**Novel Composite Membranes for Hydrogen Separation
in Gasification Processes in Vision 21 Energy Plants**

Final Technical Progress Report

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ABSTRACT

This report describes the work performed, accomplishments and conclusion obtained from the project entitled “Novel Composite Membranes for Hydrogen Separation in Gasification Processes in Vision 21 Energy Plants” under the United States Department of Energy Contract DE-FC26-01NT40973. ITN Energy Systems was the prime contractor. Team members included: the Idaho National Engineering and Environmental Laboratory; Nexant Consulting; Argonne National Laboratory and Praxair. The objective of the program was to develop a novel composite membrane structure for hydrogen separation as a key technology module within the future “Vision 21” fossil fuel plants.

The separation technology module is targeted for use within the gasification module of the “Vision 21” fossil fuel plant. The high performance and low-cost manufacturing of the proposed technology will benefit the deployment of “Vision 21” fossil fuel plant processes by improving the energy efficiency, flexibility and environmental performance of these plants. Of particular importance is that this technology will also produce a stream of pure carbon dioxide. This allows facile sequestration or other use of this greenhouse gas. These features will benefit the U.S. in allowing for the continued use of domestic fossil fuels in a more energy efficient and environmentally acceptable manner.

The program developed and evaluated composite membranes and catalysts for hydrogen separation. Components of the monolithic modules were fabricated by plasma spray processing. The engineering and economic characteristics of the proposed Ion Conducting Ceramic Membrane (ICCM) approach, including system integration issues, were also assessed. This resulted in a comprehensive evaluation of the technical and economic feasibility of integration schemes of ICCM hydrogen separation technology within Vision 21 fossil fuel plants.

Several results and conclusion were obtained during this program. In the area of materials synthesis, novel pyrochlore-based proton conductors were identified, synthesized and characterized. They exhibited conductivity as high as 0.03 S/cm at 900 °C. Long-term stability under CO₂ and H₂ atmospheres was also demonstrated.

In the area of membrane fabrication by plasma spray processing, the initial results showed that the pyrochlore materials could be processed in a spray torch. Although leak-tight membranes were obtained, cracking, most likely due to differences in thermal expansion, remained a problem. More modeling and experimental work can be used to solve this problem.

Finally the techno-economic analyses showed that the ITN ICCM approach for separating H₂ is comparable to conventional pressure swing adsorption (PSA) technology in efficiency and economics. Enhanced membrane flux and lower operating temperatures may make the ICCM approach superior to PSA.

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EXECUTIVE SUMMARY

ITN Energy Systems, along with its team members, the Idaho National Engineering and Environmental Laboratory, Nexant Consulting, Argonne National Laboratory and Praxair, propose to develop a novel composite membrane structure for hydrogen separation as a key technology module within the future “Vision 21” fossil fuel plants.

The ITN team is taking a novel approach to hydrogen separation membrane technology where fundamental engineering material development is fully integrated into fabrication designs; combining functionally graded materials, monolithic module concept and plasma spray manufacturing techniques.

The technology is based on the use of Ion Conducting Ceramic Membranes (ICCM) for the selective transport of hydrogen. The membranes are comprised of composites consisting of a proton conducting ceramic and a second metallic phase to promote electrical conductivity. Functional grading of the membrane components allows the fabrication of individual membrane layers of different materials, microstructures and functions directly into a monolithic module. Plasma spray techniques, common in industrial manufacturing, are well suited for fabricating ICCM hydrogen separation modules inexpensively, yielding compact membrane modules that are amenable to large scale, continuous manufacturing with low costs.

The ICCM hydrogen separation technology is targeted for use within the gasification module of the “Vision 21” fossil fuel plant. The high performance and low-cost manufacturing of the proposed technology will benefit the deployment of “Vision 21” fossil fuel plant processes by improving the energy efficiency, flexibility and environmental performance of these plants. Of particular importance is that this technology will also produce a stream of pure carbon dioxide. This allows facile sequestration or other use of this greenhouse gas. These features will benefit the U.S. in allowing for the continued use of domestic fossil fuels in a more energy efficient and environmentally acceptable manner.

This program developed and evaluated composite membranes and catalysts for hydrogen separation. Components of the monolithic modules were fabricated by plasma spray processing. The engineering and economic characteristics of the proposed ICCM approach, including system integration issues, were also assessed. This resulted in a comprehensive evaluation of the technical and economic feasibility of ICCM hydrogen separation for implementation within the “Vision 21” fossil fuel plant. Additionally, a patent application and an invention disclosure were submitted as part of this program.

Specific work and accomplishments in this program include:

- Development of novel, pyrochlore-based, proton-conducting materials for use in H₂ separation membranes from shifted syngas (a mixture of hydrogen and carbon dioxide);
- Demonstration of long-term stability of H₂ separation membranes under CO₂ and H₂ atmospheres;

- Development of plasma spray techniques for the manufacturing of thin-film, cermet membranes for H₂ separation;
- Demonstration of the utility of thin-film, cermet membranes for the separation of H₂ from syngas streams;
- Development of, and techno-economic analyses performed on, a reference plant and six different plant scenarios comparing ITN ICCM with pressure swing adsorption for H₂ separation; and
- Recommendations for future R&D to improve the techno-economic feasibility of the ITN ICCM approach for separating H₂.

An important finding from this program was that the ITN ICCM approach for separating H₂ is comparable to conventional pressure swing adsorption technology in efficiency and economics. While the ICCM component is only a small portion of the total IGCC plant cost, its performance, and ultimately costs, can have significant effect on the overall co-production economics. Several improvements in membrane performance were identified including: enhanced membrane flux; lower operating temperature; incorporating the water-gas-shift reaction within the membrane module; and increasing the stability of the membrane to contaminants in the coal gasification stream. These improvements in the membrane are likely to make the ICCM approach more economical than pressure swing adsorption towards H₂ separation from coal gasification streams. The best route for obtaining this goal is through additional materials discovery to enhance the membrane materials' fundamental properties.

SECTION 1. INTRODUCTION

This report describes the work performed, results obtained and conclusions drawn over the course of this program on the development and evaluation of “Novel Composite Membranes for Hydrogen Separation in Gasification Processes in Vision 21 Energy Plants.”

The technology developed in this program is based on the use of Ion Conducting Ceramic Membranes (ICCM) for the selective transport of hydrogen. The membranes are comprised of composites consisting of a proton-conducting ceramic and a second metallic phase to promote electrical conductivity. Functional grading of the membrane components allows the fabrication of individual membrane layers of different materials, microstructures and functions directly into a monolithic module. Plasma spray techniques, common in industrial manufacturing, are well suited for fabricating ICCM hydrogen separation modules inexpensively, yielding compact membrane modules that are amenable to large-scale, continuous manufacturing with low costs.

This program developed and evaluated composite membranes and for hydrogen separation. Specifically, new materials for the proton-conducting membrane were identified and characterized. Plasma spray manufacturing methods were also developed in order to determine the feasibility of this approach for membrane module manufacturing. Finally, the engineering and economic characteristics of the proposed ICCM approach, including system integration issues, were assessed. Results and conclusions from this work will now be described.

SECTION 2. EXPERIMENTAL

2.1 Task 2 - Membrane Materials Development and Evaluation

2.1.1 Experimental Apparatus

The experimental apparatus at ITN consisted primarily of two components: the equipment for fabricating ceramic powders and membranes and the equipment for determining the hydrogen flux through the membranes.

2.1.1.1 Powder and Membrane Fabrication

Equipment for the preparation of ceramic powders consisted of primarily low (1200 °C) and high (1700 °C) temperature furnaces operating in air. The low temperature furnaces used were Lindberg (BF51828C) and the high temperature furnaces were purchased from Deltech (DT-31), both with integral temperature controllers. Auxiliary laboratory equipment, such as balances, stirrers and mixers, were conventional laboratory type equipment. A vibratory sieve shaker (VRW, AS200B) was also purchased for powder preparation in this program.

Equipment for fabricating membranes consisted of a uniaxial press with heated platens (Carver, model #1617) for pressing membrane disks. Dies for the press were purchased commercially (Carver). A high temperature tube furnace (Carbolite, STF16) with an integral controller was used to sinter the membranes under a reducing atmosphere.

Powders and membranes were evaluated by x-ray diffraction (XRD) located at the Colorado School of Mines (Seimens Kristalloflex 810). Scanning electron microscopy (SEM) was also performed on the powders and membranes located at the Colorado School of Mines (JEOL JXA-A40). Optical microscopy was performed at ITN. (Zeiss Axiovert 25 and Olympus SZH). Samples for SEM and optical microscopy were prepared at ITN using standard metallographic techniques and equipment (Buehler Isomet 1000 saw and Leco VARI/POL polisher). Membrane density was determined using the standard Archimedes method.

2.1.1.2 Membrane Test Equipment

The experimental apparatus for evaluating the hydrogen flux through the sintered membranes was designed at ITN. A process flow diagram of the apparatus is shown in Figure 2-1. Membranes were sealed onto metal tubes (discussed below) and inserted into the measurement apparatus. The experimental apparatus consisted of a vertical tube furnace capable of operation up to 1200 °C. The mounted membranes were attached to an outer Inconel 600 tube using compression fittings. The sample tube was of sufficient length so that the compression fittings were out of the heating zone. Gas streams (feed, sweep, permeate and retentate) were added or removed using separate stainless steel tubing attached to the outer Inconel tube with compression fittings, all outside of the heating zone. Likewise, temperature measurement was performed with thermocouples inserted into the Inconel tube through compression fittings. In this way, the experimental chamber was separated from the outside environment. A photograph of the membrane test apparatus mounted in the furnace is shown in Figure 2-2.

Gas flows were controlled through mass flow controllers (MKS). Specific gases used included H₂, N₂, He, Ar and CO₂. Before entering the membrane test stream, the gas stream was

Figure 2-1. Process flow diagram for experimental apparatus to evaluate H₂ separation membranes.

humidified, if desired, by sparging through a heated stainless steel vessel. The vessel temperature was controlled using a heating tape and temperature control system. This allowed for controlled humidity in the feed stream. The flow of the permeate stream was measured with either a gas flow meter (MKS) or a soap film flowmeter. The composition of the permeate stream was measured using a computer controlled gas chromatograph (GC) (Shimadzu GC-8AIT). The gas stream was injected directly into the GC via an injection valve. Measurement of the composition and flow rate allowed for the calculation of hydrogen flux through the membrane.

2.1.2 Preparation of Pyrochlore Membrane Materials and Cermet Membranes

2.1.2.1 Powder Preparation Methods

Two methods were used for preparing membrane powders: the conventional solid-state method and the Pechini process. For the conventional solid-state method, the appropriate metal oxides and carbonates were used as the starting materials. These precursor powders were dry-milled in polyethylene jars with cylindrical zirconia grinding media for 12 hours. The powders were then

fired at 1200 °C for 12 hours and then ground slightly by hand. Powders were then analyzed by XRD.

The Pechini process is a technique for preparing very fine metal oxide powders.¹⁻³ In this process, stock solutions of the desired metal ions were prepared by dissolving the appropriate metal nitrate salts (obtained commercially) in water at a concentration of approximately 2M. The metal content of these solutions was determined gravimetrically by conversion of the metal nitrate solution to the metal oxide by heating at 900 °C.

To prepare the powders with this second technique, the appropriate quantities of the metal nitrate solutions were weighed out and combined in a large Pyrex beaker. An aqueous solution of citric acid and ethylene glycol in equimolar amounts was also prepared. The citric acid/ethylene glycol solution was then added slowly and with stirring to the metal nitrate solution. The molar ratio of total organics to total metal nitrate was 1:1. This combined solution was then heated slowly with stirring on a hot plate at approximately 70 °C to remove the water. The solution thickened into a gel without precipitate formation. When the bulk of the water (~75%) was removed, the gel began to decompose as evidenced by NO_x formation. This decomposition process resulted in a very fine ash. This ash was recovered and then ground by hand in a mortar and pestle. The ash was then transferred to an alumina crucible and heated to 900 °C for two hours. The final product was a very fine powder and analyzed by XRD to confirm that the proper phase had formed.

2.1.2.2 Membrane Preparation Methods

Cermet membranes were prepared at ITN by uniaxial pressing followed by sintering in a reducing atmosphere. Powders were prepared initially by the solid-state method as described above but later prepared by the Pechini process due to the higher sintered densities obtained with the latter powders.

Regardless of the powder preparation technique, cermet membranes consisting of both the ceramic powder and nickel, or nickel alloy, were prepared in a similar manner. The nickel or nickel alloy was obtained commercially as a powder with particle size <1 μm. The nickel to ceramic powder ratios were varied from 20 to 80 wt%. The metal and ceramic powders were milled together using zirconia media. Polyvinyl alcohol served as the binder and was added to

Figure 2-2. Photograph of membrane test stand showing the membrane test assembly.

the powder mixture at 2 wt% from a 20 wt% aqueous solution. The resulting slurry was dried and subsequently passed through a 50 mesh sieve. The powder was then placed in a cylindrical die and uniaxially pressed. The membranes were fired first at 300 °C in air for binder removal and then under 3% H₂/Ar for sintering. Sintering temperatures varied from 1300-1450 °C for 2-8 hr, depending upon the specific metal. Membranes prepared in this manner had sintered densities typically >90% as measured by the Archimedes' method. Total porosity was typically <10%. Figure 2-3 shows a photograph of a typical cermet membrane fabricated in this manner.

Figure 2-3. Photograph of a cermet membrane consisting of a pyrochlore proton conducting ceramic and nickel.

2.1.3 Membrane Flux Measurements

2.1.3.1 Sealing

The first step in evaluating the membranes was to seal the membrane onto a metal tube for insertion into the experimental apparatus. Several techniques for sealing were evaluated over the course of this program. Based on discussions with Argonne National Laboratory,^{4,5} we initially decided to pursue brazing as the technique for sealing the cermet membranes into the membrane test assembly. Two commercially available brazes were selected for evaluation: a gold/nickel/vanadium alloy (Gold-ABA-V, Wesgo Metals) with a liquidus point of 1090 °C and a gold/copper alloy of 35/65wt% (Lucas-Milhaupt) with a liquidus point of 1015 °C. Both of these brazes are available as foils. Brazing was performed by heating the assembly under 3% H₂/Ar to temperatures up to approximately 50 °C above the liquidus point of the individual brazes.

The initial attempts to achieve a completely leak-free seal were unsuccessful for several reasons. The initial attempt at brazing used a geometry in which the brazing foil was situated between the edge of the membrane and the inner diameter of the holder. This is illustrated in Figure 2-4A. Attempts to seal the membrane to the holder in this manner were unsuccessful due to the large gap between the membrane and the holder wall. It was also determined that the vanadium in the Gold-ABA-V braze was becoming oxidized during the brazing process. We therefore began using carbon as a getter to remove residual oxygen or water in the gas stream. This solved the vanadium oxidation problem.

We then switched to a geometry in which the braze foil was situated between the membrane face and the holder face as illustrated in Figure 2-4B. This geometry appeared to yield seals that exhibited less of a leak but still were not acceptable. The main reason for this was that there was still a large gap between the membrane face and the holder due to warping of the membrane which occurred during sintering of the membrane. To reduce the warping, a creep-flattening step

was performed. For this process, the cermet membrane was heated at 1420-1450 °C for 8 hr under 3% H₂/Ar with a ceramic weight placed on it. This drastically reduced the warping of the sample. To obtain an even flatter surface, the membranes were surface-ground after the creep-flattening step using 120 grit silicon carbide paper. The membranes resulting from this process exhibited warping of less than 100 µm which was deemed suitable for brazing.

Figure 2-4. Schematic illustration of two different brazing geometries. A: Edge sealing. B: Face sealing.

Several additional changes were also made to the sealing procedures. The most important of these was switching to braze alloys in the form of washers, rather than foils. This allowed for coverage of the brazing surface on the holder without disruptions. Additionally, ensuring that the samples were flat and situated evenly within the holder helped in obtaining gas-tight seals.

An additional important change to the sealing process was switching to a two-step process in which the membrane was first sealed to a Ni washer using a higher temperature braze (Gold-ABA-V, Wesgo Metals) and then followed by brazing the membrane/washer assembly into the holder with a lower temperature braze (gold/copper, 35/65 wt%, Lucas-Milhaupt).

Despite these changes, obtaining gas-tight seals was still not routine with an unacceptably high failure rate. Looking at the proton conducting materials, it was hypothesized that a potential mismatch in coefficients of thermal expansion (CTEs) was causing the problem. Specifically, the large volume of the holder was causing stresses in the much thinner membranes.

In order to reduce this problem, the membrane holder was redesigned as shown in Figure 2-5. Specifically, an additional part was added to separate the membrane from the holder. Here, the membrane is brazed onto the top sealing surface (A). After brazing, the stem of the membrane holder is welded into the test housing assembly C as well as to the gas manifolding system. The top of this assembly is then closed off by welding the mating component onto it as before. This whole assembly is then placed into the furnace. This change aided in obtaining lower leak seals but was still inconsistent.

Two additional modifications were made to the sealing method. The first was to use pure gold washers for sealing, rather than brazes. The second was to use a Pyrex paste as the sealing material. Both of these approaches appeared to be more consistent, particularly when used in the

new membrane holder. Most of the testing in the latter stages of the program was performed with these methods.

Figure 2-5. Drawing of re-designed sample holder.

SECTION 3. RESULTS AND DISCUSSION

3.1 Task 1 - Test Plan and Project Management

Task 1 was focused on project management. This included contract negotiations and development of the project plan. Communications, reporting, presentations and intellectual property development all fell under this Task.

Communications consisted of regular teleconference calls between ITN and the various subcontractors. Several conference calls and meetings with DOE personnel were also made. Meetings and presentations included the following:

- Kick-off meeting on December 11, 2001 at ITN with representatives from all subcontractors and DOE personnel;
- M. Schwartz presentation at the Vision 21 Contractor's Meeting in Morgantown, WV, in November 2001;
- M. Schwartz presentation at the 27th International Technical Conference on Coal Utilization & Fuel Systems, March 2002;
- M. Schwartz presentation at the Pittsburgh Coal Conference, September 2002;
- M. Schwartz presentation at the 225th American Chemical Society Meeting under the Fuel-Division-Sponsored Symposium "Advances in Membranes for Energy and Fuel Applications," March 2003;
- M. Schwartz presentation at the 14th Annual Meeting of the North American Membrane Society, May 2003;
- A meeting was held between ITN, Nexant and DOE representatives during the 20th Annual International Pittsburgh Coal Conference, September 2003;
- M. Schwartz presentation 20th Annual International Pittsburgh Coal Conference, September 2003;
- Site visit of DOE personnel to ITN in December 2003. Representatives of all subcontractors were also present;
- A site visit of ITN and Nexant personnel to DOE/NETL took place in April 2004;
- M. Schwartz presentation at the 15th Annual U.S. Hydrogen Conference, April 2004.

One U.S. patent application entitled "Novel Proton Conducting Materials and Devices Incorporating Them" was filed by ITN in May 2004. A joint ITN and INEEL invention disclosure entitled "Thermal Spray Fabrication of an Electron and Proton conducting Membrane for Hydrogen Separation" was filed November of 2003.

3.2 Task 2 - Membrane Materials Development and Evaluation

3.2.1 Preparation of Pyrochlore Materials and Cermet Membranes

3.2.1.1 Preparation of Proton-Conducting Powders

The objective of this Task was to identify and evaluate suitable proton conducting materials for incorporation into cermet, hydrogen separation membranes. Both known and new membrane materials were developed. Results obtained in this area will now be discussed.

Initial work performed under this Task focused on fabricating samples of proton conducting powders. The first sample prepared was $\text{BaCe}_{0.9}\text{Gd}_{0.1}\text{O}_{2.95}$ (BCGO) in order to continue to develop this material beyond the earlier work at Argonne National Laboratory.⁴ This material has been prepared phase-pure as shown in Figure 3-1.

Attempts were then made to prepare sintered membranes in the form of thin disks consisting of a mixture of the BCGO powder and nickel powder. Initial membrane disks were fabricated as

Figure 3-1. XRD pattern of BCGO indicating phase-purity of the material prepared at ITN.

described above but showed low sintered density, <70 %. The procedure was repeated as a function of sintering temperature until the melting point of Ni was reached but higher densities could not be obtained.

One possible reason for the problems obtaining higher sintered densities is that the BCGO powder was reacting with the aqueous binder system. Therefore, experiments were performed in which the membrane disks were fabricated by dry-pressing and using binder systems in which the solvents were alcohol rather than water. Dry-pressing proved to be unsuitable for fabricating the membrane disks as the lack of strength in the green membrane disks prevented their handling. The use of alcohol solvents proved to be an improvement. However, maximum sintered densities of only approximately 80% were obtained.

In addition to the perovskite proton-conducting phases that have already been well studied, *i.e.* BCGO, we also fabricated and evaluated pyrochlore phases that were recently identified as proton conductors.⁶⁻⁸ This class of compounds is exemplified by $\text{La}_2\text{Zr}_2\text{O}_7$. Yttrium doping of this phase has been shown to greatly enhance proton conductivity.^{8,9}

In this program, the strategy that was directing the development of new materials compositions was to vary the stoichiometry and oxygen defect concentration within the basic pyrochlore structure, $\text{A}_2\text{B}_2\text{O}_7$. For example, by changing the A and B cations, the basicity, and therefore the propensity to incorporate protons into the lattice, can be changed. This change can be made independently of the oxygen stoichiometry. The latter parameter can also be changed independently by incorporating dopants into the lattice. This also can increase the concentration of interstitial protons in the lattice which will in turn increase the proton conductivity. However, it needs to be noted that changing the A and B cations, as well as introducing dopants, will also change structural parameters of the material. This may have a negative effect on the conductivity. Of course, it is possible that changes to the structural parameters can also have a positive effect. Creating distortions in the lattice so as to enhance the mobility of charge carriers has been explored for oxide ion conductivity in these materials¹⁰⁻¹² but should be applicable to proton conductivity as well.

As described in the experimental section above, initial attempts at preparing powders of the materials were made using the conventional solid-state method. Figure 3-2 compares XRD patterns of $\text{La}_2\text{Zr}_{1.8}\text{Y}_{0.2}\text{O}_{6.9}$ powders fired at 1250 °C and 1600 °C, respectively. Even at the highest reaction temperature, the material is not phase pure as small amounts of La_2O_3 still remain (noted in Figure 3-2 by asterisks). Additional samples based on $\text{La}_2\text{Zr}_{1.8}\text{Y}_{0.2}\text{O}_{6.9}$ with different doping schemes were prepared but the XRD patterns of all these materials indicated that although they formed in the pyrochlore structure, they still contained a small amount of one of the starting materials.

The literature indicates that these materials can be prepared single-phased.⁶⁻⁸ Therefore it appeared that the problem was one of incomplete reaction, despite the high reaction temperatures. One way to achieve complete reaction is to fire the materials at even higher temperatures or longer times. However, this will result in powders with large particle size which in turn will make it difficult to achieve a high density membrane. This is particularly important in light of the problems processing BCGO as described above.

Figure 3-2. X-ray diffraction patterns of $\text{La}_2\text{Zr}_{1.8}\text{Y}_{0.2}\text{O}_{6.9}$ fired at 1250 (red) and 1600 (blue) °C. The asterisks represent peaks due to La_2O_3 .

To overcome this processing problem, we developed a second fabrication procedure based on the Pechini process. The experimental details of this powder preparation process were given above in the Experimental section. Figure 3-3 shows the XRD pattern of a typical pyrochlore sample prepared in this manner. The pattern shows no reflections due to any of the constituent metal oxides indicating that it is single-phased. Additionally, the broad reflections are indicative of a very fine powder. This should be compared to the previous results (Figure 3-2) in which the pyrochlore powders always contained some impurities. This result indicated that the Pechini process was suitable for obtaining phase pure pyrochlore powders.

An additional concern with the traditional proton-conducting materials has been the stability of the materials in the presence of carbon dioxide at high temperatures. An experiment was performed in which a proton-conducting pyrochlore powder was exposed to a gas stream of 50% carbon dioxide in hydrogen for six hours at 900 °C. XRD was performed after this exposure and no evidence for carbonate formation was found either in the diffraction pattern or through visual examination. This is not surprising as this composition does not contain any alkaline earth metals which are particularly susceptible to carbonate formation. Although this experiment was

performed under atmospheric pressure and for a relatively short time, it indicates that these materials should be stable to carbon dioxide during membrane operation.

Figure 3-3. XRD pattern of a doped pyrochlore prepared using the Pechini process.

3.2.1.2 Preparation of Cermet Membranes

Cermet membranes were fabricated as described above in the Experimental section. The use of fine powders, prepared by the Pechini process, greatly increased sintered densities. However, porosity still caused a high leakage rate. Several routes were evaluated in order to reduce this leakage.

One route was the use of a Ni/B (15 wt%) alloy in place of nickel as the metallic component in the cermet membrane. The strategy here is that this alloy has a lower melting point than nickel, 1065 vs. 1453 °C, respectively, which would promote liquid phase sintering. Additionally, Ni/B alloys are known to have better wetting properties than metals towards metal oxides and this may cause stronger bonding, and higher density, than with the pure nickel.^{13,14} However, membranes with acceptable density and porosity were not obtained.

Other alloys were also evaluated and were identified based on metallic components that oxidize easier than nickel and might therefore enhance the wetting between the metal and ceramic components of the membranes. Specific alloys investigated during this reporting period included Ni/Zr (30wt%), Ni/Al (21wt%) and Ni/Ti(10wt%). Membranes were fabricated in a manner identical to the membranes using pure nickel except that the sintering temperatures were adjusted so that they were 10-30 °C below the melting point of the specific alloy.

The membrane containing the Ni/Zr alloy shattered upon sintering. This is most likely due to volume changes associated with absorption of hydrogen by the alloy component. We were unable to fabricate membranes with low porosity with the other membranes as well. Several factors were contributing to this. In the case of the Al-containing alloys, the aluminum component leached from the metal as determined by visual inspection as well as microprobe analysis. The titanium-containing membranes appeared to melt before sintering. This work was not pursued further.

We also evaluated the use of sintering agents in attempts to get even higher densities. To overcome this problem, we began to investigate metal oxides as potential sintering agents. Sintering agents examined included SiO_2 , MgO , BaO , Na_2O and Li_2O and were all obtained from commercial sources. These sintering agents were added to the ceramic powder as either the oxide or carbonate, as appropriate, and milled thoroughly. Membranes were prepared using the procedures discussed previously including maintaining a sintering temperature of 1420-1440 °C. However, none of these sintering aids were effective in reducing membrane porosity.

We then evaluated sintering aids prepared from mixtures of these compounds and now obtained membranes with acceptable densities. Specifically, with a sintering aid mixture of 7 wt%, a sintered density 100% of theoretical was obtained. This is in contrast to a typical density of 75% of theoretical when no sintering aid is used. Additionally, the porosity of the membranes decreased from 34% to 7% when the sintering aid was used. Figure 3-4 shows a comparison of membranes fabricated with and without a sintering aid. As can be seen in the Figure, the sample with the sintering aid had a much smaller diameter, reflecting greater shrinkage and higher density.

The reason that much higher densities were obtained for membranes using the sintering aid mixture as compared to the single component sintering aid is not known. However, it is likely that the combination of Na_2O and/or Li_2O with SiO_2 leads to lower melting phases than what is obtained for the pure components. This then leads to greater sintering, although chemical interactions between the sintering aid components and the ceramic component of the membrane, resulting in a lower sintering temperature, cannot be ruled out.

Experiments were then performed to determine the lowest quantity of sintering aid that could be used and still obtain low porosity. It was determined that roughly 3 wt% was the minimum amount of sintering aid that resulted in acceptably low porosity.

Finally, an alternate sintering aid was examined. Previous sintering aids we had studied contained silica which could potentially reduce overall conductivity by segregating into the grain

Figure 3-4. Photograph comparing membranes prepared with (A) and without a sintering aid (B).

boundaries and acting as an insulating layer. This new sintering aid is based on a metal phosphate and allowed us to sinter the metal-free ceramics to greater than 95% density.

Another reason for evaluating a phosphate-based sintering aid is that recent reports in the literature have indicated that phosphates can reduce to metal phosphides which are electrically-conducting.¹⁵ This would permit the development of a second, electrically conducting phase within the pure ceramic component, resulting in a mixed ionically and electrically conducting ceramic suitable for a hydrogen separation membrane. Membranes prepared in this manner showed very low total conductivity, presumably due to a lack of electrical conductivity. Alternatively, the second phase, if segregated in the grain boundaries, could have blocked proton conductivity.

3.2.2 Hydrogen Flux Property Characterization of Cermet Membranes

Successful completion of the membrane fabrication work allowed us to begin testing membranes for hydrogen flux. Some of the first results obtained are shown in Figure 3-5. This example is of one composition which consists of a pyrochlore proton conductor with 50wt% Ni. The membrane was tested under several different atmospheres containing H₂, N₂, CO₂ and H₂O. Specific atmospheres are shown in Figure 3-5. At 50% H₂ in N₂, a flux rate greater than 0.3 ml/min-cm² (normalized to 1mm thickness) was achieved. The membrane was stable over

Figure 3-5. Plot of H₂ flux rate vs. time in different feedstreams.

operation at 900 °C for several days. The sample was also thermally cycled several times over this period with no loss of flux rate. One important point is that these membranes did not have any catalysts on the surfaces and in fact had been polished smooth during the flattening procedure. This would tend to result in lower flux rates as the surface areas would be greatly reduced by a lack of roughness.

Figure 3-5, as well as Figure 3-6, also show the effect to of the introduction of both CO₂ and H₂O. Here the flux was decreased relative to the dry H₂ in N₂ feedstream. This drop was attributed to a decrease in the H₂ concentration in the feedstream due to the presence of water and the reverse water gas shift reaction:

Measurement of the gases in the feedstream indicated that CO was being produced. We were unable to measure the exact composition of the feedstream because of Reaction 3-1. The results in Figure 3-5 also indicated that the membrane is stable over at least several days with both H₂O and CO₂ in the feedstream.

Figure 3-6. Plot of H₂ flux rate as function of the CO₂:H₂ ratio in the feedstream for atmospheres that are either dry or contain H₂O in the H₂O:H₂ ratio of 0.6.

Results from additional compositions are summarized in Figure 3-7. All measurements were taken at 900-915 °C with a feedstream of 50% H₂ in N₂ without any steam. The data shown in Figure 3-7 are all normalized to 1mm thickness.

Composition 1 in Figure 3-7 is La₂Zr_{1.8}Y_{0.2}O_{6.9} (LZYO), a composition that has previously been reported as to exhibit proton coconductivity.^{8,9} As can be seen from the figure, the novel pyrochlore phases that ITN has developed show substantially higher flux rates than LZYO.

