The overall objective of this project was to undertake the development of an integrated multi-contaminant removal process in which hydrogen sulfide (H$_2$S), carbonyl sulfide (COS), ammonia (NH$_3$), chlorides and heavy metals, including mercury, arsenic, selenium and cadmium, present in the coal-derived syngas can be removed in a single process. To accomplish this, a novel process called UCSRP-HP (University of California Sulfur Recovery Process-High-Pressure) that directly converts H$_2$S into elemental sulfur at 285 to 300 °F and at any given sour gas pressure, and removes the various contaminants was adopted. During this research, data critical to developing and evaluating UCSRP-HP technology for multi-contaminant removal from syngas derived from Illinois #6 coal was obtained.

Aspen Plus simulations were performed that indicate complete removal of ammonia (NH$_3$), hydrogen chloride (HCl) and hydrogen selenide (H$_2$Se) in the aqueous scrub step of the UCSRP-HP process. Thermodynamic considerations point to minimal formation of COS in the UCSRP-HP reactor for a CO$_2$-rich-feed as compared to a CO-rich-feed gas stream.

An economic evaluation was performed that integrated the UCSRP-HP into a nominal 550 MWe Integrated Coal Gasification Combined Cycle (IGCC) facility gasifying Illinois coal. The results indicated that the cost of electricity (COE) could be reduced by about $9.60/MWhr (9.3%) compared to conventional IGCC technology with carbon capture, transportation, and sequestration, estimated at $103.00/MWhr by DOE. The COE saving is mainly derived from a reduction in the overall capital expenses of about $123 MM (December 2006 dollars). The UCSRP-HP design emits about 22 lb/hr less SO$_2$; reducing the emissions to near zero would increase the capital by about $10 MM and have a minimal effect on the COE. With these promising results, UCSRP-HP was integrated with advanced H$_2$/CO$_2$ separation technologies. With LANL/SRI’s PBI membrane, UCSRP-HP decreased the COE from $98.20 to $94.70/MWhr. With GTI/PoroGen’s CarboLock membrane contactors, the COE decreased from $93.40 to $91.70/MWhr. While replacing Selexol with the Advance H$_2$ Membrane decreased the COE with RTI Warm Gas Cleanup from $101.60 to $89.60/MWhr, substituting UCSRP-HP for RTI resulted in an $88.80/MWhr COE.
EXECUTIVE SUMMARY

Advanced gasification systems are planned to provide synthesis gas feed for advanced, combined cycle power plants; for separation systems for hydrogen production or for separating CO₂ for sequestration purposes; or for chemical conversion plants. All of these advanced applications require that any sulfur-containing species, as well as other contaminants, be reduced to parts-per-million (ppm) or in some cases parts-per-billion (ppb) levels. For acid-gas and trace contaminant removal, technologies that are either currently available or under development include: low-temperature or refrigerated solvent-based scrubbing systems using amines, such as MDEA, or physical solvents, such as Rectisol, Selexol, Sulfinol, or high temperature sorbents, such as RTI’s HTDP and DSRP (high-temperature desulfurization system and direct sulfur recovery process). Sulfur recovery is performed by a gas-phase, catalytic Claus process or sulfuric acid production followed by a tail gas recovery process, such as SCOT. Varieties of processes are required to remove trace components, such as ammonia, hydrogen chloride, and heavy metals. These processes are capital and energy intensive, with minimal reference plants due to the small number of Integrated Coal Gasification Combined Cycle (IGCC) facilities operating worldwide.

GTI is in the process of developing the UCSRP-HP (University of California Sulfur Recovery Process-High-Pressure), an integrated multi-contaminant removal process whereby coal-derived syngas is sent to an absorbing column where chlorides and ammonia, as well as trace heavy metals (mercury, selenium, arsenic, and cadmium) are removed from the gas stream. The partially cleaned gas then passes to a reactor column at a temperature above the melting point (247 °F) and below the polymerization temperature (310 °F) of elemental sulfur and a gasification pressure of 400 psig (or any higher pressure). Hydrogen sulfide in the syngas, together with injected sulfur dioxide, dissolves in a solvent that is circulated co-currently or counter-currently in the column. The Claus reaction is carried out in the liquid phase. Sulfur is only sparingly soluble in the solvent and so forms a separate liquid phase. The solvent contains a homogeneous liquid catalyst (less than 1% by weight of the solution). The catalyst is a commonly available and inexpensive material that does not degrade nor dissolve in the sulfur. One-third of the recovered sulfur product is burned with oxygen (if an oxygen-blown gasifier is involved this would be a fraction of the oxygen requirement for the process), and fed to the reactor column. The process is ideal for syngas desulfurization at 285 to 300 °F, at any given pressure (higher the better) and offers a tighter integration with the process for removal of trace contaminants and heavy metals. This process is projected to be significantly lower in capital and operating cost compared to conventionally applied amine or physical solvent-based acid-gas removal process followed by Claus/SCOT process plus systems to remove other contaminants or warm gas cleanup systems now in development.

This project was a laboratory program to obtain critical process data for the treatment of syngas derived from Illinois basin coals. The ultimate application of this process will favor high sulfur coals, such as Illinois #6, by reducing the cleanup costs and providing an environmentally benign facility to utilize economically these coals. The specific objectives of the project include laboratory work on the removal of heavy metals,
ammonia, halogens, and carbonyl sulfide and the development of an Aspen-Plus®-based computer simulation model used to perform a techno-economic evaluation of the process applied to syngas cleanup for a 500 MWe coal-based IGCC power plant. This projected was co-funded by the U.S. DOE through National Energy Technology Laboratory (NETL) under their funding opportunity announcement number DE-PS26-04NT42249. The funding received from ICCI was used to conduct work under each task using test conditions and gas compositions representative of Illinois # 6 coal.

Tasks 1 and 2 were the experimental and theoretical work of this project. Heavy metals and ammonia/halogen removal, take place in the scrub section of the UCSRP-HP process. Aspen Plus simulations performed indicated complete removal of ammonia (NH₃), hydrogen chloride (HCl) and hydrogen selenide (H₂Se) in the aqueous scrub step of the UCSRP-HP process. Trace sulfur removal takes place in the reaction section of the process. Thermodynamic considerations point to less formation of COS in the UCSRP-HP reactor for a CO₂-rich feed as compared to a CO-rich-feed gas stream.

