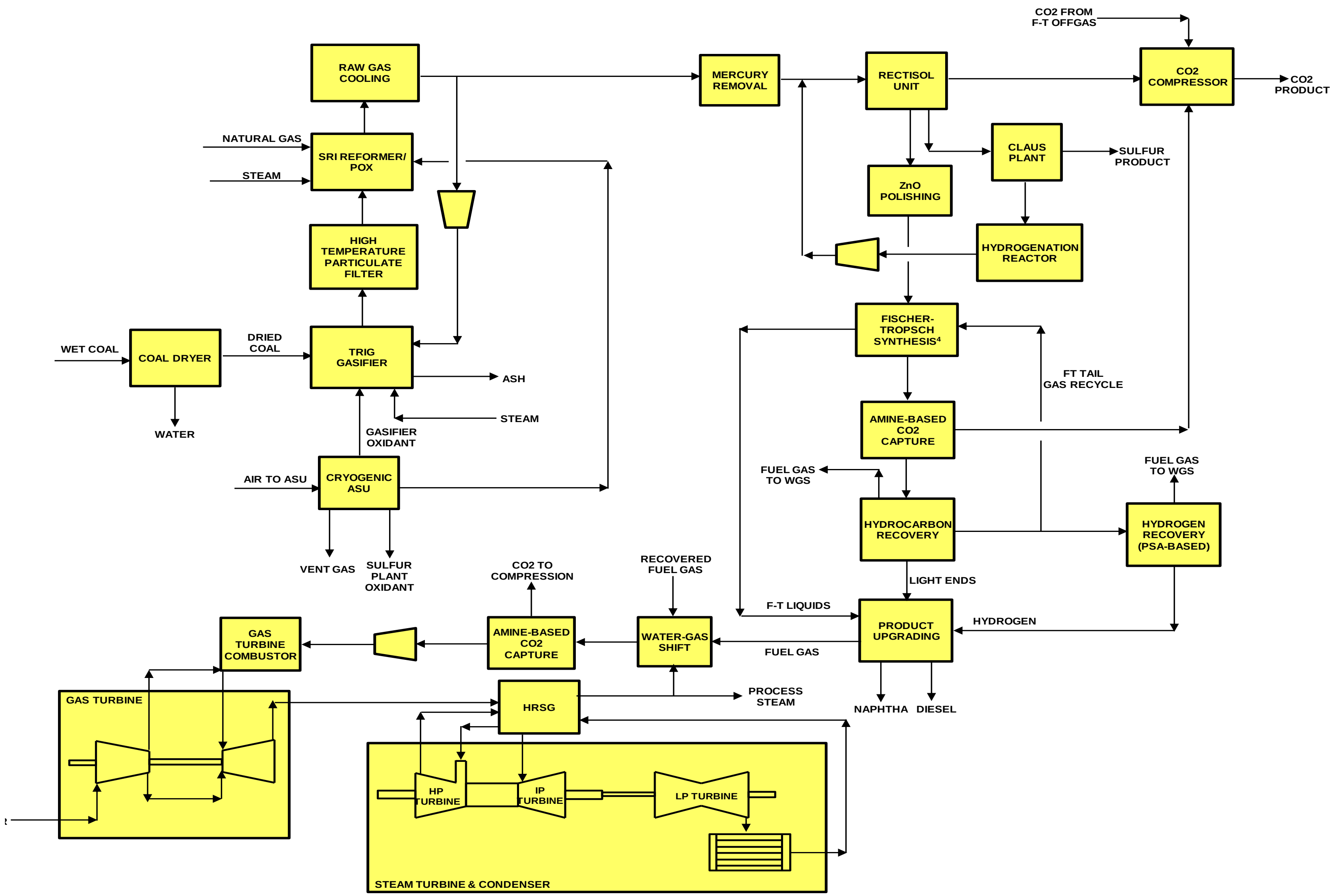
The POX Case is a preliminary conceptual design that couples the SRI POX reformer with the TRIG gasification process to produce syngas for the 50,000 BPD FT CTL plant.

Similar to the Base Case, the two-stage Rectisol process is used for AGR. However, as the POX reformer is able to convert practically all the tar and most of the methane in the raw syngas to CO and H_2, methane concentration in the FT feed gas is minimized.

The simplified BFD for the POX Case FT CTL plant is shown in Figure 2-2. It is assumed to operate with an annual on-stream factor of 90 percent.

The SRI POX block in Figure 2-2 is the differentiating technology of interest in this study. All other blocks are expected to have the same unit operations as those in the Base Case, but with different throughputs due to differing flow quantities. Costs for these systems will be scaled based on system capacities using factors given in the QGESS *Capital Cost Scaling Methodology* document wherever possible.

Figure 2-2
POX Case: TRIG Gasifier w/SRI POX FT CTL Plant - Simplified BFD



2.3 PROCESS DESIGN CRITERIA

SRI carried out a simulation of the Base and the POX FT CTL cases on ASPEN Plus to obtain the process heat and material balances (HMB). The results from SRI's simulations were then provided to Nexant. The HMB were used to estimate the overall FT CTL plant performance and cost. Cost estimation methodology is described in Section 2.4 below utilizing TRIG gasifier reference costs from DOE/NETL report 1399 and the FT/AGR/balance of plant reference costs from DOE/NETL report 1477.

2.4 CAPITAL COST ESTIMATION METHODOLOGY

2.4.1 General

For FT CTL plants with CO_2 capture, the DOE/NETL 1477 Report provided a code of accounts grouped into 14 major systems. Each of these major systems is broken down further into different subsystems. This type of code-of-accounts structure has the advantage of grouping all reasonably allocable components of a system or process into a specific system account.

The FT CTL plant in the DOE/NETL 1477 Report is based on producing syngas for the FT synthesis unit using Illinois No. 6 coal fed Shell gasifiers. Since the SRI process is based on PRB coal fed TRIG gasifiers for syngas production, the costs associated with the TRIG gasifier and its feed crushing and drying subsystems are drawn from the cost accounts of the PRB coal based TRIG gasifiers case in the DOE/NETL Report 1399.

The capital cost scaling follows the guidelines and parameters that are described in the NETL *Capital Cost Scaling Methodology* document was used to perform the cost estimation for systems that are not related to the SRI POX reformer. In general, this cost estimation methodology involves determining the scaling parameters, exponents and coefficients from the *Capital Cost Scaling Methodology*, as well as the reference costs and baseline capacities from DOE/NETL Reports 1399 and 1477. Once these have been established, the capital cost can be estimated based on the corresponding system capacities from the HMB defined by SRI's ASPEN models of the FT CTL plants.

For the SRI POX reformer cost, Nexant performed a capacity factored bottoms-up, major-equipment cost based on historical data for autothermal reformers with SRI providing some guidance on their catalyst and reactor requirements. Installed cost is factored from estimated equipment using in-house installation factors for similar type of equipment.

2.4.2 FT CTL Plant Capital Cost Estimate Criteria

The capital cost estimates for the FT CTL plant systems that are independent of the TRIG gasifier and the SRI POX reformer, mainly the FT synthesis block (systems under Account 5AA in Table 2-2 below) and the supporting utility and offsite systems (the GTCC and CW systems under Accounts 6, 7, 8, and 9 in Table 2-2 below), were factored from information in DOE/NETL Report 1477 on the FT CTL plant with CO_2 capture. While the SRI Aspen model included their own FT block simulation, there are no internal process details or equipment designs to allow cost estimation for the block. Since SRI POX is the technology focus of this study, it was decided to use the FT design and cost from DOE Report 1477. By keeping the total H_2 + CO molar flow (roughly 138,600 lbmoles/hr) to the FT synthesis block identical to that used in DOE Report 1477 for generating 50,000 BPD of FT liquid, the FT block capital cost for the SRI Base Case and POX Case were assumed to be the same as that in Report 1477. FT block cost impact to the relative COP comparison between the two SRI cases is therefore eliminated or at least minimized.

For the supporting system cost estimates, GT fuel (FT tailgas) for each case is assumed to be the same as that in Report 1477 except for FT syngas feed methane adjustment. For 50,000 BPD liquid productions using DOE once-through FT process with CO_2 capture, Report 1477 shows a FT block syngas feed containing 43 lbmoles/hr of methane, and a FT tailgas containing 3,077 lbmoles/hr of methane. It is assumed that the DOE once-through FT tailgas will include 3,077 – 43 = 3,034 lbmoles/hr of generated methane, in addition to what is in the syngas feed. For SRI Base Case, syngas feed (after scaled to 138,600 lbmoles/hr of H_2 + CO) contains roughly 11,419 lbmoles/hr of methane so its FT tailgas will contain 14,453 lbmoles/hr of methane. For SRI POX Case, syngas feed (also scaled to 138,600 lbmoles/hr of H_2 + CO) contains roughly 359 lbmoles/hr of methane so its tailgas will contain 3,393 lbmoles/hr of methane. The GE MS6001B GT is re-rated for these adjusted tailgas flows to estimate the corresponding GT output to be used for capacity factor costing of the GT output dependent systems.

For factoring the steam turbine dependent systems, it is not possible to calculate the STG performance directly from Report 1477 since overall plant steam balances was not included in Report 1477. The only GTCC information in Report 1477 is the GT output (about 112 MWe) and the STG output (about 316 MWe). A reference case GE MS6001B GTCC calculation was carried out using Report 1477 FT tailgas coupled with Nexant in-house overall steam balance data for Illinois #6 coal fed Shell gasifier with iron catalyst based slurry bed FT technology for liquid fuels production,. Nexant in-house FT reactor steam generation and FT LP steam consumption were arbitrarily varied until the reference GTCC output matches the Report 1477 GT and STG outputs. This reference case calculated FT steam generation/consumption data is then used for GTCC performance calculations for all SRI cases. Because the SRI's process design did not provide steam generation information, reference case process steam balances (other than FT which is fixed) were pro-rated based on gasifier/POX syngas flows and temperatures, as well as PRB coal sulfur and chloride content for GTCC performance estimation for the two SRI cases. The GTCC performance estimations also identify the scaling factors for the HRSG, the Surface Condenser and the Cooling Water Systems.

The costs associated with the TRIG gasifier and coal handling (crushing, conveying, drying, etc.) were factored from information in the DOE/NETL Report 1399. The costs in the reference DOE/NETL reports (Report 1399 and 1477) were adjusted for differences in unit or plant capacity according to NETL's Guidelines as described in the NETL Capital Cost Scaling Methodology QGESS document. The costs associated with the TRIG gasifier and coal handling (crushing, conveying, drying, etc.) were factored from information in the DOE/NETL Report 1399. The costs in the reference DOE/NETL reports were adjusted for differences in unit or plant capacity according to NETL's Guidelines as described in the NETL Capital Cost Scaling Methodology QGESS document. Table 2-2 shows the code of accounts for the FT CTL plant. These systems are further broken down to include the various subsystems. The scaling parameters for these BOP subsystems, as laid out by the NETL *Capital Cost Scaling Methodology* document, are also shown in this table.

Table 2-2
Code of Accounts for Report FT CTL Plant

Acct No.	Item/Description	Scaling Parameter
1	COAL & SORBENT HANDLING	
1.1	Coal Receive & Unload	Coal Feed Rate
1.2	Coal Stackout & Reclaim	Coal Feed Rate
1.3	Coal Conveyors & Yard Crush	Coal Feed Rate
1.4	Other Coal Handling	Coal Feed Rate
1.9	Coal & Sorbent Handling Foundations	Coal Feed Rate
2	COAL & SORBENT PREP & FEED	
2.1	Coal Crushing & Drying	Coal Feed Rate
2.2	Prepared Coal Storage & Feed	Coal Feed Rate
2.3	Dry Coal Injection System	Coal Feed Rate
2.4	Misc Coal Prep & Feed	Coal Feed Rate
2.9	Coal & Sorbent Feed Foundation	Coal Feed Rate
3	FEEDWATER & MISC. BOP SYSTEMS	
3.1	Feedwater System	BFW (HP only)
3.2	Water Makeup & Pretreating	Raw Water Makeup
3.3	Other Feedwater Subsystems	BFW (HP only)
3.4	Service Water Systems	Raw Water Makeup
3.5	Other Boiler Plant Systems	Raw Water Makeup
3.6	FO Supply Sys and Nat Gas	Coal Feed Rate
3.7	Waste Treatment Equipment	Raw Water Makeup
3.8	Misc Power Plant Equipment	Coal Feed Rate
4	GASIFIER & ACCESSORIES	
4.1	Gasifier, Syngas Cooler & Auxiliaries	Syngas Throughput
4.3	ASU/Oxidant Compression	O_2 Production
4.4	LT Heat Recovery and Fuel Gas Saturation	Syngas Flow
4.6	Other Gasification Equipment	Syngas Flow
4.9	Gasification Foundations	Syngas Flow
5A	GAS CLEANUP & PIPING	
5A.1	Rectisol	Gas Flow to AGR
5A.2	Elemental Sulfur Plant	Sulfur Production
5A.3	Mercury Removal	Estimated Hg Flow
5A.4	Shift Reactors	N/A
5A.5	Blowback Gas Systems	Syngas Throughput
5A.6	Fuel Gas Piping	Fuel Gas Flow
5A.9	HGCU Foundations	Sulfur Production
5AA	FT SYNTHESIS AND PRODUCT UPGRADE	
5AA.1	FT Synthesis	FT Liquid Product Flow
5AA.2	Amine CO2 Absorption	FT Liquid Product Flow
5AA.3	Amine Regeneration Section	FT Liquid Product Flow
5AA.4	Compression	FT Liquid Product Flow
5AA.5	Hydrocarbon Recovery	FT Liquid Product Flow
5AA.6	Hydrogen Recovery	FT Liquid Product Flow
5AA.7	Autothermal Reformer	FT Liquid Product Flow
5AA.8	Naphtha Hydrotreater	FT Liquid Product Flow
5AA.9	Diesel Hydrotreater	FT Liquid Product Flow
5AA.10	Wax Hydrotreater	FT Liquid Product Flow
5AA.11	HP Raw Fuel Gas Compressor	FT Liquid Product Flow
5AA.12	HP Fuel Gas to GT Compressor	FT Liquid Product Flow