One problem with the data shown in Figure 3-7 is that we were not able to reproduce the flux rate for the composition with the highest flux rate, Composition 3. A second sample of this composition was prepared and tested and showed a flux rate that was lower by over an order of magnitude. This is illustrated in Figure 3-8. The reason for this is not clear. One possibility is that it was observed that Sample A (high flux) had a much higher leak rate, 20-30%, than Sample B, 5-10%. Although the flux data has been corrected for leakage, the higher leakage rate in Sample A may indicate some microcracking. This would both increase the active surface area and decrease the effective thickness of this sample relative to a sample that showed less cracks

Figure 3-7. Plot of thickness-normalized H₂ flux rate as a function of time for several pyrochlore cermet membranes at 900 °C and 50% H₂ in N₂.

Figure 3-8. Plot of thickness-normalized H₂ flux rate as a function of time for two different samples of the same composition.

and less leakage. The overall effect is that the flux rate would appear higher for the sample with more microcracking than the alternate sample.

In addition to flux rates, it is also possible to extract conductivity data for these samples with certain assumptions. For a mixed ion-electron conductor, the flux rate is derived from the Wagner model and given by:¹⁶

$$J = \frac{\sigma_{\text{amb}} RT}{4F^2 l} \ln \frac{P_{\text{out}}}{P_{\text{in}}} \quad (3-2)$$

where J is the flux rate in mol/cm²-sec, σ_{amb} is the ambipolar conductivity in S/cm (given below), R is the Molar Gas Constant, T is the absolute temperature, F is Faraday's constant, l is the thickness and P_{out} and P_{in} are the H₂ partial pressures in the permeate and feedstreams, respectively.

The ambipolar conductivity is given by:

$$\sigma_{amb} = \frac{(\sigma_{ion})(\sigma_{elec})}{\sigma_{ion} + \sigma_{elec}} \quad (3-3)$$

where σ_{ion} and σ_{elec} are the proton and electronic conductivity, respectively. Assuming the electronic conductivity is much greater than the ionic conductivity, then σ_{amb} reduces to σ_{ion} . Since the electronic conductivity in these cermet membranes is due to the presence of Ni, we believe that this is a valid assumption. An additional assumption for Equation 3-2 to be valid is that the flux is limited by diffusion of the protons through the membrane, not surface exchange reactions. Again, we believe this to be valid because of the relatively thick membrane used in these experiments.

Table 3-1 summarizes the conductivity extracted for the samples measured. As expected, the highest flux membrane has the highest conductivity. One important observation is that the conductivity measured for LZY0 is very similar to the reported value of approximately 2×10^{-3} S/cm.⁹ This validates the above assumptions as well as our measurement techniques.

Table 3-1

Material	H₂ Flux Rate (ml/min-cm²) Normalized to 1mm Thickness	Conductivity (S/cm)
Composition 1 (LZY0)	0.05	2.6×10^{-3}
Composition 2	0.31	1.7×10^{-2}
Composition 3		
Sample A	1.00	0.15
Sample B	0.14	9.5×10^{-3}

One major issue associated with proton conducting cermet membranes is the temperature of operation. Any decrease in the operating temperature will have a beneficial impact on the overall economics of the hydrogen separation process. To this end, we evaluated the temperature dependence of the hydrogen flux rate through the pyrochlore-based membranes.

Figures 3-9 and 3-10 show the results of these experiments. Figure 3-9 shows the dependence of the proton conductivity as a function of temperature for both wet and dry atmospheres. As expected, the conductivity in a wet atmosphere is slightly higher, most likely due to the presence of additional charge carriers. Figure 3-10 shows a plot of $\ln(\sigma T)$ as a function of inverse temperature. A fit of a straight line to this data yields the activation energy (E_a) for proton conduction, assuming Arrhenius behavior. Values of 0.56 and 0.57 eV are obtained for dry and wet atmospheres, respectively. The wet and dry activation energies are similar, indicating that no major structural changes or changes in conduction mechanism are occurring upon exposure to the wet atmosphere. The observed activation energies are relatively low and similar to other

Figure 3-9. Plot of conductivity vs. temperature for a Ni (40v%)/pyrochlore proton conducting membrane.

Figure 3-10. Plot of $\ln(\sigma T)$ vs. inverse temperature for a Ni (40v%)/pyrochlore proton conducting membrane.

proton conductors.^{17,18} These results indicate that lower temperature operation of the membrane is possible without a great loss in flux rate.

To further enhance conductivity of the pyrochlore-based materials, we expanded the doping scheme discussed above to include the alkaline earth metals in the pyrochlore A lattice site and ceria in the B lattice site. All materials could be prepared single-phase as determined by XRD.

Cermet membranes with nickel were prepared from these powders and conductivity measured according to the procedures detailed above. Conductivity for the various samples were in the 0.01-0.03 S/cm range, typical of what we had observed earlier. Samples appeared to be stable under water and carbon dioxide atmospheres. The addition of water boosted conductivity slightly, as observed in earlier samples.

Following the rationale detailed above, additional material compositions were generated. Figure 3-11 shows conductivity data for a new composition with the highest conductivity observed in the program. As can be seen in the Figure, the conductivity at 900 °C is 0.061 S/cm. The conductivity increases in a wet atmosphere. The material is also stable in water and carbon dioxide atmospheres.

In addition to new compositions, we also examined the effect of metal concentration on the conductivity of the membranes. Higher metal concentrations, 70-80%, did not seem to have a major effect on the overall conductivity.

Figure 3-11. Plot of conductivity vs. time for a pyrochlore-based, cermet membrane in both wet and dry atmospheres.

3.3 Task 3 - Thin-Film Device Fabrication

The objective of this Task was to develop the techniques for fabricating thin-film cermet membranes using thermal spray technology. INEEL had responsibility for developing the spray techniques with assistance from ITN.

Initial work in this Task focused on preparing cermets on porous pure metallic substrates in the form of discs. The substrates were purchased commercially (Mott) and were stainless steel or nickel. For the ceramic, the initial work utilized yttria-stabilized zirconia (YSZ) since this material is commercially available for thermal spray and has a similar composition and crystal structure to the pyrochlore-based proton conductors which are the focus of this program.

Figure 3-12 shows an example of one of the first Ni/YSZ cermets with approximately 40 vol% Ni thermally sprayed onto a porous (0.2 μm porosity) stainless steel disk. This membrane proved to be impermeable when pressurized on one side.

Figure 3-12. Cross section micrograph (100x) of a thermally-sprayed Ni/YSZ membrane onto a porous stainless steel disk.

The success obtained for the thermally sprayed Ni/YSZ cermet allowed us to switch from YSZ to a pyrochlore proton conducting material. To this end, ITN prepared one kilogram batches of powder of a composition developed at ITN for deposition by INEEL. The initial experiments showed that cermet membranes could be fabricated similar to the Ni/YSZ membranes fabricated earlier. However, these membranes were permeable. Cross-sectional micrographs (Figure 3-13) indicated that the porosity was due to some large particles of the ceramic when compared to the membranes sprayed using commercial YSZ powder. We therefore prepared additional batches in which the powder has been sieved to obtain a more uniform particle size.

Figure 3-13. Cross-sectional micrographs of plasma-sprayed cermet membranes at 100x (top) and 200x (bottom). A.) Ni/pyrochlore membrane using pyrochlore ceramic powder prepared at ITN. B.) Ni/YSZ membrane using commercially available YSZ.

A second area pursued was in designing and fabricating new substrates for the thin-film membranes. These substrates consisted of a 41 mm diameter porous Ni disk (porosity 0.5 μm) welded into a nickel washer. With this arrangement, the cermet membrane were deposited directly onto the porous disc and washer surface and then brazed into the experimental apparatus for testing.

As noted above, one reason for the porosity in the pyrochlore-based membranes was poor powder morphology. Proton-conducting powders of the composition used previously were fabricated using the conventional solid-state synthesis technique rather than the Pechini process. Care was taken with these powders during the preparation process in order to get the best characteristics for the spray process. Specifically, powders were calcined at 1400 $^{\circ}\text{C}$ for 12 hr. Additionally, the powders had minimal post-sintering grinding and were sieved -150 mesh. Using these new powders, INEEL successfully fabricated cermet membranes from the pyrochlore proton conducting powders with a thickness of several hundred microns.

Figure 3-14 shows a typical sample from the initial batches of membranes that INEEL fabricated. This membrane consisted of ~40% Ni by volume and was 360 μm thick. The sample had been sprayed onto a porous Ni substrate that was mounted into a nickel ring. This assembly was then mounted into the ITN test system. The data presented are with a feedstream of 50% H_2 in N_2 . The sample had a leak rate of less than 2%.

Figure 3-14. Plot of H₂ flux as a function of operating time for a thermally-sprayed pyrochlore cermet membrane operating under 50% H₂/N₂.

The data in Figure 3-14 show a flux rate of approximately 0.4 ml/min-cm² for a 0.36 mm thick sample or approximately 0.1 ml/min-cm² for a 1 mm thick sample. This flux rate should be compared to a value of 0.3 ml/min-cm² (normalized for 1 mm thickness) that was observed for this same material in the form of a dense sintered disk.

To try to understand this discrepancy, micrographic analyses were performed on a cross-sectional piece of membranes from this batch. Figure 3-15 shows a typical example. Since the preparation of the cross-section can close the

Figure 3-15. Cross-sectional micrograph of a thermally-sprayed membrane on a porous Ni substrate. Note the densified layer at the substrate/membrane interface.

porosity in the substrate, this sample was subjected to an etch of 10% HNO_3 in water. The micrograph clearly shows that the porous substrate had densified at the interface between the sprayed cermet membrane and the substrate.

Upon examining the fabrication procedures, it was hypothesized that a grit-blasting step, used to promote adhesion between the sprayed film and substrate holder, was causing the densification. The sprayed membranes previous to this batch had not undergone this procedure since the membranes were sprayed onto unmounted substrates. The grit-blasting step was added when we switched to the mounted substrates. Subsequent experiments comparing substrates grit-blasted with those not grit-blasted positively identified this step as causing the substrate surface densification.

The next batch of membranes was grit-blasted through a mask that protected the substrate surface and subsequent cross-sectional analysis showed that the densification did not occur to the extent that it occurred in the previous batch (Figure 3-16). This sample had a thickness of 420 μm and showed a flux rate of 0.7 ml/min-cm^2 or 0.3 ml/min-cm^2 when normalized to 1 mm. This flux rate is now comparable to that observed for the thick, dense membrane. However, the leak rates were roughly 20%, substantially higher than the earlier, sprayed membrane samples.

To solve this problem, we used a more powerful spray gun. This allowed for denser films and results in decreased porosity. Another change made was to return to the grit-blasting procedure for enhancing adhesion between the film and porous substrate. However, as discussed above, the grit-blasting procedure closed some of the porosity of the substrate surface. Therefore, we investigated using substrates with higher porosity, 0.5-10 μm , rather than the 0.2 μm porosity substrates used to date. The strategy was that the surface of the larger porosity substrate is inherently rougher than the finer porosity substrate and therefore grit-blasting may not be necessary. Additionally, if grit-blasting the surface of the higher porosity substrate will be less likely to result in loss of porosity.

Initial experiments fabricating thin-film membranes on stainless steel porous substrates with porosity as high as 10 μm were performed and the membranes did not show leakage at room temperature. Subsequent testing at ITN for hydrogen flux showed that these membranes exhibited low leak rates, some less than 1%.

Figure 3-17 shows a plot of membrane conductivity as a function of operating time under H_2 and $\text{H}_2/\text{H}_2\text{O}$ atmospheres as well as after one thermal cycle. The membrane consisted of ~40 wt% Ni and was ~500 μm thick. The results are all at 900 $^\circ\text{C}$. As can be seen in the Figure, the

Figure 3-16. Cross-sectional micrograph of a thermally-sprayed membrane on a porous Ni substrate. Note that the interface between substrate and membrane is more porous than that shown in Figure 3-15.

conductivity of the sample was ~ 0.03 S/cm and was completely stable over the 200 hr experiment. Thermal cycling also had no effect on the membrane performance. Conductivity in a steam-containing atmosphere was higher than in the dry atmosphere, as expected for a proton conductor.

Figure 3-17. Plot of conductivity vs. time for a thermally-sprayed membrane at 900 °C. See text for the experimental details.

The effect of a CO₂-containing atmosphere, as well as the membrane leak rate, are shown in Figure 3-18. The data are reported as a thickness-normalized H₂ flux rate, rather than conductivity as in Figure 3-17, since we are not sure of the H₂ concentration due to the reaction of the CO₂ with the H₂ to produce H₂O and CO in the reverse water-gas-shift reaction (Equation 3-1).

The data shown in Figure 3-18 show that the membrane is completely stable towards CO₂-containing atmospheres over the time tested. The membrane also showed a very low leak rate, $\sim 0.5\%$, over the course of the experiment. Not only did the flux remain stable, but the leak rate as well, indicating that these pyrochlore proton-conducting materials will be stable under actual operating conditions.

A high surface area membrane, ~ 100 cm², was also prepared. A photograph of this membrane is shown in Figure 3-19. The membrane showed a high leak rate as it was prepared on a stainless steel substrate, rather than Ni, as Ni substrates were not available at the time of the experiment. The use of the stainless steel substrate required the use of stainless steel as the metal in the cermet membrane, rather than Ni.

In order to obtain thinner and lower leak membranes, and to determine the reproducibility of the process, additional membrane samples were generated by varying the deposition parameters.

Figure 3-18. Plot of normalized H₂ flux rate as a function of time for a thermally-sprayed membrane in a CO₂-containing atmosphere at 900 °C.

One important parameter investigated was the speed of the thermal spray head as it passed across the substrate surface. This has the effect of producing thinner layers per pass across the substrate. The membrane then would be built up by a greater number of thinner layers. This resulted in membranes of reduced thickness but still exhibiting very low leak rates.

Another parameter determined to be important was the use of an inert gas shroud during the spray process. This has the effect of keeping the nickel from oxidizing during the spray process. Nickel oxide present in the membrane will reduce to nickel upon exposure to hydrogen

during membrane operation. This in turn causes porosity in the membrane. Elimination of nickel oxide during the fabrication process is then expected to lower porosity and leakage rates.

A third parameter that was investigated is the ceramic powder. All thermally-sprayed membranes had been prepared using powder fabricated at ITN using conventional ceramic powder

Figure 3-19. Photograph of a thermally-sprayed, high surface area membrane.

processing techniques. Powders of an ITN proprietary pyrochlore composition were prepared commercially by Praxair Surface Technologies (Woodinville, WA). The powders were spray-granulated as part of the preparation procedure. Membranes prepared from these powders exhibited a more uniform microstructure with lower porosity than those prepared from powders fabricated at ITN (Figure 3-20). The reason for the difference is not clear but may have to do with the better flowability of the commercial powders. This leads to a more homogenous mixture entering the torch. The commercial powders are also more spherical which leads to a more homogeneous melting during the spray process. Independent of the cause, the observation of lower porosity thin films using the commercial powders is an important result in that it indicates that thinner membranes are possible with better powders. It also indicated that the compositions being investigated are amenable to large-scale manufacturing which will be important as we scale up the membrane size.

Figure 3-20. Comparison of microstructures of thermally-sprayed cermet membranes using powder prepared conventionally at ITN (A) and commercial powder (B).

A summary of most of the sprayed membranes to date is shown in Figure 3-21. Over forty membranes were fabricated and hydrogen flux rates measured. The figure shows the conductivity measured and the leak rate as expressed as a percentage. There is a variation in the conductivity measured for all the samples but an even greater variation in leak rate.

Some of the parameters that were varied include:

- Powder composition and specific batch;
- Powder preparation and morphology;
- Membrane thickness;
- Substrate porosity – 0.5, 2 and 10 μm ;
- Various torch parameters such as the specific gun, power, flow rate, gas used and deposition rastering pattern; and
- Spray gun condition.

Figure 3-21. Plot of conductivity and leak rate for a range of thermally-sprayed cermet membranes.

The data show that it is possible to obtain a conductivity of 0.03 S/cm with low leak rates in a reproducible manner.

Another important result obtained is the effect of thermal cycling on membrane performance. Figure 3-22 shows a plot of the conductivity and leak rate as a function of the number of cycles. In this experiment, the membrane was heated from room temperature to 900 °C at a rate of 5 °C/hr, held at temperature for several hours and then cooled to room temperature at the same rate. No evidence for membrane performance degradation was observed.

In attempts to fabricate membranes down to 50 μm, we fabricated membranes using a newer, higher-power torch. This work was performed with the help of Northwest Mettech Corp. They manufacture a unique thermal spray coating system called the Axial III. The 600 Series Axial III Plasma Torch was developed to withstand the harsh environment of nitrogen/hydrogen plasmas. This plasma chemistry was discovered to have very beneficial properties such as high deposition efficiencies and feed rates that yield dense, hard coatings when ceramic materials are sprayed. With the Axial III torch, powders are injected directly into the center of the plasma with little or no classification of the powder feedstock. Since the powders are virtually 100% entrained in the plasma flame, very efficient heating and melting of the feedstock materials takes place. One of the benefits of efficient particle heating and melting is higher density coatings.

After developing a set of spray parameters for the Axial III torch that produced dense cermet coatings of surrogate materials (zirconia and nickel), coatings of the pyrochlore-based ceramic with nickel were formed on 10 μm diameter porous metal substrate. Figures 3-23 and 3-24 are both cross section views of these membranes. The cross section shown in Figure 3-23 is a relatively dense and crack free portion of the membrane that is well adhered to the porous metal

Figure 3-22. Plot of conductivity and leak rate as a function of the number of thermal cycles for a thermally-sprayed, thin film membrane.

substrate. However, Figure 3-24 shows a portion of the same membrane that is separated from the substrate and cracked. This crack probably initiated at the interface due to differential thermal expansion. As coatings are built up layer-by-layer, residual stress between the substrate and coating can also build up. These stresses are generated from several different sources. As the coating is being applied, the membrane cermet material and substrate come to an elevated processing temperature together. As they cool back to room temperature, the differential thermal expansion increases the residual stress, resulting in failure. To reduce the residual stress that is a

Figure 3-23. Cross section of membrane fabricated using the Northwest Mettech Corp. Axial III torch.

result of the differential thermal expansion, Northwest Mettech attempted to control the processing temperature to a maximum of 10 to 20 °C above room temperature using cooling air jets. Although this significantly improved the integrity of the membrane, it was still not completely defect-free and only a minor improvement in the leak rate was achieved.

In addition to experiments on pyrochlore-based materials, experiments were also performed with BCGO as the ceramic portion of the cermet membrane. Initial attempts at fabricating thin-film BCGO were unsuccessful. However, later attempts, using the knowledge gained from the experiments with the pyrochlore materials, were successful in that leak-free films were produced. We believe that this was a combination of better

powder preparation and an improved deposition process. The conductivity measured for BCGO-based membranes was approximately 0.01 S/cm in a dry atmosphere and slightly higher in a wet atmosphere. This is lower than what would be expected from the membranes prepared using conventional ceramic processing techniques.

3.4 Task 4 – Engineering

This task was the responsibility of Nexant, with input from Praxair. Its primary objectives were to (1) establish various conceptual IGCC plant designs with which the application of ITN's Ion Conducting Ceramic Membranes (ICCM) for high-temperature H₂ separation as a Vision 21 technology component module can be identified and system integration issues defined; and (2) perform plant design, process heat, and material and utility balances, and

capital cost estimates to allow preliminary ICCM techno-economics to be evaluated. A current state-of-the-art IGCC conceptual plant design was chosen as a reference plant. Step-wise design modifications were implemented to the power-production-only reference plant design to allow for H₂ co-production, with and without CO₂ capture, and to facilitate the comparison of ITN ICCM membranes with the conventional H₂ separation technology of PSA. Higher temperature 'Warm Gas Clean Up' technology currently being developed by the DOE was evaluated as a complementary technology for ICCM. Process IGCC plant design models were developed to assist with the comparative studies and to ensure that the analyses are carried out on a consistent basis. Task 4 had three subtasks:

- Task 4.1 Membrane Module Design;
- Task 4.2 and 4.3 Process Integration and Preliminary Techno-Economic Analysis

Task 4.1 had the objective of identifying/quantifying the membrane properties and module design variables that are critical for successful application of ITN's membrane technology to an IGCC Vision 21 co-production facility. It was intended to provide the needed R&D targets to guide the membrane material development that are essential in achieving the performance and cost improvement required for commercialization of ITN's ICCM technology for IGCC application. The size of the membrane module, the number of membranes and their sizes, and their connectivity, *etc.* all has strong impacts in the equipment costs, and the installation and operability of the membrane facility.

Figure 3-24. Cross section of membrane fabricated with the Axial III torch showing delamination and crack caused by differential thermal expansion.

Detailed membrane module design can be very complex, and it was beyond the current scope of work. Task 4.1 was limited to preliminary one-dimensional modeling analysis to identify various membrane module criteria (*e.g.*, membrane size and configuration), operating conditions (*e.g.*, pressure effects of hydrogen partial pressure pinch, *etc.*), estimation of membrane surface area requirement that are pertinent to a commercial scale IGCC plant integration plant design and economic analysis of Tasks 4.2 and 4.3.

Tasks 4.2 and 4.3 had the objective of defining the various process integration issues and evaluating the techno-economic viability of the ITN ICCM hydrogen separation membrane as a Vision 21 technology component module for future IGCC plant designs. Collectively, Tasks 4.2 and 4.3 represent the bulk of the Nexant's efforts for the project.

3.4.1 Task 4.1 - Membrane Module Design

A spreadsheet-based model, *H2M.xls*, was developed for this purpose. The model has been used to set the design parameters for the ITN membrane model subsystem in the overall IGCC/H₂ membrane conceptual process design, integration and techno-economic analysis of Tasks 4.2 and 4.3, the results of which are reported in the next section.

3.4.1.1 Modeling Scheme, Equations and Theory

A schematic of an ion transport membrane module is shown in Figure 3-25. The following assumptions have been used in the development of the hydrogen separation membrane model:

Figure 3-25. ICCM unit schematic.

- Steady-state, 1-dimensional, finite-difference model
 - Isothermal
 - H₂ flux calculated from ΔP across membrane
 - Ion transport step rate controlling
 - Flux proportional to $\ln(P_{H_2}(\text{retentate}))/\ln(P_{H_2}(\text{permeate}))$
 - No leakage of other gases (CO, CO₂, H₂O & Inerts) across membrane
- Homogeneous WGS reaction kinetics are included, but may be turned ‘on’ or ‘off’
- Deactivation and other side reactions are ignored, including
 - Effect of contaminants
 - FT synthesis, coking, etc.

The governing differential equation in the spread sheet model is given above (Equation 3-2, page 17). The spreadsheet model can be easily recoded, in the future, to consider other mechanisms for membrane transport using other flux relationships.

A backward finite difference formulation is used to represent the governing differential equation (Equation 3-2) in the spreadsheet model. The values calculated in any given row are only dependent on values in the preceding row. Therefore, iteration is not required to achieve convergence at each point in the grid. This is particularly convenient in the spreadsheet format, since it allows the grid to be expanded by clicking on a row and dragging it downward; thus, copying all cell formulae and updating all cell references.

Model Description

The Excel™ workbook *H2M.xls* contains two permanent worksheets. These are *H2-ITM* and *CONSTANTS*. The latter contains predefined constants as Excel Named Cells, which can be accessed throughout the workbook (*e.g.* the molecular weight of H₂, *MW_H2*, one million, *MM*, and the conversion factor for psi to atm units, *psi_to_atm*). The former worksheet, *H2-ITM*, contains the actual membrane unit mathematical model. Neither of these sheets should be deleted, since this would make the model inoperable. In addition, it is recommended that once a feasible solution to a given membrane simulation is complete, it be saved separately, either as a new workbook or as a copy of the *H2-ITM* sheet. Most of the information provided in *H2-ITM* is presented in several different sets of dimensional units.

Worksheet *H2-ITM* is broken into several functionally different sections. Working from the top to the bottom of the active area of the worksheet, Block *A5:M35* summarizes the inlet and outlet temperatures, pressures, flowrates, and stream compositions, and the membrane performance parameters: CO conversion, hydrogen recovery, flux, and H₂ partial pressure approach. The model assumes a planar membrane geometry consisting of a stack of alternating membrane, membrane supports and flow channels. Block *A37:L59* contains geometry-related parameters for the membrane unit: overall plate dimensions, membrane layer and channel thickness, segment size to be used for numerical integration, feed superficial velocity, and computed size characteristics of the complete membrane stack.

Block *A61:H70* contains simplified hydraulic calculations for flow through a single retentate flow channel. Based on a specified feed superficial velocity, the actual flow velocity, pressure drop, and residence time through the unit are estimated. These calculations are used to verify that the conditions specified result in laminar flow with a reasonable pressure drop. Block *A72:J97* contains calculations for estimating the economic performance of the membrane unit. The membrane unit capital investment and annual capital charge are estimated and compared to the relative values of the feed and product streams. The unit profitability is contained in cell *I97*. Excel Solver™ can be used to adjust the superficial velocity to maximize membrane unit profitability by adjusting the number of plates per unit. Other unit profitability maximizations are also possible.

Block *A93:I110* contains membrane specific properties, such as conductivity, activation energy, and technology improvement factor. The technology improvement factor allows the user to easily adjust membrane performance. For example, Excel GoalSeek™ can be used to adjust this parameter to match either experimental results or R&D performance targets. A flag can also be set here to turn the water-gas-shift (WGS, Equation 1) reaction on or off. This flag has been provided because of uncertainty in the kinetic parameters for the WGS reaction at the operating conditions of the ITN membrane. It is also possible that the ITN membrane, currently or with future modifications, may possess catalytic activity. The flag allows sensitivity cases involving the WGS reaction to be modeled.

Finally, the worksheet rows *116* through *1163* contain the actual numerical grid used to integrate and solve the membrane model. The H_2 partial pressures and flux, and the flowrates and concentrations of all molecular species contained in the retentate are reported for each point along the membrane. The grid has been made larger than the actual dimensions of the membrane, so that the user can change the dimensions of the membrane or the step size of the grid without adding to or modifying the cells and rows in the grid. This region of the worksheet is named *PROFILES*, and built-in Excel Lookup functions are used to extract information from this region that is used in the calculation of some of the model output discussed above. Information reported in the grid cells outside of the physical dimensions of the membrane are ignored in calculating performance parameters. Additional columns of information are calculated when the WGS reaction is considered. This includes the net rate of production or consumption of H_2 by the water-gas-shift reaction.

Formulas used in the *H2-ITM* worksheet, refer to several Visual Basic for Applications™ coded functions provided with the workbook. These are *RateWGS*, *KeqWGS*, and *TeqWGS*, which are used to calculate the net rate of reaction, and the equilibrium constant and temperature, respectively, for the WGS reaction.

Input Data

These data fall into several categories:

- Stream Properties
 - 1) Feed Pressure (psia)
 - 2) Feed Temperature (°C)
 - 3) Feed Flowrate (lb mol/hr)

- 4) Feed Composition for H₂, CO, CO₂, H₂O, other gases (mole %)
- 5) Permeate Pressure (psia)
- 6) Steam Addition ratio to the feed flow (mol/mol)
- Membrane Assembly Dimensions
 - 1) Plate Width (cm)
 - 2) Plate Length (cm, assumed equal to plate length)
 - 3) Length of segment used for numerical calculations (cm)
 - 4) Thicknesses (μm) of individual components making up membrane: Feed Channel, Surface Catalyst, Ion Transport Membrane, Porous Support, and Permeate Channel

The assembly can be easily programmed for other stack arrangements.

- Hydrodynamics
 - 1) Feed Superficial Velocity per channel (cm/sec)
 - 2) Viscosity of feed stream (cP)
- Economic Parameters
 - 1) Equipment manufacturing cost per unit of membrane area (\$/sq.ft.)
 - 2) Cost of installation as percentage of inside-battery limit (ISBL) installed cost
 - 3) Outside-Battery Limit (OSBL) cost as percentage of total ISBL cost
 - 4) Other One-Time Costs (OTC) as percentage of total installed costs (ISBL + OSBL)
 - 7) Annual capital recovery factor as percentage of total investment (ISBL + OSBL + OTC)
 - 8) Other fixed costs as percentage of total investment
 - 9) Feed and product values (shadow prices): feed value (\$/MM Btu), assumed same as fuel value of waste gas, feed temperature penalty (\$/M lb) steam value (\$/M lb), permeate value (\$/MM Btu), and permeate pressure penalty (\$/MM Btu)
- Membrane Properties
 - 1) Membrane Conductivity (Ohm⁻¹-cm⁻¹) at 900°C
 - 2) Energy of Activation (electron-Volts)
 - 3) Technology Improvement Factor
 - 4) Flag for turning on/off Water-Gas-Shift (WGS) reaction

Output

The output is generally reported in several sets of dimensional units. The following information is predicted by the model:

- Permeate and Retentate Molar Flowrates and Compositions

- CO Conversion and H₂ Recovery
- Average Flux across membrane and H₂ Partial Pressure Pinch at outlet
- Membrane Unit Mechanical Design
 - 1) Number of Flow Channels
 - 2) Total Stack Height
 - 3) Total Stack Area
 - 4) Total Stack Volume
 - 5) Stack Area-to-Volume Ratio
- Hydraulics
 - 1) Average velocity at flowing conditions
 - 2) Reynolds Number
 - 3) Pressure Drop
 - 4) Average Residence Time
- Unit Economics
 - 1) Membrane Unit Cost
 - 2) Cost of Installation
 - 3) OSBL Costs
 - 4) Other One-Time Costs
 - 5) Capital Charge
 - 6) Other Fixed Costs
 - 7) Feed Costs
 - 8) Product Revenues
 - 9) Unit Profitability
- Profiles through Membrane Unit (per channel)
 - 1) H₂ Partial Pressure in Retentate (psia)
 - 2) H₂ Partial Pressure in Permeate (psia)
 - 3) H₂ Flux across membrane (g-mole/cm²-s)
 - 4) Incremental H₂ transport across membrane (g-mole/cm²)
 - 5) Incremental H₂ production from WGS reaction (g-mole/cm²)
 - 6) H₂ Flow in retentate (g-mole/-s)
 - 7) CO Flow in retentate (g-mole/-s)

- 8) CO_2 Flow in retentate (g-mole/-s)
- 9) H_2O Flow in retentate (g-mole/-s)
- 10) Flow of Inerts in retentate (g-mole/-s)
- 11) Total Retentate Flow (g-mole/-s)
- 12) Mole Fraction H_2 in Retentate
- 13) H_2 Flow in Permeate (g-mole/-s)
- 14) Cumulative H_2 Productivity, H_2 recovered / $\text{H}_2 + \text{CO}$ in feed
- 15) Total Concentration in Retentate, (g-mole/liter)
- 16) H_2 Concentration in Retentate, (g-mole/liter)
- 17) CO Concentration in Retentate, (g-mole/liter)
- 18) CO_2 Concentration in Retentate, (g-mole/liter)
- 19) H_2O Concentration in Retentate, (g-mole/liter)
- 20) Inerts Concentration in Retentate, (g-mole/liter)
- 21) Net Rate of Production of H_2 from WGS reaction, (g-mole/liter-s)

Results and Discussion

A series of model runs were made to determine the effect of the water-gas shift (WGS) reaction on the performance of the ion transport membrane. These are shown in Figures 3-26 and 3-27. Case 1 shown in Figure 3-26 is the base case using a previously shifted feed gas. No WGS

Figure 3-26. Hydrogen productivity analysis, Cases 1-5.

activity is assumed. The H_2 productivity rises gradually along the length of the membrane and begins to flatten out towards the exit of the membrane unit. Productivity is approximately 70%. Surprisingly, if the homogeneous WGS reaction does occur within the unit, it results in a lower productivity of roughly 60%, as shown by Case 2. This results from raising the temperature prior to membrane separation. At the membrane temperature, equilibrium is shifted back toward CO and water, lowering the potential hydrogen production from this feed stream. Case 3, like Case 1, assumes no WGS activity but uses an unshifted feed gas. The productivity is much lower, approximately 34%. With this feed, the productivity rises to about 40%, if the membrane unit has WGS activity (Case 4). Additional steam injection (Case 5) has very little effect on productivity for this syngas feed composition.