Task 3 is the design modeling and economic evaluation work of this project. To support development, Aspen Plus® model has been completed for the UCSRP-HP and used to integrate the process into an IGCC plant with carbon capture using Illinois #6 coal. DOE economic studies were used to determine the potential impact of the technology compared to the current gas cleanup technology as well integrated with future technologies. For the base UCSRP-HP case, the process was introduced just downstream of the first heat exchanger (HP Steam) in the Gas Cooling, BFW Heating & Knockout block where the syngas feed is at ~772 psia and 450 °F. This is downstream of the Quench and Scrubber Section and the Water Gas Shift Reactors. The DOE Case 2 is rejoined at the feed to the CO₂ removal section of the dual stage Selexol unit. This design replaces or eliminates the Mercury Removal, H₂S-removal section of the dual-stage Selexol unit, the Claus Plant, and Hydrogenation Reactor and Gas Cooler Section. At the request of DOE, two levels of sulfur removal were studied. One case was for a conventional IGCC where the feed to the turbine would be ~8ppm and the other case was where the feed to the turbine or a chemical application would be <50 ppb.

The raw syngas is further cooled within the UCSRP-HP to 165 °F and then processed in a high-pressure, co-current, down-flow Water Contactor unit to separate a large fraction of water present in the gas along with much of the NH₃ and essentially all of the halogens and heavy metals, as sulfides or water-soluble salts. Following this step, the gas is preheated to about 250 °F prior to its processing in the UCSRP-HP reactor.

The UCSRP-HP reactors are co-current, down-flow contactors. At the operating conditions of ~750 psia and 250-300 °F, the sulfur forms as a liquid, essentially immiscible in DGM/catalyst solvent and over twice as dense. This overcomes the equilibrium limitations that occur in conventional gas-phase Claus reactor systems. The sulfur is separated, filtered by a DURCO sulfur filter, and sent to a sulfur pit or to a commercial-design O₂/sulfur submerged combustion furnace, as needed to generate the required liquid SO₂ for reaction with H₂S in the UCSRP-HP reactors. Solvent and reaction water are recovered by cooling and the cleaned syngas is transferred to the CO₂
removal section of the dual-stage Selexol unit and the power island.

To achieve the <50 ppb sulfur target, a COS hydrolysis unit was introduced within the UCSRP-HP block to produce a syngas containing about 7.6 ppm H2S and 0.4 ppm COS. A zinc oxide guard bed is also added downstream of the high-pressure separator to reduce the total sulfur level to <50 ppb in the effluent gas.

The cost analysis of a nominal 550 MWe IGCC plant with carbon capture was conducted following DOE economic evaluation guidelines to calculate capital expenditures (CAPEX), operating and maintenance costs (OPEX), and the levelized cost of electricity (COE). The results compared to conventional cold-gas cleanup are shown in Table ES-I.

**Table ES-I. Comparative COE Data: DOE Case 2 vs. UCSRP-HP**

<table>
<thead>
<tr>
<th></th>
<th>DOE Case 2</th>
<th>GTI UCSRP-HP Base Case</th>
<th>GTI UCSRP-HP Chemical case</th>
</tr>
</thead>
<tbody>
<tr>
<td>Coal Feed Rate, dry, lb/hr</td>
<td>444,737</td>
<td>444,737</td>
<td>444,737</td>
</tr>
<tr>
<td>SO2 Emissions (lb/hr)</td>
<td>56</td>
<td>34</td>
<td>~0</td>
</tr>
<tr>
<td>Gas Turbine, MWe</td>
<td>464.0</td>
<td>464.0</td>
<td>464.0</td>
</tr>
<tr>
<td>Sweet Gas Expander, MWe</td>
<td>6.3</td>
<td>6.3</td>
<td>5.8</td>
</tr>
<tr>
<td>Steam Turbine, MWe</td>
<td>274.7</td>
<td>287.5</td>
<td>287.5</td>
</tr>
<tr>
<td>Total Auxiliaries, MWe</td>
<td>-189.3</td>
<td>-184.5</td>
<td>-184.5</td>
</tr>
<tr>
<td>Net Power for Sale, MWe</td>
<td>555.7</td>
<td>573.3</td>
<td>572.8</td>
</tr>
<tr>
<td>Thermal Efficiency, % (HHV)</td>
<td>32.5</td>
<td>33.5</td>
<td>33.4</td>
</tr>
<tr>
<td>Total CAPEX, $MM (Yr-2006 $)</td>
<td>1,328</td>
<td>1,205</td>
<td>1,215</td>
</tr>
<tr>
<td>Cost of Power, $/MWhr (or, mills/kWh)**</td>
<td>103.0</td>
<td>93.4</td>
<td>95.2</td>
</tr>
</tbody>
</table>

These results show:
- A net CAPEX savings of about $123 MM (Dec’06 dollars) based on the conservative design and ± 30% cost estimate basis;
- The overall thermal efficiency (HHV basis) would increase from about 32.5% for the DOE Case 2 to about 33.3% for the UCSRP-HP design;
- An increase of about 17.6 MW (~3.2 %) in net power sale;
A reduction of about $9.60/MWhr (~9.3%) in the cost of power production with carbon capture, CO₂ compression plus transport/storage/monitoring; and

A reduction in total SO₂ emission of about 22 lb/hr;

Preparing the syngas for a chemical application that requires <50 ppb sulfur in the syngas would increase the required CAPEX by about $10.1MM over the base case, and would result in a subsequent increase of $1.80/MWhr in COE if the syngas were to be used for power generation in a near zero SO₂ discharge IGCC plant.

With these promising results, UCSRP-HP was integrated with developing technologies, identified in DOE’s Current and Future Technologies: A Pathway Study Focused on Non-carbon Capture Advanced Power Systems R&D using Bituminous Coal- Volume 1 and 2.⁸,⁹ These included (1) the LANL/SRI’s PBI (polybenzimidazole), high-temperature (~250-450 °C), polymeric CO₂-removal membrane, (2) the GTI/PoroGen’s CarboLock Membrane for CO₂ removal, and (3) an advanced H₂-separation membrane that would meet specific year-2015 performance targets assumed by DOE/Noblis LLC were also examined. These cases were compared to the published data on RTI’s warm gas cleanup (WGC) process and can be seen in Table ES-II.