Acct No.	Item/Description	Scaling Parameter
5AA.13	WGS Reactor	FT Liquid Product Flow
5AA.14	Amine CO2 Absorption	FT Liquid Product Flow
5AA.15	Amine Regeneration Section	FT Liquid Product Flow
5B	CO_2 REMOVAL & COMPRESSION	
5B.2	CO_2 Compression & Drying	CO_2 Flow
6	COMBUSTION TURBINE/ACCESSORIES	
6.1	Combustion Turbine Generator	Gas Turbine Capacity
6.2	Combustion Turbine Foundations	Gas Turbine Capacity
7	HRSG, DUCTING & STACK	
7.1	Heat Recovery Steam Generator	HRSG Duty
7.3	Ductwork	HRSG Duty
7.4	Stack	HRSG Duty
7.9	HRSG, Duct & Stack Foundations	HRSG Duty
8	STEAM TURBINE GENERATOR	
8.1	Steam TG & Accessories	Turbine Capacity
8.2	Turbine Plant Auxiliaries	Turbine Capacity
8.3a	Condenser & Auxiliaries	Condenser Duty
8.3b	Air Cooled Condenser	Condenser Duty
8.4	Steam Piping	BFW (HP Only)
8.9	TG Foundations	Turbine Capacity
9	COOLING WATER SYSTEM	
9.1	Cooling Towers	Cooling Tower Duty
9.2	Circulating Water Pumps	CW Flow Rate
9.3	Circ. Water System Auxiliaries	CW Flow Rate
9.4	Circ Water Piping	CW Flow Rate
9.5	Makeup Water System	Raw Water Makeup
9.6	Component Cooling Water System	CW Flow Rate
9.9	Circ. Water System Foundations	CW Flow Rate
10	ASH/SPENT SORBENT HANDLING SYS	
10.1	Slag Dewatering & Cooling	Slag Production
10.6	Ash Storage Silos	Slag Production
10.7	Ash Transport & Feed Equipment	Slag Production
10.8	Misc. Ash Handling System	Slag Production
10.9	Ash/Spent Sorbent Foundation	Slag Production
11	ACCESSORY ELECTRIC PLANT	
11.1	Generator Equipment	Turbine Capacity
11.2	Station Service Equipment	Auxiliary Load
11.3	Switchgear & Motor Control	Auxiliary Load
11.4	Conduit & Cable Tray	Auxiliary Load
11.5	Wire & Cable	Auxiliary Load
11.6	Protective Equipment	Auxiliary Load
11.7	Standby Equipment	Total Gross Output
11.8	Main Power Transformers	Total Gross Output
11.9	Electrical Foundations	Total Gross Output
12	INSTRUMENTATION & CONTROL	
12.4	Other Major Component Control	Auxiliary Load
12.6	Control Boards, Panels & Racks	Auxiliary Load

Acct No.	Item/Description	Scaling Parameter
12.7	Computer & Accessories	Auxiliary Load
12.8	Instrument Wiring & Tubing	Auxiliary Load
12.9	Other I & C Equipment	Auxiliary Load
13	IMPROVEMENT TO SITE	
13.1	Site Preparation	Accounts 1-12
13.2	Site Improvements	Accounts 1-12
13.3	Site Facilities	Accounts 1-12
14	BUILDING & STRUCTURES	
14.1	Combustion Turbine Area	Gas Turbine Power
14.2	Steam Turbine Building	Accounts 1-12
14.3	Administration Building	Accounts 1-12
14.4	Circulation Water Pumphouse	CW Flow Rate
14.5	Water Treatment Buildings	Raw Water Makeup
14.6	Machine Shop	Accounts 1-12
14.7	Warehouse	Accounts 1-12
14.8	Other Buildings & Structures	Accounts 1-12
14.9	Waste Treating Building & Structures	Raw Water Makeup

2.4.3 Home Office, Engineering Fees and Project/Process Contingencies

Engineering and Construction Management Fees and Home Office cost, project and process contingencies were factored from the each subsystem's TFC. These were then added to the TFC to come up with the total project cost (TPC) of the system. Factors from the DOE/NETL 1477 Baseline Report were used.

2.4.4 Owner's Cost

Owner's cost was then added to TPC to come up with the total overnight cost (TOC) for the system. Owner's costs as defined in the DOE/NETL 1477 Report include the following:

- Preproduction Costs –
  - 6 months of all labor cost
  - 1 month of maintenance materials
  - 1 month of non-fuel consumables
  - 1 month of waste disposal
  - 25% of 1 month fuel cost at 100% capacity factor
  - 2% TPC

- Inventory Capital -
  - 60 day supply of fuel and consumable at 100% CF
  - 0.5% TPC

- Initial Cost for Catalyst and Chemicals per design
- Land Cost = $900,000 at 300 acres x $3,000/acre
- Other Owner's Costs at 15% TPC

- Financing Costs at 2.7% TPC

2.5 OPERATION & MAINTENANCE COSTS

The operation and maintenance (O&M) costs pertain to those charges associated with operating and maintaining the power plants over their expected life. These costs include:

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

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 power generation. Variable O&M costs were estimated based on 90% capacity factor.

2.5.1 Fixed Costs

Operating labor cost was determined based on the number of operators required to work in the plant. Other assumptions used in calculating the total fixed cost include:

• 2011 Base hourly labor rate, $/hr	$39.7
• Length of work-week, hrs	50
• Labor burden, %	30
• Administrative/Support labor, % O&M Labor	25
• Maintenance material + labor, % TPC	2.8
• Maintenance labor only, % maintenance material + labor	35
• Property Taxes and insurances, % TPC	2

2.5.2 Variable Costs

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

The unit costs for major consumables and waste disposal was selected from DOE/NETL 1477 Baseline Report, QGESS *Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases* and from the QGESS *Fuel Prices for Selected Feedstocks in NETL Studies* document.

The 2011 coal price as delivered to the Montana FT CTL plant is $19.63/ton, per the QGESS *Fuel Prices for Selected Feedstocks in NETL Studies* document.

2.5.3 CO_2 Transport and Storage Costs

As specified in DE-FOA-0000784 Attachment 2, CO_2 Transport and Storage (T&S) costs used for the Montana FT CTL plant location is $22/tonne. Per the TEA reporting requirements, the COPs are reported both with and without the cost of CO_2 T&S.

2.6 FINANCIAL MODELING BASIS

2.6.1 Cost of Production

The key measure to evaluate overall economic financial viability of the FT CTL plant is the estimation of the crude oil equivalent required selling price (RSP) of the Fischer-Tropsch liquid products. The RSP is the minimum price at which the products must be sold to recover the annual revenue requirement (ARR) of the plant. The ARR is the annual revenue needed to pay the operating costs, service the debt, and provide the expected rate of return for the investors. The FT CTL project is considered economic viable if the market price of the product is equal to or above the calculated RSP.

$$\text{RSP} = \frac{\begin{array}{c}\text{first year}\\ \text{capital charge}\end{array} + \begin{array}{c}\text{first year}\\ \text{fixed operating}\\ \text{cost}\end{array} + \begin{array}{c}\text{fist year}\\ \text{variable operating}\\ \text{cost}\end{array}}{\text{annual net Fischer Tropsch Liquid production}}$$

$$\text{RSP} = \frac{(\text{CCF})(\text{TOC}) + \text{OC}_{\text{fix}} + (\text{CF})\,(\text{OC}_{\text{var}})}{\text{annual net Fischer Tropsch Liquid production}}$$

The ARR is the sum of fuel cost, variable operating cost, fixed operating cost, and annual capital component minus the by-product credits for electric power sale revenues. The annual capital component of the ARR is determined as the product of the total overnight cost (TOC) and the capital charge factor (CCF). The CCF for evaluating the RSP is determined from NETL Power Systems Financial Model (PSFM). Commercial fuels project financial structure is best suited for the FT CTL plant CCF calculation. The estimated CCF using commercial fuels project financial structures is shown in Table 2-3. The capital charge factor of 0.218 will be used for estimating for the RSP financial analysis.

Table 2-3
Financial Parameters for CCF Estimate

Scenario	Commercial Fuels
Percent Debt	50%
Percent Equity	50%
Debt Interest Rate	8.00%
Internal Rate of Return on Equity (IRROE)	20%
After Tax Weighted Cost of Capital	12.48%
Capital Charge Factor (CCF)	0.218

The economic assumptions and finance structure/capital expenditure period is defined under High-Risk Fuels Projects in report DOE/NETL-2011/1489 September 29, 2011, (revision from DOE/NETL-401/090808) "*Recommended Project Finance Structure for the Economic Analysis of Fossil-Based Energy Project*". Listed below are the financial parameters and assumptions for the PSFM model:

- Income tax rate, % — 38
- Equity desired rate of return, % — 20
- Type of debt financing — Non-Recourse
- Repayment term of debt, years — 30
- Debt repayment grace period, years — 0
- Debt reserve fund — None
- Depreciation — 20 years, 150% declining balance
- Working capital — None
- Plant operational life, years — 30
- Plant economic life, years — 35
- Tax holiday, years — 0
- EPC escalation, % per year — 3.6
- COP (revenue) nominal escalation, % — 3.0
- Coal price nominal escalation, % — 3.0
- O&M cost nominal escalation, % — 3.0
- Duration of construction, years — 5
- First year of construction — 2011
- Construction cost distribution, %
    - Year 1 — 10%
    - Year 2 — 30%
    - Year 3 — 25%
    - Year 4 — 20%
    - Year 5 — 15%

All costs are expressed in the "first-year-of-construction" year dollars, and the resulting RSP is also expressed in "first-year-of-construction" year dollars.

The conceptual plants produce three products for sale. Those products are: (1) FT diesel fuel, (2) FT naphtha, and (3) electric power. All light gases including LPG are used within the plant. FT naphtha, although it has a similar boiling range to gasoline, has not traditionally been considered to be suitable for refining into high octane gasoline because of its high paraffinic nature. This analysis assumes that the naphtha can be sold at a discounted price compared to the diesel fuel. To express the RSP in terms of equivalent crude oil price, historically, the ratio of the price of crude oil: ultra-low sulfur diesel is 1.25 and naphtha: diesel is 0.7. The discount price is assumed to be 0.7692 (1/1.3) the value of the diesel fuel. The relative value is used to determine the equivalent diesel fuel yield from the CTL plant in terms of barrels per year.

The petroleum equivalent diesel price is calculated by taking the first year of production for diesel in $/Bbl and multiplying this value by the ratio of the lower heating values of FT diesel and petroleum diesel.

$$Petroleum\ Equivalent\ Diesel\ Price = \left(\frac{Petroleum\ Diesel\ LHV}{FT\ Diesel\ LHV}\right) * FY\ COP\ FT\ Diesel$$

- The equivalent crude oil price is then calculated by multiplying the petroleum equivalent diesel price by a factor of 0.80.

- RSP Equivalent Crude Oil = 0.80 x Petroleum Equivalent Diesel Price.

- The factor of 0.80 was calculated from data of historic spot prices provided by the EIA from June 2009 through November 2013 for various fuel types. This data was used to develop correlations between the various fuel prices and the WTI crude oil price (Crude oil: Ultra-low sulfur diesel is 1.25 and Naphtha: Diesel is 0.70). The ECO price is the minimum market price for crude oil at which the first-year RSPs will be met.

Sensitivity analyses of FT liquids products required selling price (RSP) will be performed on the following parameters:

- Capital cost of advanced technology (SRI POX cost)
- Fuel prices (Coal price)
- Sales of CO_2 at plant gate prices of $0-60/tonne
- Cost of CO_2 emissions of $0-60/tonne
- Power price for net imports/exports at $60/MWh

2.6.2 CO_2 Sales Price

As outlined in the TEA's reporting requirements, sensitivity analysis is to be done to determine the impact of CO_2 sales on FT CTL COP. The varying parameter is the CO_2 sales price at the FT CTL plant gate and is to range between $0/tonne (baseline case assuming no value to the product CO_2) and $60/tonne.

Per the reporting requirements for the TEA, the cost of capturing CO_2 shall be reported, if a reference non-capture plant is available. Since the scope of work did not specify the modeling of an analogous case without capture, CO_2 avoided cost analysis for like technology reference will not be performed.

2.6.3 Cost of CO_2 Emissions

The TEA also requires sensitivity analysis on cost of CO_2 emissions to be performed. The varying parameter is the CO_2 emissions cost. The range of the emissions cost is between $0/tonne (baseline case assuming no CO_2 emissions cost) and $60/tonne.

Because CO_2 emissions from the various FT product upgrading furnaces are not available, sensitivity to CO_2 emission cost will be based on estimated emissions from the GTCC power plant. Using only the GTCC emission should provide good approximation on the relative difference between the Base Case and the POX Case emission sensitivities since emissions from the FT product upgrading furnaces, as well as other process furnace vents, should be almost the same between the two Cases.