Figure 3-27 shows additional cases, which examine the effect of improved separation and increased WGS activity. Improving the membrane separation five fold, only marginally improves the productivity from about 40% to 46% (Case 6 vs. Case 5). However, the separation occurs more readily and the productivity gain flattens out at a much shorter length. If this membrane improvement is coupled with a nine fold increase in WGS activity (Case 7), the increase in productivity is substantial, approaching the 70% achieved in Case 1 with previously shifted feed gas. In the limit as the activity increases, WGS equilibrium is achieved (Case 8), resulting in a maximum productivity of about 72%.

Figure 3-27. Hydrogen productivity analysis, Cases 6-8.

Figure 3-28 shows the effect of increasing the membrane size on the net revenue from operation of the membrane unit. As the H_2 recovery rises rapidly, the cost of additional membrane area is more than offset by the increased hydrogen production. As the membrane is made larger, a limit is reached where the additional H_2 production cannot pay for itself and net revenue declines. The

Figure 3-28. The effect of membrane size on revenue.

maximum hydrogen recovery for this analysis was approximately 90%. No WGS activity was assumed.

Appendix A contains the derivation of the theoretical limit for H₂ recovery from membrane separation with no WGS activity, Appendix B contains the derivation of the theoretical limit for CO conversion to H₂ with no membrane separation, and Appendix C contains the derivation of the theoretical limit for H₂ productivity from membrane separation with the WGS reaction. These limits may be compared to the results of the model, which predicts actual performance given finite dimensions for the membrane unit.

In summary, it is critical that the H₂ separation membrane be optimized for any given operation. This includes considering whether the WGS reaction will be carried out upstream of this unit or internally. Theoretical limits exist to the maximum productivity that can be achieved with membrane separation. At this point in time, very little information is available on both the catalytic and non-catalytic activity of the WGS reaction in this application. Clearly, more research will be necessary to enable the accurate design of these units to be carried out.

3.4.2 Tasks 4.2 and 4.3 - Process Integration and Preliminary Techno-Economic Analysis

The objective of these subtasks was to establish various conceptual IGCC plant designs for the application of ITN's Ion Conducting Ceramic Membranes (ICCM) for high temperature H₂ separation as a Vision 21 technology component module and to evaluate its techno-economic viability in comparison with existing technologies. For H₂ separation, the current technology of pressure swing absorption (PSA) is used as the basis for comparison. A state-of-the-art IGCC conceptual plant design was chosen as a reference plant. Step-wise design modifications were implemented to the power-production-only reference plant design to allow for hydrogen co-production, with CO₂ capture, as anticipated components for a future Vision 21 IGCC facility. High temperature syngas clean up is evaluated as a complementary technology to ICCM.

Process plant design models were developed to assist with the comparative study and to ensure that the analysis is carried out on a consistent basis. Specific scope of work included (a) defining the overall process arrangements which efficiently integrate the ITN ICCM membrane into an IGCC plant to co-produce H₂ under conditions of with or without CO₂ capture; (b) performing preliminary overall heat and material and utility balances to establish overall plant performances; (c) determining system capacities or equipment sizes for capital cost estimates; (d) estimating annual revenues and operating and maintenance expenses to determine power and H₂ production costs; and (e) evaluating ITN ICCM techno-economic variability and any technology developments, if needed, to continue its commercial development. Work performed under Tasks 4.2 and 4.3 is described and summarized in the following sections:

Task 4.2.1. Study Basis

Task 4.2.1.1 Design Basis

Task 4.2.1.2 Economic Assumptions and Evaluation Methodology

Task 4.2.2. IGCC Conceptual Design Scenarios

Task 4.2.2.1 Reference Case IGCC

Task 4.2.2.2 Case 1 – IGCC Power Production with CO₂ Capture

Task 4.2.2.3 Case 2 – IGCC Hydrogen & Power Co-Production with CO₂ Capture (Conventional PSA)

Task 4.2.2.4 Case 3 – IGCC Hydrogen & Power Co-Production with CO₂ Capture (ICCM Membrane with Conventional AGR)

Task 4.2.2.5 Case 4 – IGCC Hydrogen & Power Co-Production with CO₂ Capture (ICCM Membrane with Warm Syngas Clean Up)

Task 4.2.2.6 Case 5 – Coal Based Hydrogen Production with CO₂ Capture (Conventional PSA)

Task 4.2.2.7 Case 6 – Coal Base Hydrogen Production with CO₂ Capture (ICCM Membrane)

Task 4.2.3. Results and Discussion

3.4.2.1 Study Basis

Design Basis

By agreement with DOE, the following is the design basis that was assumed for the study, along with the rationale used in selecting the key IGCC technology components that are common to the various case studies.

- Coal Feed and Design Site Conditions
 - Illinois No. 6 with typical analysis shown in Table 3-1.
 - Typical Mid-Western USA location with climatic and site conditions as specified in Table 3-2.
- Gasification and Associated High Temperature Heat Recovery Technology
 - E-GasTM gasification process.
 - The reason that E-GasTM gasifier was chosen for the study over other advanced oxygen-blow entrained flow gasifiers was based on design yield and cost information availability

in a recently completed DOE study [DE-AC26-99FT40342] of “Gasification Plant Cost and Performance Optimization.”

- Air Separation Technology
 - Commercial technology of cryogenic separation.
 - Advanced Ion Transport Membrane as a developing technology for O₂ production, which potentially can be a key component in a future Vision 21 IGCC energy plant, is not considered in order to minimize any design and cost uncertainty that may compromise the results of the current comparative study.
- Power Generation
 - GE 7FA+e (210 MW) turbine as described in the cited DOE study DE-AC26-99FT40342 above.
 - Integration onto the air separation unit for improved overall IGCC efficiency was not considered. A fully-integrated ASU/gasification/gas turbine design will result in better overall IGCC efficiency but it would not affect the result of our comparison study.
- Product Specifications
 - Hydrogen, with compositions and battery limits of temperature and pressure as shown in
 - Recovered CO₂ of typical pipeline compositions, battery temperature and pressure limits as shown in Table 3-3

Table 3-1 – Illinois #6 Coal Analysis

Ultimate as Received (AR) Analysis, Wt%

Carbon	59.87
Hydrogen	4.27
Nitrogen	1.11
Sulfur	2.21
Oxygen	7.07
Chlorine	0.11
Ash	10.86
<u>Moisture</u>	<u>14.50</u>
Total Wt%	100.00
HHV, Btu/lb (AR)	10,900
LHV, Btu/lb (AR)	10,295

Table 3-2 – Study Site Conditions

Atmospheric Pressure, psia	14.4
Ambient Dry Bulb Temperatures, °F:	
Maximum	93
Minimum	-20
Average	60
Ambient Wet Bulb Temperatures, °F:	
Summer Design	65
Ambient Air Composition, Vol%:	
Nitrogen	77.27
Argon	0.93
Oxygen	20.73
Water	1.04
<u>Carbon Dioxide</u>	<u>0.03</u>
Total Vol%	100.00
Relative Humidity, % Saturation	60

Table 3-3 – Product Specifications and Battery Limits

H ₂ Product	
Maximum Vol% H ₂	99.0
Minimum Vol% CO ₂	Not Available
B/L Pressure, psig	1000
B/L Temperature, °F	100 Maximum
Recovered CO ₂	
Minimum Vol% CO ₂	97
Maximum H ₂ O Content ppmw	50
B/L/ Pressure, psig	2000
B/L Temperature, °F	100 Maximum

Task 4.2.1.2 Economic Assumptions and Evaluation Methodology

For the study, economic evaluations and ICCM techno-economic feasibility comparison is based on the break-even analysis; with the resultant product price defined as the break-even price that will balance the plant's annual incomes against the annual expenditures. Table 3-4 shows a set of economic assumptions and assumed unit costs that were used for the study, and Table 3-5 a set of annual expenditure and income accounts used for the analysis.

As an example, for break-even H₂ pricing, Hydrogen Sales (line n, Table 3-5.) is calculated so that the Pre-Income Tax Total (line o) Annual Expenses is equal to Annual Incomes. Also, there will be no Federal (line p.) and State (line q.) income tax charges because there is no net income.

Investment Repayment (line j.) is an amortized annual repayment on the Initial Investments over life of the plant at the specified interest rates (return on investment). These annual repayments are equivalent to the depreciation charges on capital investment plus the interest expenses. Total initial capital investment will be developed for the various case studies and will have the breakdown shown in Table 3-6.

Cost of land, which is a part of the initial project investment, is excluded from the analysis. It is assumed that land will appreciate at the same rate as the return on investment so that land cost (plus interests) will be salvaged at end of the plant life.

Capital cost estimate for the various case studies were developed based on sized major equipment from process heat and mass balances to the maximum extent possible. Direct labor and bulk material costs are factored from and added to the major equipment cost to generate the total direct cost. Field indirects are added based on Nexant's in-house experience to generate the total field cost. Home office charge and contingency are then added to obtain a total installed cost for economic analysis.

Table 3-4 – Economic Assumptions

Plant Life, years	25
Initial Operating Funds, months	3
Land Cost, \$/ft ²	Excluded
Onstream Factor (@ Full Capacity), %	80
Coal Feed Cost, \$/Short Ton(As Received)	29.4 (Note 1)
Electricity Price, \$/kW-hr:	To be determined
Export	
Import (required for H ₂ production only analysis)	0.04 (Note 1)
Raw Makeup Water (River) Cost, \$/1000 gal	0.25
Byproduct Values:	
Recovered CO ₂ Credit, \$/ST	To Be Determined
Sulfur/Sulfuric Acid, \$/ST Sulfur Equiv.	30/60
Slag & Ash, \$/ST (Dry Basis)	- 10
Waste Water, \$/1000 gallons	0
Annual Maintenance Costs:	
Material, % of Total Plant Cost	1
Labor, % of Total Plant Cost	2
Labor Salary & Benefits:	
Average Salary, \$/Year	70,000
Benefits & Overheads, % of Salary	35
Insurance Costs, % of Total Plant Cost	1
Interest or Return on Investment, %	10
Annual Taxes:	
Local & Property Taxes, % of Total Plt Cost	1
State Income Tax (SIT), % Net Income less FIT	10
Fed Income Tax (FIT), % Net Income	30

Note(1) - Hydrogen product pricing sensitivity analysis will be performed for these parameters.

Table 3-5 – Annual Income vs. Expenditure Accounts Used for Break-even Analysis

	Annual Expenses \$MM/Yr	Annual Incomes \$MM/Yr
a. Coal Feed	XXXX	
b. Purchased Power	XXXX	
c. Raw Makeup Water	XXXX	
d. Catalysts & Chemical Consumed	XXXX	
e. Ash & Slag Disposal	XXXX	
f. Maintenance Labor & Materials	XXXX	
g. Admin Salaries & Overheads	XXXX	
h. Insurances	XXXX	
i. Local & Property Taxes	XXXX	
j. Amortized Investment Repayment	XXXX	
k. Power Sales		XXXX
l. Sulfur/Sulfuric Acid Sales		XXXX
m. CO ₂ Sequestration Credit		XXXX
n. <u>Hydrogen Sales</u>	<u> </u>	<u>XXXX</u>
o. Pre-Income Tax Total	XXXX	
p. Federal Income Tax	XXXX	
q. <u>State Income Tax</u>	<u>XXXX</u>	<u> </u>
r. After Income Tax Total	XXXX	XXXX

3.4.2.2 Task 4.2.2 IGCC Conceptual Design Scenarios

Various IGCC conceptual designs were developed and modeled with gradual increasing degree of complexity in order to fully assess the techno-economic feasibility of ITN's ICCM. Application of the ICCM to IGCC was believed to be most viable under conditions of H₂ and power co-production with CO₂ capture. Under this scenario, two main products, H₂ and power, plus a recovered CO₂ stream, will be produced from an IGCC plant. Determining a break-even cost (or fair-market value) for H₂ production to assess/compare ICCM techno-economics would require first defining the break-even costs for power and CO₂. Otherwise, the H₂ production cost will be skewed because it will be either subsidizing or being subsidized by power sale and/or CO₂ credit. For relative comparisons, it is important to determine the various product break-even costs for all IGCC co-production schemes by using a same consistent methodology and design

Table 3-6 – Breakdown for Total Initial Capital Investment

	Capital Cost \$MM	Initial Investment \$MM/Yr
Total Installed Cost (TIC)	xxxx	
Home Office Cost, y% of TIC	xxxx	
<u>Contingency, w% of TIC</u>	<u>xxxx</u>	
Total Plant Cost (TPC)	xxxx	
Initial Operating Capital	xxxx	
<u>Initial Investments</u>		<u>xxxx</u>
Total Initial Cost/Investment	xxxx	xxxx

basis. The following step-wise conceptual design scenarios were developed and analyzed for this purpose. Sensitivity analysis was also carried out to see if ICCM would offer any significant techno-economic advantage for a coal-gasification-based facility producing only H₂ as the main product.

- **Reference Case IGCC**

This is essentially a single-train, state-of-the-art IGCC conceptual design based on E-GasTM gasifier and a GE 7FA+e gas turbine as described in the previously cited 2003 DOE DE-AC26-99FT40342 study. It is modeled based on the performance and cost estimation information extracted from the DOE study. It defines the base case performance and a break-even cost for IGCC power production, without CO₂ recovery.

- **Case 1 – IGCC Power Production with CO₂ Capture**

Case 1 represents a revised IGCC design, incorporating pre-combustion CO₂ capture via a selective Selexol acid gas removal process for both H₂S and CO₂ removal. Assuming no net credit for the recovered CO₂, this case defines the break-even cost of electricity with CO₂ recovery. Alternatively, it can be used to define the break-even credit required for the recovered CO₂ to maintain the break-even electricity price established by the Reference Case IGCC design.

- **Case 2 – IGCC H₂ and Power Co-Production with CO₂ Capture (Conventional PSA Technology)**

Case 1 is modified for hydrogen co-production, using conventional PSA Technology. This case is used to establish the break-even cost of co-producing H₂ with PSA based on exporting the co-produced electricity at its fair-market value established by the Reference Case, and on exporting (or taking credit for) the recovered CO₂ at its fair-market value established by the Case 1. This case requires two E-GasTM gasifiers to generate enough syngas to produce significant quantity of H₂ and to support the combustion of one GE 7FA+e gas turbine.

Due to the low-pressure purge gas produced from the PSA, this scheme has the option to use cheaper but less efficient boiler-based power plant for co-producing power. This option is investigated as a sensitivity analysis for a coal-gasification-based facility producing only H₂.

- **Case 3 – IGCC H₂ and Power Co-Production with CO₂ Capture (ITN ICCM Technology with Conventional AGR)**

Integration requirement for ICCM into the IGCC will be described in more detail later. Ideally, the high-temperature ICCM membrane is best suited for integrating into an IGCC plant for H₂ and power co-production, whereby the pressurized hot residual (retentate) syngas can be sent directly as fuel to the gas turbine.

Because of the uncertainty with ‘Hot (*e.g.* 950 to 1800 °F)’ and ‘Warm (*e.g.* 300 to 750 °F) Clean Up’ technology development as a high-temperature, acid-gas removal process, it was decided to retain the conventional low temperature (LT) acid gas removal (AGR) process for selective removal of H₂S and CO₂, thus placing ICCM in a disadvantaged position of directly comparing with a commercially-proven technology of PSA. In doing so, however, it minimizes the potential design and cost uncertainties of comparing multiple developing technologies at one time.

- **Case 4 – IGCC H₂ and Co-Production with CO₂ Capture (ITN ICCM Technology with Warm Syngas Clean Up)**

This case investigates the techno-economics of ICCM technology with a needed complementary higher temperature AGR process of Warm Syngas Clean Up that is currently under development by DOE.

- **Case 5 – Coal-Based H₂ Production with CO₂ Capture (Conventional PSA)**

Cases 5 and 6 were developed to assess/compare the techno-economic of ICCM technology under a coal-based hydrogen-production-only scenario. This scenario is designed to maximize H₂ production/extraction from the produced syngas and minimize power generation. It was believed that in this scenario, ICCM may have distinct cost advantages over PSA, especially with low electricity costs. With these scenarios, onsite power generation facility is still provided as a means of disposal the residual purge gas to generate as much power as possible for in-plant consumption.

Case 5 investigated coal-based H₂ production using PSA technology. A two-stage sour shift is used to convert 90%+ of the CO in the syngas to H₂, followed by a conventional selective Selexol for H₂S and CO₂ removal. A gas-fired boiler is used to dispose the low-pressure PSA purge gas for power generation. Power is to be imported as required to satisfy all in-plant needs.

- **Case 6 – Coal Based Hydrogen Production with CO₂ Capture (ICCM Membrane)**

In contrast to Case 5, residual purge gas produced from the ICCM will be at pressure and temperature. A GE 6B gas turbine is used to burn it for power generation. Power will also have to be imported to satisfy all in-plant needs. Results are to be compared directly with that of PSA of Case 5.

Required CO₂ Capture – For the co-production case scenarios of Case 1 to Case 4, it was assumed that government regulation requires capturing 85% of produced CO₂ from the IGCC

facility. Sensitivity analysis was performed with selected case scenarios, to compare their economics with an allowable credit for the recovered CO₂.

Design Oxygen Purity - For the IGCC cast studies involving co-production, H₂ recovery from shifted syngas is not maximized in order to leave sufficient energy to meet the 7FA+e gas turbine fuel demands. Thus the slightly higher syngas H₂ partial pressure from using 99.5% oxygen offers very little advantage for H₂ extraction; the shifted syngas volumetric flow reduction from using 99.5% vs. 95% oxygen is approximately 1% on dry basis, and 0.5% on wet basis. There is very little equipment size savings from using 99.5% vs. 95% oxygen. Therefore, these designs were based on using a 95% purity oxygen ASU unit to take advantage of the lower oxygen plant's power consumption and cost.

For the coal-based H₂ production only scenarios of Cases 5 and 6, H₂ recovery from the shifted syngas is maximized to minimize onsite generation of power, a lower value commodity. Therefore, gasification designs for these schemes were based on 99.5% oxygen purity.

Task 4.2.2.1 Reference Case IGCC Plant Design

A reference case was chosen to represent a current state-of-the-art IGCC conceptual plant design with which the application of ITN hydrogen separation membrane is to be investigated and compared with conventional PSA technology for H₂ co-production. A computer model was developed to facilitate the comparative analysis, drawing upon various in-house process engineering and equipment cost estimation database from previous Nexant studies. The reference case chosen has a configuration with major IGCC technology components similar to the Subtask 1.5A Optimized Single-Train Coal Based IGCC Design as described in the 2003 DOE study [DE-AC26-99FT40342] entitled "Gasification Plant Cost and Performance Optimization." Key technology components include the use of an E-GasTM gasifier for coal gasification and a GE 7FA+e combustion turbine (210 MW) for power production. The GE 7FA+e turbine design represents an improvement over the existing 7FA machine with a higher firing temperature and increased pressure ratio. The 7FA+e turbine has a higher capacity and thermal efficiency as well as lower NO_x and CO emissions.

Reference Case IGCC – Configuration Figure 3-29 shows the simplified block flow diagram of the Reference Case IGCC design. It contains the following major processing steps:

A.) High pressure E-GasTM technology, with all the associated coal feed handling and preparation facilities, is used for coal gasification. The E-GasTM gasifier is equipped with hot cyclones, radiant boiler, dry char filter, followed with BFW (boiler feed water) preheaters to remove and recycle chars, and to maximize sensible heat recovery from the hot syngas. Water wash of the cooled syngas is used for chloride removal. The E-GasTM gasifier can achieve a 99% carbon conversion, with a slag byproduct containing only 1% unconverted carbon.

A maximum single gasifier capacity of 3000 STPD (dry basis) is assumed for the study. Per DOE's "Gasification Plant Cost and Performance Optimization" Study, an installed spare gasifier is included to achieve an 80% on-stream availability factor.

B.) Conventional cryogenic air separation unit (ASU) is used for the production of a 95% purity oxygen stream required for gasification.

C.) Low temperature gas cooling is provided to cool the syngas prior to feeding it to the downstream acid gas removal process.

D.) Selective MDEA is chosen as the AGR (acid gas removal) process to remove all of the H₂S and a small portion of the CO₂ from the syngas prior to sending it to the gas turbine, and to produce a H₂S-rich acid gas feed for the Sulfur Recovery Unit (SRU). For this study, the H₂S-rich acid gas feed to the SRU is designed to have approximately 45% H₂S.

E.) Claus sulfur recovery with Tail Gas Treating process is used to convert the H₂S acid gas into sulfur for export.

F.) Combined cycle power generation is based on a single GE's 7FA+e syngas fired gas turbine, with a 1500 psig HRSG and steam turbine generators.

G.) Major off-plot and utility support facilities including: water treating; waste water treating; plant and instrument air; relief & flare system; interconnecting pipeway; electrical distribution system; fire protection facilities; shops, warehouses, control room and admin buildings; plant DCS and communication systems; and cooling water system.

Reference Case IGCC - Plant Performance The computer model was developed to facilitate process heat and material balances and the overall balance of plant. The model contains enough details to allow most of the process equipment to be sized and cost estimation made as part of the overall modeling effort. Appendix D shows the modeling summary outputs for the six case studies performed.

All process streams with detailed compositions and operation conditions are presented in the model output. Selected stream flows corresponding to the stream numbers shown on the simplified overall block flow diagram of Figure 3-29 are extracted and presented in Table 3-7 to show the major input and output stream to and from the plant.

Table 3-8 presents the overall performance of the Reference Case IGCC design. This single-train Reference Case IGCC design consumes about 2,800 tons per day of Illinois #6 and produces approximately 325 MWe of power at an overall plant efficiency of 38%. In comparison, the 2003 DOE Optimized Single-Train Coal Based IGCC Design of Subtask 1.5A, which has a similar plant configuration, has a somewhat higher efficiency of 39%. The Subtask 1.5A design is fully optimized and thus it is expected to be more efficient. With the current Reference Case modeling, it is not worth the effort to further optimize the design, since its development is only intended to investigate the relative techno-economic merit of co-producing H₂ using ITN technology vs. PSA technology. The front-end IGCC performance is largely cancelled out so that the small overall reference IGCC plant in-efficiency should not impact the result of the study.

Reference Case - Conventional IGCC Design (No CO2 Capture)

Figure 3-29. Block diagram for reference case.

Table 3-7

Reference Case IGCC Major Stream Flows

Stream Number	1.	2.	3.	4.	5.	6.	7.	8
Stream Description	As-Rec Coal Feed	Oxygen from ASU	Syngas from Gasifier (Include Recycle)	Syngas from HT Gas Cool. (Include Recycle)	Syngas from LT Gas Cool.	Sweet Syngas from AGR	Flue Gas from IGCC/Blr	Sulfur Byproduct
Vapor Composition, Vol%:	-----	-----	-----	-----	-----	-----	-----	-----
N2		3.79	1.11	1.09	1.39	1.42	66.60	
O2		95.00	0.00	0.00	0.00	0.00	11.89	
CO2		0.00	13.25	12.98	16.50	15.83	8.30	
Ar		1.21	0.36	0.35	0.44	0.45	0.85	
H2		0.00	25.71	25.17	32.01	32.71	0.00	
CO		0.00	35.97	35.21	44.79	45.77	0.00	
H2S		0.00	0.70	0.68	0.87	0.00	0.00	
COS		0.00	0.03	0.03	0.04	0.00	0.00	
CH4		0.00	3.00	2.94	3.74	3.82	0.00	
SO2		0.00	0.00	0.00	0.00	0.00	0.00	
H2O		0.00	19.86	21.55	0.22	0.00	12.36	
-----	-----	-----	-----	-----	-----	-----	-----	-----
Total Vol%	---	100.00	100.00	100.00	100.00	100.00	100.00	---
Molar Flow Rate, lbmoles/	---	5,437	24,290	24,812	17,557	17,181	135,963	—
Mass Flow Rate: lb/hr	231,844	173,669	514,948	524,349	389,133	371,762	3899475	5,103
" " " : STPD	2,782	2,084	6,179	6,292	4,670	4,461	46,794	61
Vol Flow Rate, MMSCFD	---	49.5	221.2	226.0	159.9	156.5	1238.3	—
HHV, MMBtu/Hr	2,527	0	2,241	2,250	1,936	1,897	315	20
Pressure, psig	0.0	565.0	465.0	391.0	361.0	351.0	0.0	0.0
Temperature, deg F	60	240	1,850	306	95	95	260	60

Table 3-8

Reference Case IGCC Plant Performance

A <u>Import or Feeds:</u>		
	Coal Feed, STPD (AR)	2,782
	95% Oxygen, STPD	2,084
	Make Up Water, GPM	2,643
B <u>Export or Products:</u>		
	Electrical Power, MWe	281
	Hydrogen, MMSCFD	0
	Recovered CO ₂ , STPD	0
	Sulfur, STPD	61
	Slag & Ash (Incl. Carbon), STPD Dry	319
	Process Waste Water, GPM	1,049
C <u>Carbon Balance, STPD Equivalent CO₂:</u>		
	Coal Feed	6,103
	Power Plant Combustion Air	20
	<u>Total CO₂ in:</u>	6,123
	Power Plant Stack Emission	5,956
	SRU/TGTU Stack Emission	106
	Equivalent CO ₂ in Slag & Ash	61
	Recoverable CO ₂ from AGR	0
	<u>Total CO₂ out:</u>	6,123
D <u>Power Balance, MWe</u>		
	Gas Turbine	209.3
	Steam Turbine	116
	<u>Total Generation:</u>	325.3
	Coal Prep & Gasification	3.0
	ASU & O ₂ Compression	27.5
	Sour Shift & LT Syngas Cooling	0.0
	AGR (Acid Gas Removal)	0.4
	H ₂ Recovery & Compression	0.0
	Sulfur Recovery	0.3
	CO ₂ Compression	0.0
	Flue Gas Compression	0.0
	HRSG/Boiler/BFW Treating	3.9
	BOP Consumption	8.8
	<u>Total Consumption:</u>	43.8
E Overall Power + H₂ Efficiency, %HHV		
		38.0

Reference Case IGCC - Preliminary Economics Developing a Reference Case IGCC design has the objective of providing a firm basis upon which the process integration of the ITN H₂ separation membrane can be clearly identified and its techno-economic viability can be assessed. Since full application of the H₂ separation membrane is to be explored under different IGCC co-production scenarios, it is also beneficial to clearly establish a baseline IGCC power production cost, so that H₂ production economics can be determined.

Table 3-9 shows the preliminary capital cost estimates for the Reference Case IGCC design. For comparison, The DOE 2003 Subtask 1.5A Optimized IGCC Design has an estimated mid-2000 installed cost of \$375 MM (excluding contingency) vs. the estimated 2003 cost of \$389 MM for this Reference Case design.

Table 3-10 summarizes the annual revenues and expenses balance for the Reference Case IGCC, allowing a break-even cost of electricity production to be determined.

Table 3-9
Reference Case IGCC Preliminary Capital Cost Estimates

Coal Feed, STPD (AR)	2,782
Net Power Export, MWe	281
<u>Installed Cost, \$MM</u>	
Coal Prep & Gasification	102.0
ASU & O ₂ Compression	44.1
Sour Shift & LT Syngas Cooling	6.4
AGR (Acid Gas Removal)	9.5
H ₂ Recovery & Compression	0.0
Sulfur Recovery	5.3
CO ₂ Drying & Compression	0.0
Fuel Gas Compression	0.4
Gas Turbine/Generator	71.1
Steam Turbine/Generator	19.5
HRSG/Boiler/BFW Treating/Stack	21.2
BOP (Balance of Plant)	91.4
<u>Total Field Cost:</u>	370.9
Home Office Cost	18.3
Contingency, 15%	55.6
<u>Total Installed Cost:</u>	444.8
Installed Cost, \$/Net kW	1581

Table 3-10

Reference Case IGCC Annual Revenue & Expense Estimates

Coal Feed, STPD (AR)	2,782
% CO ₂ Capture	0
Net Power Export, MWe	281
Hydrogen Export, MMSCFD	0
CO ₂ Recovered, STPD	0
Installed Cost, \$MM	445
<u>Installed Operating Capital, \$MM</u>	16
Total Initial Investment, \$MM	461
<u>Annual Expenses, \$MM/Yr:</u>	
Coal Feed Cost (29.4/ST AR)	23.9
Power Inport Cost	0.0
Raw Water Cost	0.3
Catalysts & Chemical Cost	6.3
Ash & Slag Disposal Cost	0.9
Maintenance Labor & Material	13.3
Adminstration Salaries & Overheads	11.8
Insurance & Local Taxes	8.8
Amortized (Over 25 Yr) Investment Repayme	50.8
Federal & State Income Taxes	0.0
Total Expense	116.3
<u>Annual Revenues, \$MM/Yr:</u>	
Power Sale (by Difference)	115.8
Hydrogen Sale	0
Sulfur Byproduct Sale	0.5
Credit for Recovered CO ₂	0.0
Total Revenue:	116.3

Cost of Electricity and Sensitivity The cost of electricity (or break-even cost) for the Reference Case IGCC as shown in Table 3-10 is thus:

$$115,800,000 / (281 \times 24 \times 365 \times 0.8) = \$58.7 \text{ per MW-Hr}$$

where :

115,800,000	=	Annual electricity sales revenue from Table 3-10
281	=	Power Export, MWe (from Table 3-8)
24	=	Hours per day
365	=	Days per year
0.8	=	On-stream factor

This is the minimum electricity sales price to justify the construction of the Reference Case IGCC power plant based on the economic parameters defined in Task 4.2.1.2 (pg 42) of this report. This minimum price changes with changes to each parameter. Figures 3-30 through 3-33 show the sensitivity of the cost of electricity to the various selected economic parameters.

Effect of Price of As-Received Coal: The cost of electricity increases by about 15% when the coal cost is increased from \$29.4/ST to \$50/ST, and decreases by about 21% if the feed coal is free.