**Table ES-II. Comparative Data for COE (including CO₂ capture, compression, transport, plus storage/monitoring)**

<table>
<thead>
<tr>
<th>Design Case</th>
<th>H₂S Removal Option</th>
<th>CO₂ Removal Option</th>
<th>COE, $/MWhr</th>
<th>COE Differential, %</th>
</tr>
</thead>
<tbody>
<tr>
<td>DOE-1</td>
<td>Selexol</td>
<td>None</td>
<td>77.8</td>
<td>Base</td>
</tr>
<tr>
<td>DOE-2</td>
<td>Selexol</td>
<td>Selexol</td>
<td>103</td>
<td>32.4</td>
</tr>
<tr>
<td>A</td>
<td>UCSRP-HP</td>
<td>Selexol</td>
<td>93.4</td>
<td>20.1</td>
</tr>
<tr>
<td>B</td>
<td>UCSRP-HP</td>
<td>GTI/ PoroGen CarboLock Membrane Contactor w/Selexol</td>
<td>91.7</td>
<td>17.9</td>
</tr>
<tr>
<td>C</td>
<td>Selexol</td>
<td>LANL/ SRI PBI Membrane</td>
<td>98.2</td>
<td>26.2</td>
</tr>
<tr>
<td>D</td>
<td>UCSRP-HP</td>
<td>LANL/ SRI PBI Membrane</td>
<td>94.7</td>
<td>21.7</td>
</tr>
<tr>
<td>E</td>
<td>RTI</td>
<td>Selexol</td>
<td>101.6</td>
<td>30.6</td>
</tr>
<tr>
<td>F</td>
<td>RTI</td>
<td>Advanced H₂ Membrane</td>
<td>89.6</td>
<td>15.2</td>
</tr>
<tr>
<td>G</td>
<td>UCSRP-HP</td>
<td>Advanced H₂ Membrane</td>
<td>88.8</td>
<td>14.2</td>
</tr>
</tbody>
</table>
OBJECTIVES

The overall objective of this project was the development of an integrated multi-contaminant removal process in which H₂S, NH₃, HCl, and heavy metals including Hg, As, Se, and Cd present in the coal-derived syngas were removed to specified levels in a single process step. To accomplish this, a high pressure University of California Sulfur Recovery Process (hereafter referred to as UCSRP-HP) that directly converts H₂S into elemental sulfur at 285 to 300 °F and at any given sour gas pressure was used. The process also removes contaminants such as NH₃, HCl, Hg, Se, and other trace contaminants.

The specific objectives of the proposed projects include (i) gathering data for further verification of the process concept for trace component removal and recovery, measuring contaminant removal efficiencies, reaction kinetics, and optimization of process parameters for testing in the bench-scale unit, (ii) incorporating the data and process design into the Aspen-Plus based computer simulation model established during Phases 1 and 2, and (iii) updating the techno-economic evaluation of the overall process applied to syngas cleanup for a 500 MWe, Illinois-coal-based IGCC power plant. This project was co-funded by the U.S. DOE through National Energy Technology Laboratory (NETL) under an extension of their funding opportunity announcement number DE-FC26-05NT42458. The scope of work for the DOE project concentrated on the further development in the bench-scale unit. The funding received from ICCI was used to conduct work under each task using test conditions and gas compositions representative of Illinois #6 coal.

INTRODUCTION AND BACKGROUND

When coal is gasified, the syngas produced may contain not only H₂S, but also NH₃ and HCl and heavy metals, such as As, Cd, Hg, and Se. Before this syngas is used as fuel for a gas turbine or further processing to methane, liquid hydrocarbons, or hydrogen, all of the above should be reduced to very low values.

In advanced gasification applications where a low-temperature absorption process such as Rectisol or Selexol is employed to scrub the gas and remove the sulfur compounds, the sulfur-containing species such as H₂S and COS are recovered as an acid gas, which then requires a sulfur-recovery process. The modified Claus process coupled with a tail gas treatment (TGT) process such as SCOT are typically used to recover elemental sulfur and produce a dischargeable plant tail gas. Some plant designs produce sulfuric acid rather than elemental sulfur. Mercury removal is achieved by chemisorption with sulfur-impregnated carbon.

Research at the University of California, Berkeley (UCB), coupled with experimental work at Gas Technology Institute (GTI), is leading to the development of an integrated multi-contaminant removal process whereby syngas is sent to a reactor column at a temperature above the melting point (247 °F) and below the polymerization temperature (310 °F) of elemental sulfur, at a gasification pressure of 400 psig or higher. Details of
the UCSRP-HP have been presented previously.\textsuperscript{1}

H\textsubscript{2}S in the syngas, together with injected SO\textsubscript{2}, dissolves in a solvent that is circulated in the column. The Claus reaction where H\textsubscript{2}S reacts with SO\textsubscript{2} to form sulfur and water occurs in the liquid phase. The sulfur formed is only sparingly soluble in the solvent and therefore forms a separate liquid phase. The solvent contains 3-Pyridinemethanol, a homogeneous liquid catalyst at less than 1\% by weight of the solution. This catalyst is a commonly available and inexpensive material that does not degrade nor dissolve in the sulfur. One-third of the sulfur product is burned with oxygen (if an oxygen-blown gasifier is involved this would be a fraction of the oxygen requirement for the process), and recycled to the reactor column as the SO\textsubscript{2} source. The remaining sulfur product is sent to the sulfur pit for export. Means are also provided to remove COS, HCl, NH\textsubscript{3}, and trace heavy metals. The treated gas leaving the system is expected to meet the strict specifications set for H\textsubscript{2}S and the other contaminants for turbines, fuel cells, and catalytic processes.

The process is ideal for syngas desulfurization at 285 to 300 °F and at any given pressure (the higher the better) and offers a tighter integration with the process for removal of trace contaminants and heavy metals. It is expected to be significantly lower in capital and operating cost compared to conventionally applied amine or physical solvent-based acid-gas removal process followed by Claus/SCOT process. Testing done at GTI has shown negligible chemical consumption (including catalyst), unlike typical chemicals costs of $300 - $1000 per ton sulfur removed found in competing processes. There is much less need for stainless steels in the process, and no apparent cut-off point in terms of sulfur handling at which Claus/SCOT becomes more economical.

This process differs from liquid redox processes in important ways. There is no need for filtering a solid sulfur paste with attendant handling problems and loss of solvent. The sulfur quality can be as good as Claus sulfur due to the low solubility of the solvent in the liquid sulfur, and to the large density difference and ease of liquid/liquid separation in the process. The process can operate at significantly higher temperatures than the liquid redox or CrystaSulf processes, which is of value in IGCC applications. No foaming of the solution occurs since the solvent is non-aqueous and has no surfactant properties. No sticky or solid sulfur is present anywhere in the system so the problems of liquid redox plugging and pump wear would not be present.