Section 3 Base Case: TRIG FT CTL with Rectisol-Based CO_2 Capture

3.1 PROCESS OVERVIEW

The Base Case FT CTL plant is a Montana PRB coal fired TRIG gasifier based FT CTL plant designed to produce 50,000 BPD of FT diesel and naphtha. The tail gas from FT synthesis block is used as fuel for the GTG. The gas turbine generator (GTG) used is a GE MS6001B class turbine with a nominal ISO gross GT output of approximately 42 MWe. The GTCC power plant is equipped with a heat recovery steam generator (HRSG) and steam turbines to maximize power recovery.

The syngas exiting the TRIG gasifier is at 1,800 ^oF and has a H_2/CO ratio of 0.8. The gasifier syngas is cooled and cleaned in particulate filters and water scrubbed before it is sent for mercury removal. The cleaned syngas is then processed in the Rectisol acid gas removal (AGR) unit to remove H_2S and CO_2 from the syngas. The treated syngas is fed to the Fischer-Tropsch synthesis section with essentially all of the sulfur and contaminants removed from the syngas. Additional CO_2 removal is required from the FT synthesis product gas and the FT product upgrading tail gas to meet the less than 10% carbon emission requirement. Recovered CO_2 is purified and compressed to sequestration pressure of 2,200 psig at the plant battery limit. H_2S recovered from the Rectisol unit is converted into elemental sulfur in the Claus plant.

The nominal net Base Case FT CTL power export capacity after accounting for the auxiliary loads which include CO_2 capture and compression is approximately 350 MWe.

The cost/performance prototype for the FT synthesis/upgrading section is based on DOE/NETL Report 1477 and has the following key design parameters:

- The FT feed gas H_2+CO flow rate is 137,600 lbmoles/hr and has a H_2/CO ratio of 0.7
- The FT upgrading process is configured to provide 50,000 BPD of FT liquid fuels with a 70:30 yield of FT diesel to FT naphtha
- CO_2 is removed by amine absorption from the FT synthesis product gas
  CO_2 is removed by WGS/amine absorption from the FT product upgrading unit tail gas.
- Part of the FT product upgrading unit tail gas is sent to the PSA unit for hydrogen recovery for hydrotreating.

For cost estimating purpose, the FT feed H_2+CO flow rate from the SRI ASPEN HMB is normalized to the DOE/NETL Report 1477 FT feed H_2+CO flow rate for producing 50,000 BPD of FT liquid fuels with a 70:30 yield of diesel: naphtha.

3.2 PERFORMANCE RESULTS

The SRI-modelled Base Case FT CTL plant with CO_2 capture consumes 35,900 TPD of PRB coal at the Montana site to produce 50,000 BPD of FT diesel and naphtha and produces a net output of 350 MWe. Estimated overall performance for the Base Case FT CTL plant based on SRI's simulation is summarized in Table 3-1.

Table 3-1
SRI Base Case TRIG FT CTL Performance

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	SRI Base Case FT Synthesis,TRIG PRB Coal
Gas Turbine Power	459,354
Steam Turbine Power	454,424
TOTAL POWER, kWe	**913,778**
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling	3,004
Coal Milling	13,928
Slag Handling	3,383
Air Separation Unit Auxiliaries	4,780
Air Separation Unit Main Air Compressor	281,370
Oxygen Compressor	42,430
Nitrogen Compressors	6,932
CO_2 Compressor	62,128
Boiler Feedwater Pumps	21,647
Condensate Pump	585
Quench Water Pump	0
Syngas Recycle Compressor	4,253
Circulating Water Pump	8,436
Ground Water Pumps	199
Cooling Tower Fans	4,416
Scrubber Pumps	1,181
Acid Gas Removal	24,952
Gas Turbine Auxiliaries	16,464
Steam Turbine Auxiliaries	576
Claus Plant/TGTU Auxiliaries	507
Claus Plant TG Recycle Compressor	426
FT Power Requirement	31,951
Miscellaneous Balance of Plant	20,483
Transformer Losses	5,390
TOTAL AUXILIARIES, kWe	**559,422**
NET POWER, kWe	**354,356**
CONSUMABLES	
As-Received Coal Feed, lb/hr	2,987,983
Condenser Duty, MMBtu/hr	2,575
Raw Water Withdrawal, gpm	2,162

3.3 EQUIPMENT LIST

As the Base Case TRIG FT CTL is based on the DOE/NETL Reports 1399 (TRIG gasifier) and 1477 (FT synthesis and upgrade), the reader should refer to the FT CTL equipment list in the DOE/NETL 1399 and 1477 reports.

3.4 CAPITAL COST

For cost estimating purpose, the FT feed H_2+CO flow rate from the SRI ASPEN HMB is normalized to the DOE/NETL Report 1477 FT feed H_2+CO flow rate for producing 50,000 BPD of FT liquid fuels with a 70:30 yield of diesel to naphtha. The coal rate is then adjusted to correspond to the normalized FT feed H_2+CO rate. The adjustment is shown in Table 3-2:

Table 3-2
Base Case TRIG FT CTL H_2+CO Adjustments

		DOE/NETL 1477	SRI Base Case FT CTL ASPEN Simulation	Adjusted for Cost Estimated
AR Coal	T/D	21,006	35,757	35,856
Coal Type		Illinis No. 6	PRB	PRB
FT Synthesis Feed				
H2+CO	lbmols/hr	137,598	137,220	137,598
H2/CO				
FT Liquid Products				
Diesel	BPD	35,234		35,234
Naphtha	BPD	14,762		14,762
Total	BPD	49,996		49,996
Vol% Diesel	%	70%		70%
Vol% Naphtha	%	30%		30%
Total	%	100%		100%

Table 3-3 shows the cost breakdown of the Base Case TRIG FT CTL with Rectisol-based AGR, consistent with the Code of Accounts format as expressed in the DOE/NETL 1477 report.

Table 3-4 shows the calculation and addition of owner's costs to determine the TOC, used to calculate COP.

Table 3-3
SRI Base Case TRIG FT CTL Total Plant Cost Summary

SRI FT Base Case With CO2 Sequestration Total Plant Cost Details (June 2011 Cost Basis)														
	Coal Type	PRB		Coal Feed (AR) =	2,987,983	lbs/hr		FT Product Slate =	50,000	Bbl/Day Total Liquid		100%		
	Gasifier	TRIG			35,856	tons/day			35,234	Bbl/Day Diesel		70%		
				Coal HHV (AR) =	8,564	Btu/lb			14,762	Bbl/Day Naptha		30%		
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	$	$/BPD	$/BPD$_{ECO}$	$/BPD$_{EPD}$
1	COAL & SORBENT HANDLING	$53,381	$9,360	$40,828	$0	$0	$103,569	$9,179	$0	$22,551	$135,299	$3,591	$3,302	$4,128
2	COAL & SORBENT PREP & FEED	$277,781	$22,546	$47,040	$0	$0	$347,367	$31,822	$0	$78,410	$457,599	$12,145	$11,168	$13,960
3	FEEDWATER & MISC BOP SYSTEMS	$6,625	$4,240	$6,619	$0	$0	$17,484	$1,624	$0	$4,630	$23,738	$630	$579	$724
4	GASIFIER & ACCESSORIES													
4.1	Gasifier, Syngas Cooler & Auxiliaries (TRIG)	$293,469	$0	$126,093	$0	$0	$419,562	$37,462	$96,734	$84,850	$638,608	$16,950	$15,586	$19,482
4.2	SRI POX	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.3	ASU/Oxidant Compression	$285,347	$0	$0	$0	$0	$285,347	$27,659	$0	$31,301	$344,308	$9,138	$8,403	$10,504
4.4	LT Heat Recovery & FG Saturation	$21,197	$0	$8,003	$0	$0	$29,199	$2,850	$0	$6,410	$38,460	$1,021	$939	$1,173
4.X	Other Gasification Equipment	$0	$22,674	$13,073	$0	$0	$35,747	$3,287	$0	$9,578	$48,613	$1,290	$1,186	$1,483
	SUBTOTAL 4.	$600,013	$22,674	$147,168	$0	$0	$769,856	$71,259	$96,734	$132,140	$1,069,989	$28,399	$26,114	$32,643
5A	GAS CLEANUP & PIPING	$427,406	$13,467	$355,582	$0	$0	$796,455	$75,053	$143,052	$203,123	$1,217,683	$32,319	$29,719	$37,149
5AA	FT SYNTHESIS AND PRODUCT UPGRADE													
5AA.1	FT Synthesis	$220,390	$0	$0	$0	$0	$220,390	$21,157	$59,505	$75,264	$376,316	$9,988	$9,184	$11,480
5AA.X	Hydrocarbon Upgrading	$254,952	$0	$0	$0	$0	$254,952	$24,400	$61,371	$83,669	$424,392	$11,264	$10,358	$12,947
5AA.Y	Amine CO2 Removal	$94,742	$0	$0	$0	$0	$94,742	$9,095	$25,580	$32,356	$161,773	$4,294	$3,948	$4,935
	SUBTOTAL 5AA.	$570,083	$0	$0	$0	$0	$570,083	$54,653	$146,456	$191,288	$962,481	$25,546	$23,490	$29,363
5B.2	CO2 REMOVAL & COMPRESSION	$70,092	$0	$23,764	$0	$0	$93,856	$8,752	$0	$20,523	$123,131	$3,268	$3,005	$3,756
6	COMBUSTION TURBINE/ACCESSORIES	$50,897	$423	$4,096	$0	$0	$55,416	$11,996	$12,788	$15,949	$96,149	$2,552	$2,347	$2,933
7	HRSG, DUCTING & STACK													
7.1	Heat Recovery Steam Generator	$62,999	$0	$12,200	$0	$0	$75,198	$6,974	$0	$8,217	$90,390	$2,399	$2,206	$2,758
7.X	Ductwork & Stack	$10,283	$7,286	$9,540	$0	$0	$27,108	$2,457	$0	$4,800	$34,365	$912	$839	$1,048
	SUBTOTAL 7.	$73,281	$7,286	$21,739	$0	$0	$102,307	$9,431	$0	$13,017	$124,754	$3,311	$3,045	$3,806
8	STEAM TURBINE GENERATOR	$74,560	$1,643	$21,685	$0	$0	$97,889	$8,457	$0	$15,190	$121,536	$3,226	$2,966	$3,708
9	COOLING WATER SYSTEM	$11,504	$15,280	$13,064	$0	$0	$39,848	$3,572	$0	$9,197	$52,616	$1,397	$1,284	$1,605
10	ASH/SPENT SORBENT HANDLING SYS	$70,479	$4,735	$34,677	$0	$0	$109,891	$10,272	$0	$12,994	$133,157	$3,534	$3,250	$4,062
11	ACCESSORY ELECTRIC PLANT	$50,087	$25,002	$43,938	$0	$0	$119,028	$10,019	$0	$25,200	$154,246	$4,094	$3,765	$4,706
12	INSTRUMENTATION & CONTROL	$14,998	$3,040	$9,861	$0	$0	$27,899	$2,483	$1,395	$5,329	$37,106	$985	$906	$1,132
13	IMPROVEMENTS TO SITE	$8,369	$4,815	$21,664	$0	$0	$34,847	$3,450	$0	$11,490	$49,788	$1,321	$1,215	$1,519
14	BUILDINGS & STRUCTURES	$0	$14,300	$15,570	$0	$0	$29,870	$2,656	$0	$5,211	$37,736	$1,002	$921	$1,151
	Total Cost	$2,359,558	$148,812	$807,295	$0	$0	$3,315,665	$314,676	$400,426	$766,241	$4,797,008	$127,320	$117,076	$146,345

Table 3-4
SRI Base Case TRIG FT CTL Total Plant Cost Summary

Owner's Costs		$1,000	$/Bbl FT Liquids
Preproduction Costs			
6 months All Labor		$47,109	$1,250
1 Month Maintenance Materials		$7,986	$212
1 Month Non-Fuel Consumables		$1,596	$42
1 Month Waste Disposal		$2,631	$70
25% of 1 Months Fuel Cost at 100% CF		$5,352	$142
2% of TPC		$95,940	$2,546
Total		$160,614	$4,263
Inventory Capital			
60 day supply of fuel at 100% CF		$0	$0
60 day supply of non-fuel consumables at 100% CF		$305	$8
0.5% of TPC (spare parts)		$23,985	$637
Total		$24,290	$645
Owner's Cost			
Initial Cost for Catalyst and Chemicals		$26,969	$716
Land		$900	$24
Other Owner's Cost		$719,551	$19,098
Financing Costs		$129,519	$3,438
Total		$876,939	$17,539
Total Overnight Costs (TOC)		$5,858,852	$155,503

3.5 OPERATING COSTS

Table 3-5 shows the operating cost breakdown for the Base Case TRIG FT CTL plant.