Cost of Electricity vs. Price of AR Coal Feed

Figure 3-30. Reference Case sensitivity of the cost of electricity as a function of the coal price.

Effect of % Return on Investment: The cost of electricity increases by about 15% when the ROI is increased from 10% to 25%, and decreases by about 15% when the ROI is reduced from 10% to 5%.

Figure 3-31. Reference Case sensitivity of the cost of electricity as a function of the return on investment.

Figure 3-32. Reference Case sensitivity of the cost of electricity as a function of the plant life.

Figure 3-33. Reference Case sensitivity of the cost of electricity as a function of the capital cost contingency.

Task 4.2.2.2 Case 1 - IGCC Power Production with CO₂ Capture

Case 1 incorporates design changes to the Reference Case IGCC plant for CO₂ capture. It is used to establish a break-even CO₂ captured cost. This design retains the key technology features of using an E-GasTM gasifier for supplying the fuel to support a 7FA+e gas turbine power production.

Case 1 Configuration Figure 3-34 shows the simplified block flow diagram for the Case 1 design. It contains the same major processing steps as the Reference Case IGCC except for the following:

- Addition of a two-stage catalytic sour shift unit ahead of the low temperature gas cooling system to produce a predominantly H₂-CO₂ rich syngas gas.
- Utilizing Selexol acid gas removal process to remove all of the H₂S and bulk of the CO₂ from the syngas. The Selexol process produces a H₂S-rich acid gas feed to Sulfur Recovery plus a separate bulk CO₂ gas for sequestration in regeneration. For this study, it is assumed that the sweetened syngas will have a residual CO₂ partial pressure of 5 psia and the H₂S-rich acid gas will contain approximately 55 vol% CO₂.
- Addition of a CO₂ dehydration and compression unit to dry and compressed the recovered CO₂ to 2000 psig at the battery limit ready for sequestration.

Case 1 - IGCC with CO2 Capture

Figure 3-34. Block diagram for Case 1 – IGCC Power Production with CO₂ capture.

Table 3-11 lists the flow and composition of selected major streams shown on the simplified overall block flow diagram.

TABLE 3-11
Case 1 IGCC (with CO₂ Recovery) - Major Stream Flows

Stream Number	1.	2.	3.	4.	5.	6.	7.	8	9
Stream Description	As-Rec Coal Feed	Oxygen from ASU	Syngas from Gasifier (Include Recycle)	Syngas from HT Gas Cool. (Include Recycle)	Syngas from LT Gas Cool.	Sweet Syngas from AGR	Flue Gas from IGCC/Blr	CO ₂ to to Sequest. (Impurity free Basis)	Sulfur Byproduct
Vapor Composition, Vol%:	-----	-----	-----	-----	-----	-----	-----	-----	-----
N ₂		3.79	1.12	1.10	0.98	1.66	67.06		
O ₂		95.00	0.00	0.00	0.00	0.00	11.64		
CO ₂		0.00	13.26	13.07	40.98	1.35	1.14	100.00	
Ar		1.21	0.36	0.35	0.31	0.53	0.86		
H ₂		0.00	25.73	25.35	51.93	88.08	0.00		
CO		0.00	36.00	35.47	2.30	3.90	0.00		
H ₂ S		0.00	0.70	0.69	0.61	0.00	0.00		
COS		0.00	0.03	0.03	0.03	0.00	0.00		
CH ₄		0.00	3.00	2.96	2.64	4.48	0.00		
SO ₂		0.00	0.00	0.00	0.00	0.00	0.00		
H ₂ O		0.00	19.80	20.97	0.22	0.00	19.30		
Total Vol%	---	100.00	100.00	100.00	100.00	100.00	100.00	100.00	---
Molar Flow Rate, lbmoles/hr	---	5,739	25,624	26,003	26,250	15,476	135,062	10,361	---
Mass Flow Rate: lb/hr	244,740	183,330	543,287	550,111	728,981	75,172	3624137	456,002	5,387
" " " : STPD	2,937	2,200	6,519	6,601	8,748	902	43,490	5,472	65
Vol Flow Rate, MMSCFD	---	52.3	233.4	236.8	239.1	141.0	1230.1	94	---
HHV, MMBtu/Hr	2,668	0	2,365	2,372	2,057	2,015	489	0	21
Pressure, psig	0.0	670.0	570.0	496.0	361.0	351.0	0.0	2000.0	0.0
Temperature, deg F	60	240	1,850	321	95	95	260	91	60

Case 1 Performance & Cost Analysis Following the same methodology as with the Reference Case design, Case 1 performance, preliminary cost estimates, and annual revenue/expenditure cost breakdowns leading to the estimate of a CO₂ capture cost are summarized in Tables 3-12 to 3-14, respectively.

Assuming no net revenue credit for the recovered CO₂, the estimated cost of electricity (or break-even cost) for the Case 1 IGCC as shown in Table 3-14 is then:

$$133,200,000 / (223 \times 24 \times 365 \times 0.8) = \$85.4 \text{ per MW-Hr}$$

where :

$$133,200,000 = \text{Annual electricity sales revenue from Table 3-14}$$

$$223 = \text{Power Export, Mwe}$$

$$24 = \text{Hours per day}$$

$$365 = \text{Days per year}$$

$$0.8 = \text{Onstream factor}$$

Table 3-12
Case 1 Plant Performance

A <u>Import or Feeds:</u>		
Coal Feed, STPD (AR)		2,937
95% Oxygen, STPD		2,200
Make Up Water, GPM		2,370
B <u>Export or Products:</u>		
Electrical Power, MWe		223
Hydrogen, MMSCFD		0
Recovered CO ₂ , STPD		0
Sulfur, STPD		65
Slag & Ash (Incl. Carbon), STPD Dry		337
Process Waste Water, GPM		982
C <u>Carbon Balance, STPD Equivalent CO₂:</u>		
Coal Feed		6,443
Power Plant Combustion Air		20
<u>Total CO₂ in:</u>		6,463
Power Plant Stack Emission		815
SRU/TGTU Stack Emission		112
Equivalent CO ₂ in Slag & Ash		64
Recoverable CO ₂ from AGR		5,472
<u>Total CO₂ out:</u>		6,463
D <u>Power Balance, MWe</u>		
Gas Turbine		209.1
Steam Turbine		82.6
<u>Total Generation:</u>		291.7
Coal Prep & Gasification		3.0
ASU & O ₂ Compression		29.6
Sour Shift & LT Syngas Cooling		0.5
AGR (Acid Gas Removal)		7.5
H ₂ Recovery & Compression		0.0
Sulfur Recovery		0.3
CO ₂ Drying & Compression		16.1
Flue Gas Compression		0.0
HRSG/Boiler/BFW Treating		3.9

Table 3-13

Case 1 - Preliminary Capital Cost Estimates

Coal Feed, STPD (AR)	2,937
CO2 Capture, %	85
Net Power Export, MWe	223
<u>Installed Cost, \$MM</u>	
Coal Prep & Gasification	105.9
ASU & O2 Compression	45.8
Sour Shift & LT Syngas Cooling	17.5
AGR (Acid Gas Removal)	48.9
H2 Recovery & Compression	0.0
Sulfur Recovery	5.5
CO2 Drying & Compression	28.2
Fuel Gas Compression	0.3
Gas Turbine/Generator	71.0
Steam Turbine/Generator	18.8
HRSG/Boiler/BFW Treating/Stack	15.6
BOP (Balance of Plant)	87.8
<u>Total Field Cost:</u>	445.3
Home Office Cost	24.0
Contingency, 15%	66.8
<u>Total Installed Cost:</u>	536.1
Installed Cost, \$/Net kW	2,410

Table 3-14

Case 1 - Annual Revenue & Expense Estimates

Coal Feed, STPD (AR)	2,937
% CO ₂ Capture	85
Net Power Export, MWe	223
Hydrogen Export, MMSCFD	0
CO ₂ Recovered, STPD	5,472
Installed Cost, \$MM	536
<u>Installed Operating Capital, \$MM</u>	18
Total Initial Investment, \$MM	554
<u>Annual Expenses, \$MM/Yr:</u>	
Coal Feed Cost (29.4/ST AR)	25.2
Power Inport Cost	0.0
Raw Water Cost	0.2
Catalysts & Chemical Cost	7.6
Ash & Slag Disposal Cost	1.0
Maintenance Labor & Material	16.1
Adminstration Salaries & Overheads	11.8
Insurance & Local Taxes	10.8
Amortized (Over 25 Yr) Investment Repayme	61.1
Federal & State Income Taxes	0.0
Total Expense	133.8
<u>Annual Revenues, \$MM/Yr:</u>	
Power Sale (by Difference)	133.2
Hydrogen Sale	0
Sulfur Byproduct Sale	0.6
Credit for Recovered CO ₂	0.0
Total Revenue:	133.8

This corresponds to 145% of the Reference Case electricity production cost, and represents the minimum electricity sales price that would need to justify the construction of an IGCC power plant that recover up to 85% of the feed carbon as CO₂ for potential sequestration. Figures 3-35

through 3-38 shows the sensitivity of the estimated electricity production cost, for both with and without CO₂ capture, as a function of the following economic variables:

- Price of As-Received Coal
- % Return on Investment
- Plant Life
- Capital Cost Contingency

In an IGCC co-production scheme, it is important that the correct break-even power production cost corresponding to the specific variable changes is used to estimate the H₂ break-even cost.

Figure 3-35. Case 1 sensitivity of the cost of electricity as a function of the coal price.

Figure 3-36. Case 1 sensitivity of the cost of electricity as a function of the return on investment.

Figure 3-37. Case 1 sensitivity of the cost of electricity as a function of the plant life.

Figure 3-38. Case 1 sensitivity of the cost of electricity as a function of the capital cost contingency.

Task 4.2.2.3 Case 2 - IGCC H₂ and Power Co-Production with CO₂ Capture (Conventional PSA)

Figure 3-39 shows a simplified block flow diagram of the Case 2 configuration. It contains the same major gasification and power train blocks of that of Case 1 with the additional of the following processing:

- A PSA unit to extract hydrogen from the sweetened syngas downstream of the AGR;
- A hydrogen compressor system to compress the product hydrogen from about 300 psig to the required B/L pressure of 1000 psig; and
- A gas compression system to compress the low pressure PSA purge gas from about 5 psig to 340 psig for use as fuel in the GE 7FA+e gas turbine.

Table 3-15 lists the flow and composition of selected major streams as shown on the simplified block flow diagram for this configuration. Overall plant performance, preliminary cost estimates and annual revenue/expenditure cost breakdown are summarized in Table 3-16 to 3-18, respectively.

Assuming no net revenue credit for the captured CO₂ and a \$85.4/MW-hr cost for IGCC power production as defined by Case 1, the break-even hydrogen production cost for this PSA-based conventional IGCC co-production with CO₂ capture case is:

$$175,600,000 / (212,400 \times 365 \times 0.8) = \$2.83 \text{ per 1000 SCF}$$

where :

175,600,000	=	Annual H ₂ sales revenue
212.4	=	H ₂ Export, MMSCFD
24	=	Hours per day
365	=	Days per year
0.8	=	On-stream factor

This estimated \$2.83/1000 SCF hydrogen cost represents the minimum selling price needed to justify the construction of an IGCC co-production plant that would require capturing up to 85% of the produced CO₂ for sequestration.

Case 2 - IGCC Hydrogen Co-Production with CO2 Capture

Figure 3-39. Block diagram for Case 2.

Table 3-15
Case 2 - IGCC Co-Production with CO2 Capture (Conventional PSA)

Stream Number	1.	2.	3.	4.	5.	6.	7.	8.	9.	10.	11.
Stream Description	As-Rec Coal Feed	Oxygen from ASU	Syngas from Gasifier (Include Recycle)	Syngas from HT Gas Cool. (Include Recycle)	Syngas from LT Gas Cool.	Sweet Syngas from AGR	Flue Gas from IGCC/Blr	Hydrogen for Export	H2 Recovery Purge	CO2 to to Sequest. (Impurity free Basis)	Sulfur Byproduct
Vapor Composition, Vol%:	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
N2		3.79	1.11	1.10	0.98	1.66	66.86	0.01	4.44		
O2		95.00	0.00	0.00	0.00	0.00	11.40	0.00	0.00		
CO2		0.00	13.25	13.04	40.97	1.49	2.72	0.00	4.00	100.00	
Ar		1.21	0.36	0.35	0.31	0.53	0.94	0.00	1.43		
H2		0.00	25.72	25.31	51.93	87.97	0.00	99.99	67.71		
CO		0.00	36.00	35.42	2.29	3.88	0.00	0.00	10.43		
H2S		0.00	0.70	0.69	0.61	0.00	0.00	0.00	0.00		
COS		0.00	0.03	0.03	0.03	0.00	0.00	0.00	0.00		
CH4		0.00	3.00	2.95	2.64	4.47	0.00	0.00	11.99		
SO2		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		
H2O		0.00	19.82	21.11	0.24	0.00	18.08	0.00	0.00		
-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
Total Vol%	---	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	---
Molar Flow Rate, lbmoles/hr	---	13,746	61,438	62,443	62,947	37,157	135,983	23,318	13,838	24,787	---
Mass Flow Rate: lb/hr	586,205	439,113	1302466	1320568	1816940	182,485	3699665	47,070	135,415		12,902
" " " : STPD	7,034	5,269	15,630	15,847	21,803	2,190	44,396	565	1,625	13,091	155
Vol Flow Rate, MMSCFD	---	125.2	559.6	568.7	573.3	338.4	1238.5	212.4	126.0	225.8	---
HHV, MMBtu/Hr	6,390	0	5,670	5,689	4,931	4,831	461	2,867	1,963	0	51
Pressure, psig	0.0	635.0	535.0	461.0	326.0	316.0	0.0	1000.0	340.0	2000.0	0
Temperature, deg F	60	240	1,850	316	95	95	260	91	191	91	60

Table 3-16

Case 2 - IGCC Co-Production with CO2 Capture Plant Summary

A <u>Import or Feeds:</u>		
Coal Feed, STPD (AR)		7,034
95% Oxygen, STPD		5,269
Make Up Water, GPM		4,217
B <u>Export or Products:</u>		
Electrical Power, MWe		123
Hydrogen, MMSCFD		212
Recovered CO2, STPD		13,091
Sulfur, STPD		155
Slag & Ash (Incl. Carbon), STPD Dry		806
Process Waste Water, GPM		1,929
C <u>Carbon Balance, STPD Equivalent CO2:</u>		
Coal Feed		15,432
Power Plant Combustion Air		20
<u>Total CO2 in:</u>		15,452
Power Plant Stack Emission		1,951
SRU/TGTU Stack Emission		256
Equivalent CO2 in Slag & Ash		154
Recoverable CO2 from AGR		13,091
<u>Total CO2 out:</u>		15,452
D <u>Power Balance, MWe</u>		
Gas Turbine		209.3
Steam Turbine		92.2
<u>Total Generation:</u>		301.5
Coal Prep & Gasification		7.6
ASU & O2 Compression		70.5
Sour Shift & LT Syngas Cooling		1.4
AGR (Acid Gas Removal)		17.1
H2 Recovery & Compression		13.8
Sulfur Recovery		0.6
CO2 Drying & Compression		38.4
Flue Gas Compression		9.8
HRSG/Boiler/BFW Treating		5.2
BOP Consumption		13.7
<u>Total Consumption:</u>		178.1
E Overall Power + H2 Efficiency, %HHV		
		51.5

Table 3-17
Case 2 - Preliminary Capital Cost Estimates

Coal Feed, STPD (AR)	7,034	
CO2 Capture, %	85	
Net Power Export, MWe	123	
H2 Production, MMSCF	212	
<u>Installed Cost, \$MM</u>		<u># Train</u>
Coal Prep & Gasification	221.2	2+1 spare gasifier
ASU & O2 Compression	103.9	2
Sour Shift & LT Syngas Cooling	46.3	2
AGR (Acid Gas Removal)	110.9	2
H2 Recovery & Compression	48.0	1
Sulfur Recovery	10.2	1
CO2 Drying & Compression	52.0	1
Fuel Gas Compression	5.1	
Gas Turbine/Generator	71.1	1
Steam Turbine/Generator	23.3	1
HRSG/Boiler/BFW Treating/Stack	17.2	
BOP (Balance of Plant)	112.2	
<u>Total Field Cost:</u>	821.4	
Home Office Cost	41.6	
Contingency, 15%	123.2	
<u>Total Installed Cost:</u>	986.2	
Intalled Cost, \$/Net kW	2410 ⁽¹⁾	
Intalled Cost, \$/SCFD H2	3.24 ⁽²⁾	

Note:

(1) Capital cost associated with exporting power is assumed to be identical to that of Case 1

(2) Capital cost associated with exporting H2. \$/SCFD = (TIC - Net MWe export x 2,410)/(MMSCFD H2 produced)

Table 3-18**Case 2 - Annual Revenue & Expense Estimates**

Coal Feed, STPD (AR)	7,034
% CO2 Capture	85
Net Power Export, MWe	123
Hydrogen Export, MMSCFD	212
CO2 Recovered, STPD	13,091
Installed Cost, \$MM	986
<u>Installed Operating Capital, \$MM</u>	35
Total Initial Investment, \$MM	1,021
<u>Annual Expenses, \$MM/Yr:</u>	
Coal Feed Cost (29.4/ST AR)	60.4
Power Inport Cost	0.0
Raw Water Cost	0.4
Catalysts & Chemical Cost	14.0
Ash & Slag Disposal Cost	2.4
Maintenance Labor & Material	29.6
Adminstration Salaries & Overheads	11.8
Insurance & Local Taxes	19.8
Amortized (Over 25 Yr) Investment Repayme	112.4
Federal & State Income Taxes	0.0
Total Expense	250.7
<u>Annual Revenues, \$MM/Yr:</u>	
Power Sale (\$85.4/MWH)	73.7
Hydrogen Sale (By Difference)	175.6
Sulfur Byproduct Sale	1.4
Credit for Recovered CO2	0.0
Total Revenue:	250.7

Task 4.2.2.4 Case 3 – IGCC H₂ Co-Production with CO₂ Capture (ITN ICCM Membrane with Conventional AGR)

As part of a preliminary engineering assessment exercise at the onset of the project, a set of performance and operational conditions for ITN's ICCM was identified to develop some understanding of the current state-of-the-art of the technology, performance that can be realistically achieved within the research and development work proposed, and potential for improvement. These estimated performance criteria are shown in Table 3-19 below. The flux shown are achievable instantaneous rate at the given hydrogen differential partial pressure. It was agreed that conceptual engineering design analysis should be based on an estimated achievable commercial performance with sensitivity analysis performed within a reasonable +/- range, as needed.

Table 3-19

	<u>Current</u>	<u>Commercial</u>
Flux Rate (cm/min-cm ²)	2	50
Temperature (°C/°F)	900/1652	815/1500
Hydrogen Purity (%)	95	99.99
Hydrogen Recovery (%)	75	85
Pressure Differential (psi)	400	200

There are other ICCM membrane related design issues that can affect its integration and application to an IGCC-based energy plant. Examples include the need of internal membrane water-gas-shift activity to increase the ICCM H₂ flux, as pointed out in the preliminary membrane module design work of Task 4.1, and the membrane's stability to steam, sulfur, chloride, and other trace contaminants expected from the gasification of coal. These effects, unfortunately, are not very well defined as yet. They are not considered in the current IGCC conceptual design application analysis.

Case 3 - Integration of ITN ICCM Module Integration is mainly focused on aligning/balancing the ICCM operating temperature with the other needed IGCC process technologies. It involved feed/effluent heat exchanging. Sensitivity analyses were performed to investigate the effect of the ICCM H₂ production pressure on the overall system cost. Internal ICCM membrane water-gas-shift activity was not assumed. Thus for the case studies undertaken for IGCC H₂ co-production, a two-stage catalytic sour shift unit upstream of the ICCM is included in order to produce sufficient H₂ with a high enough partial pressure for effective downstream ICCM H₂ separation. The H₂ partial pressure is further increased by removing the CO₂ from the shifted syngas upstream of the ICCM via a selective Selexol unit. The Selexol unit also removes the H₂S from the shifted syngas prior to sending the H₂-rich feed stream to the ICCM module. With co-production, balance of plant is required to produce a syngas purge stream from the ICCM with sufficient heating content to sustain the operation of the GE 7FA+e turbine.

The overall ITN ICCM Process Plant consists of three major sections:

- Feed/effluent heat exchange;
- ITN ICCM module which potentially consisting of numerous monolithic units with dimensions as shown in Figure 3-25 (pg 31); and
- Hydrogen product compression.

Feed/Effluent Heat Exchange Section Figure 3-40 shows a simplified process flow diagram of the ITN process plant. Shifted and sweetened syngas from the Selexol AGR unit is delivered to the ICCM process plant battery limit at approximately 95 °F and 370 psig. It is heated to 1050 °F in a series of two heat exchange with the permeate and retentate leaving the ICCM module. The 1050 °F feed is then heated to 1500 °F by burning part of the feed with small amount of oxygen in a partial oxidation (POX) type in-line burner, F-1, before being sent to the ITN ICCM module for H₂ extraction. The produced H₂ is sent to the product compression section where it is cooled and compressed to the required battery limit pressure for export.

The ICCM retentate purge gas, at 345 psig and 860 °F, is sent directly to the combined cycle plant for combustion in the gas turbine (GT) without further cooling or compression; the 345 psig pressure allows a 5 psi piping pressure drop while still meeting the GE 7FA+e GT operation requirement.

ICCM Membrane Module Section The 1500 °F shifted syngas feed from the Preheating Section is sent onto the ICCM Module (R-1) where H₂ is extracted from the syngas side (the retentate side) across the membrane to the H₂ product side (the permeate side) based on the H₂ partial pressure (PP) differential between the two sides, as governed by the flux equation as shown in Equation 3-2 (pg 17). The module was modeled as a series of membrane elements.

For Case 3 IGCC Hydrogen Co-Production design, the syngas feed to the ITN process unit battery limit contains approximately 88 mole% H₂ (dry basis) which is reduced to about 85% entering the R-1 after the POX preheating step. This yields a H₂ partial pressure of approximately 315 psia for the retentate side at the inlet to R-1. This partial pressure decreases along each sequential membrane element within the R-1 as hydrogen is being extracted.

The H₂ concentration in the syngas exiting R-1 is set by the desired H₂ recovery, which in part is governed by economics, as an increasingly larger membrane surface would be required to recover the same quantity of H₂ as its partial pressure in the retentate decreases. In addition, the H₂ recovery is also set by the fuel gas requirement for the downstream gas turbine, unless one is willing to import natural gas as a supplementary fuel. For Case 3 co-production, approximately 70% of the H₂ in the shifted syngas can be recovered. The resulting R-1 retentate side exit H₂ concentration is approximately 61 mole% which gives an outlet H₂ partial pressure of approximately 225 psia.

Figure 3-40. Case 3 simplified process flow diagram for the ITN ICCM process plant.

Hydrogen production pressure - Setting the H₂ production pressure of the ICCM Module is a trade-off between the membrane capital cost vs. overall system operating costs. A lower hydrogen production (ICCM permeate side) pressure would reduce the total membrane area required to produce the same quantity of product, but will require higher compression costs (a large compressor and higher power consumption). Sensitivity analysis was performed. Figure 3-41 shows the result that there is an optimum ICCM permeate H₂ production pressure for a specific process design, which is also somewhat dependent on economic assumptions.

Figure 3-41. Plot of relative cost of ICCM module as a function of the membrane permeate pressure.

ICCM membrane area estimation - Membrane cost is a major component of the total ITN ICCM Process Plant capital cost. It is estimated based on the amount of surface area required for the desired H₂ recovery, based on the flux Equation 3-2 (pg 17) with an assumed operating temperature of 1500 °F, H₂ flux of 50 ml/minute/cm² at a H₂ partial pressure differential of 200 psi. Detailed discussion on the transport properties and equations are cited in Section 3.4.1 Task 4.1 (pg 31 ff.).

To estimate the membrane area requirement, the total H₂ recovery is modeled with 50 membrane elements in series each recovering an equal quantity (1/50) of the total H₂ production. As the H₂ partial pressure in the retentate side of the ICCM elements decreases, the membrane surface area required increases as shown in Figure 3-42.

Figure 3-42 shows each ICCM membrane sectional area requirement for Case 3 of which it has a calculated total membrane area of 243,000 ft².

Figure 3-42. Plot of ICCM membrane sectional area requirement as a function of the H₂ extracted for Case 3.

H₂ Compression Section For the Case 3 ICCM process diagram as shown in Figure 3-40, H₂ extracted from the ICCM Module Section is delivered to the H₂ compression section at approximately 400 °F after heat exchanged against cold syngas feed to the ICCM Module. It is then cooled by cooling water to about 90 °F and sent through a KO Drum, C-1, to remove any condensed water. Vapor from the KO Drum is compressed from 80 psig to 1000 psig through the multi-stage H₂ product compressor package, K-1. The compressed H₂ is cooled to less than 100 °F with cooling water in E-4 before delivered to the battery limit for export. The exact number of compression stages is selected to limit the discharge temperature from each stage to no higher than 270 °F with interstage feed cooling. To minimize potential fire and explosion hazards associated with H₂ compression, compressor discharge temperature is typically limited to 270-300 °F.

Case 3 IGCC Co-Production Plant Configuration Figure 3-43 shows the simplified block flow diagram of the Case 3 IGCC Co-Production configuration using ITN ICCM for H₂ separation. The flow scheme has the same major processing steps as the Case 1 IGCC configuration except for the following additional processing steps:

- An ITN Membrane Unit to extract H₂ in the sweetened syngas from AGR; and
- A H₂ compressor system to compress the product hydrogen from about 100 psig to the required B/L pressure of 1000 psig.

**Case 3 - IGCC Hydrogen Co-Production with CO2 Capture
(ICCM with Conventional AGR)**

Figure 3-43. Block diagram for Case 3.

Table 3-20 lists the flow and composition for selected major streams shown on the simplified overall block flow diagram.

Case 3 - Plant Performance and Cost Analysis Table 3-21 summarizes the overall plant performance of this case with estimated costs and itemized annual revenue/expenditure breakdown as shown in Tables 3-22 and 3-23, respectively. Assuming no net revenue credit for the captured CO₂ and \$85.4/MW-hr for exported power, break-even H₂ production for this Case 3 configuration is \$2.96 per 1000 SCF.

Approximately 60% of the ITN ICCM Process Plant cost is attributed to the membrane module (assumed \$100/ft² as quoted in previous DOE analysis), and the balance for feed preheating and H₂ compression.

Table 3-20
Case 3 - IGCC Co-Production with CO2 Capture (ITN ICCM with Conventional AGR)

Stream Number	1.	2.	3.	4.	5.	6.	7.	8.	9.	10.	11.
Stream Description	As-Rec Coal Feed	Oxygen from ASU	Syngas from Gasifier (Include Recycle)	Syngas from HT Gas Cool. (Include Recycle)	Syngas from LT Gas Cool.	Sweet Syngas from AGR	Flue Gas from IGCC/Blr	Hydrogen for Export	H2 Recovery Purge	CO2 to to Sequest. (Impurity free Basis)	Sulfur Byproduct
Vapor Composition, Vol%:	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
N2		3.79	1.12	1.10	0.98	1.66	67.25	0.00	4.42		
O2		95.00	0.00	0.00	0.00	0.00	11.72	0.00	0.00		
CO2		0.00	13.26	13.08	40.98	1.26	2.67	0.00	3.92	100.00	
Ar		1.21	0.36	0.35	0.31	0.53	0.95	0.00	1.42		
H2		0.00	25.73	25.39	51.94	88.16	0.00	99.99	61.50		
CO		0.00	36.01	35.54	2.30	3.90	0.00	0.00	9.62		
H2S		0.00	0.70	0.69	0.61	0.00	0.00	0.00	0.00		
COS		0.00	0.03	0.03	0.03	0.00	0.00	0.00	0.00		
CH4		0.00	3.00	2.96	2.64	4.48	0.00	0.00	11.03		
SO2		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		
H2O		0.00	19.80	20.85	0.20	0.00	17.41	0.00	8.09		
-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
Total Vol%	---	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	---
Molar Flow Rate, lbmoles/hr	---	14,448	61,417	62,237	62,921	37,071	135,244	22,669	14,546	24,868	---
Mass Flow Rate: lb/hr	586,205	461,540	1302096	1316869	1787305	178,838	3689475	45,751	155,514		12,902
" " " : STPD	7,034	5,538	15,625	15,802	21,448	2,146	44,274	549	1,866	13,133	155
Vol Flow Rate, MMSCFD	---	131.6	559.4	566.9	573.1	337.6	1231.8	206.5	132.5	227	---
HHV, MMBtu/Hr		0	5,670	5,685	4,931	4,831	441	2,788	1,907	0	51
Pressure, psig	0.0	695.0	595.0	521.0	386.0	376.0	0.0	1000.0	340.0	2,000	0
Temperature, deg F	60	240	1,850	324	95	95	260	91	862	91	60

Table 3-21

Case 3 - IGCC Co-Production with CO2 Capture Plant Summary

A <u>Import or Feeds:</u>		
Coal Feed, STPD (AR)		7,034
95% Oxygen, STPD		5,538
Make Up Water, GPM		4,335
B <u>Export or Products:</u>		
Electrical Power, MWe		119
Hydrogen, MMSCFD		206
Recovered CO2, STPD		13,134
Sulfur, STPD		155
Slag & Ash (Incl. Carbon), STPD Dry		806
Process Waste Water, GPM		1,979
C <u>Carbon Balance, STPD Equivalent CO2:</u>		
Coal Feed		15,432
Power Plant Combustion Air		20
<u>Total CO2 in:</u>		15,452
Power Plant Stack Emission		1,908
SRU/TGTU Stack Emission		256
Equivalent CO2 in Slag & Ash		154
Recoverable CO2 from AGR		13,134
<u>Total CO2 out:</u>		15,452
D <u>Power Balance, MWe</u>		
Gas Turbine		209.3
Steam Turbine		97.4
<u>Total Generation:</u>		306.7
Coal Prep & Gasification		7.6
ASU & O2 Compression		74.9
Sour Shift & LT Syngas Cooling		1.4
AGR (Acid Gas Removal)		18.5
H2 Recovery & Compression		26.5
Sulfur Recovery		0.6
CO2 Drying & Compression		38.5
Flue Gas Compression		0.0
HRSG/Boiler/BFW Treating		5.3
BOP Consumption		14.4
<u>Total Consumption:</u>		187.7
E Overall Power + H2 Efficiency, %HHV		
		50.0

Table 3-22

Case 3 - Preliminary Capital Cost Estimates

Coal Feed, STPD (AR)	7,034	
CO2 Capture, %	85	
Net Power Export, MWe	119	
H2 Production, MMSCF	206	
<u>Installed Cost, \$MM</u>		<u># Train</u>
Coal Prep & Gasification	221.2	2+1 spare gasifier
ASU & O2 Compression	107.6	2
Sour Shift & LT Syngas Cooling	48.2	2
AGR (Acid Gas Removal)	111.1	2
H2 Recovery & Compression	44.1	1
Sulfur Recovery	10.2	1
CO2 Drying & Compression	52.1	1
Fuel Gas Compression	0.4	
Gas Turbine/Generator	71.1	1
Steam Turbine/Generator	23.2	1
HRSG/Boiler/BFW Treating/Stack	18.1	
BOP (Balance of Plant)	113.9	
<u>Total Field Cost:</u>	821.2	
Home Office Cost	45.0	
Contingency, 15%	123.2	
<u>Total Installed Cost:</u>	989.4	
Installed Cost, \$/Net kW	2410 ⁽¹⁾	
Installed Cost, \$/SCFD H2	3.40 ⁽²⁾	

Note:

(1) Capital cost associated with exporting power is assumed to be identical to that of Case 1

(2) Capital cost associated with exporting H2, \$/SCFD = (TIC - Net MWe export x 2,410)/(MMSCFD H2 produced)

Table 3-23**Case 3 - Annual Revenue & Expense Estimates**

Coal Feed, STPD (AR)	7,034
% CO2 Capture	85
Net Power Export, MWe	119
Hydrogen Export, MMSCFD	206
CO2 Recovered, STPD	13,133
Installed Cost, \$MM	989
<u>Installed Operating Capital, \$MM</u>	35
Total Initial Investment, \$MM	1,024
<u>Annual Expenses, \$MM/Yr:</u>	
Coal Feed Cost (29.4/ST AR)	60.4
Power Inport Cost	0.0
Raw Water Cost	0.5
Catalysts & Chemical Cost	14.0
Ash & Slag Disposal Cost	2.4
Maintenance Labor & Material	29.7
Adminstration Salaries & Overheads	11.8
Insurance & Local Taxes	19.8
Amortized (Over 25 Yr) Investment Repayme	112.7
Federal & State Income Taxes	0.0
Total Expense	251.3
<u>Annual Revenues, \$MM/Yr:</u>	
Power Sale (\$85.4/MWH)	71.1
Hydrogen Sale (By Difference)	178.8
Sulfur Byproduct Sale	1.4
Credit for Recovered CO2	0.0
Total Revenue:	251.3

Task 4.2.2.5 Case 4 – IGCC Hydrogen Co-Production with CO₂ Capture (ITN ICCM Membrane with Warm Gas Clean Up)

Instead of the conventional low temperature (LT) chemical solvent (e.g., amine) or physical solvent (e.g., Selexol) type of acid gas removal processes, Case 4 uses an advance regenerative solid absorbent (e.g., ZnO-TiO₂-based) process for H₂S removal from syngas at around 800 °F (425 °C). This type of Warm Gas Clean Up technology, currently under development by the DOE, offers potential performance and economic improvements to an IGCC process for syngas desulphurization. The process has the advantage that the syngas does not have to be cooled in order to remove H₂S. When coupled into a high temperature H₂ separation membrane system such as ICCM, the syngas cooling and reheating cycle can be minimized. When compared with a LT AGR process such as Selexol, the regenerative solid (ZnO-TiO₂) absorbent Warm Gas Clean Up process currently being developed does not offer selectively removal of CO₂. However, it is envisioned that a separate solid (e.g., CaO based) absorption/regeneration bed can be incorporated for that purpose.