Although there is no minimum (or maximum) pressure at which the process can operate, the flow of solvent is reduced and reaction rates are increased at higher pressures. This permits the use of smaller equipment and lowers operating costs; hence, the process is projected to be attractive for the treatment of H\textsubscript{2}S-containing gases at high pressure. This technology offers great advantages for Illinois basin coals. The ability to reduce the cost for sulfur and other coal impurities will lead to a greater usage of the high sulfur Illinois coals that are not currently competitive for power generation.
EXPERIMENTAL PROCEDURES

The overall approach taken in this project is to use simulated coal-derived syngas for laboratory testing of the UCSRP-HP process, with supporting computer simulations. The gas mixtures are complex, consisting of a number of components down to the parts per million (ppm) level. Therefore, modeling results have been used to design experiments, simplify gas mixtures used in the experiments, and capture the important aspects of the process.

Trace component (elemental mercury, $\text{Hg}^0$) removal

Mercury will react with $\text{H}_2\text{S}$ to form $\text{HgS}$ in both gas and liquid phases. The $\text{HgS}$ in the solvent will be a suspended solid. Figure 1 shows the experimental set-up for studying the removal of elemental mercury from the gas stream. The apparatus consists of the following main components:

- 3000 ml Pyrex glass reactor vessel fitted with a variable speed stirrer and a drain valve for liquid sample acquisitions;
- heating jacket around the glass reactor to heat the water to 90 and 165 °F (not shown);
- mercury permeation tube assembly (permeation devices have constant emission rate of the component at a fixed temperature); and
- sodium hydroxide and sulfur impregnated activated carbon trap.

Contaminants in the syngas include ammonia and hydrogen chloride that are highly water-soluble. Simulation results have shown that ammonia is removed either as a chloride or as a carbamate. In other words, ammonia and hydrogen chloride is expected to be completely removed from the gas phase during the aqueous scrub step of the process. With that in mind, an aqueous solution of ammonium chloride is used as the absorption medium in the glass reactor. Since the actual syngas feed will be $\text{CO}_2$ rich, and mercury will precipitate as sulfides in the presence of hydrogen sulfide, the carrier gas for mercury vapor contains both carbon dioxide and hydrogen sulfide.
Figure 1 Experimental set-up for trace metal (elemental mercury) removal

Trace sulfur removal

The fate of COS in the process is critical for the success of the UCSRP-HP scheme. This necessitates the need to study the mechanism of COS formation and consumption under the process conditions. The experimental set-up for studying the trace sulfur components is given in Figure 2.

The apparatus consists of a 4 L steel reaction vessel equipped with a variable speed stirrer. The vessel is also equipped with a bottom drain valve that is used to remove periodically sulfur produced during a side reaction. The reaction vessel was designed to be operated from 120-160 °C at pressures up to 500 psia. The exit gas stream passes through the gas-washing bottle to condense the water produced in the reaction and then to a second gas-washing bottle filled with 50% sodium hydroxide (NaOH) solution to consume all of the untreated H₂S before it is vented. H₂S and SO₂ leak detectors are placed near the setup to detect any leaks. The flow rates of H₂S and SO₂ are controlled by using two Brooks Mass Flow Controllers (MFC) and a back pressure regulator is used to maintain a constant reactor pressure of 400 psia. The outlet flow is measured with a bubble flow meter. The gas streams are analyzed using a Varian CP-4900 Micro Gas Chromatograph, equipped with a Thermal Conductivity Detector (TCD) and calibrated with known gas standards.
RESULTS AND DISCUSSION

Task 1 – Trace Component Removal and Recovery Experiments

In the UCSRP-HP process one of the key steps is the conversion of raw syngas (at 400-800 psia and 400 °F) from the Quench/Scrubber section to a H₂ (+ CO₂) rich gas in a WGS (Water Gas Shift) reactor using sour shift catalyst. Heat is recovered and a major portion of the residual water vapor is removed from the effluent gas from the WGS reactor. The various contaminants [e.g., H₂S, NH₃, HCl, As (AsH₃), Se (H₂Se), Hg, HCN] in the syngas are also removed along with the water stream.

Phase equilibrium calculations for the system using Aspen Plus (v. 7.1, Aspen Tech) have been performed for establishing a theoretical basis for the separation. The property method used was the electrolyte NRTL (eNRTL) suitable for electrolyte systems. The feed for the calculations is the stream after the sour gas shift reactor. Equilibrium distribution of species for various temperatures and pressures of interest have been calculated using the process simulators. Additionally, the distribution of species in the solvents (water and DGM) has been calculated at the Claus reactor conditions where H₂S is reacted with SO₂ in the catalyzed DGM solvent.

The feed gas composition used in the simulation studies is given in Table 1. It can be considered as a base case scenario with no minor contaminants present (e.g., H₂Se, HCl, Hg, AsH₃, HCN).
### Table 1 Feed Gas Composition (after WGS reactor)

<table>
<thead>
<tr>
<th>COMPONENTS</th>
<th>Formula</th>
<th>Volume %</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ammonia</td>
<td>NH₃</td>
<td>0.11</td>
</tr>
<tr>
<td>Hydrogen sulfide</td>
<td>H₂S</td>
<td>0.48</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>N₂</td>
<td>2.05</td>
</tr>
<tr>
<td>Carbon dioxide</td>
<td>CO₂</td>
<td>31.13</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>H₂</td>
<td>43.06</td>
</tr>
<tr>
<td>Water</td>
<td>H₂O</td>
<td>23.17</td>
</tr>
<tr>
<td><strong>Total = 100</strong></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

The various reactions in aqueous phase that were taken into account in the simulation are given below:

Dissociation of water

\[2H₂O \rightleftharpoons H₃O^+ + OH^-\]

Dissociation of hydrogen sulfide

\[H₂S + H₂O \rightleftharpoons H₃O^+ + HS^-\]
\[HS^- + H₂O \rightleftharpoons H₃O^+ + S^{2-}\]

Dissociation of ammonia

\[NH₃ + H₂O \rightleftharpoons NH₄^+ + OH^-\]

Formation of ammonium bisulfide and ammonium sulfide

\[NH₄^+ + HS^- \rightleftharpoons NH₄SH\]
\[2NH₄^+ + S^{2-} \rightleftharpoons (NH₄)_2S\]

HCl addition

\[HCl + H₂O \rightleftharpoons H₃O^+ + Cl^-\]
\[NH₄Cl(s) \rightleftharpoons NH₄^+ + Cl^-\]