Table 3-5
SRI Base Case Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES				
Case:	SRI Base Case - TRIG Gasifier FT CTL with Rectisol based AGR for CO2 Capture			
Plant Size (BPSD FT Liquid Fuels)	**49996**		**Heat Rate (Btu/kWh):**	
Primary/Secondary Fuel:	**PRB**	**Coal**	**Fuel Cost ($/MMBtu):**	
Design/Construction	5 years		**Book Life (yrs):**	20
TPC (Plant Cost) Year	June 2011		**TPI Year:**	2016
Capacity Factor (%)	90.00		**CO2 Captured (STPD)**	37,730

OPERATING & MAINTENANCE LABOR					
Operating Labor					
Operating Labor Rate (base):	$39.70	$/hr			
Operating Labor Burden:	30.00	% of base			
Labor Overhead Charge	25.00	% of labor			
Operating Labor Requirements per Shift	**units/mod**	mod	**Total Plant**		
Skilled Operator	2.0	4	8.0		
Operator	10.0	4	40.0		
Foreman	1.0	4	4.0		
Lab Tech's etc	3.0	4	12.0		
TOTAL Operating Jobs	16.0		64.0		
				Annual Cost $	Annual Unit Cost $/(Bbl/day)
Annual Operating Labor Cost				$28,934,630	578.740
Maintenance Labor Cost				$46,439,837	928.873
Administration & Support Labor				$18,843,617	376.903
Property Taxes and Insurance				$95,940,165	1,918.962
TOTAL FIXED OPERATING COSTS				**$190,158,250**	3,803.479

VARIABLE OPERATING COSTS						
						$/Bbl
Maintenance Material Cost					**$86,245,412**	$5.2513
Consumables	Consumption Initial	Consumption /Day	Unit Cost	Initial Fill Cost		
Water(/1000 gallons)	0	1,557	1.67	**$0**	**$856,002**	$0.0521
Chemicals						
MU & WT Chem (lb)	0	9274	0.27	$0	$816,038	$0.0497
Carbon (Hg Removal) (lb)	513803	879	1.63	$837,499	$470,697	$0.0287
FT Catalyst (m3)	1010447	3533	7.15	$7,224,696	$8,297,075	$0.5052
Water Gas Shift Catalyst (ft3)	0	0.00	771.99	$0	$0	$0.0000
Methanol (Rectisol Solution) (tons)	1882	45.27	300.00	$564,644	$4,461,174	$0.2716
Amine Solution (gal)	473928	151.07	36.79	$17,435,794	$1,825,801	$0.1112
Hydrotreating Catalyst (ft3)	773	0.71	700.00	$540,879	$163,278	$0.0099
Naphtha Reforming Catalyst (ft3)	233	0.21	900.00	$209,543	$61,702	$0.0038
Isomerization Catalyst (ft3)	311	0.28	500.00	$155,740	$46,619	$0.0028
Claus Catalyst (ft3)	w/equip	3.55	203.15	$0	$236,982	$0.0144
Subtotal Chemicals				**$26,968,795**	**$16,379,367**	$0.9973
Other						
Butane (tons)	0	0	651.34	$0	$0	$0.0000
Gases, N2 etc.(/100scf)	0	0	0.00	$0	$0	$0.0000
LP Steam (/1000 lbs)	0	0	0.00	$0	$0	$0.0000
Subtotal Other				**$0**	**$0**	$0.0000
Waste Disposal:						
Spent Mercury Catalyst (lb)	0	879	0.65	$0	$187,701	$0.0114
Flyash (ton)	0	0.00	0.00	$0	$0	$0.0000
Slag (ton)	0	3421.73	25.11	$0	$28,224,582	$1.7185
Subtotal Waste Disposal				**$0**	**$28,412,283**	$1.7300
By-products & Emissions						
Sulfur (tons) / CO2 Product (Tonnes/day)	0	34,229	0.00	$0	$0	$0.0000
Supplemental Electricity (for sale) (MWh)		8505	58.59	$0	-$163,685,169	-$9.9664
Subtotal By-Products				**$0**	**-$163,685,169**	-$9.9664
TOTAL VARIABLE OPERATING COSTS				**$26,968,795**	**-$31,792,105**	-$1.9358
Coal (tons)	**0**	**35,856**	**19.63**	**$0**	**$231,214,517**	$14.0781
Natural Gas (1000 CF)	**0**	**0**	**5.13**	**$0**	**$0**	$0.0000
Total Feed					**$231,214,517**	$14.0781

3.6 COST OF PRODUCTION

Table 3-6 shows a summary of the power output, CAPEX, OPEX, COP and cost of CO_2 capture for the Base Case TRIG FT CTL with Rectisol-based AGR and CO_2 capture. The Base Case FT CTL COP is estimated to be $116/Bbl of FT Diesel

Table 3-6
SRI Base Case TRIG FT CTL Plant Performance and Economic Summary

	TRIG FT CTL Base Case
CAPEX, $MM	
Total Installed Cost (TIC)	$3,316
Total Plant Cost (TPC)	$4,797
Total Overnight Cost (TOC)	$5,859
OPEX, $MM/yr (90% Capacity Factor Basis)	
Fixed Operating Cost (OC_{Fix})	$190
Variable Operating Cost Less Fuel (OC_{VAR})	$132
Fuel Cost (OC_{Fuel})	$231
Power Export Credit	($164)
FT Products, BPD	
FT Diesel	14,762
FT Naphtha	35,234
Total	49,996
Power Production, MWe	
Gas Turbine	459
Steam Turbine	454
Auxiliary Power Consumption	559
Net Power Output	354
COP FT Diesel, excl CO2 TS&M, $/bbl FT diesel	115.4
COP FT Diesel, incl CO2 TS&M, $/bbl FT diesel	136.8
COP FT EPD, excl CO2 TS&M, $/bbl EPD	125.3
COP FT EPD, incl CO2 TS&M, $/bbl EPD	146.6
COP FT ECO, excl CO2 TS&M, $/bbl ECO	100.2
COP FT ECO, incl CO2 TS&M, $/bbl ECO	121.5
COP FT Naphtha, excl CO2 TS&M, $/bbl FT Naphtha	80.3
COP FT Naphtha, incl CO2 TS&M, $/bbl FT Naphtha	101.7

Section 4 POX Case: TRIG Gasifier with SRI POX Reformer FT CTL

4.1 PROCESS OVERVIEW

The POX Case FT CTL power plant, like the Base Case plant, is a Montana PRB coal-fired TRIG-based FT CTL plant designed to produce 50,000 BPD of FT diesel and naphtha. The tail gas from FT synthesis block is used as fuel for the GTG. The gas turbine generator (GTG) used is a GE MS6001B class turbine with a nominal ISO gross GT output of approximately 42 MWe. The GTCC power plant is equipped with a heat recovery steam generator (HRSG) and steam turbines to maximize power recovery.

The syngas exiting the TRIG gasifier is at 1,800 ^oF and has a H_2/CO ratio of 0.8. The raw syngas leaving the TRIG gasifier contains a significant amount of methane and potentially some tar as well. In the Base Case, the fate of the tar is unknown while the methane remains unconverted throughout the syngas heat recovery and cleaning processes. The addition of the SRI POX Reformer to the TRIG FT CTL plant increases the conversion of tars and methane into CO and H_2 which reduces the coal feed requirement for producing 50,000 BPD of FT liquid fuels. The resulting syngas exits the POX reformer at about the same temperature as the raw syngas leaving the TRIG gasifier (1,800ºF) and has a higher H_2/CO ratio of 1.06. The syngas exiting the POX reformer is cooled and cleaned in particulate filters and water scrubbed before it is sent for mercury removal. The cleaned syngas is then processed in the Rectisol acid gas removal (AGR) unit to remove H_2S and CO_2 from the syngas.

The treated syngas from the AGR is fed to the Fischer-Tropsch synthesis section with essentially all of the sulfur and contaminants removed from the syngas. Additional CO_2 removal is required from the FT synthesis product gas and the FT product upgrading tail gas to meet the less than 10% carbon emission requirement. Recovered CO_2 is purified and compressed to sequestration pressure of 2,200 psig at the plant battery limit. H_2S recovered from the Rectisol unit is converted into elemental sulfur in the Claus plant.

4.2 PERFORMANCE RESULTS

The SRI-modeled POX Case FT CTL plant with CO_2 capture consumes 26,400 TPD of PRB coal at the Montana site to produce 50,000 BPD of FT diesel and naphtha and produces a net power import of approximately 47 MWe. Overall performance for the POX Case FT CTL plant is summarized in Table 4-1, which includes auxiliary power requirements.

Table 4-1
SRI POX Case TRIG FT CTL Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	SRI POX Case FT Synthesis,TRIGP RB Coal
Gas Turbine Power	123,557
Steam Turbine Power	231,706
TOTAL POWER, kWe	**355,263**
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling	2,211
Coal Milling	10,251
Slag Handling	2,490
Air Separation Unit Auxiliaries	3,583
Air Separation Unit Main Air Compressor	210,928
Oxygen Compressor	31,821
Nitrogen Compressors	6,347
CO_2 Compressor	49,262
Boiler Feedwater Pumps	613
Condensate Pump	185
Quench Water Pump	0
Syngas Recycle Compressor	3,130
Circulating Water Pump	5,612
Ground Water Pumps	130
Cooling Tower Fans	2,938
Scrubber Pumps	1,044
Acid Gas Removal	17,574
Gas Turbine Auxiliaries	4,429
Steam Turbine Auxiliaries	294
Claus Plant/TGTU Auxiliaries	373
Claus Plant TG Recycle Compressor	314
FT Power Requirement	31,951
Miscellaneous Balance of Plant	15,075
Transformer Losses	2,096
TOTAL AUXILIARIES, kWe	**402,650**
NET POWER, kWe	**-47,386**
CONSUMABLES	
As-Received Coal Feed, lb/hr	2,199,072
Condenser Duty, MMBtu/hr	813
Raw Water Withdrawal, gpm	1,413

4.3 CAPITAL COST

For cost estimating purpose, the FT feed H_2+CO flow rate from the SRI ASPEN HMB is normalized to the DOE/NETL Report 1477 FT feed H_2+CO flow rate for producing 50,000 BPD of FT liquid fuels with a 70:30 yield of diesel to naphtha. The adjustments are as follows:

Table 4-2
SRI POX Case TRIG FT CTL H_2+CO Adjustments

		DOE/NETL 1477	SRI POX FT CTL ASPEN Simulation	Adjusted for Cost Estimated
AR Coal	T/D	21,006	28,372	26,389
Coal Type		Illinis No. 6	PRB	PRB
FT Synthesis Feed				
H2+CO	lbmols/hr	137,598	147,940	137,598
H2/CO				
FT Liquid Products				
Diesel	BPD	35,234		35,234
Naphtha	BPD	14,762		14,762
Total	BPD	49,996		49,996
Vol% Diesel	%	70%		70%
Vol% Naphtha	%	30%		30%
Total	%	100%		100%

Table 4-3 shows the cost breakdown of the POX Case FT CTL utilizing the TRIG gasifier coupled with the SRI POX reformer-based FT CTL, expressed in a consistent format with the Code of Accounts in the DOE/NETL 1477 report.

Table 4-4 shows the calculation and addition of owner's costs to determine the TOC, used to calculate COP.