On the premise that an IGCC H₂ co-production facility in the future would certainly require CO₂ capture, the following two Warm Gas Clean Up integration scenarios were screened to see if there is any merit to continue with this option:

- a) Warm Gas Clean Up coupled with ICCM membrane for IGCC H₂ co-production with CO₂ capture – **without water-gas-shift, and relying on post-combustion flue gas CO₂ capture,**
- b) Warm Gas Clean Up coupled with ICCM membrane for IGCC H₂ co-production with CO₂ capture – **with water-gas-shift, and a potential CaO-based absorption/regenerative process for pre-combustion CO₂ removal.**

Preliminary screening results indicated that arrangement (b) above is not feasible. The anticipated cost savings for not having to reheat the shifted and sweetened gas from near ambient temperature up to the Warm Gas Clean Up temperature was overshadowed by the additional costs associated with heating the large amount of excess water in the shifted gas from 800 °F to the ICCM membrane operating temperature of 1500 °F. In addition, this large amount of excess water also dilutes the ICCM purge gas heating value way below the GT minimum requirement of 150 Btu/SCF. To increase this ICCM purge gas heating value would require cooling the purge gas to near ambient temperature, dropping off some excess water, before feeding it to the turbine. This would make arrangement (b) similar to the ICCM with conventional AGR option of Case 3. This, coupled with the design uncertainty of the suggested warm CaO absorption/regenerative process for CO₂ removal, arrangement (b) will not be further investigated.

Techno-economic evaluation of arrangement (a) was carried out to see if it has any merit, recognizing that the process scheme may forfeit any seemingly efficiency advantage associated with pre-combustion CO₂ capture. Process design and cost evaluation of the Warm Gas Clean Up process is outside of the current scope of work. For our case study, it is assumed that it has the same capital cost as a conventional LT AGR process. This was done on the premise that if the overall techno-economic analysis of arrangement (a) shows promise, future sensitivity analysis can be performed with, perhaps, a more realistic set of cost data.

Case 4 – IGCC Hydrogen Co-Production Plant Configuration Figure 3-44 is a simplified overall block flow diagram for this case scenario – *i.e.*, scenario a) cited above. Its IGCC configuration is similar to that of Case 3 with the exception of the following:

- Sour Shift and Low temperature Syngas Cooling are eliminated;
- A solid adsorbent bed is used downstream of the gasifier to removal chloride from the syngas, instead of water wash;
- Regenerative ZnO-TiO₂ type of solid absorbent is used for H₂S removal from syngas, instead of a Selexol unit;
- Sulfuric acid plant is used instead of Claus SRU to recover sulfur for export in the form of sulfuric acid; and
- An amine absorption unit is added for CO₂ capture from flue gas.

Without water-gas-shift, H₂ production is limited to its content in the syngas from gasification, which for the current E-GasTM design is only about 26%. Thus, it is assumed that the ITN ICCM membrane can be operated at a H₂ partial pressure differential as low as 3 to 5 psi. Due to the limited H₂ recovery possible for this case, syngas from one large E-GasTM gasifier is sufficient to recover the H₂ and supplying the required fuel demand for one 7FA+e gas turbine.

The overall process, as shown in Figure 3-44 and Table 3-24, involves generating the syngas from one 3000 STPD (dry) E-GasTM gasifier. The syngas is then cleaned and cooled to 800 °F going through the two-stage hot cyclone/radiant boiler/dry char filter, before routing to the Warm Gas Clean Up unit for H₂S removal. The 800 °F sulfur-free syngas, without any shift, is then sent on to the ITN ICCM Process Plant where H₂ is recovered. The 1500 °F purge gas leaving the ICCM is heat exchanged and cooled before sending to the 7FA+e turbine as fuel. CO₂ is captured from the flue gas by a commercial amine absorption/regeneration process.

Case 4 - Plant Performance Summary and Cost Analysis Table 3-25 summarizes the overall plant performance of this case with estimated costs and itemized annual revenue/expenditure breakdown as shown in Tables 3-26 and 3-27, respectively. Assuming no net revenue credit for the captured CO₂ and \$85.4/MW-hr for exported power, break-even hydrogen production for this Case 4 configuration is \$5.05 per 1000 SCF.

**Case 4 - IGCC Hydrogen Co-Production with CO2 Capture
(ICCM with Warm Gas Clean Up)**

Figure 3-44. Block diagram for Case 4.

Table 3-24

Case 4 - IGCC Co-Production with CO2 Capture (ITN ICCM with Warm Gas Clean Up)

Stream Number	1.	2.	3.	4	5	6	7	8	9	10
Stream Description	As-Rec Coal Feed	Oxygen from ASU	Syngas from Gasifier (Include Recycle)	Cooled Syngas	Sweet Syngas from AGR	Flue Gas from IGCC/Blr	Hydrogen for Export	H2 Recovery Purge	CO2 to to Sequest. (Impurity free Basis)	Sulfur Byproduct
Vapor Composition, Vol%:	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
N2		3.79	1.12	1.12	1.13	67.95	0.00	1.42		
O2		95.00	0.00	0.00	0.00	12.60	0.00	0.00		
CO2		0.00	13.28	13.28	13.36	10.85	0.00	17.40	100.00	
Ar		1.21	0.36	0.36	0.36	0.89	0.00	0.45		
H2		0.00	25.77	25.77	25.96	0.00	99.99	8.24		
CO		0.00	36.06	36.06	36.33	0.00	0.00	43.70		
H2S		0.00	0.70	0.70	0.00	0.00	0.00	0.00		
COS		0.00	0.03	0.03	0.00	0.00	0.00	0.00		
CH4		0.00	3.01	3.01	3.03	0.00	0.00	3.65		
SO2		0.00	0.00	0.00	0.00	0.00	0.00	0.00		
H2O		0.00	19.67	19.67	19.82	7.71	0.00	25.14		
-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
Total Vol%	---	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	---
Molar Flow Rate, lbmoles/hr	---	7,092	30,645	27,580	27,373	133,379	5,126	22,292	12,445	---
Mass Flow Rate: lb/hr	293,102	226,552	649,875	584,887	577,534	3946107	10,347	574,182	547,726	6,451
" " " : STPD	3,517	2,719	7,798	7,019	6,930	47,353	124	6,890	6,573	77
Vol Flow Rate, MMSCFD	---	64.6	279.1	251.2	249.3	1214.8	46.7	203.0	113	---
HHV, MMBtu/Hr	3,195	0	2,832	2,549	2,500	193	630	1,824	0	26
Pressure, psig	0.0	565.0	465.0	415.0	405.0	0.0	1000.0	340.0	2,000	0
Temperature, deg F	60	240	1,850	800	800	216	91	995	91	60

Table 3-25

Case 4 - IGCC Co-Production with CO2 Capture Plant Summary

A <u>Import or Feeds:</u>		
Coal Feed, STPD (AR)		3,517
95% Oxygen, STPD		2,719
Make Up Water, GPM		7,333
B <u>Export or Products:</u>		
Electrical Power, MWe		152
Hydrogen, MMSCFD		47
Recovered CO2, STPD		6,573
Sulfur, STPD		77
Slag & Ash (Incl. Carbon), STPD Dry		403
Process Waste Water, GPM		2,648
C <u>Carbon Balance, STPD Equivalent CO2:</u>		
Coal Feed		7,716
Power Plant Combustion Air		20
<u>Total CO2 in:</u>		7,736
Power Plant Stack Emission		1,070
SRU/TGTU Stack Emission		16
Equivalent CO2 in Slag & Ash		77
Recoverable CO2 from AGR		6,573
<u>Total CO2 out:</u>		7,736
D <u>Power Balance, MWe</u>		
Gas Turbine		209.5
Steam Turbine		52.4
<u>Total Generation:</u>		261.9
Coal Prep & Gasification		3.8
ASU & O2 Compression		36.0
Sour Shift & LT Syngas Cooling		1.4
AGR (Acid Gas Removal)		0.6
H2 Recovery & Compression		10.6
Sulfur Recovery		0.1
CO2 Drying & Compression		26.3
Flue Gas Compression		0.0
HRSG/Boiler/BFW Treating		4.4
Flue Gas CO2 Recovery		5.0
BOP Consumption		23.0
<u>Total Consumption:</u>		111.2
E Overall Power + H2 Efficiency, %HHV		
		36.0

Table 3-26

Case 4 - Preliminary Capital Cost Estimates

Coal Feed, STPD (AR)	3,517	
CO2 Capture, %	85	
Net Power Export, MWe	152	
H2 Production, MMSCF	47	
<u>Installed Cost, \$MM</u>		<u># Train</u>
Coal Prep & Gasification	120.2	1+1 spare gasifier
ASU & O2 Compression	53.1	1
Sour Shift & LT Syngas Cooling	4.1	1
AGR (Acid Gas Removal)	9.4	1
H2 Recovery & Compression	21.6	1
Sulfur Recovery	20.5	1
CO2 Drying & Compression	39.0	1
Fuel Gas Compression	0.5	1
Gas Turbine/Generator	71.1	1
Steam Turbine/Generator	10.4	1
HRSG/Boiler/BFW Treating/Stack	22.2	1
Flue Gas CO2 Recovery	53.5	2
BOP (Balance of Plant)	117.8	
<u>Total Field Cost:</u>	543.4	
Home Office Cost	35.1	
Contingency, 15%	81.5	
<u>Total Installed Cost:</u>	660.0	
Installed Cost, \$/Net kW	2,410 ⁽¹⁾	
Installed Cost, \$/SCFD H2	6.27 ⁽²⁾	

Note:

(1) Capital cost associated with exporting power is assumed to be identical to that of Case 1

(2) Capital cost associated with exporting H2, \$/SCFD = (TIC - Net MWe export x 2,410)/(MMSCFD H2 produced)

Table 3-27

Case 4 - Annual Revenue & Expense Estimates

Coal Feed, STPD (AR)	3,517
% CO2 Capture	85
Net Power Export, MWe	152
Hydrogen Export, MMSCFD	47
CO2 Recovered, STPD	6,573
Installed Cost, \$MM	660
<u>Installed Operating Capital, \$MM</u>	22
Total Initial Investment, \$MM	682
<u>Annual Expenses, \$MM/Yr:</u>	
Coal Feed Cost (29.4/ST AR)	30.2
Power Inport Cost	0.0
Raw Water Cost	0.8
Catalysts & Chemical Cost	9.2
Ash & Slag Disposal Cost	1.2
Maintenance Labor & Material	19.8
Adminstration Salaries & Overheads	11.8
Insurance & Local Taxes	13.2
Amortized (Over 25 Yr) Investment Repayme	75.1
Federal & State Income Taxes	0.0
Total Expense	161.3
<u>Annual Revenues, \$MM/Yr:</u>	
Power Sale (\$85.4/MWH)	91.1
Hydrogen Sale (By Difference)	68.8
Sulfur Byproduct Sale	1.4
Credit for Recovered CO2	0.0
Total Revenue:	161.3

Task 4.2.2.6 Case 5 – Coal-Based Hydrogen Production with CO₂ Capture (Conventional PSA)

Cases 5 and 6 were developed to assess/compare the techno-economics of ITN's ICCM technology under a coal-gasification-based, H₂-production-only scenario. These scenarios are designed to maximize H₂ production/extraction from the produced syngas and minimize the generation of power, which is a low-value commodity. It was believed that under this type of scenario, ICCM might have a cost advantage over PSA, especially with low electricity costs. With these scenarios, onsite power generation facility is still provided as a means of disposal of the residual purge gas to generate as much power as possible for in-plant consumption. Both design configurations will include CO₂ capture. Thus the techno-economics of coal-based hydrogen production will include CO₂ captured costs.

Case 5 investigates coal-based hydrogen production using conventional PSA technology. A two-stage, sour shift is used to convert 90%+ of the CO in the syngas to H₂, followed by a conventional selective Selexol for H₂S and CO₂ removal. A gas-fired boiler is used to dispose of the low-pressure PSA purge gas for power generation. Power is to be imported as required to satisfy all in-plant needs.

Case 5 - Coal Based Hydrogen Production Plant Configuration Figure 3-45 shows a simplified block flow diagram for this coal-gasification-based, H₂ production plant designed for 85% CO₂ capture. The design is similar to Case 2 of IGCC H₂ co-production scheme discussed in Task 4.2.2.3 (pg 64 ff.) except for the following:

- Has only one E-GasTM gasification train, with a spare gasifier included; and
- Boiler-based power plant is used instead of a gas-turbine-based, combined cycle plant.

Table 3-28 lists the flow and composition for selected major streams identified on the simplified overall block flow diagram.

Case 5 - Plant Performance Summary and Cost Analysis

Table 3-29 summarizes the overall plant performance of this case with estimated costs and the itemized annual revenue/expenditure breakdown as shown in Tables 3-30 and 3-31, respectively. Assuming no net revenue credit for the captured CO₂ and \$40/MW-hr for imported power, break-even H₂ production for this Case 5 scheme is \$3.55 per 1000 SCF.

Sensitivity analyses were performed as a function of various cost and economic scenarios, results of which will be discussed in Section 3.4.2.3 (pp 104-112), in conjunction with the results of Case 6.

Case 5 - Coal Gasification Based Hydrogen Production with CO2 Capture (Conventional PSA)

Figure 3-45. Simplified block diagram for Case 5.

Table 3-28

Case 5 - Coal Gasification Based Hydrogen Production with CO2 Capture (Conventional PSA)

Stream Number	1.	2.	3.	4.	5.	6.	7.	8.	9.	10.	11.
Stream Description	As-Rec Coal Feed	Oxygen from ASU	Syngas from Gasifier (Include Recycle)	Syngas from HT Gas Cool. (Include Recycle)	Syngas from LT Gas Cool.	Sweet Syngas from AGR	Flue Gas from Boiler	Hydrogen for Export	H2 Recovery Purge	CO2 to to Sequest. (Impurity free Basis)	Sulfur Byproduct
Vapor Composition, Vol%:	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
N2		0.38	0.27	0.26	0.23	0.40	67.15	0.01	2.01		
O2		99.50	0.00	0.00	0.00	0.00	1.63	0.00	0.00		
CO2		0.00	13.41	13.19	41.36	1.49	10.50	0.00	7.66	100.00	
Ar		0.12	0.08	0.08	0.07	0.13	0.93	0.00	0.65		
H2		0.00	26.01	25.60	52.42	89.47	0.00	99.99	45.95		
CO		0.00	36.40	35.82	2.33	3.97	0.00	0.00	20.39		
H2S		0.00	0.70	0.69	0.62	0.00	0.00	0.00	0.00		
COS		0.00	0.03	0.03	0.03	0.00	0.00	0.00	0.00		
CH4		0.00	3.04	2.99	2.66	4.55	0.00	0.00	23.35		
SO2		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		
H2O		0.00	20.04	21.33	0.28	0.00	19.78	0.00	0.00		
Total Vol%	---	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	---
Molar Flow Rate, lbmoles/hr	---	6,562	30,354	30,850	31,151	18,250	17,437	14,697	3,554	12,388	---
Mass Flow Rate: lb/hr	293,102	209,948	640,191	649,121	874,276	81,457	486,351	29,666	51,791		6,451
" " " : STPD	3,517	2,519	7,682	7,789	10,491	977	5,836	356	621	6,542	77
Vol Flow Rate, MMSCFD	---	59.8	276.5	281.0	283.7	166.2	158.8	133.9	32.4	113	---
HHV, MMBtu/Hr	3,195	0	2,833	2,843	2,464	2,414	65	1,807	607	0	26
Pressure, psig	0.0	635.0	535.0	461.0	326.0	316.0	0.0	1000.0	45.0	2,000	0
Temperature, deg F	60	240	1,850	317	100	95	260	91	176	91	60

Table 3-29

Case 5 - Coal Gasification Based Hydrogen Production with CO2 Capture Plant Summary

A <u>Import or Feeds:</u>		
Coal Feed, STPD (AR)		3,517
99.5% Oxygen, STPD		2,519
Power Import, MW		18
Make Up Water, GPM		2,157
B <u>Export or Products:</u>		
Electrical Power, MWe		0
Hydrogen, MMSCFD		134
Recovered CO2, STPD		6,542
Sulfur, STPD		77
Slag & Ash (Incl. Carbon), STPD Dry		403
Process Waste Water, GPM		1,014
C <u>Carbon Balance, STPD Equivalent CO2:</u>		
Coal Feed		7,716
Power Plant Combustion Air		3
<u>Total CO2 in:</u>		7,719
Power Plant Stack Emission		966
SRU/TGTU Stack Emission		132
Equivalent CO2 in Slag & Ash		77
Recoverable CO2 from AGR		6,542
<u>Total CO2 out:</u>		7,717
D <u>Power Balance, MWe</u>		
Gas Turbine		-0.9
Steam Turbine		75.4
<u>Total Generation:</u>		74.5
Coal Prep & Gasification		3.8
ASU & O2 Compression		38.1
Sour Shift & LT Syngas Cooling		0.7
AGR (Acid Gas Removal)		8.5
H2 Recovery & Compression		8.7
Sulfur Recovery		0.3
CO2 Drying & Compression		19.2
Flue Gas Compression		2.6
HRSG/Boiler/BFW Treating		2.1
Flue Gas CO2 Recovery		0.0
BOP Consumption		7.9
<u>Total Consumption:</u>		91.9
E Overall Power + H2 Efficiency, %HHV		
		54.7

Table 3-30

Case 5 - Preliminary Cost Estimation Summary
(Conventional PSA)

Coal Feed, STPD (AR)	3,517
Oxygen Purity, vol%	99.5
CO2 Capture, %	85
Net Power Export, MWe	-17.6
H2 Production, MMSCF	134
<u>Installed Cost, \$MM</u>	
Coal Prep & Gasification	120.2
ASU & O2 Compression	52.0
Sour Shift & LT Syngas Cooling	19.6
AGR (Acid Gas Removal)	55.4
H2 Recovery & Compression	34.4
Sulfur Recovery	6.3
CO2 Drying & Compression	32.0
Fuel Gas Compression	3.2
Gas Turbine/Generator	0.0
Steam Turbine/Generator	14.5
HRSG/Boiler/BFW Treating/Stack	14.4
Flue Gas CO2 Recovery	0.0
BOP (Balance of Plant)	72.3
<u>Total Field Cost:</u>	424.3
Home Office Cost	22.0
Contingency, 15%	63.6
Total Installed Cost:	509.9
Installed Cost, \$/SCFD H2	3.81

Table 3-31

Case 5 - Annual Revenue & Expense Estimate Breakdown
(Conventional PSA)

Coal Feed, STPD (AR)	3,517
% CO2 Capture	85
Net Power Export, MWe	-17.6
Hydrogen Export, MMSCFD	134
CO2 Recovered, STPD	6,542
Installed Cost, \$MM	510
<u>Installed Operating Capital, \$MM</u>	21
Total Initial Investment, \$MM	531
<u>Annual Expenses, \$MM/Yr:</u>	
Coal Feed Cost (29.4/ST AR)	30.2
Power Import Cost	4.9
Raw Water Cost	0.2
Catalysts & Chemical Cost	7.2
Ash & Slag Disposal Cost	1.2
Maintenance Labor & Material	15.3
Administration Salaries & Overheads	11.8
Insurance & Local Taxes	10.2
Amortized (Over 25 Yr) Investment Repayme	58.5
Federal & State Income Taxes	0.0
Total Expense	139.5
<u>Annual Revenues, \$MM/Yr:</u>	
Power Sale (\$85.4/MWH)	0
Hydrogen Sale (By Difference)	138.8
Sulfur Byproduct Sale	0.7
Credit for Recovered CO2	0.0
Total Revenue:	139.5

Task 4.2.2.7 Case 6 – Coal Based Hydrogen Production with CO₂ Capture (ITN ICCM Membrane)

Case 6 Coal Gasification Based H₂ Production with ITN Membrane – Plant Configuration

Figure 3-46 shows a simplified block flow diagram of this Case 6, coal-gasification, H₂ production plant, designed for 85% CO₂ capture. A two-stage sour shift is included to convert 90%+ of CO in the syngas to H₂, followed by conventional low temperature AGR (Selexol) to remove H₂S and capture CO₂, prior to sending the clean syngas to the ICCM membrane for H₂ separation. Onsite power generation is provided with a combined cycle plant based on a GE 6B gas turbine, as a means of disposing the residue gas from the ICCM plant. The plant requires small additional amount of import power to meet all in-plant demands. This Case 6 has a similar configuration as Case 3 with the exception of the following:

- Only one E-GasTM gasification train is used, with a spare gasifier; and
- The power generation combined cycle plant is based on a GE 6B instead of the 7FA+e gas turbine.

Table 3-32 lists the flow and composition for selected major streams identified on Figure 3-46.

Case 6 - Plant Performance Summary and Cost Analysis Table 3-33 summarizes the overall plant performance of this case with estimated costs and the itemized annual revenue/expenditure breakdown as shown in Tables 3-34 and 3-35, respectively. Assuming no net revenue credit for the captured CO₂ and \$40/MW-hr for imported power, break-even H₂ production for this Case 6 scheme is \$3.61 per 1000 SCF.

Coal-based H₂ production cost sensitivity analyses are performed as a function of the following economic variables, and compared with the results of Case 5 as shown in Section 3.4.2.3 (pp 104-112):

- Price of imported power;
- Price of as-received coal;
- % return on investment;
- Plant life; and
- Capital cost contingency.

Case 6 - Coal Gasification Based Hydrogen Production with
CO2 Capture (ICCM Membrane)

Figure 3-46. Simplified block diagram for Case 6.

Table 3-32

Case 6 - Coal Gasification Based Hydrogen Production with CO2 Capture (ITN ICCM)

Stream Number	1.	2.	3.	4.	5.	6.	7.	8.	9.	10.	11.
Stream Description	As-Rec Coal Feed	Oxygen from ASU	Syngas from Gasifier (Include Recycle)	Syngas from HT Gas Cool. (Include Recycle)	Syngas from LT Gas Cool.	Sweet Syngas from AGR	Flue Gas from IGCC/Blr	Hydrogen for Export	H2 Recovery Purge	CO2 to to Sequest. (Impurity free Basis)	Sulfur Byproduct
Vapor Composition, Vol%:	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
N2		0.38	0.27	0.26	0.23	0.40	67.61	0.00	2.06		
O2		99.50	0.00	0.00	0.00	0.00	12.18	0.00	0.00		
CO2		0.00	13.41	13.20	41.38	1.45	4.29	0.00	8.66	100.00	
Ar		0.12	0.08	0.08	0.07	0.13	0.86	0.00	0.66		
H2		0.00	26.02	25.62	52.44	89.50	0.00	99.99	31.53		
CO		0.00	36.41	35.85	2.33	3.97	0.00	0.00	19.60		
H2S		0.00	0.70	0.69	0.62	0.00	0.00	0.00	0.00		
COS		0.00	0.03	0.03	0.03	0.00	0.00	0.00	0.00		
CH4		0.00	3.04	2.99	2.67	4.55	0.00	0.00	22.44		
SO2		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		
H2O		0.00	20.04	21.26	0.23	0.00	15.07	0.00	15.06		
-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----	-----
Total Vol%	---	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	---
Molar Flow Rate, lbmoles/hr	---	6,870	30,352	30,825	31,136	18,242	42,705	14,708	3,416	12,395	---
Mass Flow Rate: lb/hr	293,102	219,787	640,145	648,670	891,049	81,118	1186377	29,683	58,369	545,509	6,451
" " " : STPD	3,517	2,637	7,682	7,784	10,693	973	14,237	356	700	6,546	77
Vol Flow Rate, MMSCFD	---	62.6	276.4	280.8	283.6	166.1	389.0	134.0	31.1	113	---
HHV, MMBtu/Hr	3,195	0	2,833	2,842	2,464	2,414	121	1,809	517	0	26
Pressure, psig	0.0	645.0	545.0	471.0	336.0	326.0	0.0	1000.0	285.0	2,000	0
Temperature, deg F	60	240	1,850	318	95	95	260	91	1,066	91	60

Table 3-33

Case 6 - Coal Gasification Based Hydrogen Production with CO2 Capture Plant Summary
(ITN ICCM with Conventional AGR)

A <u>Import or Feeds:</u>		
Coal Feed, STPD (AR)		3,517
99.5% Oxygen, STPD		2,637
Power Imported, MW		26
Make Up Water, GPM		2,098
B <u>Export or Products:</u>		
Electrical Power, MWe		0
Hydrogen, MMSCFD		134
Recovered CO2, STPD		6,546
Sulfur, STPD		77
Slag & Ash (Incl. Carbon), STPD Dry		403
Process Waste Water, GPM		977
C <u>Carbon Balance, STPD Equivalent CO2:</u>		
Coal Feed		7,716
Power Plant Combustion Air		6
<u>Total CO2 in:</u>		7,722
Power Plant Stack Emission		966
SRU/TGTU Stack Emission		132
Equivalent CO2 in Slag & Ash		77
Recoverable CO2 from AGR		6,547
<u>Total CO2 out:</u>		7,722
D <u>Power Balance, MWe</u>		
Gas Turbine		51.7
Steam Turbine		33
<u>Total Generation:</u>		84.7
Coal Prep & Gasification		3.8
ASU & O2 Compression		40.0
Sour Shift & LT Syngas Cooling		0.7
AGR (Acid Gas Removal)		8.7
H2 Recovery & Compression		28.3
Sulfur Recovery		0.3
CO2 Drying & Compression		19.2
Fuel Gas Compression		0.0
HRSG/Boiler/BFW Treating		2.0
BOP Consumption		7.4
<u>Total Consumption:</u>		110.4
E Overall Power + H2 Efficiency, %HHV		
		53.9

Table 3-34

Case 6 - Preliminary Plant Cost Estimates

Coal Feed, STPD (AR)	3,517
CO2 Capture, %	85
Net Power Export, MWe	-25.7
H2 Export, MMSCF	134
<u>Installed Cost, \$MM</u>	
Coal Prep & Gasification	120.3
ASU & O2 Compression	53.6
Sour Shift & LT Syngas Cooling	20.6
AGR (Acid Gas Removal)	55.4
H2 Recovery & Compression	24.8
Sulfur Recovery	6.3
CO2 Drying & Compression	32.0
Fuel Gas Compression	0.1
Gas Turbine/Generator	20.2
Steam Turbine/Generator	11.6
HRSG/Boiler/BFW Treating/Stack	6.8
BOP (Balance of Plant)	72.0
<u>Total Field Cost:</u>	423.7
Home Office Cost	23.7
Contingency, 15%	63.6
<u>Total Installed Cost:</u>	511.0
Installed Cost, \$/SCFD H2	3.81

Table 3-35

Case 6 - Annual Estimated Revenues & Expenditure Breakdown

Coal Feed, STPD (AR)	3,517
% CO2 Capture	85
Net Power Export, MWe	-25.7
Hydrogen Export, MMSCFD	134
CO2 Recovered, STPD	6,546
Installed Cost, \$MM	511
<u>Installed Operating Capital, \$MM</u>	22
Total Initial Investment, \$MM	533
<u>Annual Expenses, \$MM/Yr:</u>	
Coal Feed Cost (29.4/ST AR)	30.2
Power Inport Cost	7.2
Raw Water Cost	0.2
Catalysts & Chemical Cost	7.2
Ash & Slag Disposal Cost	1.2
Maintenance Labor & Material	15.3
Adminstration Salaries & Overheads	11.8
Insurance & Local Taxes	10.2
Amortized (Over 25 Yr) Investment Repayme	58.8
Federal & State Income Taxes	0.0
Total Expense	142.1
<u>Annual Revenues, \$MM/Yr:</u>	
Power Sale	0
Hydrogen Sale (By Difference)	141.4
Sulfur Byproduct Sale	0.7
Credit for Recovered CO2	0.0
Total Revenue:	142.1

Case 6 – Coal-Based Hydrogen Production Cost Sensitivity Analysis

Price of Electricity Figures 3-47 compares the break-even H_2 production cost as a function of imported electricity price for Case 6 (ITN membrane) vs. Case 5 (PSA) configuration scenario.