H₂Se addition

\[H₂Se + H₂O \rightleftharpoons H₃O^+ + HSe^-\]
\[HSe^- + H₂O \rightleftharpoons H₃O^+ + Se^{2-}\]

Hg addition

\[Hg^{2+} + 2H₂O \rightleftharpoons H₃O^+ + HgOH^+\]
\[HgOH^+ + 2H₂O \rightleftharpoons H₃O^+ + Hg(OH)₂\]
\[Hg^{2+} + Hg \rightleftharpoons Hg₂^{2+}\]
AsH₃ addition
\[ \text{AsH}_3(g) \leftrightarrow \text{AsH}_3(aq) \]

The equilibrium distribution of species (H₂S, NH₃, H₂O) at various temperatures and pressures in water are given in Figure 3. As seen, H₂S is mostly present in the vapor phase unlike ammonia that is largely dissolved in aqueous phase. Effect of pressure and temperature on the equilibrium distribution is as expected. Increase in pressure increases the gas solubility and an increase in temperature decreases the gas solubility.

The effect of the addition of contaminants (H₂Se, HCl, Hg, and AsH₃) on the distribution of ammonia and hydrogen sulfide is considered next. The contaminants are introduced one at a time to the base gas mixture in the simulation. The temperature and pressure conditions chosen were respectively 250 °F and 800 psi. Two cases are considered for each contaminant, one is the addition of 1/10th the volume percentage of ammonia and the other is an addition equal to the amount of ammonia present in the mixture. In other words, the volume percentage of the contaminant will be 0.011% and 0.11% for the simulation runs. It may be noted that the actual concentrations of these contaminants are much less (of the order of 1/50th the concentration of ammonia).

It can be seen from Figure 4A that HCl and H₂Se addition greatly reduces the amount of ammonia in the vapor phase. The enhanced distribution in the aqueous phase is simply
due to the acid-base chemistry involved between ammonia (base) and the various hydracids present (e.g., HCl, H₂S, H₂Se). It may be noted that H₂S and H₂Se are the hydracids of Group VI elements (S, Se) with H₂Se being a stronger acid. In other words, increased solubility for H₂Se in the aqueous phase can be expected. Equilibrium speciation results predict the formation of bisulfide (HS⁻) and biselenide (HSe⁻) in the aqueous phase.

In the case of hydrogen sulfide (Figure 4B), the effect of the contaminant addition on the equilibrium distribution is seen to be minimal. The reactions considered do not include the precipitation of heavy metals (mercury and arsenic), as sulfides. Both mercury and arsenic can form sulfides. The consideration of these precipitation reactions will alter the distribution of H₂S in the vapor phase.

It may be pointed that in the case of arsine (AsH₃), the proton affinity is very low (compared to ammonia) and the tendency to form the onium ions MH₄⁺ (M = N, As) is very little resulting lower solubility. NH₃ and AsH are the trihydrides of Group V elements nitrogen (N) and arsenic (As).

It needs to be mentioned that the equilibrium calculations (Figure 3 and Figure 4) did not take into consideration the effect of reactions due to ammonia and carbon dioxide. The various reactions can be given as follows:

\[
\begin{align*}
CO_2 + 2H_2O & \rightleftharpoons H_3O^+ + HCO_3^- \\
HCO_3^- + H_2O & \rightleftharpoons H_3O^+ + CO_3^{2-} \\
NH_3 + HCO_3^- & \rightleftharpoons H_2O + NH_2COO^- \\
NH_4HCO_3_{(s)} & \rightleftharpoons NH_4^+ + HCO_3^- \\
NH_4COONH_2_{(s)} & \rightleftharpoons NH_4^+ + NH_2COO^- 
\end{align*}
\]

The major effect of CO₂ addition is on the distribution of ammonia (NH₃). The calculations show the formation of a solid phase (ammonium carbamate) resulting in the removal of the contaminant from the solvent phase.
Figure 4 Effect of trace component addition on the distribution of ammonia and hydrogen sulfide (Aspen simulation, T = 250 °F, P = 800 psi, Base gas composition: Table 1), A. Ammonia B. Hydrogen sulfide
Task 2 – Trace Sulfur Removal Laboratory Experiments

In the reactor section of the UCSRP-HP process, hydrogen sulfide reacts with sulfur dioxide to form water and liquid sulfur. The reaction proceeds in the liquid phase (DGM solvent) containing a homogeneously dissolved catalyst. Under these conditions, there are paths where trace sulfur components such as carbonyl sulfide (COS) and carbon disulfide could form. The fate of trace sulfur components is of critical importance to the success of the UCSRP-HP scheme. The various reactions involving the trace sulfur components are given below.

Formation and consumption reactions for carbonyl sulfide (COS)

\[
CO_{(g)} + S_{(l)} \rightleftharpoons COS_{(g)}
\]
\[
CO_{(g)} + H_2O_{(g)} \rightleftharpoons CO_2(g) + H_2(g)
\]
\[
2CO_2_{(g)} + S_{(l)} \rightleftharpoons 2CO_{(g)} + SO_{2g}
\]
\[
H_2S_{(g)} + CO_{2(g)} \rightleftharpoons H_2O_{(g)} + COS_{(g)}
\]
\[
2CO_{2(g)} + 2S_{(l)} \rightleftharpoons COS_{(g)} + CO_{(g)} + SO_{2g}
\]
\[
CO_{2(g)} + H_2S_{(g)} \rightleftharpoons COS_{(g)} + H_2O_{(g)}
\]
\[
2CO_{2(g)} + 3S_{(l)} \rightleftharpoons 2COS_{(g)} + SO_{2g}
\]
\[
CO_{(g)} + H_2S_{(g)} \rightleftharpoons H_2(g) + COS_{(g)}
\]

Presence of hydrocarbons (e.g., methane) and formation of carbon disulfide (CS$_2$)

\[
CH_4_{(g)} + 2S_{2(g)} \rightleftharpoons CS_2_{2(g)} + 2H_2S_{(g)}
\]
\[
CS_{2(g)} + H_2O_{(g)} \rightleftharpoons COS_{(g)} + H_2S_{(g)}
\]
\[
CS_{2(g)} + CO_{2(g)} \rightleftharpoons 2COS_{(g)}
\]
\[
2CS_{2(g)} + SO_{2(g)} \rightleftharpoons 2COS_{(g)} + 3S_{(l)}
\]