Table 4-3
SRI TRIG FT CTL POX Case Total Plant Cost Summary

SRI FT POX Case With CO2 Sequestration Total Plant Cost Details (June 2011 Cost Basis)														
	Coal Type	PRB		Coal Feed (AR) =	2,199,072	lbs/hr		FT Product Slate =	50,000	Bbl/Day Total Liquid		100%		
	Gasifier	TRIG			26,389	tons/day			35,234	Bbl/Day Diesel		70%		
				Coal HHV (AR) =	8,564	Btu/lb			14,762	Bbl/Day Naptha		30%		
Acct		**Equipment**	**Material**	**Labor**		**Sales**	**Bare Erected**	**Eng'g CM**	**Contingencies**		**TOTAL PLANT COST**			
No.	**Item/Description**	**Cost**	**Cost**	**Direct**	**Indirect**	**Tax**	**Cost $**	**H.O & Fee**	**Process**	**Project**	**$**	**$/BPD**	**$/BPD$_{ECO}$**	**$/BPD$_{EPD}$**
1	COAL & SORBENT HANDLING	$44,141	$7,740	$33,761	$0	$0	$85,641	$7,590	$0	$18,647	$111,879	$2,982	$2,742	$3,427
2	COAL & SORBENT PREP & FEED	$226,898	$18,416	$38,423	$0	$0	$283,738	$25,993	$0	$64,047	$373,778	$9,962	$9,160	$11,450
3	FEEDWATER & MISC BOP SYSTEMS	$5,476	$3,521	$5,344	$0	$0	$14,341	$1,334	$0	$3,799	$19,474	$519	$477	$597
4	GASIFIER & ACCESSORIES													
4.1	Gasifier, Syngas Cooler & Auxiliaries (TR	$256,134	$0	$110,051	$0	$0	$366,185	$32,696	$84,428	$74,056	$557,364	$14,854	$13,659	$17,074
4.2	SRI POX	$72,732	$0	$24,344	$97,125	$162,815	$357,016	$66,532	$90,523	$84,710	$598,781	$15,958	$14,674	$18,343
4.3	ASU/Oxidant Compression	$257,230	$0	$0	$0	$0	$257,230	$24,934	$0	$28,217	$310,381	$8,272	$7,606	$9,508
4.4	LT Heat Recovery & FG Saturation	$18,751	$0	$7,079	$0	$0	$25,830	$2,521	$0	$5,670	$34,021	$907	$834	$1,042
4.X	Other Gasification Equipment	$0	$20,058	$11,564	$0	$0	$31,622	$2,908	$0	$8,473	$43,003	$1,146	$1,054	$1,317
	SUBTOTAL 4.	$604,847	$20,058	$153,039	$97,125	$162,815	$1,037,883	$129,591	$174,950	$201,126	$1,543,550	$41,137	$37,827	$47,284
5A	GAS CLEANUP & PIPING	$324,604	$6,976	$266,102	$0	$0	$597,682	$56,362	$108,183	$152,611	$914,838	$24,381	$22,420	$28,024
5AA	FT SYNTHESIS AND PRODUCT UPGRADE													
5AA.1	FT Synthesis*	$220,390	$0	$0	$0	$0	$220,390	$21,157	$59,505	$75,264	$376,316	$10,029	$9,222	$11,528
5AA.X	Hydrocarbon Upgrading	$254,952	$0	$0	$0	$0	$254,952	$24,400	$61,371	$83,669	$424,392	$11,310	$10,400	$13,001
5AA.Y	Amine CO2 Removal	$94,742	$0	$0	$0	$0	$94,742	$9,095	$25,580	$32,356	$161,773	$4,311	$3,965	$4,956
	SUBTOTAL 5AA.	$570,083	$0	$0	$0	$0	$570,083	$54,653	$146,456	$191,288	$962,481	$25,651	$23,587	$29,484
5B.2	CO2 Compression & Drying	$57,622	$0	$19,536	$0	$0	$77,158	$7,195	$0	$16,871	$101,224	$2,698	$2,481	$3,101
6	COMBUSTION TURBINE/ACCESSORIES	$50,897	$423	$4,096	$0	$0	$55,416	$11,996	$12,788	$15,949	$96,149	$2,562	$2,356	$2,945
7	HRSG, DUCTING & STACK													
7.1	Heat Recovery Steam Generator	$13,562	$0	$2,626	$0	$0	$16,189	$1,501	$0	$1,769	$19,459	$519	$477	$596
7.X	Ductwork & Stack	$2,214	$1,569	$2,054	$0	$0	$5,836	$529	$0	$1,033	$7,398	$197	$181	$227
	SUBTOTAL 7.	$15,776	$1,569	$4,680	$0	$0	$22,025	$2,030	$0	$2,802	$26,857	$716	$658	$823
8	STEAM TURBINE GENERATOR	$39,307	$1,012	$10,313	$0	$0	$50,633	$4,448	$0	$6,930	$62,011	$1,653	$1,520	$1,900
9	COOLING WATER SYSTEM	$8,668	$11,930	$10,146	$0	$0	$30,744	$2,755	$0	$7,127	$40,626	$1,083	$996	$1,245
10	ASH/SPENT SORBENT HANDLING SYS	$58,108	$4,001	$28,593	$0	$0	$90,701	$8,478	$0	$10,744	$109,923	$2,930	$2,694	$3,367
11	ACCESSORY ELECTRIC PLANT	$34,492	$21,600	$37,642	$0	$0	$93,734	$7,957	$0	$20,284	$121,975	$3,251	$2,989	$3,736
12	INSTRUMENTATION & CONTROL	$14,371	$2,913	$9,449	$0	$0	$26,732	$2,379	$1,337	$5,106	$35,553	$948	$871	$1,089
13	IMPROVEMENTS TO SITE	$8,323	$4,788	$21,546	$0	$0	$34,657	$3,431	$0	$11,428	$49,517	$1,320	$1,213	$1,517
14	BUILDINGS & STRUCTURES	$0	$14,011	$15,261	$0	$0	$29,272	$2,603	$0	$5,107	$36,981	$986	$906	$1,133
	Total Cost	$2,063,614	$118,957	$657,930	$97,125	$162,815	$3,100,441	$328,794	$443,715	$733,866	$4,606,816	$122,776	$112,897	$141,122

Table 4-4
SRI TRIG FT CTL POX Case Total Plant Cost Summary

Owner's Costs		$1,000	$/Bbl FT Liquids
Preproduction Costs			
6 months All Labor		$45,958	$1,225
1 Month Maintenance Materials		$7,669	$204
1 Month Non-Fuel Consumables		$1,514	$40
1 Month Waste Disposal		$1,936	$52
25% of 1 Months Fuel Cost at 100% CF		$3,939	$105
2% of TPC		$92,136	$2,456
Total		$153,153	$4,082
Inventory Capital			
60 day supply of fuel at 100% CF		$0	$0
60 day supply of non-fuel consumables at 100% CF		$200	$5
0.5% of TPC (spare parts)		$23,034	$614
Total		$23,234	$619
Owner's Cost			
Initial Cost for Catalyst and Chemicals		$25,505	$680
Land		$900	$24
Other Owner's Cost		$691,022	$18,416
Financing Costs		$124,384	$3,315
Total		$841,811	$16,836
Total Overnight Costs (TOC)		$5,625,013	$149,912

4.4 OPERATING COSTS

Table 4-5 shows the operating cost breakdown for the POX Coupled TRIG/SRI POX Reformer-based FT CTL.

Table 4-5
SRI TRIG FT CTL POX Case Initial and Annual O&M Costs

INITIAL & ANNUAL O&M EXPENSES

Case:	SRI POX Case - TRIG Gasifier FT CTL / SRI POX with Rectisol based AGR for CO2 Capture			
Plant Size (BPSD FT Liquid Fuels)	**49996**		**Heat Rate (Btu/kWh):**	
Primary/Secondary Fuel:	**PRB**	**Coal**	**Fuel Cost ($/MMBtu):**	
Design/Construction	5 years		**Book Life (yrs):**	20
TPC (Plant Cost) Year	June 2011		**TPI Year:**	2016
Capacity Factor (%)	90.00		**CO2 Captured (STPD)**	29,916

OPERATING & MAINTENANCE LABOR

Operating Labor

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

Operating Labor Requirements per Shift	**units/mod**	mod	**Total Plant**
Skilled Operator	2.0	4	8.0
Operator	10.0	4	40.0
Foreman	1.0	4	4.0
Lab Tech's etc	3.0	4	12.0
TOTAL Operating Jobs	16.0		64.0

	Annual Cost $	Annual Unit Cost $/(Bbl/day)
Annual Operating Labor Cost	$28,934,630	578.740
Maintenance Labor Cost	$44,598,585	892.045
Administration & Support Labor	$18,383,304	367.696
Property Taxes and Insurance	$92,136,319	1,842.879
TOTAL FIXED OPERATING COSTS	**$184,052,839**	3,681.361

VARIABLE OPERATING COSTS

						$/Bbl
Maintenance Material Cost					**$82,825,944**	$5.0431

Consumables	Consumption Initial	Consumption /Day	Unit Cost	Initial Fill Cost	Annual Cost	$/Bbl
Water(/1000 gallons)	0	1,017	1.67	**$0**	**$559,296**	$0.0341
Chemicals						
MU & WT Chem (lb)	0	6060	0.27	$0	$533,185	$0.0325
Carbon (Hg Removal) (lb)	378145	647	1.63	$616,376	$346,420	$0.0211
FT Catalyst (m3)	1010447	3533	7.15	$7,224,696	$8,297,075	$0.5052
Water Gas Shift Catalyst (ft3)	0	0.00	771.99	$0	$0	$0.0000
Methanol (Rextisol Solution (tons)	1886	45.35	300.00	$565,691	$4,469,440	$0.2721
Amine Solution (gal)	439994	140.26	36.79	$16,187,373	$1,695,072	$0.1032
Hydrotreating Catalyst (ft3)	773	0.71	700.00	$540,879	$163,278	$0.0099
Naphtha Reforming Catalyst (ft3)	236	0.21	900.00	$212,171	$62,476	$0.0038
Isomerization Catalyst (ft3)	315	0.29	500.00	$157,693	$47,204	$0.0029
Claus Catalyst (ft3)	w/equip	2.61	203.15	$0	$174,411	$0.0106
Subtotal Chemicals				**$25,504,879**	**$15,788,561**	$0.9613
Other						
Butane (tons)	0	0	651.34	$0	$0	$0.0000
Gases, N2 etc.(/100scf)	0	0	0.00	$0	$0	$0.0000
LP Steam (/1000 lbs)	0	0	0.00	$0	$0	$0.0000
Subtotal Other				**$0**	**$0**	$0.0000
Waste Disposal:						
Spent Mercury Catalyst (lb)	0	647	0.65	$0	$138,143	$0.0084
Flyash (ton)	0	0.00	0.00	$0	$0	$0.0000
Slag (ton)	0	2518.30	25.11	$0	$20,772,499	$1.2648
Subtotal Waste Disposal				**$0**	**$20,910,642**	$1.2732
By-products & Emissions						
Sulfur (tons) / CO2 Product (Tonnes/day)	0	27,140	0.00	$0	$0	$0.0000
Supplemental Electricity (for sale) (MWh)		-1137	58.59	$0	$21,888,933	$1.3328
Subtotal By-Products				**$0**	**$21,888,933**	$1.3328
TOTAL VARIABLE OPERATING COSTS				**$25,504,879**	**$141,973,377**	$8.6445
Coal (tons)	**0**	**26,389**	**19.63**	**$0**	**$170,167,390**	$10.3611
Natural Gas (1000 CF)	**0**	**0**	**5.13**	**$0**	**$0**	$0.0000
Total Feed					**$170,167,390**	$10.3611

4.5 COST OF PRODUCTION

Table 4-6 shows a summary of the FT liquid products output, power output, CAPEX, OPEX, COP and cost of CO_2 capture for the POX Coupled TRIG/SRI POX Reformer-based FT CTL. The COP for the POX Case FT CTL in $119/Bbl of FT diesel is shown in Table 4-6.

Table 4-6
SRI TRIG FT CTL POX Case Plant Performance and Economic Summary

	TRIG FT CTL POX Case
CAPEX, $MM	
Total Installed Cost (TIC)	$3,100
Total Plant Cost (TPC)	$4,607
Total Overnight Cost (TOC)	$5,625
OPEX, $MM/yr (90% Capacity Factor Basis)	
Fixed Operating Cost (OC_{Fix})	$184
Variable Operating Cost Less Fuel (OC_{VAR})	$120
Fuel Cost (OC_{Fuel})	$170
Power Export Credit	$22
FT Products, BPD	
FT Diesel	14,762
FT Naphtha	35,234
Total	49,996
Power Production, MWe	
Gas Turbine	124
Steam Turbine	232
Auxiliary Power Consumption	403
Net Power Output	(47)
COP FT Diesel, excl CO2 TS&M, $/bbl FT diesel	119.3
COP FT Diesel, incl CO2 TS&M, $/bbl FT diesel	137.5
COP FT EPD, excl CO2 TS&M, $/bbl EPD	129.5
COP FT EPD, incl CO2 TS&M, $/bbl EPD	147.7
COP FT ECO, excl CO2 TS&M, $/bbl ECO	103.6
COP FT ECO, incl CO2 TS&M, $/bbl ECO	121.8
COP FT Naphtha, excl CO2 TS&M, $/bbl FT Naphtha	83.0
COP FT Naphtha, incl CO2 TS&M, $/bbl FT Naphtha	101.2

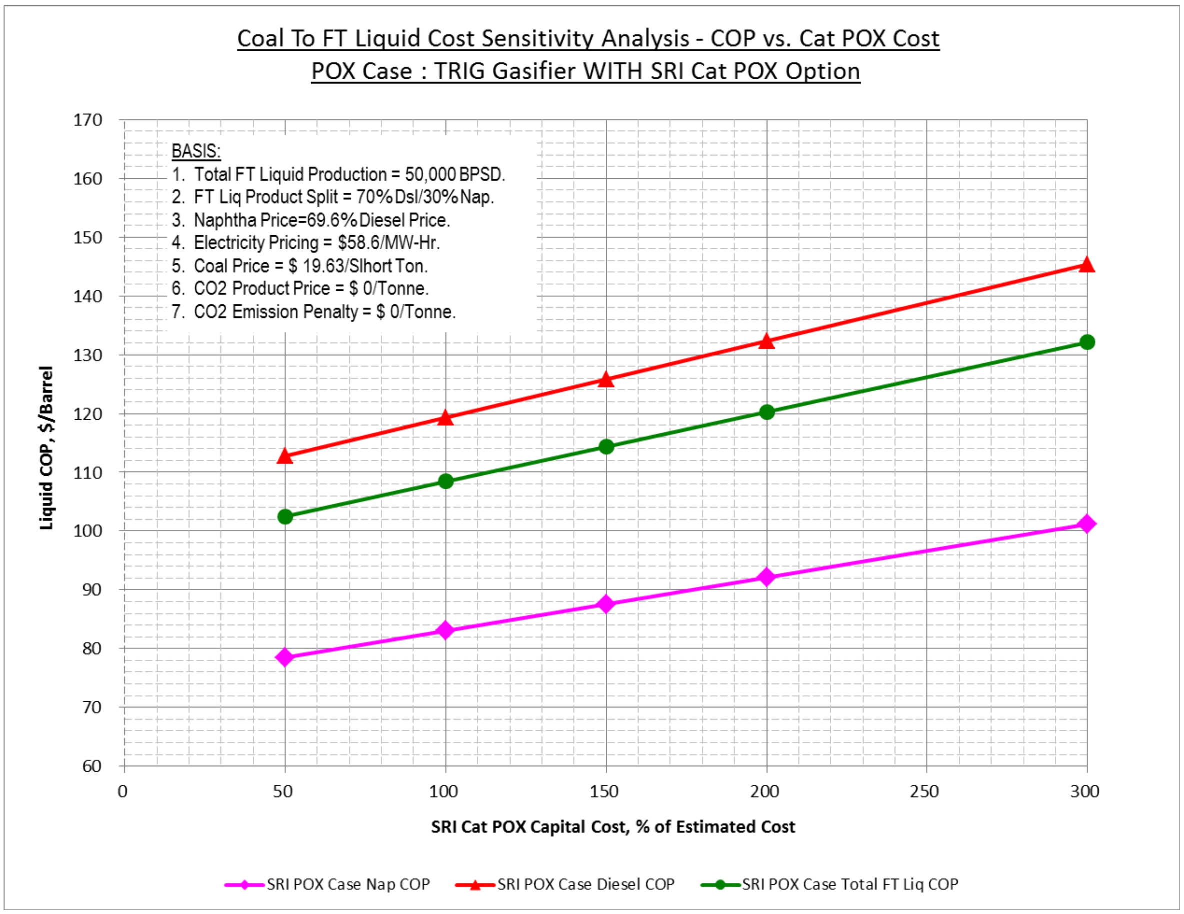
Section 5 Sensitivity Analysis

Sensitivity analysis was carried out to determine the effects of various cost parameters on the overall COP of the SRI TRIG gasifier based FT CTL Plant. The parameters investigated include: POX Reformer cost, feedstock cost, FT CTL plant capacity factor, price of electricity, CO_2 sales price, and cost of CO_2 emissions.