Figure 3-47. Plot of H_2 cost as a function of cost of imported electricity.

Price of Coal Figure 3-48 compares the break-even H_2 production cost as a function of feed coal price for Case 6 (ITN membrane) vs. Case 5 (PSA) configuration scenario.

Figure 3-48. Plot of H_2 production cost as a function of cost of coal feed.

Return On Investment Figure 3-49 shows the sensitivity of the break-even H_2 production cost to the % return-on-investment.

Figure 3-49. Plot of H_2 production cost as a function of return on investment.

Plant Life Figure 3-50 shows the sensitivity of the break-even H_2 production cost to the plant life.

Capital Cost Contingency - Figure 3-51 shows the sensitivity of the break-even H_2 cost to the Capital Cost Contingency for the PSA-based and the ITN-based H_2 production plants.

Figure 3-51. Plot of H_2 production cost as a function of cost contingency.

3.4.2.3 Tasks 4.2 and 4.3 Results and Discussion

IGCC H_2 Power Co-Production – ICCM versus PSA for H_2 Separation

Tables 3-36 to 3-38 show the estimated performance and costs of the three IGCC co-production flow schemes investigated, providing a side-by-side comparison with each other as well as with that of the Reference Case and Case 1 IGCC design. The objective is to compare the ICCM membrane technology with PSA for hydrogen separation from syngas within the context of an IGCC hydrogen co-production facility, with CO_2 capture. As cited earlier, the Reference Case was chosen to represent a current state-of-the-art IGCC design. With the Case 1 results, they collectively define a baseline performance and cost estimates with which the techno-economics of the ICCM membrane can be evaluated and compared with PSA.

The reference case was developed based on a configuration, with all the major IGCC technology components, similar to that of Subtask 1.5A Optimized Single-Train Coal Based IGCC Design as described in the 2003 DOE study [DE-AC26-99FT40342] entitled “Gasification Plant Cost and Performance Optimization.” While the Reference Case design was developed independently, its estimated performance and costs are comparable with the cited 2003 DOE study. It needs to be emphasized, however, that the developmental objective of both the Reference and Case 1 designs was not to establish an engineering cost estimates for the construction and operation of an IGCC H_2 power co-production facility, but rather to obtain a reference point usable for a

systemic and consistent comparison of the ICCM versus PSA technology. While integration needs and specific product slate and CO₂ capture requirements necessitate the using of slightly different IGCC plant configurations for comparison, key IGCC technology components (*e.g.*, gasification and power train) were by and large being kept the same.

The IGCC configuration and process description of the six design scenarios were summarized in Section 3.4.2.2 above (pp 44-47) and specific details followed (pp 47-99) and will not be repeated here. Key economic assumptions used in estimating their power and hydrogen production costs, as cited in Task 4.2.1.2 (pp 42-44), include:

- \$29.3 per short ton for Illinois #6 coal on as-received basis,
- Estimated plant capital (installed) cost includes 15% contingency,
- Estimated total investment cost includes 3 months of operating expense, and
- For break-even analysis, capital investment is amortized over the 25-year life of the plant, at an assumed interest rate of 10%.

The Reference Case establishes a baseline installed cost of \$1,581 per kW for IGCC, with a net efficiency of 38%; and under the economic assumptions and methodology used for the current study, has a break-even power production cost of \$58.7 per MWe-hr. By comparing the Case 1 design performance and costs with that of the Reference Case, IGCC plant efficiency and cost penalty incurred for capturing 85% of the produced CO₂ were determined. The estimated break-even pre-combustion IGCC CO₂ capture cost is about \$26.0 per ton of CO₂, with an incurred efficiency loss, reducing the overall plant efficiency from 38% to 28.5%. The overall efficiency loss is mainly attributed to:

- Loss in gross power production due to a lower syngas heating value and steam consumed by the water-gas-shift unit,
- Increased in power consumption with the use of a selective Selexol for pre-combustion CO₂ capture, and
- Captured CO₂ compression.

As shown in Table 3-36, of the three IGCC H₂ co-production schemes (Case 2 to Case 4) investigated, the two designs using ICCM membrane are not as attractive in comparison with that of using PSA. The overall plant efficiency decreases from 51.5% for Case 2 (using PSA) to 50.0% with ICCM. Case 4 of using ICCM with Warm Gas Clean Up suffers an even larger decrease in efficiency, from 51.5% to 36.0% that is attributed mainly to the inclusion of a post-combustion flue gas CO₂ recovery unit.

IGCC Co-Production Efficiency In comparing these IGCC co-production schemes (as well as with that of the Reference Case) in terms of overall plant efficiency, it is worthy to take note that the overall efficiency is defined as the ratio of the net power export plus H₂ product HHV divided by the coal HHV. For direct comparison of the Reference Case and the Case 1 power-production-only IGCC efficiency, a hydrogen-to-power convention efficiency, which depends on the power cycle assumed, needs to be defined. Table 3-39 below shows how the H₂-to-electric power conversion efficiency may impact the overall estimated efficiency for the various IGCC schemes evaluated.

Table 3-36

Overall IGCC Co-Production Case Performance & Efficiency Compared

	Reference Case IGCC	Case 1 IGCC w/CO ₂ Capture	Case 2 <-- IGCC Hydrogen Co-Production with 85% CO ₂ Capture --> [with PSA]	Case 3 [ICCM ⁽¹⁾]	Case 4 [ICCM ⁽²⁾]
Coal Feed, STPD (AR)	2,782	2,937	7,034	7,034	3,517
Coal Feed HHV, MW	740.4	781.6	1,872	1,872	936.1
Coal Feed Carbon, STPD CO ₂ Equivalent	6,103	6,433	15,432	15,432	7,716
CO ₂ Captured, STPD	0	5,472	13,091	13,133	6,573
% Recovered	0	85	85	85	85
Gross Power Output, MWe					
Gas Turbine	209.3	209.1	209.3	209.3	209.5
Steam Turbine	116.0	82.6	92.2	97.4	52.4
Total Generation:	325.3	291.7	301.5	306.7	261.9
Power Consumed, MWe					
Coal Prep & Gasification	3.0	3.2	7.6	7.6	3.8
ASU & O ₂ Compression	27.5	29.6	70.5	74.9	36.0
Sour Shift & LT Syngas Cooling	0.0	0.5	1.4	1.4	0
AGR (Acid Gas Removal)	0.4	7.5	17.1	18.5	0.6
H ₂ Recovery & Compression	0.0	0	13.8	26.6	10.6
Sulfur Recovery	0.3	0	0.6	0.6	0.1
CO ₂ Compression	0.0	16.1	38.4	38.50	26.3
Fuel Gas Compression	0.0	0	9.8	0	0
HRSG/Boiler/BFW Treating	3.9	3.9	5.2	5.3	4.4
Flue Gas CO ₂ Recovery	0.0	0	0	0	5.0
BOP Consumption	8.8	8.1	13.7	14.4	23.0
Total Consumption:	43.9	69.2	178.1	187.8	109.8
Net Power Export, MWe	281.4	222.6	123.3	118.9	152.3
Hydrogen Export, MMSCFD	N/A	N/A	212.4	206.5	46.7
Net Power + H ₂ Efficiency, %HHV	38.0	28.5	51.5	50.0	36.0

NOTE - (1) ICCM membrane w/conventional low temperature AGR
(2) ICCM membrane w/Warm Syngas Clean Up

Estimated Capital and Break-Even Production Costs Table 3-37 shows the cost comparison for the various cases. Co-production designs of both Case 2 (with PSA) and Case 3 (with ICCM) have a comparable capital cost of roughly \$990MM; Case 4 (ICCM with Warm Gas Clean Up) has only one gasification train, but includes two flue gas CO₂ capture plants which cost about \$53.5MM.

Break-even H₂ production costs are estimated with and without CO₂ captured credit. To calculate these production costs without any CO₂ recovery credit, it is assumed that power can be exported at the Case 1 break-even production cost of \$85.4/MWe-hr. With CO₂ recovery credit, it is assumed that the difference in power production (sale) revenue for Case 1 (with 222.6 MWe net export) between its estimated break-even cost of \$85.4/MWe-hr and that of \$58.7/MWe-hr for the Reference Case is the break-even cost associated with CO₂ capture. Case 1 recovered 5,472 tons per day of CO₂, yielding a break-even CO₂ recovery credit cost of \$26.0 per ton.

As shown in Table 3-38, Case 2 of using PSA for IGCC H₂ co-production has a lower H₂ cost at \$2.83 per 1000 SCF. Case 3 using ICCM membrane have a 5% higher production cost at about \$2.96/1000 SCF because of its lower efficiency due to higher O₂ and H₂ compression power demands. Case 4 using ICCM with Warm Gas Clean Up has a much higher H₂ production cost at about \$5.05/1000 SCF mainly due to higher capital cost associated with the inclusion of post-combustion flue gas CO₂ capture.

Table 3-37

IGCC Co-Production Case Capital Cost Compared

	Reference Case	Case 1	Case 2	Case 3	Case 4
	IGCC	IGCC	<-- IGCC H ₂ Co-Production with 85% CO ₂ Capture -->		
		w/CO ₂ Capture	[with PSA]	[ICCM ⁽¹⁾]	[ICCM ⁽²⁾]
Coal Feed, STPD (AR)	2,782	2,937	7,034	7,034	3,517
Net Power Export, MWe	281.4	222.6	123.3	118.9	152.3
Hydrogen Export, MMSCFD	N/A	N/A	212.4	206.5	46.7
CO ₂ Captured, STPD	0	5,472	13,091	13,133	6,573
% Recovered	0	85	85	85	85
Installed Cost, \$1,000,000					
Coal Prep & Gasification	102.0	105.9	221.2	221.2	120.2
ASU & O ₂ Compression	44.1	45.8	103.9	107.6	53.1
Sour Shift & LT Syngas Cooling	6.4	17.5	46.3	48.2	4.1
AGR (Acid Gas Removal)	9.5	48.9	110.9	111.1	9.4
H ₂ Recovery & Compression	0.0	0	48	44.1	21.6
Sulfur Recovery	5.3	6	10.2	10.2	20.5
CO ₂ Compression	0.0	28.2	52	52.00	39
Fuel Gas Compression	0.4	0.3	5.1	0.4	0.5
Gas Turbine/Generator	71.1	71.1	71.1	71.1	71.1
Steam Turbine/Generator	19.5	18.8	23.3	23.2	10.4
HRSG/Boiler/BFW Treating	21.2	15.6	17.2	18.1	22.2
Flue Gas CO ₂ Recovery	0.0	0	0	0	53.5
BOP Consumption	91.4	87.8	112.2	113.9	117.8
Total Field Cost	370.9	445.4	821.4	821.1	543.4
Home Office Cost	18.3	24.0	41.6	45.0	35.1
Contingency	55.6	66.8	123.2	123.2	81.5
Total Installed Cost	444.8	536.2	986.2	989.3	660.0
Installed Cost, \$/Net kW	1,581	2,410	2,410 ⁽³⁾	2,410 ⁽³⁾	2,410 ⁽³⁾
Installed Cost, \$/SCFD H ₂	N/A	N/A	3.24	3.40	6.27

OTE - (1) ICCM membrane w/conventional low temperature AGR

(2) ICCM membrane w/Warm Syngas Clean Up

(3) For co-production cases, assumed 'Case 1 Power Production CO₂ Capture' cost

With an estimated CO₂ recovery credit of \$26.0 per ton, the break-even H₂ production cost reduces to \$1.60, \$1.68 and \$3.47 for Case 2, Case 3 and Case 4 respectively.

Case 4 (ICCM with Warm Gas Clean Up) bears a higher cost penalty because of flue gas CO₂ capture. A separate run was carried out for Case 4 assuming that CO₂ capture is not required. Its H₂ production cost drops drastically, from \$5.04 to \$1.95 per thousand standard cubic feet. In comparison, corresponding production cost decrease for both Case 2 and Case 3 by only 3%, mainly due to the elimination of CO₂ compression. With Case 4, eliminating flue gas CO₂ capture and CO₂ compression saves about \$120MM in capital cost (both ISBL and BOP) and about 30 MW in power consumption. In addition, it generates 70 MWe more power from eliminating the CO₂ recovery regeneration steam demands. Under this scenario, ICCM application seems economically viable.

Table 3-38
IGCC Co-Production Case Product Break-Even Cost Compared

	Reference Case	Case 1	Case 2	Case 3	Case 4
	IGCC	IGCC w/CO ₂ Capture	<-- IGCC H ₂ Co-Production with 85% CO ₂ Capture --> [with PSA]	[ICCM ⁽¹⁾]	[ICCM ⁽²⁾]
Coal Feed, STPD (AR)	2,782	2,937	7,034	7,034	3,517
Net Power Export, MWe	281.4	222.6	123.3	118.9	152.3
Hydrogen Export, MMSCFD	N/A	N/A	212.4	206.5	46.7
CO ₂ Captured, %	0	85	85	85	85
Installed Cost, \$MM	445.0	536.0	986.0	989.0	660.0
Initial Operating Capital, \$MM	16.0	18.0	35.0	35.0	22.0
Total Initial Investment, \$MM	461.0	554.0	1021.0	1024.0	682.0
Annual Expenses, \$MM/Yr:					
Coal Feed Cost (29.4/ST AR)	23.9	25.2	60.4	60.4	30.2
Raw Water Cost	0.3	0.2	0.4	0.5	0.8
Power Import Cost	0.0	0.0	0.0	0.0	0.0
Catalysts & Chemical Cost	6.3	7.6	14.0	14.0	9.2
Ash & Slag Disposal Cost	0.9	1.0	2.4	2.4	1.2
Maintenance Labor & Material	13.3	16.1	29.6	29.7	19.8
Administration Salaries & Overheads	11.8	11.8	11.8	11.8	11.8
Insurance & Local Taxes	8.8	10.8	19.8	19.8	13.2
Amortized (Over 25 Yr) Investment Repayment	50.8	61.1	112.4	112.7	75.1
Federal & State Income Taxes	0.0	0.0	0.0	0.0	0.0
Total Expenses	116.1	133.8	250.7	251.3	161.3
Annual Revenues, \$MM/Yr:					
Power Sale	115.6	133.2	73.7	71.1	91.1
Hydrogen Sale	0	0	175.6	178.8	68.8
Sulfur Byproduct Sale	0.5	0.6	1.4	1.4	1.4
Credit for Recovered CO ₂	0.0	0.0	0.0	0.0	0.0
Total Revenue	116.1	133.8	250.7	251.3	161.3
Break-Even Cost of Power, \$/MWe-hr	58.7	85.4	85.4 ⁽³⁾	85.4 ⁽³⁾	85.4 ⁽³⁾
Break-Even Cost of Hydrogen, \$/1000 SCF	N/A	N/A	2.83	2.96	5.05

NOTE - (1) ICCM membrane w/conventional low temperature AGR
(2) ICCM membrane w/Warm Syngas Clean Up
(3) For co-production cases, assumed 'Case 1 Power Production CO₂ Capture' cost

Table 3-39

	Assumed H ₂ -to-kW Efficiency	
<u>Overall IGCC Plant Efficiency</u>	<u>@100%</u>	<u>@50%</u>
Reference Case	38.0%	38.0%
Case 1	28.5%	28.5%
Case 2	51.5%	29.0%
Case 3	50.0%	28.2%
Case 4	36.0%	26.1%

Coal Gasification Based H₂ Production - ICCM versus PSA for H₂ Separation

Tables 3-40 to 3-42 show the estimated performance and cost comparison of the two coal-gasification-based H₂ production plants, comparing ICCM directly against PSA. Both schemes were designed to maximize H₂ production to the maximum extent possible, and generate power by burning the purge gas from the ICCM or the PSA H₂ separation unit. The PSA purge gas is of low pressure; hence a conventional boiler-based power train is used to minimize compression cost. With the ICCM, the purge gas is at a pressure, 285 psia, which can be sent directly to the gas turbine.

The ICCM membrane ‘design basis’ operating temperature of 1500 °F used for the study imposes severe restrictions of effectively integrating the ICCM module into a coal gasification design for H₂ production. The envisioned thermal efficiency improvement with coupling a hydrogen separation membrane capable of operating at a higher temperature immediately downstream of the gasifier is not realized. Some of the issues include:

- The ICCM module needs a corresponding high-temperature syngas clean up process ahead of it for sulfur removal. The E-GasTM syngas contains close to 0.7 % of H₂S and the ICCM is known to have sulfur stability problem. Without a complementary high temperature syngas clean up process, the ICCM has to be integrated at a lower temperature.
- The ICCM membrane needs internal WGS activity for effective H₂ separation. The syngas generated from the E-GasTM gasifier contains only about 26% H₂, with a partial pressure only about 120 psi, which would not support an ICCM module based IGCC design with sufficient flux for large scale H₂ production from coal.
- With the ICCM study designs, it was necessary to shift the syngas to increase its H₂ content. A split-feed, 2-stage sour WGS unit was used to convert over 90+% of the CO in the syngas into H₂, increasing its content to about 53%, with a partial pressure of approximately 180 psi. The shifted syngas needs to be heated back up to the ICCM operating temperature of 1500 °F with a POX and additional heat exchanger equipment in attempt to minimize overall efficiency loss. If the ICCM membrane is capable of internal shift, capital savings from eliminating the external shift process will certainly improve its economics.
- A concomitant process is needed, within or in conjunction of the ICCM module, to remove the CO₂ from the shifted syngas in order to (1) further increase the H₂ partial pressure for effective ICCM separation and (2) capture the produced CO₂ as envisioned it would be required in a future IGCC plant. With the ICCM based designs, a low-temperature Selexol acid gas removal process had to be used for this purpose, with the anticipated suffering in the overall plant efficiency.

Without any complementary technology or enhanced ICCM membrane performance characteristics cited above, it is not likely that any envisioned IGCC thermal efficiency improvement can be achieved with the inclusion of an ICCM membrane. From that perspective, the comparative results of Case 5 and Case 6 are quite encouraging. Despite these limitations, Case 6 shows that an ICCM-based, coal H₂ production plant, using conventional WGS and AGR can be comparative with that of PSA; overall efficiency is only 0.8% lower and capital costs are comparative. The estimated break-even H₂ production cost using ICCM is about 2% higher, attributable mostly to the lower overall plant efficiency.

Table 3-40

Coal Gasification Based Hydrogen Production Case - Performance & Efficiency Compared

	Case 5	Case 6
	PSA	ICCM
	<u>w/Boiler</u>	<u>w/GE 6B Turbine</u>
Coal Feed, STPD (AR)	3,517	3,517
Coal Feed HHV, MW	935.8	935.8
Net Hydrogen Production, MMSCFD	133.9	134.0
Power Import, MWe	17.4	25.7
CO2 Captured, %	85	85
Fuel Gas Pressure to Power Train, Psig	65	285
Gross Power Output, MWe		
Gas Turbine	N/A	51.7
Steam Turbine (minus boiler fan consumption)	74.5	33
Total Generation:	74.5	84.7
Power Consumed, MWe		
Coal Prep & Gasification	3.8	3.8
ASU & O2 Compression	38.1	40
Sour Shift & LT Syngas Cooling	0.7	0.7
AGR (Acid Gas Removal)	8.5	8.7
H2 Recovery & Compression	8.7	28.3
Sulfur Recovery	0.3	0.3
CO2 Compression	19.2	19.2
Fuel Gas Compression	2.6	0
HRSG/Boiler/BFW Treating	2.1	2
BOP Consumption	7.9	7.4
Total Consumption:	91.9	110.4
Overall Efficiency, % HHV	54.7	53.9

Table 3-41

Coal Gasification Based Hydrogen Production Case - Capital Cost Compared

	Case 5 PSA <u>w/Boiler</u>	Case 6 ICCM <u>w/GE 6B Turbine</u>
Coal Feed, STPD (AR)	3,517	3,517
Net Hydrogen Production, MMSCFD	133.9	134.0
Power Import, MWe	17.4	25.7
CO2 Captured, %	85	85
Installed Cost, \$1,000,000		
Coal Prep & Gasification	120.2	120.2
ASU & O2 Compression	52.0	53.6
Sour Shift & LT Syngas Cooling	19.6	20.6
AGR (Acid Gas Removal)	55.4	55.4
H2 Recovery & Compression	34.4	24.8
Sulfur Recovery	6.3	6.3
CO2 Compression	32.0	32.0
Fuel Gas Compression	3.2	0.1
Gas Turbine/Generator	0	20.2
Steam Turbine/Generator	14.5	11.6
HRSG/Boiler/BFW Treating	14.4	6.8
BOP Consumption	72.3	72
Total Field Cost	424.3	423.6
Home Office Cost	22.0	23.7
Contingency	63.6	63.6
Total Installed Cost	509.9	510.9

Table 3-42

Coal Gasification Based Hydrogen Production Case - Breakeven Cost Compared

	Case 5 PSA <u>w/Boiler</u>	Case 6 ICCM <u>w/GE 6B Turbine</u>
Coal Feed, STPD (AR)	3,517	3,517
Net Hydrogen Production, MMSCFD	133.9	134.0
Power Import, MWe	17.4	25.7
CO2 Captured, %	85	85
Installed Cost, \$MM	510	511
Initial Operating Capital, \$MM	21	22
Total Initial Investment, \$MM	531	533
Annual Expenses, \$MM/Yr:		
Coal Feed Cost (29.4/ST AR)	30.2	30.2
Raw Water Cost	0.2	0.2
Power Import Cost	4.9	7.2
Catalysts & Chemical Cost	7.2	7.2
Ash & Slag Disposal Cost	1.2	1.2
Maintenance Labor & Material	15.3	15.3
Administration Salaries & Overheads	11.8	11.8
Insurance & Local Taxes	10.2	10.2
Amortized (Over 25 Yr) Investment Repayment	58.5	58.7
Total Expenses	139.5	142.0
Annual Revenues, \$MM/Yr:		
Hydrogen Sale	138.8	141.4
Sulfur Byproduct Sale	0.7	0.7
Credit for Recovered CO2	0	0
Total Revenue	139.5	142.1
Break-Even Cost of Hydrogen, \$/1000 SCF	3.55	3.61

SECTION 4. CONCLUSIONS

The overall objective of this program was to determine the technical and economic feasibility of using ICCM technology for H₂ production in “Vision 21” power plants. The program had three areas of emphasis: materials development, manufacturing process development and process engineering and economics. Most of the major conclusions from the work were given in the Results and Discussions sections for the individual Tasks and will only be reviewed here with some implications for technical feasibility.

In the area of materials development, novel pyrochlore-based proton conductors were identified, synthesized and characterized. A conductivity of 0.03 S/cm at 900 °C was obtained as the best result. The initial focus on pyrochlore materials for use in ICCMs was validated in this program as the materials appear to be stable at high temperatures under CO₂ and H₂O atmospheres. The refractory nature of these materials, while useful for long-term durability at high temperatures, is also problematic for membrane fabrication. High temperatures, or low-temperature processes, such as the Pechini process, will be required to fabricate structures from these materials.

The value for conductivity of 0.03 S/cm at 900 °C is lower than a value of 1 S/cm (p 74) which was used in the process flow schemes developed in Task 4. A goal for flux or conductivity is only meaningful within a techno-economic analysis. The results of the comparisons for Cases 3, 5 and 6 (PSA or ICCM H₂ separation) indicate the ICCM will be competitive if a conductivity close to 1 S/cm can be achieved. This of course is made under a number of assumptions about membrane cost, durability, *etc.* and more detailed calculations will be required as the physical properties of the membrane materials are further investigated.

In the area of membrane fabrication by plasma spray processing, the initial results were promising. Powders acceptable for spray processing we prepared both in-house at ITN and purchased commercially, indicating that the pyrochlore materials could be processed in a spray torch. Although some leak-tight membranes were obtained, the issue of loss of porosity in the metal substrate may have covered over some cracking occurring in the membrane. Here, more modeling and experimental work needs to be performed to reduce cracking associated with the mismatch in thermal expansion coefficients arising from the metallic and ceramic components of the cermet membrane. Another issue is the membrane thickness. The minimum membrane thickness obtained during the program was ~200 μm. Ideally, the membrane will need to be in the 50 μm range based on the economic calculations.

An extensive techno-economic analysis was done to compare the utility of ITN ICCM with PSA in a prototype “Vision 21” plant. Six different plant scenarios, as well as a reference plant, were developed and compared. The effect of CO₂ capture, an important aspect of future power plants, was also incorporated. Given the current estimated achievable performance and operating conditions of the membrane, and under the various IGCC H₂ co-production configurations investigated, the ITN ICCM membrane is only slightly less efficient/economical than the conventional PSA technology. These results are encouraging; with suitable improvements in membrane properties, tailored for more efficient IGCC integration, ICCM may well be a viable technology.

The membrane design characteristics under the current state of development impose limitations with integrating the ICCM module into an IGCC facility to take full advantage of its intended

overall thermal efficiency improvement. Future R&D efforts need to be focused on the development of complementary technologies and/or improved membrane properties to reduce or eliminate these efficiency-limiting impediments. Examples of some of these needed technologies/improvements include:

- Complementary high-temperature syngas clean-up technology, or membrane, that can be both stable and functional under a raw coal syngas environment;
- A high-temperature selective, acid gas removal process for simultaneous H_2S and CO_2 removal;
- Membrane with concomitant in-situ WGS and H_2 separation activity, allowing the H_2 to be continuously extracted from the membrane as it is formed; and/or
- Complementary high-temperature/high-pressure CO_2 separation technology, to work in conjunction with H_2 ICCM.

While the ICCM membrane is only a small portion of the total IGCC plant cost, its costs can have significant effect on the overall co-production economics. The membrane cost is directly related to its required surface area, which for a desired level of recovery is a function of the achievable flux that can be obtained. Figure 4-1 shows the H_2 production cost for the Case 3 IGCC co-production configuration, as a function of the ICCM membrane permeate pressure, at different flux rate. The effect of flux rate on H_2 production economics is quite significant. Case 3 design was based on a projected ITN ICCM hydrogen flux of 50 ml/min-cm^2 . If the membrane flux can be increased to 200 ml/min-cm^2 , as some developers seem to suggest that it can be done, ICCM economics can be significantly improved. Clearly for the ICCM approach to be economically competitive with PSA, a lower temperature of operation is crucial. The best route for obtaining this is further materials discovery for new compositions with high conductivity.

Figure 4-1. Plot of H_2 break-even price as a function of the H_2 permeate pressure.

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SECTION 6. APPENDICES

6.1 Appendix A (Task 4.1) - Theoretical Limit for Hydrogen Recovery from Membrane Separation (no WGS Activity)

Nomenclature:

- F_{Ri} - total molar flowrate of retentate at inlet
- F_{Ro} - total molar flowrate of retentate at outlet
- F_{Po} - total molar flowrate of permeate at outlet
- P_R - pressure of retentate stream
- P_P - pressure of permeate stream
- $y(\text{H}_2)_{Ri}$ - mole fraction H_2 in retentate at inlet (feed gas)
- $y(\text{H}_2)_{Ro}$ - mole fraction H_2 in retentate at outlet
- $\Delta P(\text{H}_2)_o$ - hydrogen partial pressure pinch at outlet

Assumptions:

- 1) No inlet flow to permeate side of membrane unit
- 2) Pressure drop is negligible in direction of flow for both retentate and permeate
- 3) Membrane is impermeable to all species other than hydrogen
- 4) Permeate is pure hydrogen
- 5) No reactions occur within membrane unit
- 6) Isothermal conditions maintained within membrane unit
- 7) Ideal gas behavior, fugacities are equal to partial pressures

The last assumption is justified at higher temperature range of membrane operation ($\sim 900^\circ\text{C}$) and reasonably accurate at lower temperatures ($200\text{--}250^\circ\text{C}$)

Material Balances:

$$\text{Overall:} \quad F_{Ri} = F_{Ro} + F_{Po} \quad (\text{A1})$$

$$\text{Hydrogen:} \quad y(\text{H}_2)_{Ri} F_{Ri} = y(\text{H}_2)_{Ro} F_{Ro} + F_{Po} \quad (\text{A2})$$

Substitution of A1 into A2 and rearrangement yields:

$$(1 - y(\text{H}_2)_{Ri}) F_{Ri} = (1 - y(\text{H}_2)_{Ro}) F_{Ro} \quad (\text{A3})$$

$$\frac{F_{Ro}}{F_{Ri}} = \frac{(1 - y(\text{H}_2)_{Ri})}{(1 - y(\text{H}_2)_{Ro})} \quad (\text{A4})$$

Pinch Specification:

$$y(\text{H}_2)_{Ro} P_R - P_P = \Delta P(\text{H}_2)_o \quad (\text{A5})$$

$$y(\text{H}_2)_{Ro} = \frac{P_P + \Delta P(\text{H}_2)_o}{P_R} \quad (\text{A6})$$

Hydrogen Recovery:

$$\text{H}_2 \text{ Recovery} \equiv \frac{F_{Po}}{y(\text{H}_2)_{Ri} F_{Ri}} \quad (\text{A7})$$

Substitution of A2 into A7:

$$\text{H}_2 \text{ Recovery} = \frac{y(\text{H}_2)_{Ri} F_{Ri} - y(\text{H}_2)_{Ro} F_{Ro}}{y(\text{H}_2)_{Ri} F_{Ri}} = 1 - \frac{y(\text{H}_2)_{Ro} F_{Ro}}{y(\text{H}_2)_{Ri} F_{Ri}} \quad (\text{A8})$$

Substitution of A4 into A8 and rearrangement yields:

$$\text{H}_2 \text{ Recovery} = 1 - \frac{y(\text{H}_2)_{Ro} (1 - y(\text{H}_2)_{Ri})}{y(\text{H}_2)_{Ri} (1 - y(\text{H}_2)_{Ro})} \quad (\text{A9})$$

$$\text{H}_2 \text{ Recovery} = 1 - \frac{(1 - y(\text{H}_2)_{Ri})}{y(\text{H}_2)_{Ri}} \frac{y(\text{H}_2)_{Ro}}{(1 - y(\text{H}_2)_{Ro})} \quad (\text{A10})$$

Substitution of A6 into A10:

$$\text{H}_2 \text{ Recovery} = 1 - \frac{(1 - y(\text{H}_2)_{Ri})}{y(\text{H}_2)_{Ri}} \frac{(P_P + \Delta P(\text{H}_2)_o)}{(P_R - (P_P + \Delta P(\text{H}_2)_o))} \quad (\text{A11})$$

Thus, given the hydrogen mole fraction at the inlet, the membrane operating pressures, and a specified hydrogen partial pressure pinch, the hydrogen recovery can be calculated. However, the membrane area required for this level of recovery is still unknown and can only be determined by solving the ordinary differential equation governing this system.