Figure 5 shows the equilibrium constants (log K) of the various reactions at 135 °C. The thermodynamic quantity is a good indicator of the favorability or feasibility of a reaction although it gives no information on the kinetics. It can be seen that thermodynamic considerations point to COS formation is unfavorable for a CO$_2$ rich feed as compared to a CO rich gas stream. On a relative basis, the most favorable reactions are the ones where COS is consumed. Further, presence of hydrocarbons, such as methane, would have to be present for the formation of yet another sulfur component (carbon disulfide, CS$_2$).
Simulation results predict high solubility for CO₂ in DGM solvent compared to water. The equilibrium distribution ratios \( \left( \frac{y_{eq}}{x_{eq}} \right) \) at the reactor conditions (\( T = 270^\circ F, P = 770 \) psi) in DGM and water are respectively about 5 and 140. The high CO₂ solubility has the potential to form COS through reaction with H₂S. However, with low CO levels and SO₂ presence in the solvent, the \( 2CO_2(g) + 3S(l) \leftrightarrow 2COS(g) + SO_2(g) \) reaction is expected to be driven to the left, consuming any COS formed. Experimentation is required to quantify COS formation.

**Task 3 – Computer Simulation Modeling and Economics**

**UCSRP-HP Base Case in IGCC Applications with CO₂ Capture**

The UCSRP-HP Aspen Plus® model was completed. The computer model was used for process optimization and economic analysis. The model will also be used for designing and planning of pilot-scale unit for later phases.

The process design for the evaluation of UCSRP-HP (Case A) was based on DOE Case 2 of “Cost and Performance Baseline for Fossil Energy Plants”¹ for an IGCC plant with a net nominal 550-MW power generation. This is the GE gasifier for Illinois coal in IGCC
with carbon capture. At the request of DOE, a chemicals application (with a more stringent cleanup required to protect fuel cells or conversion catalysts from being poisoned) was also prepared as Design Case A-1. Table 2 shows the contaminant targets for the two cases. The UCSRP-HP cases were designed for the Chemicals specifications in both cases except for the H2S level. In the IGCC case, Design Case A, the H2S concentration in the feed gas to the gas turbine was ~8 ppm. In the chemical case, Design Case A-1, it was <50 ppb. The low level required the addition of a COS hydrolysis reactor and a guard bed added to the IGCC case as noted later.

The UCSRP-HP Design Case A keeps the DOE Case 2 design up through the first heat exchanger (HP Steam) in the Gas Cooling, BFW Heating & Knockout block where the syngas feed is at ~772 psia and 450 °F. This is downstream of the Quench and Scrubber Section and the Water Gas Shift Reactors. The DOE Case 2 is rejoined at the feed to the CO2 removal section of the dual-stage Selexol unit. This design replaces or eliminates the Mercury Removal, H2S-removal section of the dual stage Selexol unit, the Claus Plant, and Hydrogenation Reactor and Gas Cooler Section. Schematics of the DOE Case 2 and UCSRP-HP Base Case A are shown in Figure 6 and Figure 7, respectively. For these designs, the total SO2 emission from the UCSRP-HP IGCC power plant is about 34 lb/hr vs. 56 lb/hr specified in the DOE Case 2.

<table>
<thead>
<tr>
<th>Contaminant</th>
<th>Maximum after Cleanup</th>
</tr>
</thead>
<tbody>
<tr>
<td>Application</td>
<td>IGCC</td>
</tr>
<tr>
<td>Case</td>
<td>A</td>
</tr>
<tr>
<td>H2S</td>
<td>0.0128 lb SO2/MM Btu</td>
</tr>
<tr>
<td></td>
<td>50 ppbw</td>
</tr>
<tr>
<td>NH3</td>
<td>0.1 vol%</td>
</tr>
<tr>
<td>HCl</td>
<td>1 ppm</td>
</tr>
<tr>
<td>Hg</td>
<td>&gt;90% Capture</td>
</tr>
<tr>
<td>Se</td>
<td>0.2 ppm</td>
</tr>
<tr>
<td>As</td>
<td>5 ppb</td>
</tr>
<tr>
<td>Cd</td>
<td>30 ppb</td>
</tr>
</tbody>
</table>

The syngas stream from the HP Steam heat exchanger is further cooled to 165 °F and then processed in a high-pressure, co-current, down-flow Water Contactor unit to separate a large fraction of water present in the gas along with much of the NH3 and essentially all of the halogens and heavy metals, as sulfides or water-soluble salts. The sour gas feed enters Water Contactor, where it is contacted with a stream of circulating water. At the pressure, temperature, and water content of the syngas, the circulating water will have a steady-state content of NH3 and H2S. As a result, the HCl content of the feed gas will be absorbed very effectively to form highly soluble NH4Cl. A small but significant concentration of NH4HS will also be present in the liquid phase, and the heavy metals As, Cd and Hg will be absorbed to form their respective, insoluble sulfides. Selenium will be present in the syngas as H2Se and will be absorbed to form highly soluble (NH4)2Se under these conditions. At the bottom of the scrub contactor, the water stream is withdrawn and circulated by pump back to the top after dissolved gases are flashed and returned to the feed gas stream. A slipstream of the water stream will be
withdrawn for filtration and other treatment to remove the accumulated impurities, and then sent to the sour water stripper for water treatment. Following this step, the gas is preheated to about 250 °F prior to its processing in the UCSRP-HP reactor. The key objective for the removal of a large fraction of the water prior to the UCSRP-HP reactor is to minimize the cost of separation of water from DGM solvent used in the reactor.

The Aspen Plus® simulation UCSRP-HP model was used to identify a co-current, down-flow contactor reactor design that is simple and less expensive to build compared to the original counter-current designs. At the operating conditions of ~750 psia and 250-300 °F, the sulfur forms as a liquid, essentially immiscible in DGM/catalyst solvent and over twice as dense. This, along with the lower operating temperatures, overcomes the equilibrium limitations that occur in conventional gas-phase Claus reactor systems. The H₂S/SO₂ reaction is run at about 10% excess to drive the reaction to completion for the other component. A DGM slipstream is treated by hydrocyclones to remove any precipitated heavy metal salts that may not have been removed by the water filter. To be conservative, the design cases assume that some COS is formed within the UCSRP-HP reactor system.