5.1 SRI POX REFORMER COST

The TPC for the SRI POX reformer is about 13% of the total SRI TRIG POX CTL plant TPC as shown in Table 4-3 in section 4. The impact of the SRI POX reformer TPC on the COP is shown in Figure 5-1. The COP increases by approximately $1.2/Bbl of FT diesel for every 10% increase in the estimated POX TPC.

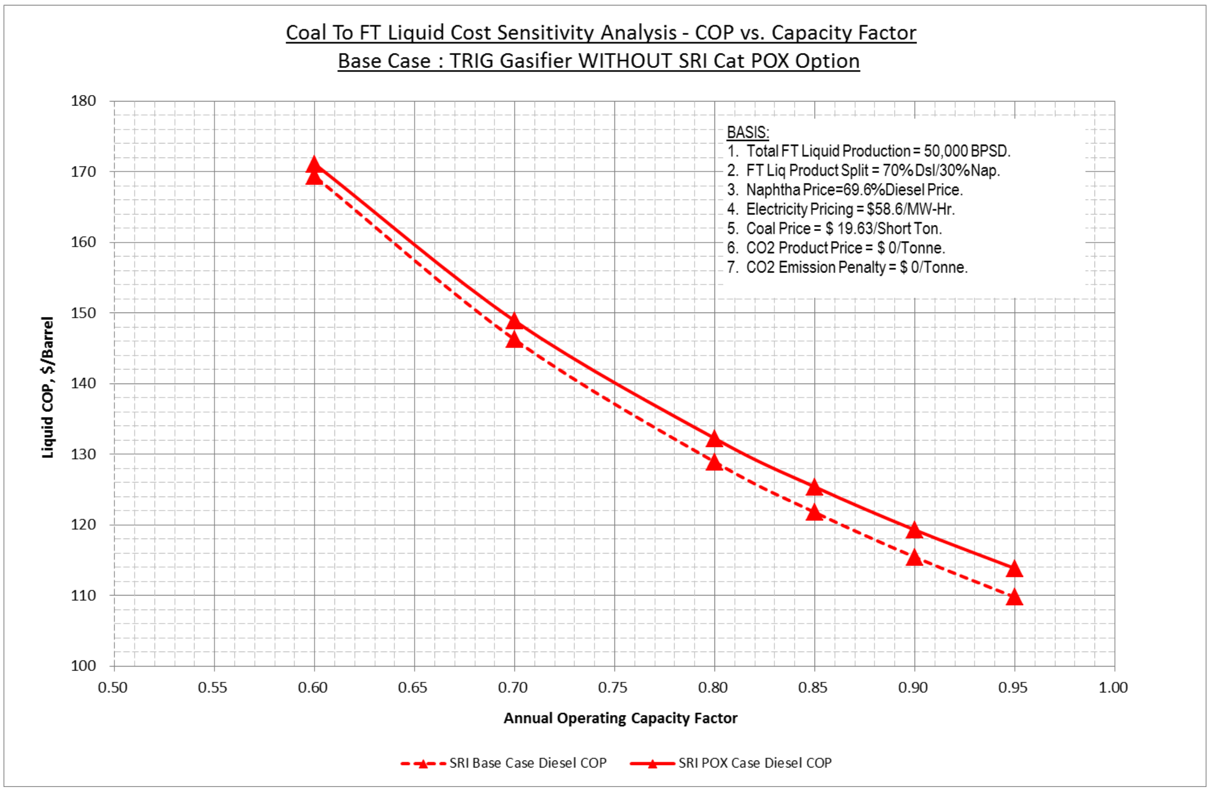
Figure 5-1
Sensitivity Analysis – COP vs POX Reformer Cost



5.2 CAPACITY FACTOR

The baseline FT CTL plant capacity factor used in this study is 0.9 (90%) which is the same as the DOE reference FT CTL case (DOE/NETL Report 1477). Figure 5-2 shows the impact of the plant capacity factor on the COP as it varies from 0.60 to 0.95. For every 0.1(10%) increase in the annual operating factor, the COP is decreased by approximately $14/Bbl FT diesel.

Figure 5-2
Sensitivity Analysis – COP vs FT CTL Plant Capacity Factor

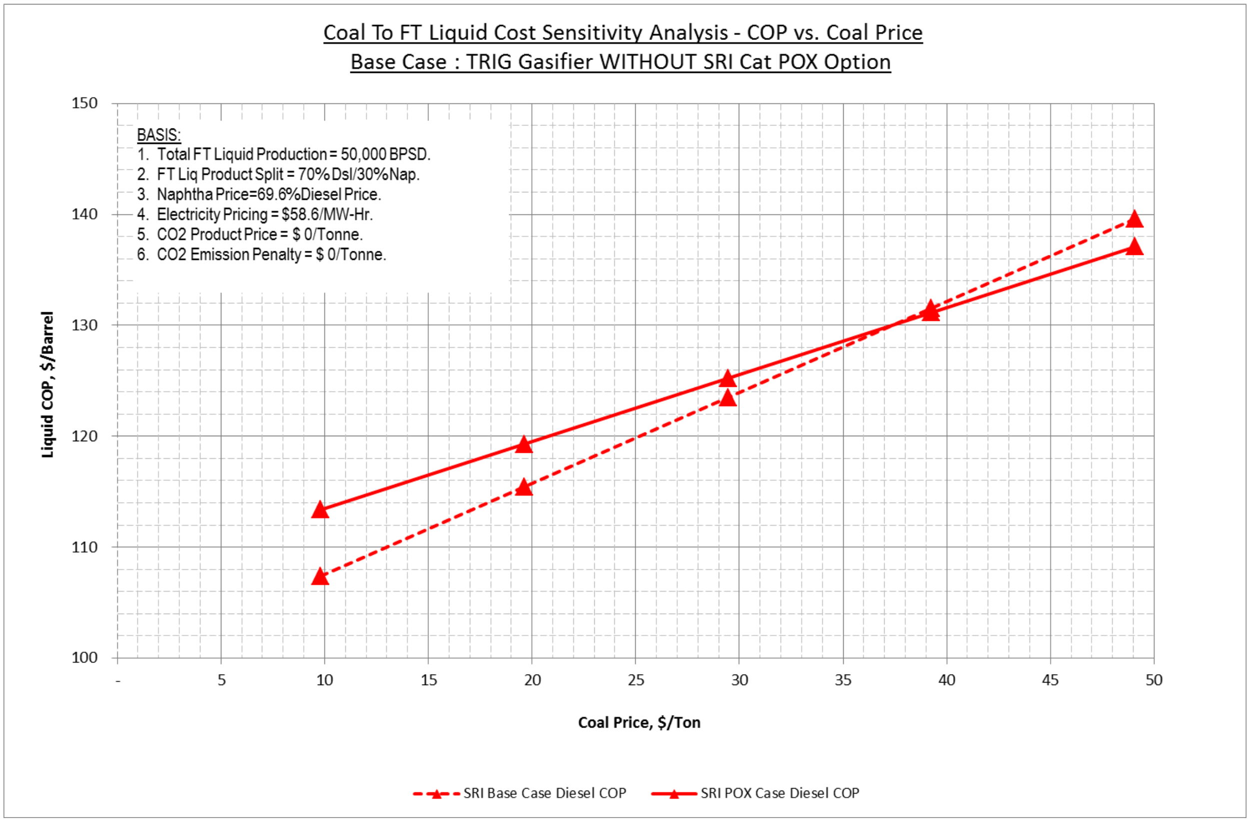


5.3 FEEDSTOCK PRICE

The baseline FT CTL plant PRB coal feedstock price used in this study is $19.63/ton. Figure 5-3 shows the change in COP with coal price as it varies from $10/ton to $60/ton.

As shown in Figure 5-3, for each $1/ton increase in coal price, the Base Case TRIG FT CTL plant COP increases by $0.85/Bbl FT diesel and for each $1/ton increase in coal price, the POX Reformer Case FT CTL COP increases by $0.60/Bbl FT diesel.

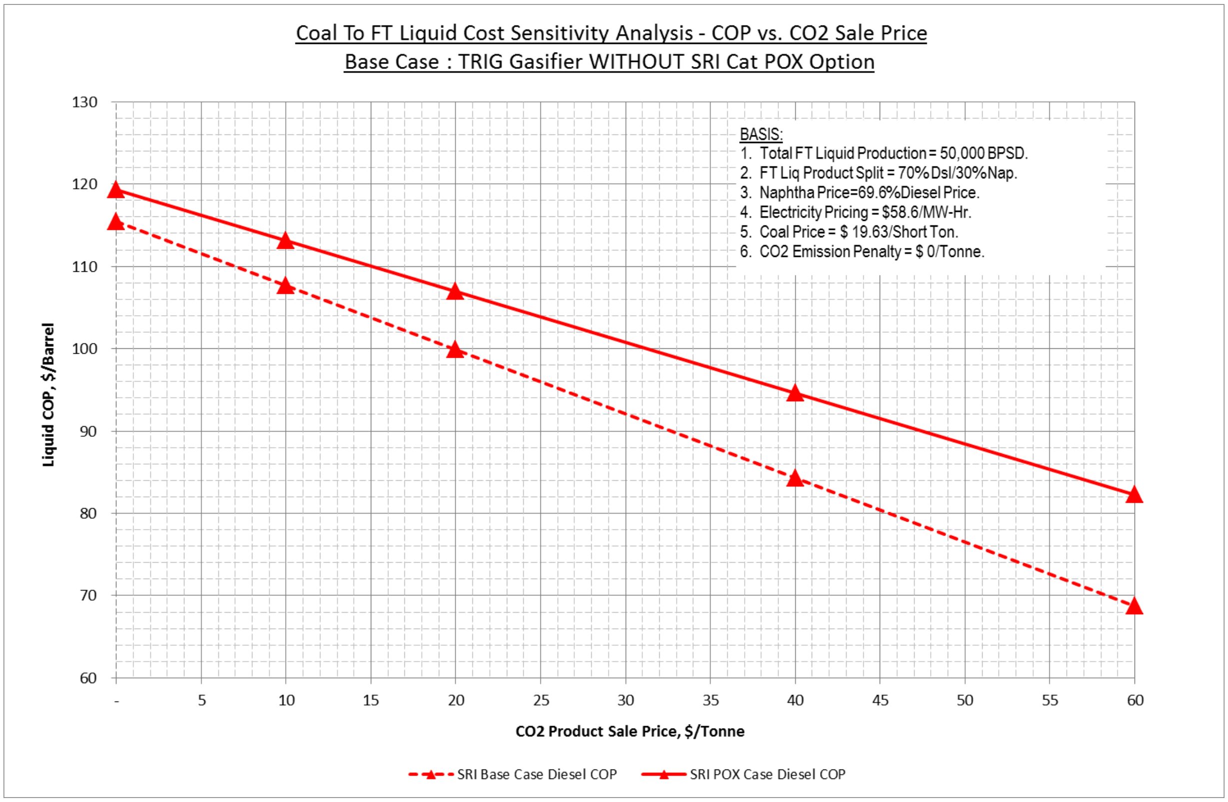
Figure 5-3
Sensitivity Analysis – COP vs Coal Price



5.4 CO_2 SALES PRICE

Sensitivity to CO_2 sales at plant gate prices is shown in Figure 5-4. The baseline case assumes that the CO_2 product carries no value ($0/tonne). The sales price is subsequently varied to a maximum of $60/tonne to determine its effect on the FT CTL plant's COP.

Figure 5-4
Sensitivity Analysis – COP vs CO_2 Sales Price



Due to the higher rate of CO_2 capture for the Base Case (37,700 tons/day CO_2 capture) over the POX Case (29,900 tons/day CO_2 capture), the rate of change in COP is higher for the Base Case as compared to the POX case as CO_2 sales price increases.