In the limit as the hydrogen partial pressure pinch approaches zero (*i.e.* the area approaches infinity), the hydrogen recovery approaches:

$$\text{H}_2 \text{ Recovery} = 1 - \frac{(1 - y(\text{H}_2)_{Ri})}{y(\text{H}_2)_{Ri}} \frac{P_P}{(P_R - P_P)} \quad (\text{A12})$$

This is the maximum hydrogen recovery achievable by membrane separation without reaction.

6.2 Appendix B (Task 4.1) - Theoretical Limit for CO Conversion to Hydrogen (no Membrane Separation)

Nomenclature:

F	- total molar flowrate
n	- component molar flowrate
T	- temperature
ΔT	- approach temperature
y	- mole fraction
ϵ	- molar reaction coordinate or extent of reaction

Subscripts:

eq	- at equilibrium
i	- at inlet
o	- at outlet
R	- reactor
T	- total

Assumptions:

- 1) No separation occurs in the unit, which is a simple reactor
- 2) WGS reaction approaches equilibrium within reactor
- 3) Isothermal conditions maintained within reactor
- 4) Ideal gas behavior, fugacities are equal to partial pressures

The last assumption is justified at higher temperature range of membrane operation (~900 °C) and reasonably accurate at lower temperatures (200-250 °C)

Water-Gas-Shift Reaction:

Stoichiometric Relationships:

$$y(\text{CO})_{Ro} = (n(\text{CO})_{Ri} - \epsilon) / n_{T,Ro} \quad (\text{B1})$$

$$y(\text{H}_2\text{O})_{Ro} = (n(\text{H}_2\text{O})_{Ri} - \epsilon) / n_{T,Ro} \quad (\text{B2})$$

$$y(\text{CO}_2)_{Ro} = (n(\text{CO}_2)_{Ri} + \epsilon) / n_{T,Ro} \quad (\text{B3})$$

$$y(\text{H}_2)_{Ro} = (n(\text{H}_2)_{Ri} + \epsilon) / n_{T,Ro} \quad (\text{B4})$$

Total molar flowrate at inlet:

$$n_{T,Ri} = n(\text{CO})_{Ri} + n(\text{H}_2\text{O})_{Ri} + n(\text{CO}_2)_{Ri} + n(\text{H}_2)_{Ri} \quad (\text{B5})$$

Total molar flowrate at outlet:

$$n_{T,Ro} = n(\text{CO})_{Ro} + n(\text{H}_2\text{O})_{Ro} + n(\text{CO}_2)_{Ro} + n(\text{H}_2)_{Ro} \quad (\text{B6})$$

And, based on B1-B4:

$$\begin{aligned} n_{T,Ro} &= (n(\text{CO})_{Ri} - \varepsilon) + (n(\text{H}_2\text{O})_{Ri} - \varepsilon) + (n(\text{CO}_2)_{Ri} + \varepsilon) + (n(\text{H}_2)_{Ri} + \varepsilon) \\ &= n_{T,Ri} \end{aligned} \quad (\text{B7})$$

Also, $y_{\varepsilon,Ri}$ is defined by:

$$y_{\varepsilon,Ri} \equiv \varepsilon / n_{T,Ri} \quad (\text{B8})$$

Reaction Equilibrium:

Concentration dependence of the equilibrium constant:

$$K_{eq} = \frac{y(\text{CO}_2)y(\text{H}_2)}{y(\text{CO})y(\text{H}_2\text{O})} \quad (\text{B9})$$

Substitution of B1-B4 into B9 and rearrangement yields:

$$K_{eq} = \frac{(n(\text{CO}_2)_{Ri} + \varepsilon)(n(\text{H}_2)_{Ri} + \varepsilon)}{(n(\text{CO})_{Ri} - \varepsilon)(n(\text{H}_2\text{O})_{Ri} - \varepsilon)} \quad (\text{B10})$$

Upon dividing both the numerator and denominator by $n_{T,Ri}^2$:

$$K_{eq} = \frac{(y(\text{CO}_2)_{Ri} + y_{\varepsilon,Ri})(y(\text{H}_2)_{Ri} + y_{\varepsilon,Ri})}{(y(\text{CO})_{Ri} - y_{\varepsilon,Ri})(y(\text{H}_2\text{O})_{Ri} - y_{\varepsilon,Ri})} \quad (\text{B11})$$

Rearranging:

$$\begin{aligned} (1 - K_{eq})y_{\varepsilon,Ri}^2 + (y(\text{CO}_2)_{Ri} + y(\text{H}_2)_{Ri} + K_{eq}(y(\text{CO})_{Ri} + y(\text{H}_2\text{O})_{Ri}))y_{\varepsilon,Ri} \\ + (y(\text{CO}_2)_{Ri}y(\text{H}_2)_{Ri} - K_{eq}y(\text{CO})_{Ri}y(\text{H}_2\text{O})_{Ri}) = 0 \end{aligned} \quad (\text{B12})$$

Several expressions for the temperature dependence of K_{eq} can be found in the literature; one relatively simple expression is^{B1}:

$$K_{eq}(T_{eq}) = \exp(4577.8/T_{eq} - 4.33) \quad (\text{B13})$$

where the temperature is in Kelvin.

Given the equilibrium temperature, B13 can be used to predict K_{eq} . Then, K_{eq} and the inlet concentrations can be used in B12 to solve for $y_{\varepsilon,Ri}$. Only one of the two roots of the quadratic equation is realistic, the other yields mole fractions which are either less than zero or greater than one. Realistic limits on $y_{\varepsilon,Ri}$ are:

$$-\min(y(\text{CO}_2)_{Ri}, y(\text{H}_2)_{Ri}) < y_{\varepsilon, Ri} < \min(y(\text{CO})_{Ri}, y(\text{H}_2\text{O})_{Ri}) \quad (\text{B14})$$

For reactors of finite length, equilibrium will not be reached, and T_{eq} will not be equal to the true temperature T . The temperature T_{eq} in B13 can be based on an effective temperature given by:

$$T_{eq} = T + \Delta T \quad (\text{B15})$$

where ΔT is the “approach” to equilibrium. ΔT can either be determined empirically or from rigorous modeling of the reactor.

Material Balances:

$$\text{Overall:} \quad F_{Ri} = F_{Ro} \quad (\text{B16})$$

Which is equivalent to B17.

$$\text{Hydrogen:} \quad (y(\text{H}_2)_{Ri} + y_{\varepsilon, Ri})F_{Ri} = y(\text{H}_2)_{Ro}F_{Ro} \quad (\text{B17})$$

Note that the flowrates F are equivalent to the n_T used above.

CO Conversion to Hydrogen:

$$\text{CO Conversion} \equiv \frac{y(\text{H}_2)_{Ro}F_{Ro} - y(\text{H}_2)_{Ri}F_{Ri}}{y(\text{CO})_{Ri}F_{Ri}} \quad (\text{B18})$$

Substitution of B17 into B18:

$$\text{CO Conversion} = \frac{y_{\varepsilon, Ri}}{y(\text{CO})_{Ri}} \quad (\text{B19})$$

Which is equivalent to the net yield of hydrogen. Thus, given the mole fractions of hydrogen, water, carbon dioxide, and carbon monoxide at the inlet of the reactor, the fractional extent of reaction (from the solution of B12 using $K_{eq}(T+\Delta T)$), the CO conversion can be calculated. However, the reactor volume required for this conversion is still unknown and can only be determined by solving the ordinary differential equation governing this system.

In the limit as the temperature approach to equilibrium approach zero (*i.e.* the reactor volume approaches infinity), $y_{\varepsilon, Ri}$ is determined from B12 using $K_{eq}(T)$. The CO conversion calculated from B19 with this value of $y_{\varepsilon, Ri}$ is the maximum conversion achievable in a simple reactor.

6.2.1 References

B1. J.M. Moe, Chem. Eng. Prog., 58, 313 (1977).

6.3 Appendix C (Task 4.1) - Theoretical Limit for Hydrogen Productivity from Membrane Separation with Reaction

Nomenclature:

F	- total molar flowrate
n	- component molar flowrate
P	- pressure
ΔP	- pressure pinch
T	- temperature
ΔT	- approach temperature
y	- mole fraction
ϵ	- molar reaction coordinate or extent of reaction

Subscripts:

eq	- at equilibrium
i	- at inlet
o	- at outlet
P	- permeate
R	- retentate
T	- total

Assumptions:

- 1) No inlet flow to permeate side of membrane unit
- 2) Pressure drop is negligible in direction of flow for both retentate and permeate
- 3) Membrane is impermeable to all species other than hydrogen
- 4) Permeate is pure hydrogen
- 5) WGS reaction approaches equilibrium within membrane unit**
- 6) Isothermal conditions maintained within membrane unit
- 7) Ideal gas behavior, fugacities are equal to partial pressures

The last assumption is justified at higher temperature range of membrane operation (~900 °C) and reasonably accurate at lower temperatures (200-250 °C)

Water-Gas-Shift Reaction:

Stoichiometric Relationships:

$$y(\text{CO})_{Ro} = (n(\text{CO})_{Ri} - \varepsilon) / n_{T,Ro} \quad (\text{C1})$$

$$y(\text{H}_2\text{O})_{Ro} = (n(\text{H}_2\text{O})_{Ri} - \varepsilon) / n_{T,Ro} \quad (\text{C2})$$

$$y(\text{CO}_2)_{Ro} = (n(\text{CO}_2)_{Ri} + \varepsilon) / n_{T,Ro} \quad (\text{C3})$$

$$y(\text{H}_2)_{Ro} = \frac{n(\text{H}_2)_{Ro}}{n_{T,Ro}} = \frac{P_p + \Delta P(\text{H}_2)_o}{P_R} \quad (\text{C4})$$

Equations C1-C3 are the same as B1-B3; B4 has been replaced with A6, the pinch specification for H₂.

Total molar flowrate at inlet:

$$n_{T,Ri} = n(\text{CO})_{Ri} + n(\text{H}_2\text{O})_{Ri} + n(\text{CO}_2)_{Ri} + n(\text{H}_2)_{Ri} \quad (\text{C5})$$

Total molar flowrate at outlet:

$$n_{T,Ro} = n(\text{CO})_{Ro} + n(\text{H}_2\text{O})_{Ro} + n(\text{CO}_2)_{Ro} + n(\text{H}_2)_{Ro} \quad (\text{C6})$$

And, based on C1-C4:

$$\begin{aligned} n_{T,Ro} &= (n(\text{CO})_{Ri} - \varepsilon) + (n(\text{H}_2\text{O})_{Ri} - \varepsilon) + (n(\text{CO}_2)_{Ri} + \varepsilon) + y(\text{H}_2)_{Ro} n_{T,Ro} \\ &= n_{T,Ri} - (n(\text{H}_2)_{Ri} + \varepsilon) + y(\text{H}_2)_{Ro} n_{T,Ro} \end{aligned} \quad (\text{C7})$$

Rearranging:

$$\frac{n_{T,Ro}}{n_{T,Ri}} = \frac{(1 - y(\text{H}_2)_{Ri} - y_{\varepsilon,Ri})}{(1 - y(\text{H}_2)_{Ro})} \quad (\text{C8})$$

Where $y_{\varepsilon,Ri}$ is defined by:

$$y_{\varepsilon,Ri} \equiv \varepsilon / n_{T,Ri} \quad (\text{C9})$$

Reaction Equilibrium:

As before, the concentration dependence of the equilibrium constant:

$$K_{eq} = \frac{y(\text{CO}_2)y(\text{H}_2)}{y(\text{CO})y(\text{H}_2\text{O})} \quad (\text{C10})$$

Substitution of C1-C3 into C10 and rearrangement yields:

$$K_{eq} = \frac{(n(\text{CO}_2)_{Ri} + \varepsilon)(y(\text{H}_2)_{Ro} n_{T,Ro})}{(n(\text{CO})_{Ri} - \varepsilon)(n(\text{H}_2\text{O})_{Ri} - \varepsilon)} \quad (\text{C11})$$

Upon dividing both the numerator and denominator by $n_{T,Ri}^2$ and substituting C8:

$$K_{eq} = \frac{y(H_2)_{Ro}}{(1 - y(H_2)_{Ro})} \frac{(y(CO_2)_{Ri} + y_{\epsilon,Ri})(1 - y(H_2)_{Ri} - y_{\epsilon,Ri})}{(y(CO)_{Ri} - y_{\epsilon,Ri})(y(H_2O)_{Ri} - y_{\epsilon,Ri})} \quad (C12)$$

Rearranging:

$$\begin{aligned} (1 + K'_{eq})y_{\epsilon,Ri}^2 + (y(CO_2)_{Ri} + (y(H_2)_{Ri} - 1) - K'_{eq}(y(CO)_{Ri} + y(H_2O)_{Ri}))y_{\epsilon,Ri} \\ + (y(CO_2)_{Ri}(y(H_2)_{Ri} - 1) + K'_{eq}y(CO)_{Ri}y(H_2O)_{Ri}) = 0 \end{aligned} \quad (C13)$$

Where K'_{eq} is defined by:

$$K'_{eq} \equiv K_{eq} \frac{(1 - y(H_2)_{Ro})}{y(H_2)_{Ro}} = K_{eq} \frac{(P_R - (P_P + \Delta P(H_2)_{Ro}))}{(P_P + \Delta P(H_2)_{Ro})} \quad (C14)$$

The expression given previously B13, can be used for calculating K_{eq} based on the equilibrium temperature T_{eq} calculated from B15. Then, K'_{eq} can be calculated from C14 and the operating pressures P_R and P_P and approach $\Delta P(H_2)_o$. Finally, K'_{eq} and the inlet concentrations can be used in C13 to solve for $y_{\epsilon,Ri}$. As before, only one of the two roots of the quadratic equation is realistic.

Material Balances:

$$\text{Overall:} \quad F_{Ri} = F_{Ro} + F_{Po} \quad (C15)$$

$$\text{Hydrogen:} \quad (y(H_2)_{Ri} + y_{\epsilon,Ri})F_{Ri} = y(H_2)_{Ro}F_{Ro} + F_{Po} \quad (C16)$$

Note that the flowrates F are equivalent to the n_T used above.

Substitution of C15 into C16 and rearrangement yields:

$$(1 - y(H_2)_{Ri} - y_{\epsilon,Ri})F_{Ri} = (1 - y(H_2)_{Ro})F_{Ro} \quad (C17)$$

$$\frac{F_{Ro}}{F_{Ri}} = \frac{(1 - y(H_2)_{Ri} - y_{\epsilon,Ri})}{(1 - y(H_2)_{Ro})} \quad (C18)$$

Which is equivalent to C8.

Pinch Specification:

This relationship is unchanged from Appendix A:

$$y(H_2)_{Ro} = \frac{P_P + \Delta P(H_2)_o}{P_R} \quad (C19)$$

Hydrogen Productivity:

$$\text{H}_2 \text{ Productivity} \equiv \frac{F_{Po}}{(y(H_2)_{Ri} + y(CO)_{Ri})F_{Ri}} \quad (C20)$$

Substitution of C16 into C20:

$$\begin{aligned} \text{H}_2 \text{ Productivity} &= \frac{(y(\text{H}_2)_{Ri} + y_{\varepsilon,Ri})F_{Ri} - y(\text{H}_2)_{Ro}F_{Ro}}{(y(\text{H}_2)_{Ri} + y(\text{CO})_{Ri})F_{Ri}} \\ &= \frac{(y(\text{H}_2)_{Ri} + y_{\varepsilon,Ri})}{(y(\text{H}_2)_{Ri} + y(\text{CO})_{Ri})} - \frac{y(\text{H}_2)_{Ro}F_{Ro}}{(y(\text{H}_2)_{Ri} + y(\text{CO})_{Ri})F_{Ri}} \end{aligned} \quad (\text{C21})$$

Substitution of C18 into C21 and rearrangement yields:

$$\text{H}_2 \text{ Productivity} = \frac{(y(\text{H}_2)_{Ri} + y_{\varepsilon,Ri})}{(y(\text{H}_2)_{Ri} + y(\text{CO})_{Ri})} - \frac{(1 - y(\text{H}_2)_{Ri} - y_{\varepsilon,Ri})y(\text{H}_2)_{Ro}}{(y(\text{H}_2)_{Ri} + y(\text{CO})_{Ri})(1 - y(\text{H}_2)_{Ro})} \quad (\text{C22})$$

Substitution of C19 into C22:

$$\text{H}_2 \text{ Productivity} = \frac{(y(\text{H}_2)_{Ri} + y_{\varepsilon,Ri})}{(y(\text{H}_2)_{Ri} + y(\text{CO})_{Ri})} - \frac{(1 - y(\text{H}_2)_{Ri} - y_{\varepsilon,Ri})(P_P + \Delta P(\text{H}_2)_o)}{(y(\text{H}_2)_{Ri} + y(\text{CO})_{Ri})(P_R - (P_P + \Delta P(\text{H}_2)_o))} \quad (\text{C23})$$

Thus, given the mole fractions of hydrogen, water, carbon dioxide, and carbon monoxide at the inlet of the membrane reactor, the fractional extent of reaction (from the solution of C13 using $K'_{eq}(T+\Delta T)$), the membrane operating pressures, and a specified hydrogen partial pressure pinch, the hydrogen productivity can be calculated. However, the membrane area required for this productivity is still unknown and can only be determined by solving the ordinary differential equations governing this system.

In the limit as the hydrogen partial pressure pinch and the temperature approach to equilibrium approach zero (*i.e.* the membrane area approaches infinity), the hydrogen productivity approaches:

$$\text{H}_2 \text{ Productivity} = \frac{(y(\text{H}_2)_{Ri} + y_{\varepsilon,Ri})}{(y(\text{H}_2)_{Ri} + y(\text{CO})_{Ri})} - \frac{(1 - y(\text{H}_2)_{Ri} - y_{\varepsilon,Ri})}{(y(\text{H}_2)_{Ri} + y(\text{CO})_{Ri})} \frac{P_P}{(P_R - P_P)} \quad (\text{C24})$$

where $y_{\varepsilon,Ri}$ is determined from C13 using $K'_{eq}(T)$. This is the maximum hydrogen productivity achievable in a membrane reactor.

6.4 Appendix D – Model Output Summaries (Tasks 4.2 and 4.3)

REFERENCE CASE IGCC

1S7X0G09 = No H2 Recov w 100% Fd Bypass, None Shift, 58# resid CO2 PP, 0 MPH HRSF Firing, w Stm as Nox Diluent. (\$0/ST recov CO2 credit & \$0/ST CO2 emission tax)

PROCESS ECONOMIC EVALUATION SUMMARY:

OVERALL (H2 + Mwe) EFFICIENCIES:			
HHV Basis	38.0 % HHV	Breakeven H2 Price, \$/1000 SCF	N/A
LHV Basis	40.2 % LHV	Brakeven H2 Price, \$/MMBtu(HHV)	N/A
(HHV w S)	38.8 % HHV		
(LHV w S)	41.1 % LHV		

PRODUCTS: (LHV w S) 41.1 % LHV
 TPC Payback Periods, Years 8.8
 % Return On Invest (Excl Land) 11.02

Stream Flows, MPH:

Stream	1.	2.	3.	4.	5.	6.	7.	8.	OVERALL (H2 + Mwe) EFFICIENCIES:
N2	271	271	244	244	244	0	0	0.0	Hydrogen, MMSCFD
O2	0	0	0	0	0	0	0	281	Power Export, MWe
CO2	3,220	3,220	2,898	2,720	2,720	0	0	61	Sulfur, STPD
Ar	87	87	78	78	78	0	0	319	Slag & Ash, STPD Dry
H2	6,244	6,244	5,620	5,620	5,620	0	0	0	Waste Water, GPM
CO	8,737	8,737	7,864	7,864	7,864	0	0	1,049	Recovered CO2, STPD
H2S	169	169	152	0	0	0	0	0	
COS	8	8	0	0	0	0	0	0	
CH4	729	729	656	656	656	0	0	2,782	As Rec'd Coal, STPD
C2H6	0	0	0	0	0	0	0	0	Fluxant, STPD Dry
Steam	4,825	5,347	38	0	0	0	8,647	0	Power Import, MW
Total	24,290	24,812	17,557	17,181	17,181	0	8,647	2,643	Raw Water, GPM
								2,084	95.0% O2, STPD

CONSUMABLES:

As Rec'd Coal, STPD	2,782	Royalties	0.0
Fluxant, STPD Dry	0	Mainten Labor & Mat	13.3
Power Import, MW	0	Admin & Labor Salary	8.8
Raw Water, GPM	2,643	Overheads & Benefits	3.1
95.0% O2, STPD	2,084	Insurances	4.4
		Local Taxes	4.4

PROCESS DATA:

Power Sale, \$58.7/MMH	115.7
Sulfur Sale	0.5
Ash & Slag Disposal	0.9
Waste Water Disposal	0.0
CO2 Sequestn Credit	0.0
CO2 Emission Penalty	0.0
Bank Loan Repayment	40.6
Investment Recovery	10.2
Hydrogen Sales	0.0
Pre Income Tax Total	116.3
Fed Income Tax (FIT)	0.0
Stat Income Tax (SIT)	0.0
Post Income Tax Total	116.3

Cap Cost Summary \$MM(2001)	# of Trains	Power Balance, MW	FG LHV, Btu/SCF:
Coal Handling/Feeding	28.7	1.3	270.6
Gasification	57.0	1.7	270.6
Spare Gasifiers (Total 1 Spare)	16.3	---	180.0
Air Separation	44.1	17.6	180.0
O2 Compression (565 PSIG)	---	9.9	---
Shift & LT Gas Cooling	6.4	0.0	---
Acid Gas Removal & SWS	9.5	0.3	---
Sulfur Recovery	5.3	0.3	---
H2 Recovery	0.0	0.1	---
H2 Compression (1000 PSIG)	0.0	0.0	---
CO2 Dehydr'n & Comp (2000 PSIG)	0.0	0.0	---
Fuel Gas Compression (345 PSIG)	0.4	0.0	---
N2 Compression (290 PSIG)	0.0	0.0	---
Gas Turb Generator System	71.1	209.3	---
HRSG/Boiler Plt/BFW/DW/Cond	19.5	3.9	---
Steam Turb Generator System	21.2	116.0	---
CW Pumps & CT Fans	10.7	4.2	---
Flue Gas CO2 Recovery	---	---	---
BOP	80.7	4.6	---
Total Installed Cost (TIC)	370.9	325.3	43.8
Home Office @ 4.9 % of TIC	18.3		
Contingency @ 15.0 % of TIC	55.6		
Total Plant Cost (TPC)	444.8		

Cap Cost Summary \$MM(2001)	# of Trains	Power Balance, MW	FG LHV, Btu/SCF:
Coal Handling/Feeding	28.7	1.3	270.6
Gasification	57.0	1.7	270.6
Spare Gasifiers (Total 1 Spare)	16.3	---	180.0
Air Separation	44.1	17.6	180.0
O2 Compression (565 PSIG)	---	9.9	---
Shift & LT Gas Cooling	6.4	0.0	---
Acid Gas Removal & SWS	9.5	0.3	---
Sulfur Recovery	5.3	0.3	---
H2 Recovery	0.0	0.1	---
H2 Compression (1000 PSIG)	0.0	0.0	---
CO2 Dehydr'n & Comp (2000 PSIG)	0.0	0.0	---
Fuel Gas Compression (345 PSIG)	0.4	0.0	---
N2 Compression (290 PSIG)	0.0	0.0	---
Gas Turb Generator System	71.1	209.3	---
HRSG/Boiler Plt/BFW/DW/Cond	19.5	3.9	---
Steam Turb Generator System	21.2	116.0	---
CW Pumps & CT Fans	10.7	4.2	---
Flue Gas CO2 Recovery	---	---	---
BOP	80.7	4.6	---
Total Installed Cost (TIC)	370.9	325.3	43.8
Home Office @ 4.9 % of TIC	18.3		
Contingency @ 15.0 % of TIC	55.6		
Total Plant Cost (TPC)	444.8		

Cap Cost Summary \$MM(2001)	# of Trains	Power Balance, MW	FG LHV, Btu/SCF:
Coal Handling/Feeding	28.7	1.3	270.6
Gasification	57.0	1.7	270.6
Spare Gasifiers (Total 1 Spare)	16.3	---	180.0
Air Separation	44.1	17.6	180.0
O2 Compression (565 PSIG)	---	9.9	---
Shift & LT Gas Cooling	6.4	0.0	---
Acid Gas Removal & SWS	9.5	0.3	---
Sulfur Recovery	5.3	0.3	---
H2 Recovery	0.0	0.1	---
H2 Compression (1000 PSIG)	0.0	0.0	---
CO2 Dehydr'n & Comp (2000 PSIG)	0.0	0.0	---
Fuel Gas Compression (345 PSIG)	0.4	0.0	---
N2 Compression (290 PSIG)	0.0	0.0	---
Gas Turb Generator System	71.1	209.3	---
HRSG/Boiler Plt/BFW/DW/Cond	19.5	3.9	---
Steam Turb Generator System	21.2	116.0	---
CW Pumps & CT Fans	10.7	4.2	---
Flue Gas CO2 Recovery	---	---	---
BOP	80.7	4.6	---
Total Installed Cost (TIC)	370.9	325.3	43.8
Home Office @ 4.9 % of TIC	18.3		
Contingency @ 15.0 % of TIC	55.6		
Total Plant Cost (TPC)	444.8		

1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat. H2 Prod HHV Btu/SCF = 324 GTG/6

Run ID: 1S7X0G09 = No H2 Recov w 100% Fd Bypass, None Shift, 58# resid CO2 PP, 0 MPH HRSG Firing, w Stm as NOx Diluent.

INCOMES & EXPENSES:		Expenses		Incomes		INVESTMENT & CAPITAL RECOVERY:		Economic Analysis Basis:	
		\$MM/Yr	\$MM/Yr						
Coal Feed		23.9				Total Plant Capital Cost, \$MM	444.8	As Rec'd Coal Cost, \$/ST	29.4
Fluxant Feed		0.0				Initial Operating Capital, \$MM	16.4	Fluxant Cost, \$/ST Dry	10
Power Import		0.0				Land Cost @ \$0,000/Acre, \$MM	0.0	Import Power Cost, \$/MWH	58.679
Raw Water Import		0.3						Raw Water Cost, \$/1000 Gal	0.25
Cat & Chem Consumptn		6.3				Total Initial Investment, \$MM	461.2		
Royalties		0				Owner Initial Investment, \$MM	92.2		
Mainten Labor & Mat		13.3				Bank Loan, \$MM	369.0	Hydrogen Worth, \$/1000 SCF	To be Calc
Admin & Labor Salary		8.8						Export Power Worth, \$/MWH	58.679
Overheads & Benefits		3.1				Construction Period, Years	0	Sulfur Worth, \$/ST	30
Insurances		4.4				Plant Life, Years	25	Slag & Ash Worth, \$/ST Dry	-10
Local Taxes		4.4						Waste H2O Worth, \$/1000 Gal	0
Power Sales			115.7			Bank Loan Interest Rate, % per Year	10	Recovered CO2 Worth, \$/ST	0
Sulfur Sales			0.5			Bank Loan Repayment Period, Years	10	CO2 Emission Penalty, \$/ST	0
Ash & Slag Disposal		0.9				Bank Loan Repayment, \$MM/Yr	25	Annual Maint Labor+Mat, % TPC	3
Waste Water Disposal		0.0				Annual Bank Loan Repayment, \$MM/Yr	40.6	Employees, 1xAdmin + 4xOper	125
CO2 Sequestn Credit			0.0			Equival Depreciatn & Interest Expenses:		Avg Salaries, \$/Year	70,000
CO2 Emission Penalty		0.0				Bank Loan S/L Depreciatn, \$MM/Yr	14.8	OVRDs & Benefits, % Labor	35
Bank Loan Repayment		40.6				Bank Loan Interest, \$MM/Year	25.9		
Investment Recovery		10.2						Construction Period, Years	0
						Owner's Investment ROI, % per Year	10	Land Cost, \$/Ft2	0
Hydrogen Sales			0.0			Owner's Invest Repayment Period, Yrs	25	Initial Oper Fund, Months	3
						Annual Capital Recovery, \$MM/Yr	10.2	Plant Life, Years	25
Total Pre Income Taxes		116.3	116.3			Equival Depreciatn & Interest Expenses:		Annual Onstream Factor, %	80
Fed Income Tax (FIT)		0.0				Equival S/L Depreciatn, \$MM/Yr	3.7	Owner's Investment, % TPC	20
Stat Income Tax (SIT)		0.0				Equival Loan Interest, \$MM/Yr	6.5	Owner Invest Repay Period, Yrs	25
								Owner's ROI, % per Year	10
Total Post Income Tax		116.3	116.3			Overall Ht Bal, MM(lHV)/Hr:		Bank Loan Interest Rate, %/Yr	10
						In	Out	Insurances, % TPC	1
								FIT, % Net Income	30
								SIT, % Net Income less FIT	10
								Local (Property) Taxes, % TPC	1
								Royalties:	
								Gasification, STPD MF Coal	0
								Sour Shift, \$/MMSCFD Feed	0
								AGR, \$/MMSCFD Acid Gas	0
								Sulfur Plt, \$/STPD Sulfur	0
								TGTU, \$/STPD Sulfur	0
								PSA, \$/MMSCFD H2	0
								Catalysts & Chemicals, % TIC:	1.7
								Memb Perf Fact, x 20ml/min/cm2:	2.5

PROCESS ECONOMIC EVALUATION SUMMARY:

Breakeven H2 Price, \$/1000 SCF	N/A
Brkeven H2 Price, \$/MMBtu(HHV)	N/A
TPC Payback Periods, Years	8.8
% Return On Invest (Excl Land)	11.02

CASE 1

No Credit For Recovered CO₂

1S7X2G09 = No H2 Recov w 100% Fd Bypass, 2-Stg Sour Shift, 5# resid CO2 PP, 0 MPH HRSG Firing, w Stm as NOx Diluent. (\$0/ST recov CO2 credit & \$0/ST CO2 emission)

OVERALL (H2 + MWe) EFFICIENCIES: PROCESS ECONOMIC EVALUATION SUMMARY:

Stream Flows, MPH:										HHV Basis										28.5 % HHV										Breakeven H2 Price, \$/1000 SCF										N/A																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																																								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Run ID: 1S7X2G09 = No H2 Recov w 100% Pd Bypass, 2-Stg Sour Shift, 5# resid CO2 PP, 0 MPH HRSG Firing, w Stm as Nox Diluent.