The sulfur is separated, filtered by a DURCO sulfur filter, and sent to a sulfur pit or to a commercial-design O₂/sulfur submerged combustion furnace, as needed to generate the required liquid SO₂ for reaction with H₂S in the UCSRP-HP reactors. Ammonia from the DGM distillation unit is also fed to the sulfur furnace and converted to N₂ and H₂O as it passes through the furnace. The presence of S₂ vapor prevents any NOₓ formation. The combustion gas raises steam in the boiler and then passes through the condenser, where liquid sulfur is collected. The wet SO₂ gas then flows to a cooler, where liquid water, saturated with dissolved SO₂, is condensed. The SO₂ stream leaving the cooler is converted to liquid in another condenser, and then pressurized to the pressure of the reactor column by a pump.

The product syngas from the UCSRP-HP reactor is cooled to about 90 °F for (1) heat integration and (2) minimization of the loss of DGM solvent with the product syngas delivered to the IGCC plant. The cooled gas is sent to a high-pressure separator to recover DGM solvent that is processed in a distillation unit to remove the water (1) formed in the reactor due to the reaction of H₂S and SO₂, (2) present in the syngas feed to the reactors and (3) provide a lean DGM supply to the reactor.

For the chemicals case, to achieve the <50 ppb sulfur target, a COS hydrolysis unit is introduced in the UCSRP-HP reactor system to produce a syngas containing about 7.6 ppm H₂S and 0.4 ppm COS. A zinc oxide guard bed is also added downstream of the high-pressure separator to reduce the total sulfur level to <50 ppb in the effluent gas. If COS is shown not to form, the COS hydrolysis unit can be eliminated and the zinc oxide guard bed can be made smaller.

To be conservative, there are three means to collect the heavy metals in the process, namely (a) primarily by use of water filters in the Water Removal unit, and two other backup means: (b) the DURCO liquid sulfur filter, and (c) hydrocyclone/precipitation
filters to process a slip-stream of the recycle DGM solvent from the UCSRP-reactor. If experimentation shows that all of the heavy metals are collected in the Water Removal unit as solid sulfides, the backup means can be eliminated.

The sweet, cooled syngas is then transferred, as in the DOE Case 2 design, to the CO₂ removal section of the dual-stage Selexol unit and the power island. Designs were prepared to integrate UCSRP-HP with various higher-temperature, carbon capture technologies in development. For these cases, heat exchangers are used to bring the sweet syngas to the desired process temperatures.

**Process Economics**

We have followed the methodology given in the referenced DOE report to evaluate the UCSRP-HP process. Relative to the results for the DOE Case 2 that uses conventional cold gas cleanup scheme with Selexol/Claus/Tail Gas type H₂S removal processes, our study shows significant economic and environmental advantages (see Table 3) for the UCSRP-HP Base Case design:

- A net CAPEX savings of about $123 MM (Dec’06 dollars) based on the conservative design and ± 30% cost estimate basis;
- The overall thermal efficiency (HHV basis) would increase from about 32.5% for the DOE Case 2 to about 33.3% for the UCSRP-HP design;
- An increase of about 17.6 MW (~3.2 %) in net power sale;
- A reduction of about $9.60/MWhr (~9.3%) in the cost of power production with carbon capture, CO₂ compression plus transport/storage/monitoring; and
- A reduction in total SO₂ emission of about 22 lb/hr.
- In addition, preparing the syngas for a chemical application that requires <50 ppb sulfur in the syngas would increase the required CAPEX by about $10.1MM and the COE would increase by about $1.80/MWhr if the syngas was used for power generation in a near zero SO₂ discharge IGCC plant.

**Economic Potential for the Integration of Key Membrane Technologies for the Removal of CO₂ and H₂ with the UCSRP-HP Sulfur Removal Process in IGCC Applications**

The economic potential of integrating UCSRP-HP with (1) the LANL/SRI’s PBI (polybenzimidazole), high-temperature (~250-450 °C), polymeric CO₂-removal membrane, (2) the GTI/PoroGen’s CarboLock Membrane for CO₂ removal, and (3) an advanced H₂-separation membrane that would meet specific year-2015 performance targets assumed by DOE/Noblis LLC were also examined. For these economic evaluations, it is assumed that the CO₂ product would need to meet the Kinder-Morgan specifications for existing commercial CO₂ pipelines. These cases were compared to the published data on RTI’s warm gas cleanup (WGC) process. The COE data for these integration options are compared in Table 4 with those derived from literature information, namely:
Case 1  **DOE Case 1:** This case represents a nominal net 550 MW IGCC plant with no CO$_2$ capture using sulfur impregnated carbon beds for mercury removal, single-stage Selexol for H$_2$S removal, and Claus/Tail Gas for sulfur recovery.

Case 2  **DOE Case 2:** This case represents a nominal net 550 MW IGCC plant with CO$_2$ capture using the conventional cold gas cleanup (CGCU) of scheme sulfur impregnated carbon beds for mercury removal, dual-stage Selexol process for the removal of H$_2$S and CO$_2$, and Claus/Tail Gas for sulfur recovery.

Case A  This case uses Case 2 and incorporates UCSRP-HP for multicontaminant removal with conventional Selexol process for CO$_2$ removal as explained in the previous section of this report.

Case B  This case uses Case A (UCSRP-HP) with GTI/PoroGen’s CarboLock membrane contactor with Selexol for CO$_2$ removal rather than conventional columns.

Case C  This case is the published SRI study that uses LANL/SRI’s PBI membrane for hydrogen/acid gas separation and purification of the CO$_2$ stream by a single-stage Selexol/Claus/Tail Gas process for sulfur recovery.

Case D  This case uses Case C and incorporates UCSRP-HP for multicontaminant removal downstream of LANL/SRI’s PBI membrane rather than the Selexol/Claus/Tail Gas trains.

Case E  This case is the published DOE/Noblis report that uses RTI’s warm gas cleanup (WGCU) process for multicontaminant removal and a single-stage Selexol process for CO$_2$ capture.

Case F  This case is the published DOE/Noblis report that uses RTI’s WGCU process for multicontaminant removal and an Advanced H$_2$ Membrane (performance projections by DOE/Noblis) for CO$_2$ removal.

Case G  This case uses Case A and incorporates an Advanced H$_2$ Membrane for H$_2$/CO$_2$ separation.

Figure 8 shows how various developmental technologies affect the cost of electricity. The horizontal line marked “1” is the COE without carbon capture. Point “2” is the base case for carbon capture using “conventional” capture technology. The line 2-A-B-G represents incorporating UCSRP-HP for multicontaminant removal with (“A”) Selexol for CO$_2$ removal, (“B”) GTI/PoroGen’s CarboLock membrane contactor with Selexol rather than conventional columns for CO$_2$ removal, and (“G”) an Advanced H$_2$ membrane for CO$_2$ removal. Line 2-C-D represents incorporating the PBI membrane for

These cases show that UCSRP-HP can make a positive impact with new developmental technologies and could be a viable alternative to competing multicomponent cleanup technologies.