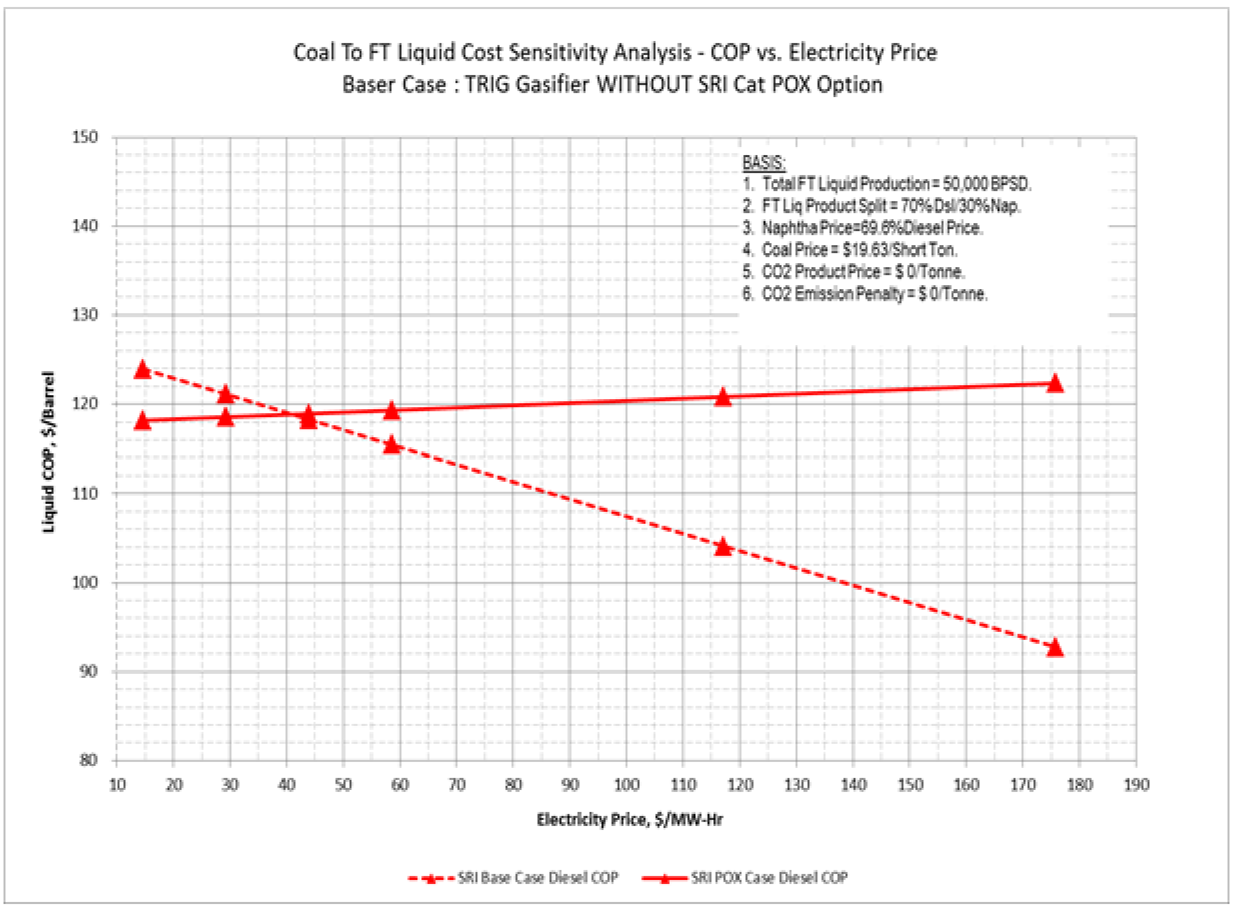
5.5 PRICE OF ELECTRICITY

As shown in the performance summaries in Tables 3-1 and 4-1, Base Case exports 354 MWe of power and the POX case imports 47 MWe of power. The net output varies among the FT CTL cases because of the methane content differences in FT tailgas fueling the gas turbines.

As electric price increases, the electricity sale reduces the Base Case COP and the electricity import increases the COP for the POX case.

For the Base Case, for each $10/MWhr increase in electricity price, the COP is reduced by $1.9/Bbl FT diesel. For the POX case, for each $10/MWhr increase in electricity price, the COP is increased by $0.2/Bbl FT diesel. At electricity price of $40/MWhr, the two cases have the same COP of $119/Bbl of FT diesel.

Figure 5-5
Sensitivity Analysis – COP vs Electricity Price



5.6 COST OF CO_2 EMISSIONS

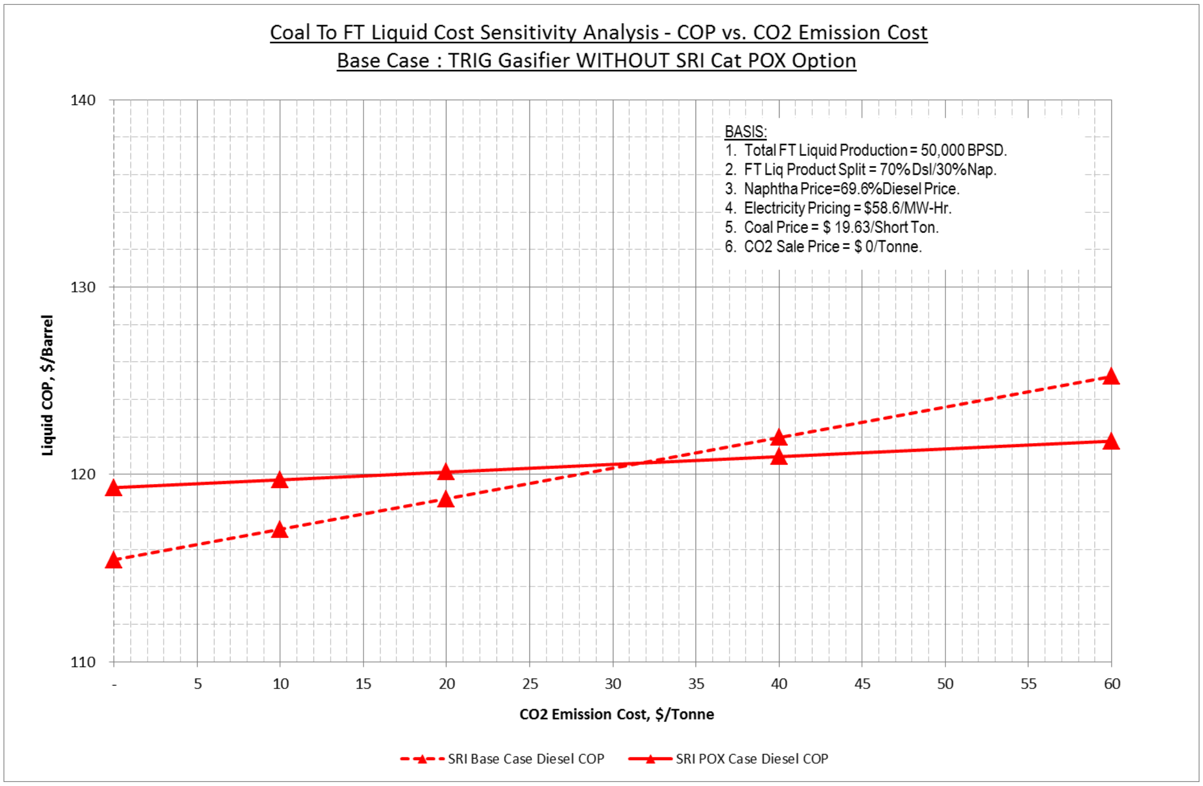
The sensitivity to CO_2 emissions costs is shown in Figure 5-6. The baseline case assumes that there are no costs associated with venting CO_2 to the atmosphere ($0/tonne). The cost of CO_2 emissions is subsequently varied to a maximum of $60/tonne to determine its effect on the FT CTL plant's COP.

Because CO_2 emissions from the various FT product upgrading furnaces are not available, sensitivity to CO_2 emission cost is based on estimated emissions from the GTCC power plant. Using only the GTCC emission should provide good approximation on the relative difference between the Base Case and the POX Case emission sensitivities since emissions from the FT product upgrading furnaces, as well as other process furnace vents, should be almost the same between the two Cases. Estimated CO_2 emission in the GTCC flue gas is approximately 7,166 tons/day for the Base Case and 1,816 tons/day for the POX Case. The difference in GTCC emission is due mainly to the higher methane content in the FT Tailgas, 14,453 lbmoles/hr for the Base Case versus 3,392 lbmoles/hr for the POX Case. These directly reflect the methane content difference in the syngas feeds to the FT Synthesis Block, 11,419 lbmoles/hr for the Base Case and 359 lbmoles/hr for the POX Case. The remaining additional 3,034 lbmole/hr of methane in the tail gases for both Cases is generated by the FT Synthesis/Upgrading reactions at 50,000 BPD FT liquid production.

For the same reason as explained in Section 5.4 the slopes are different between the Base Case and the POX Case due to the different CO_2 emission rates. The POX Case, which has less methane in the syngas feed to the FT plant, vents less CO_2 to the atmosphere, so it has a smaller increase in COP as cost of CO_2 emission cost increases.

The COP's are less sensitive to emissions cost than to CO_2 sales price because much more CO_2 is captured by the FT CTL plant than is vented (90% vs 10%), hence the COP is about 9 times more sensitive to CO_2 sales price than to cost of CO_2 emissions.

Figure 5-6
Sensitivity Analysis – COP vs Cost of CO_2 Emissions



Section 6 Conclusions

6.1 CASE CONFIGURATIONS

The two FT CTL configurations studied in this report are identified in the FT CTL case study matrix shown in Table 6-1.

Table 6-1
Case Study Matrix for FT CTL with CO_2 Capture

	Base Case[1]	POX Case[2]
Gasification Technology		
TRIG Gasifier	✓	✓
SRI POX Reformer		✓
Gas Cleanup		
Two-Stage Rectisol for CO_2 and Sulfur Removal[2]	✓	✓
Water Gas Shift		
Sour Shift	✓	✓
GE MS6001B Gas Turbine	✓	✓
CO_2 Drying and Compression (to 2,200 psig)	✓	✓

[1] Base Case modeled by SRI based on simulation of the DOE/NETL 2011/1477: Cost and Performance Baseline for Fossil Energy Plants, Volume 4: Bituminous Coal to Liquid via Fischer Tropsch Synthesis, May 12, 2014
[2] POX Case modeled by SRI which added SRI POX Reformer unit operation to the overall FT CTL process

Both plant configurations were evaluated based on installation at a greenfield site (Montana, 3,400 ft elevation. The study capacity factor (CF) of 90% was chosen to reflect the maximum availability demonstrated by FT CTL plants.

6.2 RESULTS SUMMARY

Table 6-2 shows the performance comparison and Table 6-3 show summary comparison of the capital expenditure (CAPEX), operating expenditure (OPEX), power production, and cost of production (COP) for the two cases for making 50,000 BPD of FT liquid products.

As presented in SRI's Base Case H&MB, the syngas from the TRIG gasifier contains significant amount of methane and tars due to its operating conditions. And as presented in SRI's POX Case H&MB, the addition of the sulfur-tolerant Catalytic POX unit reformed roughly 97% of the methane (from 11,419 lbmoles/hr for the Base Case down to 359 lbmoles/hr for the POX Case) and essentially all of the tars from the TRIG gasifier into additional H_2 + CO components available for the FT liquid production. Assuming constant FT liquid production at 50,000 BPD for both cases, the FT syngas feed H_2 + CO content for both SRI cases is fixed at 138,600 lbmoles/hr as specified in DOE/NETL Report 1477. The corresponding PRB coal feeds to generate this required amount of H_2 + CO are 35,856 STPD for the Base Case and 26,389 STPD for the POX Case, as pro-rated from the corresponding SRI's H&MB. Power consumptions for the coal handling and gasification/POX section corresponding to these coal rates are summarized in Table 6-2 together with those required by the other systems in the CTL plant.

As discussed in Section 2.4.2, methane in the FT feed syngas is assumed to pass through the FT reactors. After combining with the additional 3,034 lbmoles/hr of methane generated in the FT synthesis, the total FT tailgas is burned in the GTCC power plant to generate power to meet overall plant power demands. Excess power will be exported while deficit will be imported. While the Base Case tailgas total methane content is roughly 4.3 times that in the POX Case tailgas, total tailgas HHV ratio is reduced to about 3.7 times after the inclusion of equal amount of un-reacted purge H_2 and CO from FT synthesis. This corresponds to the GT output ratio between the two SRI cases as shown in Table 6-2. As shown in Table 6-2, STG output for the Base Case is only 2.0 times the POX STG output. The lowered STG ratio is due to almost the same amount of steam is generated by syngas cooling for the two cases (slightly less the coal feed ratio of 1.4). Total power generation for the Base Case is roughly 2.6 times that for the POX Case. As shown in Table 6-2, net result is that the Base Case will export about 354 MWe of power while the POX Case will import 47 MWe after meeting internal load demands.

With total FT liquid production being fixed at 50,000 BPD for both SRI Cases, and with a electrical power price of roughly $60/MW-hr, the Base Case with its higher power export has a slightly lower diesel COP than the POX Case, $115.4/Bbl vs. $ 119.3/Bbl as shown in Table 6-3. This advantage will disappear if the export power price is reduced to about $40/MW-Hr, as shown in Figure 6-1.

Because of the higher amount of methane in the FT tailgas being burned in the GTCC, the Base Case has higher CO_2 emission than the POX Case. Its COP advantage over the POX Case as shown in Table 6-3 assumes no penalty for CO_2 emission. If CO_2 emission penalty is above $33/Tonne, Base Case diesel COP will be higher than that for the POX Case, as shown in Figure 6-2.

In conclusion, for FT liquid production utilizing the low H_2/CO DOE once-through FT technology, COP for TRIG gasification of PRB coal with SRI POX is not competitive against TRIG gasification without SRI POX at electricity price at $60/MW-Hr. It becomes competitive when the electricity price drops below $40/MW-Hr, or if CO_2 emission penalty is more than $33/tonne. If export power credit is not allowed, then the SRI POX option will probably be more attractive based on COP.

Table 6-2
SRI TRG FT CTL Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	SRI Base Case FT Synthesis,TRIG PRB Coal	SRI POX Case FT Synthesis,TRIG PRB Coal
Gas Turbine Power	459,354	123,557
Steam Turbine Power	454,424	231,706
TOTAL POWER, kWe	**913,778**	**355,263**
AUXILIARY LOAD SUMMARY, kWe		
Coal Handling	3,004	2,211
Coal Milling	13,928	10,251
Slag Handling	3,383	2,490
Air Separation Unit Auxiliaries	4,780	3,583
Air Separation Unit Main Air Compressor	281,370	210,928
Oxygen Compressor	42,430	31,821
Nitrogen Compressors	6,932	6,347
CO_2 Compressor	62,128	49,262
Boiler Feedwater Pumps	21,647	613
Condensate Pump	585	185
Quench Water Pump	0	0
Syngas Recycle Compressor	4,253	3,130
Circulating Water Pump	8,436	5,612
Ground Water Pumps	199	130
Cooling Tower Fans	4,416	2,938
Scrubber Pumps	1,181	1,044
Acid Gas Removal	24,952	17,574
Gas Turbine Auxiliaries	16,464	4,429
Steam Turbine Auxiliaries	576	294
Claus Plant/TGTU Auxiliaries	507	373
Claus Plant TG Recycle Compressor	426	314
FT Power Requirement	31,951	31,951
Miscellaneous Balance of Plant	20,483	15,075
Transformer Losses	5,390	2,096
TOTAL AUXILIARIES, kWe	**559,422**	**402,650**
NET POWER EXPORT, kWe	**354,356**	**-47,386**
CONSUMABLES		
As-Received Coal Feed, STPD	35,856	26,389
Condenser Duty, MMBtu/hr	2,575	813
Raw Water Withdrawal, gpm	2,162	1,413

Table 6-3
SRI TRG FT CTL Results Summary

	TRIG FT CTL Base Case	TRIG FT CTL POX Case
CAPEX, $MM		
Total Installed Cost (TIC)	$3,316	$3,100
Total Plant Cost (TPC)	$4,797	$4,607
Total Overnight Cost (TOC)	$5,859	$5,625
OPEX, $MM/yr (90% Capacity Factor Basis)		
Fixed Operating Cost (OC_{Fix})	$190	$184
Variable Operating Cost Less Fuel (OC_{VAR})	$132	$120
Fuel Cost (OC_{Fuel})	$231	$170
Power Export Credit	($164)	$22
FT Products, BPD		
FT Diesel	14,762	14,762
FT Naphtha	35,234	35,234
Total	49,996	49,996
Power Production, MWe		
Gas Turbine	459	124
Steam Turbine	454	232
Auxiliary Power Consumption	559	403
Net Power Output	354	(47)
COP FT Diesel, excl CO2 TS&M, $/bbl FT diesel	115.4	119.3
COP FT Diesel, incl CO2 TS&M, $/bbl FT diesel	136.8	137.5
COP FT EPD, excl CO2 TS&M, $/bbl EPD	125.3	129.5
COP FT EPD, incl CO2 TS&M, $/bbl EPD	146.6	147.7
COP FT ECO, excl CO2 TS&M, $/bbl ECO	100.2	103.6
COP FT ECO, incl CO2 TS&M, $/bbl ECO	121.5	121.8
COP FT Naphtha, excl CO2 TS&M, $/bbl FT Naphtha	80.3	83.0
COP FT Naphtha, incl CO2 TS&M, $/bbl FT Naphtha	101.7	101.2

Figure 6-1
Sensitivity Analysis – COP vs Electricity Price

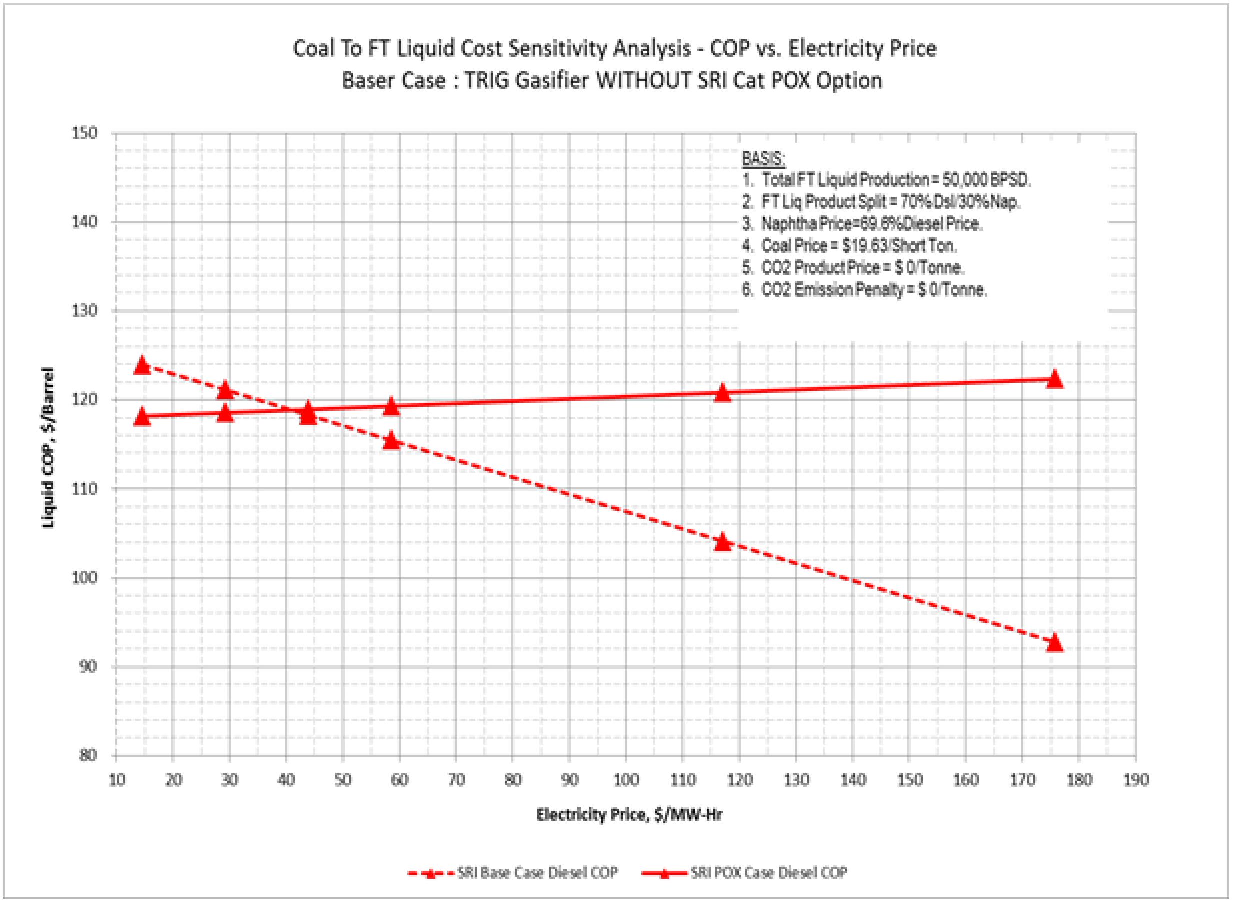
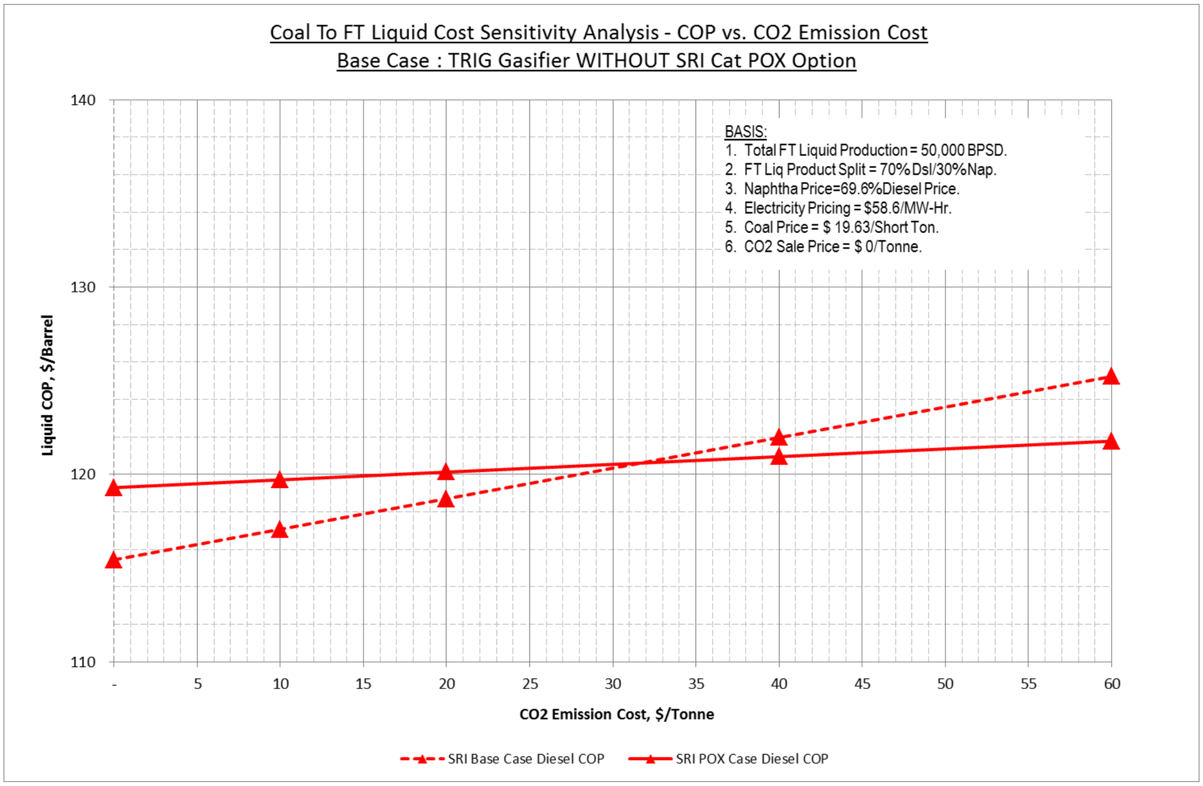


Figure 6-2
Sensitivity Analysis – COP vs CO_2 Emission Cost



°F	Degree Fahrenheit
AGR	Acid Gas Removal
AOI	Area of Interest
AR	As Received
ARR	Annual Revenue Requirement
ASU	Air Separation Unit
Bbl	Barrels
BEC	Bare Erected Cost
BFD	Block Flow Diagram
BFW	Boiler Feed Water
BOP	Balance of Plant
BPD	Barrels per Day
BPSD	Barrels per Stream Day
Btu	British Thermal Unit
CAPEX	Capital Expenditure
CCF	Capital Charge Factor
CF	Capacity Factor
CH_4	Methane
CO	Carbon Monoxide
CO_2	Carbon Dioxide
COE	Cost of Electricity
COP	Cost of Production
CTL	Coal-to-Liquids
CW	Cooling Water
DOE	Department of Energy
EPC	Engineering, Procurement, Construction
FOA	Funding Opportunity Announcement
ft	feet
FT	Fischer Tropsch
GE	General Electric
GT	Gas Turbine
GTCC	Gas Turbine Combined Cycle
GTG	Gas Turbine Generator
H_2	Hydrogen
H_2O	Water
H_2S	Hydrogen Sulfide
HHV	Higher Heating Value

HMB	Heat and Material Balance
HP	High Pressure
hr	Hour
HRSG	Heat Recovery Steam Generator
IGCC	Integrated Gasification Combined Cycle
ISO	International Organization for Standardizations
IOU	Investor Owned Utility
IRROE	Internal Rate of Return on Equity
kWe	Kilowatt electric
kWh	kilowatt hour
lb	Pound Mass
LHV	Lower Heating Value
LP	Low Pressure
max	Maximum
ME	Major Equipment
MEC	Major Equipment Cost
min	Minimum
Misc	Miscellaneous
MM	million
MU	Makeup
MWe	Megawatt electric
MWh/MWhr	megawatt hour
N_2	Nitrogen
NETL	National Energy Technology Laboratory
O&M	Operating and Maintenance
O_2	Oxygen
PFD	Process Flow Diagram
POX	Partial Oxidation
ppmv	Parts per Million by Volume
ppmW, ppmw	Parts per Million by Weight
PRB	Powder River Basin
PSFM	Power Systems Financial Model
psi	Pounds Per Square Inch
psia	Pounds Per Square Inch, absolute
psig	Pounds Per Square Inch, gauge
QGESS	Quality Guidelines for Energy System Studies
RSP	Required Selling Price
SC	Supercritical
SO_2	Sulfur Dioxide
SOPO	Statement of Project Objectives

SRI	Southern Research Institute
STG	Steam Turbine Power Generation
T&S	Transportation and Storage
TDC	Total Direct Cost
TEA	Techno-Economic Analysis
TFC	Total Field Cost
TG	Turbine Generator
TGTU	Tail Gas Treatment Unit
TIC	Total Installed Cost
TOC	Total Overnight Cost
TPC	Total Plant Cost
TPD	tons per day
TRIG	Transport Gasifier
US, USA	United States of America
vol%	Percentage by Volume
WT	Waste Treatment
WTI	West Texas Intermediate