### Table 3 Comparative COE Data: DOE Case 2 vs. UCSRP-HP

<table>
<thead>
<tr>
<th></th>
<th>DOE Case 2</th>
<th>GTI UCSRP-HP Base Case</th>
<th>GTI UCSRP-HP Chemical case</th>
</tr>
</thead>
<tbody>
<tr>
<td>Coal Feed Rate, dry, lb/hr</td>
<td>444,737</td>
<td>444,737</td>
<td>444,737</td>
</tr>
<tr>
<td>SO₂ Emissions (lb/hr)</td>
<td>56</td>
<td>34</td>
<td>~0</td>
</tr>
<tr>
<td>Gas Turbine, MWe</td>
<td>464.0</td>
<td>464.0</td>
<td>464.0</td>
</tr>
<tr>
<td>Sweet Gas Expander, MWe</td>
<td>6.3</td>
<td>6.3</td>
<td>5.8</td>
</tr>
<tr>
<td>Steam Turbine, MWe</td>
<td>274.7</td>
<td>287.5</td>
<td>287.5</td>
</tr>
<tr>
<td>Total Auxiliaries, MWe</td>
<td>-189.3</td>
<td>-184.5</td>
<td>-184.5</td>
</tr>
<tr>
<td>Net Power for Sale, MWe</td>
<td>555.7</td>
<td>573.3</td>
<td>572.8</td>
</tr>
<tr>
<td>Thermal Efficiency, % (HHV)</td>
<td>32.5</td>
<td>33.5</td>
<td>33.4</td>
</tr>
<tr>
<td>Total CAPEX, $MM (Yr-2006 $)</td>
<td>1,328</td>
<td>1,205</td>
<td>1,215</td>
</tr>
<tr>
<td>Cost of Power, $/MWhr (or, mills/kWh)**</td>
<td>103.0</td>
<td>93.4</td>
<td>95.2</td>
</tr>
</tbody>
</table>

** DOE Economic Model to determine LCOE: Levelized Cost of Electricity**
Figure 6 DOE Case 2 Process Flow Diagram, GEE IGCC with CO$_2$ Capture (Total SO$_2$ Emission: 56 lb/hr)
Figure 7 GTI UCSRP-HP Base Case (Design Case A) with CO₂ Capture (Total SO₂ Emission: 34 lb/hr)
Table 4 Comparative Data for COE (including CO₂ capture, compression, transport, plus storage/monitoring)

<table>
<thead>
<tr>
<th>Design Case</th>
<th>H₂S Removal Option</th>
<th>CO₂ Removal Option</th>
<th>COE, $/MWhr</th>
<th>COE Differential, %</th>
</tr>
</thead>
<tbody>
<tr>
<td>DOE-1</td>
<td>Selexol</td>
<td>None</td>
<td>77.8</td>
<td>Base</td>
</tr>
<tr>
<td>DOE-2</td>
<td>Selexol</td>
<td>Selexol</td>
<td>103</td>
<td>32.4</td>
</tr>
<tr>
<td>A</td>
<td>UCSRP-HP</td>
<td>Selexol</td>
<td>93.4</td>
<td>20.1</td>
</tr>
<tr>
<td>B</td>
<td>UCSRP-HP</td>
<td>GTI/ PoroGen CarboLock Membrane Contactor w/Selexol</td>
<td>91.7</td>
<td>17.9</td>
</tr>
<tr>
<td>C</td>
<td>Selexol</td>
<td>LANL/ SRI PBI Membrane</td>
<td>98.2</td>
<td>26.2</td>
</tr>
<tr>
<td>D</td>
<td>UCSRP-HP</td>
<td>LANL/ SRI PBI Membrane</td>
<td>94.7</td>
<td>21.7</td>
</tr>
<tr>
<td>E</td>
<td>RTI</td>
<td>Selexol</td>
<td>101.6</td>
<td>30.6</td>
</tr>
<tr>
<td>F</td>
<td>RTI</td>
<td>Advanced H₂ Membrane</td>
<td>89.6</td>
<td>15.2</td>
</tr>
<tr>
<td>G</td>
<td>UCSRP-HP</td>
<td>Advanced H₂ Membrane</td>
<td>88.8</td>
<td>14.2</td>
</tr>
</tbody>
</table>
Figure 8 Comparative Data for COE Using Multiple Technology Pathways and Integrations

** See Table 4 for related design cases
CONCLUSIONS AND RECOMMENDATIONS

Conclusions reached in this project are:

- Aspen Plus simulations indicate complete removal of ammonia (NH$_3$), hydrogen chloride (HCl) and hydrogen selenide (H$_2$Se) in the aqueous scrub step of the UCSRP-HP process.

- Thermodynamic considerations point to minimal formation of COS in the UCSRP-HP reactor for a CO$_2$ rich feed as compared to a CO-rich-feed gas stream.

- For economic evaluations of various novel technologies for sulfur removal and carbon capture in IGCC applications, estimates for COE would depend on overall thermal efficiency for power generation as well as the capital cost requirements.
  - As an example, the RTI WGCU technology for sulfur removal has demonstrated relatively high thermal efficiency; however, based on the DOE/NETL estimates, the COE with carbon capture for Case E (RTI WGCU with single-stage Selexol for carbon capture) has been estimated at only ~1.4% lower relative to the DOE Case 2 (CGCU with two-stage Selexol).
  - In contrast, for Case A (using the UCSRP-HP technology for sulfur removal and the single-stage Selexol for carbon capture), a COE reduction of ~9.1% relative to the DOE Case 2 is estimated. This is primarily due to a significant savings in capital cost for the UCSRP-HP process.

- As indicated in the Design Case G, integration of the UCSRP-HP process for sulfur removal with an Advanced H$_2$ Membrane for H$_2$/CO$_2$ separation could lead to a COE value that would be only about 15% higher than DOE’s baseline cost projections (in DOE Case 1) for a no carbon capture IGCC plant.
  - The COE for Case G is similar (at about 15% increase relative to the no carbon capture case) to that estimated by DOE/Noblis for the Case F using the RTI process for sulfur removal and the Advanced H$_2$ Membrane process for CO$_2$ removal.

- Further laboratory- and bench-scale testing is required in preparation to a slip-stream test of UCSRP-HP on actual coal-derived syngas.

REFERENCES


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