INCOMES & EXPENSES:		Expenses		Incomes		INVESTMENT & CAPITAL RECOVERY:		Economic Analysis Basis:		
	\$MM/Yr	\$MM/Yr	\$MM/Yr							
Coal Feed	25.2			Total Plant Capital Cost, \$MM	536.1	As Rec'd Coal Cost, \$/ST				29.4
Fluxant Feed	0.0			Initial Operating Capital, \$MM	18.2	Fluxant Cost, \$/ST Dry				10
Power Import	0.0			Land Cost @ \$0,000/Acre, \$MM	0.0	Import Power Cost, \$/MWH				85.351
Raw Water Import	0.2					Raw Water Cost, \$/1000 Gal				0.25
Cat & Chem Consumptn	7.6			Total Initial Investment, \$MM	554.3	Hydrogen Worth, \$/1000 SCF				To be Calc
Royalties	0			Owner Initial Investment, \$MM	110.9	Export Power Worth, \$/MWH				85.351
Mainten Labor & Mat	16.1			Bank Loan, \$MM	443.4	Sulfur Worth, \$/ST				30
Admin & Labor Salary	8.8					Slag & Ash Worth, \$/ST Dry				-10
Overheads & Benefits	3.1			Construction Period, Years	0	Waste H2O Worth, \$/1000 Gal				0
Insurances	5.4			Plant Life, Years	25	Recovered CO2 Worth, \$/ST				0
Local Taxes	5.4					CO2 Emission Penalty, \$/ST				0
Power Sales		133.1		Bank Loan Interest Rate, % per Year	10	Annual Maint Labor+Mat, % TPC				3
Sulfur Sales		0.6		Bank Loan Repayment Period, Years	25	CO2 Emission Penalty, \$/Year				125
Ash & Slag Disposal	1.0			Annual Bank Loan Repayment, \$MM/Yr	48.9	Avg Salaries, \$/Year				70,000
Waste Water Disposal	0.0			Equival Depreciatn & Interest Expenses:		OVHDS & Benefits, % Labor				35
CO2 Sequestn Credit		0.0		Bank Loan S/L Depreciation, \$MM/Yr	17.7	Construction Period, Years				0
CO2 Emission Penalty	0.0			Bank Loan Interest, \$MM/Year	31.1	Land Cost, \$/Ft2				0
Bank Loan Repayment	48.9					Initial Oper Fund, Months				3
Investment Recovery	12.2			Owner's Investment ROI, % per Year	10	Plant Life, Years				25
				Owner's Invest Repayment Period, Yrs	25	Annual Onstream Factor, %				80
Hydrogen Sales		(0.0)		Annual Capital Recovery, \$MM/Yr	12.2	Owner's Investment, % TPC				20
Total Pre Income Taxes	133.7	133.7		Equival Depreciatn & Interest Expenses:		Owner Invest Repay Period, Yrs				25
Fed Income Tax (FIT)	0.0			Equival S/L Depreciatn, \$MM/Yr	4.4	Owner's ROI, % per Year				10
Stat Income Tax (SIT)	0.0			Equival Loan Interest, \$MM/Yr	7.8	Bank Loan Interest Rate, %/Yr				10
Total Post Income Tax	133.7	133.7				Insurances, % TPC				1
				Overall Ht Bal, MM(LHV)/Hr:	In Out	FIT, % Net Income				30
						SIT, % Net Income less FIT				10
						Local (Property) Taxes, % TPC				1
						Royalties:				
						Gasification, STPD MF Coal				0
						Sour Shift, \$/MMSCFD Feed				0

PROCESS ECONOMIC EVALUATION SUMMARY:

Breakeven H2 Price, \$/1000 SCF N/A
 Breakeven H2 Price, \$/MMBtu(HHV) N/A

CASE 1

\$26/ST Credit for Recovered CO₂

1S7X2G09 = No H2 Recov w 100% Fd Bypass, 2-Stg Sour Shift, 5# resid CO2 PP, 0 MPH HRSG Firing, w Stm as NOx Diluent. (\$26/ST recov CO2 credit & \$0/ST CO2 emission)

OVERALL (H2 + MWe) EFFICIENCIES: PROCESS ECONOMIC EVALUATION SUMMARY:

Stream Flows, MPH:	1.	2.	3.	4.	5.	6.	7.	8.	OVERALL (H2 + MWe)	EFFICIENCIES:	TPC Payback Periods, Years	% Return On Invest (Excl Land)
N2	286	286	257	257	257	0	0	0	28.5 % HHV	HHV Basis	8.8	11.02
O2	0	0	0	0	0	0	0	0	30.2 % LHV	LHV Basis		
CO2	3,399	3,399	10,758	209	209	0	0	0	29.3 % HHV	(HHV w S)		
Ar	91	91	82	82	82	0	0	0	31.0 % LHV	(LHV w S)		
H2	6,592	6,592	13,632	13,632	13,632	0	0	0				
CO	9,224	9,224	603	603	603	0	0	0				
H2S	179	179	161	0	0	0	0	0				
COS	9	9	8	0	0	0	0	0				
CH4	770	770	693	693	693	0	0	0				
C2H6	0	0	0	0	0	0	0	0				
Steam	5,075	5,454	57	0	0	0	9,827	0				
Total	25,624	26,003	26,250	15,476	15,476	0	9,827	0				

INCOMES & EXPENSES:	INCOMES	EXPENSES	INCOMES
AR Coal @ \$29.4/ST	5,472	25.2	\$MM/Yr
Fluxant Feed	0.0	0.0	
Pwr Import, \$58.7/MWH	0.2	0.0	
Raw Water Import	0.2	0.0	
Cat & Chem Consumptn	7.6	0.0	
Royalties	0.0	0.0	
Mainten Labor & Mat	16.1	0.0	
Admin & Labor Salary	8.8	0.0	
Overheads & Benefits	3.1	0.0	
Insurances	5.4	0.0	
Local Taxes	5.4	0.0	
Power Sale, \$58.7/MWH	91.5	0.0	
Sulfur Sale	0.6	0.0	
Ash & Slag Disposal	1.0	0.0	
Waste Water Disposal	0.0	0.0	
CO2 Sequestn Credit	41.6	0.0	
CO2 Emission Penalty	48.9	0.0	
Bank Loan Repayment	12.2	0.0	
Investment Recovery	(0.0)	0.0	
Hydrogen Sales	133.7	133.7	
Pre Income Tax Total	133.7	133.7	
Fed Income Tax (FIT)	0.0	0.0	
Stat Income Tax (SIT)	0.0	0.0	
Post Income Tax Total	133.7	133.7	

CONSUMABLES:	CONSUMABLES	CONSUMABLES	CONSUMABLES
As Rec'd Coal, STPD	2,937	0.0	
Fluxant, STPD Dry	0	0.0	
Power Import, MW	0	0.0	
Raw Water, GPM	2,370	0.0	
95.0% O2, STPD	2,200	0.0	

PROCESS DATA:	PROCESS DATA	PROCESS DATA	PROCESS DATA
Gasifier	Destec	0.0	
Air Separation	Conventional	0.0	
Shift: Type	2-Stg Sour	0.0	
Overall Bypass	0.0 % Fd	0.0	
1st Stg Bypass	50.0 % Fd	0.0	
Stm/CO Ratio:	0.59	0.0	
Syngas Feed	5.00	0.0	
1st Stg Fd	77.98	0.0	
1st Stg Prod	4.41	0.0	
2nd Stg Fd	4.41	0.0	

AGR: Type	AGR: Type	AGR: Type	AGR: Type
CO2 Removal	Selexol	98.1 %	
Sweet Gas CO2 PP	4.9 psia	0.0	
SRU: Type	Claus	45 vol% Dry	
AG Fd H2S Conc	45 vol% Dry	0.0	

H2 Rec: Type	H2 Rec: Type	H2 Rec: Type	H2 Rec: Type
Fd Bypass	No	100.0 %	
H2 Recovery	90.0 %	0.0	
Net H2 Recov	0.0 %	0.0	

Power Plt: Type	Power Plt: Type	Power Plt: Type	Power Plt: Type
EGT	GE 7FA CC	1061 deg F	
FG H2 Conc	88 vol% Dry	260 deg F	
Stack Temp	927 STPD	85 %	

HHV, Btu/#	HHV, Btu/#	HHV, Btu/#	HHV, Btu/#
LHV, Btu/#	10,295	10,295	
Stack CO2 Emission	927 STPD	85 %	
% C Sequestered	85 %	85 %	

COAL PROPERTIES:	COAL PROPERTIES	COAL PROPERTIES	COAL PROPERTIES
Type	Illin #6	As Rec'd	
Analysis:	C	59.87	
H	4.27	4.27	
N	1.11	1.11	
S	2.21	2.21	
O	7.07	7.07	
Chlorine	0.11	0.11	
Ash	10.86	10.86	
Moisture	14.50	14.50	
Total Wt%	100.00	100.00	

HHV, Btu/#	HHV, Btu/#	HHV, Btu/#	HHV, Btu/#
LHV, Btu/#	10,900	10,900	
Stack CO2 Emission	927 STPD	85 %	
% C Sequestered	85 %	85 %	

Power Balance, MW	Power Balance, MW	Power Balance, MW	Power Balance, MW
Produced	95	95	
Consumed	346	346	
Net	250	250	

Cap Cost Summary	Cap Cost Summary	Cap Cost Summary	Cap Cost Summary
\$MM(2001)	\$MM(2001)	\$MM(2001)	\$MM(2001)
Coal Handling/Feeding	29.8	29.8	
Gasification	59.2	59.2	
Spare Gasifiers (Total 1 Spare)	16.9	16.9	
Air Separation	45.8	45.8	
Incl ASU	17.5	17.5	
O2 Compression (670 PSIG)	48.9	48.9	
Shift & LT Gas Cooling	5.5	5.5	
Acid Gas Removal & SWS	0.0	0.0	
Sulfur Recovery	0.0	0.0	
H2 Recovery	0.0	0.0	
H2 Compression (1000 PSIG)	28.2	28.2	
CO2 Dehydr'n & Comp (2000 PSIG)	0.3	0.3	
Fuel Gas Compression (345 PSIG)	0.0	0.0	
N2 Compression (290 PSIG)	71.0	71.0	
Gas Turb Generator System	18.8	18.8	
HRSG/Boiler Plt/BFW/DW/Cond	15.6	15.6	
Steam Turb Generator System	9.6	9.6	
CW Pumps & CT Fans	78.2	78.2	
Flue Gas CO2 Recovery	445.3	445.3	
BOP	24.0	24.0	
Total Installed Cost (TIC)	66.8	66.8	
Home Office @ 5.4 % of TIC	3.6	3.6	
Contingency @ 15.0 % of TIC	536.1	536.1	

Total Plant Cost (TPC)	Total Plant Cost (TPC)	Total Plant Cost (TPC)	Total Plant Cost (TPC)
1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.	1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.	1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.	1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.
H2 Prod HHV Btu/SCF = 324	H2 Prod HHV Btu/SCF = 324	H2 Prod HHV Btu/SCF = 324	H2 Prod HHV Btu/SCF = 324
GTG/6	GTG/6	GTG/6	GTG/6

Process Design Program	Process Design Program	Process Design Program	Process Design Program
IGH2CC9C	IGH2CC9C	IGH2CC9C	IGH2CC9C

03-Dec-2004	03-Dec-2004	03-Dec-2004	03-Dec-2004
03-Dec-2004	03-Dec-2004	03-Dec-2004	03-Dec-2004

R.Chu	R.Chu	R.Chu	R.Chu
R.Chu	R.Chu	R.Chu	R.Chu

Run ID: 1STX2G09 = No H2 Recov w 100% Fd Bypass, 2-Stg Sour Shift, 5# resid CO2 PP, 0 MPH HRSG Firing, w Scm as Nox Diluent.

INCOMES & EXPENSES:		Expenses		Incomes		INVESTMENT & CAPITAL RECOVERY:		Economic Analysis Basis:	
		\$MM/Yr	\$MM/Yr						
Coal Feed		25.2				Total Plant Capital Cost, \$MM	536.1	As Rec'd Coal Cost, \$/ST	29.4
Fluxant Feed		0.0				Initial Operating Capital, \$MM	18.2	Fluxant Cost, \$/ST Dry	10
Power Import		0.0				Land Cost @ \$0,000/Acre, \$MM	0.0	Import Power Cost, \$/MWH	58.679
Raw Water Import		0.2						Raw Water Cost, \$/1000 Gal	0.25
Cat & Chem Consumptn		7.6				Total Initial Investment, \$MM	554.3		
Royalties		0				Owner Initial Investment, \$MM	110.9		
Mainten Labor & Mat		16.1				Bank Loan, \$MM	443.4	Hydrogen Worth, \$/1000 SCF	To be Calc
Admin & Labor Salary		8.8						Export Power Worth, \$/MWH	58.679
Overheads & Benefits		3.1				Construction Period, Years	0	Sulfur Worth, \$/ST	30
Insurances		5.4				Plant Life, Years	25	Slag & Ash Worth, \$/ST Dry	-10
Local Taxes		5.4						Waste H2O Worth, \$/1000 Gal	0
Power Sales			91.5			Bank Loan Interest Rate, % per Year	10	Recovered CO2 Worth, \$/ST	26.038
Sulfur Sales			0.6			Bank Loan Repayment Period, Years	10	CO2 Emission Penalty, \$/ST	0
Ash & Slag Disposal		1.0				Annual Bank Loan Repayment, \$MM/Yr	25	Annual Maint Labor+Mat, % TPC	3
Waste Water Disposal		0.0				Equival Depreciatn & Interest Expenses:	48.9	Employees, 1xAdmin + 4xOper	125
CO2 Sequestn Credit			41.6			Bank Loan S/L Depreciation, \$MM/Yr	17.7	Avg Salaries, \$/Year	70,000
CO2 Emission Penalty		0.0				Bank Loan Interest, \$MM/Year	31.1	OVRDs & Benefits, % Labor	35
Bank Loan Repayment		48.9						Construction Period, Years	0
Investment Recovery		12.2				Owner's Investment ROI, % per Year	10	Land Cost, \$/Ft2	0
			(0.0)			Owner's Invest Repayment Period, Yrs	25	Initial Oper Fund, Months	3
Hydrogen Sales						Annual Capital Recovery, \$MM/Yr	12.2	Plant Life, Years	25
Total Pre Income Taxes		133.7	133.7			Equival Depreciatn & Interest Expenses:		Annual Onstream Factor, %	80
Fed Income Tax (FIT)		0.0				Equival S/L Depreciatn, \$MM/Yr	4.4	Owner's Investment, % TPC	20
Stat Income Tax (SIT)		0.0				Equival Loan Interest, \$MM/Yr	7.8	Owner Invest Repay Period, Yrs	25
								Owner's ROI, % per Year	10
Total Post Income Tax		133.7	133.7			Overall Ht Bal, MM(LHV)/Hr:	In Out	Bank Loan Interest Rate, %/Yr	10
								Insurances, % TPC	1
								FIT, % Net Income	30
								SIT, % Net Income less FIT	10
								Local (Property) Taxes, % TPC	1
								Royalties:	
								Gasification, STPD MF Coal	0
								Sour Shift, \$/MMSCFD Feed	0
								AGR, \$/MMSCFD Acid Gas	0
								Sulfur Plt, \$/STPD Sulfur	0
								TGTU, \$/STPD Sulfur	0
								PSA, \$/MMSCFD H2	0
								Catalysts & Chemicals, % TIC:	1.7
								Memb Perf Fact, x 20ml/min/cm2:	2.5

PROCESS ECONOMIC EVALUATION SUMMARY:

Breakeven H2 Price, \$/1000 SCF	N/A
Brkeven H2 Price, \$/MMBtu(HHV)	N/A
TPC Payback Periods, Years	8.8
% Return On Invest (Excl Land)	11.02

CASE 2

No Credit For Recovered CO₂

2572F2G09 = PSA H2 Recov w 21% Fd Bypass, 2-Stg Sour Shift, 5# resid CO2 PP, 0 MPH HRSG Firing, w Stm as NOx Diluent. (\$0/ST recov CO2 credit & \$0/ST CO2 emission)

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1.	2.	3.	4.	5.	6.	7.	8.	OVERALL (H2 + Mwe)	EFFICIENCIES:	PROCESS ECONOMIC EVALUATION SUMMARY:
=====										

Stream Flows, MPH:

[illegible]

Run ID: 2S7P2G09 = PSA H2 Recov w 21% Fd Bypass, 2-Stg Sour Shift, 5# resid CO2 PP, 0 MPH HRSG Firing, w Sctm as NOx Diluent.

INCOMES & EXPENSES:		Expenses Incomes		INVESTMENT & CAPITAL RECOVERY:		Economic Analysis Basis:	
	\$MM/Yr	\$MM/Yr	\$MM/Yr				
Coal Feed	60.4			Total Plant Capital Cost, \$MM	986.2	As Rec'd Coal Cost, \$/ST	29.4
Fluxant Feed	0.0			Initial Operating Capital, \$MM	34.6	Fluxant Cost, \$/ST Dry	10
Power Import	0.0			Land Cost @ \$0,000/Acre, \$MM	0.0	Import Power Cost, \$/MWH	85.351
Raw Water Import	0.4					Raw Water Cost, \$/1000 Gal	0.25
Cat & Chem Consumptn	14.0			Total Initial Investment, \$MM	1,020.7		
Royalties	0			Owner Initial Investment, \$MM	204.1		
Mainten Labor & Mat	29.6			Bank Loan, \$MM	816.6	Hydrogen Worth, \$/1000 SCF	To be Calc
Admin & Labor Salary	8.8					Export Power Worth, \$/MWH	85.351
Overheads & Benefits	3.1			Construction Period, Years	0	Sulfur Worth, \$/ST	30
Insurances	9.9			Plant Life, Years	25	Slag & Ash Worth, \$/1000 Gal	-10
Local Taxes	9.9					Recovered CO2 Worth, \$/ST	0
Power Sales		73.7		Bank Loan Interest Rate, % per Year	10	CO2 Emission Penalty, \$/ST	0
Sulfur Sales		1.4		Bank Loan Repayment Period, Years	10	Annual Maint Labor+Mat, % TPC	0
Ash & Slag Disposal	2.4			Bank Loan Repayment, \$MM/Yr	25	Employees, 1xAdmin + 4xOper	3
Waste Water Disposal	0.0			Annual Bank Loan Repayment, \$MM/Yr	25	Avg Salaries, \$/Year	125
CO2 Sequestn Credit	0.0			Equival Depreciatn & Interest Expenses:	90.0	OVRDs & Benefits, % Labor	70,000
CO2 Emission Penalty	0.0			Bank Loan S/L Depreciation, \$MM/Yr	32.7		35
Bank Loan Repayment	90.0			Bank Loan Interest, \$MM/Year	57.3	Construction Period, Years	0
Investment Recovery	22.5					Land Cost, \$/Pt2	0
				Owner's Investment ROI, % per Year	10	Initial Oper Fund, Months	3
Hydrogen Sales		175.6		Owner's Invest Repayment Period, Yrs	25	Plant Life, Years	25
				Annual Capital Recovery, \$MM/Yr	22.5	Annual Onstream Factor, %	80
Total Pre Income Taxes	250.7		250.7	Equival Depreciatn & Interest Expenses:		Owner's Investment, % TPC	20
				Equival S/L Depreciatn, \$MM/Yr	8.2	Owner Invest Repay Period, Yrs	25
Fed Income Tax (FIT)	0.0			Equival Loan Interest, \$MM/Yr	14.3	Owner's ROI, % per Year	10
Stat Income Tax (SIT)	0.0					Bank Loan Interest Rate, %/Yr	10
				Overall Ht Bal, MM(LHV)/Hr:	In Out	Insurances, % TPC	1
Total Post Income Tax	250.7		250.7			FIT, % Net Income	30
						SIT, % Net Income less FIT	10
						Local (Property) Taxes, % TPC	1
						Royalties:	
						Gasification, STPD MF Coal	0
						Sour Shift, \$/MMSCFD Feed	0
						AGR, \$/MMSCFD Acid Gas	0
						Sulfur Plt, \$/STPD Sulfur	0
						TGTU, \$/STPD Sulfur	0
						PSA, \$/MMSCFD H2	0
						Catalysts & Chemicals, % TIC:	1.7
						Memb Perf Fact, x 20ml/min/cm2:	2.5

PROCESS ECONOMIC EVALUATION SUMMARY:

Breakeven H2 Price, \$/1000 SCF	2.83
Breakeven H2 Price, \$/MMBtu(HHV)	8.74
TPC Payback Periods, Years	8.8
% Return On Invest (Excl Land)	11.02

CASE 2

\$26/ST Credit for Recovered CO₂

2S7P2G09 = PSA H2 Recov w 21% Fd Bypass, 2-Stg Sour Shift, 5# resid CO2 PP, 0 MPH HRSG Firing, w Stm as NOx Diluent. (\$26/ST recov CO2 credit & \$0/ST CO2 emission

OVERALL (H2 + MWe) EFFICIENCIES: PROCESS ECONOMIC EVALUATION SUMMARY:

Stream Flows, MPH:					
Syngas from Destec Gasifiers	Quenched Shift	Shifted Syngas to Acid Gas Removal	Treated Syngas to H2 Recovery Plant	Saturated Tail Gas to Power	Hydrogen Product to Compress'n
N2	685	616	616	614	2
O2	0	0	0	0	0
CO2	8,141	25,790	554	554	0
Ar	220	198	198	198	0
H2	15,803	32,686	32,686	9,370	23,316
CO	22,118	1,443	1,443	1,443	0
H2S	428	385	0	0	0
COS	21	19	0	0	0
CH4	1,844	1,660	1,660	1,660	0
C2H6	0	0	0	0	0
Steam	12,178	13,183	0	0	10,676
Total	61,438	62,443	37,157	13,838	23,318
Temperature, deg F	1,850	316	95	191	95
Pressure, psig	535	461	326	311	291
Cap Cost Summary \$MM(2001)					
Coal Handling/Feeding	67.6	(2)	---	3.4	Without Diluent
Gasification	134.4	(2)	---	4.2	After Sat w Dil
Spare Gasifiers (Total 1 Spare)	19.2	(1)	---	---	With Diluent
Air Separation	103.9	(2)	---	44.5	Target Max w Dil
O2 Compression (635 PSIG)	Incl ASU	---	---	26.0	---
Shift & LT Gas Cooling	46.3	(2)	---	1.4	---
Acid Gas Removal & SWS	110.9	(2)	---	17.1	---
Sulfur Recovery	10.2	(1)	---	0.6	---
H2 Recovery	41.9	(1)	---	0.1	---
H2 Compression (1000 PSIG)	6.1	(1)	---	13.7	---
CO2 Dehydr'n & Comp (2000 PSIG)	52.0	(1)	---	38.4	---
Fuel Gas Compression (345 PSIG)	5.1	(1)	---	9.8	---
N2 Compression (290 PSIG)	0.0	(1)	---	0.0	---
HRSG/Boiler Plt/BFW/DW/Cond	71.1	(1)	---	209.3	---
Steam Turb Generator System	23.3	(1)	---	5.2	---
CW Pumps & CT Fans	17.2	(1)	---	92.2	---
Flue Gas CO2 Recovery	14.0	---	---	6.2	---
BOP	98.2	---	---	7.6	---
Total Installed Cost (TIC)	821.4	Total	---	301.5	---
Home Office @ 5.1 % of TIC	41.6	---	---	178.2	---
Contingency @ 15.0 % of TIC	123.2	---	---	---	---
Total Plant Cost (TPC)	986.2	---	---	---	---
1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.					
FG LHV, Btu/SCF: 327.7					
Overall Bypass 0.0 % Fd					
1st Stg Bypass 50.0 % Fd					
Stm/CO Ratio: Syngas Feed 0.60					
1st Stg Fd 5.00					
1st Stg Prod 77.99					
2nd Stg Fd 4.42					
AGR: Type Selextol					
CO2 Removal 97.9 %					
Sweet Gas CO2 PP 4.9 psia					
SRU: Type Claus					
AG Fd H2S Conc 45 vol%Dry					
H2 Rec: Type PSA					
Fd Bypass 20.7 %					
H2 Recovery 90.0 %					
Net H2 Recov 71.3 %					
Chlorine 0.11					
Ash 10.86					
Moisture 14.50					
Power Plt: Type GE 7FA CC					
EGT 1054 deg F					
FG H2 Conc 68 vol%Dry					
Stack Temp 260 deg F					
HHV,Btu/# 10,900					
LHV,Btu/# 10,295					
% C Sequestered 85 %					
1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.					
H2 Prod HHV Btu/SCF = 324					
GTG/T7					

Run ID: 2S7P2G09 = PSA H2 Recov w 21% Fd Bypass, 2-Stg Sour Shift, 5# resid CO2 PP, 0 MPH HRSG Firing, w Stm as NOx Diluent.

INCOMES & EXPENSES:		Expenses Incomes		INVESTMENT & CAPITAL RECOVERY:		Economic Analysis Basis:	
\$MM/Yr	\$MM/Yr	\$MM/Yr	\$MM/Yr				
Coal Feed	60.4			Total Plant Capital Cost, \$MM	986.2	As Rec'd Coal Cost, \$/ST	29.4
Fluxant Feed	0.0			Initial Operating Capital, \$MM	34.6	Fluxant Cost, \$/ST Dry	10
Power Import	0.0			Land Cost @ \$0,000/Acre, \$MM	0.0	Import Power Cost, \$/MWH	58.679
Raw Water Import	0.4					Raw Water Cost, \$/1000 Gal	0.25
Cat & Chem Consumptn	14.0			Total Initial Investment, \$MM	1,020.7		
Royalties	0			Owner Initial Investment, \$MM	204.1	Hydrogen Worth, \$/1000 SCF	To be Calc
Mainten Labor & Mat	29.6			Bank Loan, \$MM	816.6	Export Power Worth, \$/MWH	58.679
Admin & Labor Salary	8.8					Sulfur Worth, \$/ST	30
Overheads & Benefits	3.1			Construction Period, Years	0	Slag & Ash Worth, \$/ST Dry	-10
Insurances	9.9			Plant Life, Years	25	Waste H2O Worth, \$/1000 Gal	0
Local Taxes	9.9					Recovered CO2 Worth, \$/ST	26.038
Power Sales	50.7			Bank Loan Interest Rate, % per Year	10	CO2 Emission Penalty, \$/ST	0
Sulfur Sales	1.4			Bank Loan Repayment Period, Years	25	Annual Maint Labor+Mat, % TPC	3
Ash & Slag Disposal	2.4			Bank Loan Repayment, \$MM/Yr	90.0	Employees, 1xAdmin + 4xOper	135
Waste Water Disposal	0.0			Annual Bank Loan Repayment, \$MM/Yr	90.0	Avg Salaries, \$/Year	70,000
CO2 Sequestn Credit	0.0			Equival Depreciatn & Interest Expenses:		OVHDS & Benefits, % Labor	35
CO2 Emission Penalty	0.0			Bank Loan S/L Depreciation, \$MM/Yr	32.7		
Bank Loan Repayment	90.0			Bank Loan Interest, \$MM/Year	57.3	Construction Period, Years	0
Investment Recovery	22.5					Land Cost, \$/Ft2	0
				Owner's Investment ROI, % per Year	10	Initial Oper Fund, Months	3
Hydrogen Sales				Owner's Invest Repayment Period, Yrs	25	Plant Life, Years	25
				Annual Capital Recovery, \$MM/Yr	22.5	Annual Onstream Factor, %	80
Total Pre Income Taxes	250.7			Equival Depreciatn & Interest Expenses:		Owner's Investment, % TPC	20
				Equival S/L Depreciatn, \$MM/Yr	8.2	Owner Invest Repay Period, Yrs	25
Fed Income Tax (FIT)	(0.0)			Equival Loan Interest, \$MM/Yr	14.3	Owner's ROI, % per Year	10
Stat Income Tax (SIT)	(0.0)					Bank Loan Interest Rate, %/Yr	10
Total Post Income Tax	250.7			Overall Ht Bal, MM(LHV)/Hr:	In Out	Insurances, % TPC	1
						FIT, % Net Income	30
						SIT, % Net Income less FIT	10
						Local (Property) Taxes, % TPC	1
						Royalties:	
						Gasification, STPD MF Coal	0
						Sour Shift, \$/MMSCFD Feed	0
						AGR, \$/MMSCFD Acid Gas	0
						Sulfur Plt, \$/STPD Sulfur	0
						TGTU, \$/STPD Sulfur	0
						PSA, \$/MMSCFD H2	0
						Catalysts & Chemicals, % TIC:	1.7
						Memb Perf Fact, x 20ml/min/cm2:	2.5

PROCESS ECONOMIC EVALUATION SUMMARY:

Breakeven H2 Price, \$/1000 SCF	1.60
Breakeven H2 Price, \$/MMBtu (HHV)	4.93
TPC Payback Periods, Years	8.8
% Return On Invest (Excl Land)	11.02

CASE 3

No Credit For Recovered CO₂

2S7I2G09 = ITN Membrane H2 Recov w 0% Fd Bypass, 2-Stg Sour Shift, 5# resid CO2 PP, 0 MPH HRSG Firing, w Stm as NOx Diluent. (\$0/ST recov CO2 credit & \$0/ST CO2 e

OVERALL (H2 + MWe) EFFICIENCIES: PROCESS ECONOMIC EVALUATION SUMMARY:

1. 2. 3. 4. 5. 6. 7. 8. OVERALL (H2 + MWe) EFFICIENCIES: PROCESS ECONOMIC EVALUATION SUMMARY:

Stream Flows, MPH: Syngas Quenched Shifted Treated Saturated Hydrogen GT
from Syngas to Syngas Tail Gas Product NOx
Destec Acid Gas to H2 to Power to Diluent
Gasifiers Shift Removal Recovery Plant Compress'n

685 685 616 616 643 0 0 0
0 0 0 0 0 0 0 0
8,141 8,141 25,786 469 570 0 0 0
220 220 198 198 206 0 0 0
15,803 15,803 32,682 8,946 22,666 0 0 0
22,118 22,118 1,447 1,399 1 0 0 0
428 428 385 0 0 0 0 0
COS 21 19 0 0 0 0 0 0
CH4 1,844 1,844 1,660 1,605 1 0 0 0
C2H6 0 0 0 0 0 0 0 0
Steam 12,158 12,978 128 0 1,177 0 8,994 0
Total 61,417 62,237 62,921 37,071 14,546 22,669 8,994

Temperature, deg F 1,850 324 95 862 91 550
Pressure, psig 595 521 386 371 340 290

Cap Cost Summary \$MM(2001) Trains # of FG LHV, Btu/SCF:

Coal Handling/Feeding 67.6 (2) 3.4 299.4
Gasification 134.4 (2) 4.2 299.4
Spare Gasifiers (Total 1 Spare) 19.2 (1) 185.0
Air Separation 107.6 (2) 46.8 185.0
Shift & LT Gas Cooling Incl ASU 48.2 (2) 28.1
Sulfur Recovery 111.1 (2) 18.5
H2 Recovery (243382 Ft2 Memb) 30.8 (1) 0.0
H2 Compression (1000 PSIG) 13.3 (1) 26.5
CO2 Dehydr'n & Comp (2000 PSIG) 52.1 (1) 38.5
Fuel Gas Compression (345 PSIG) 0.4 (1) 0.0
N2 Compression (290 PSIG) 0.0 (1) 0.0
Gas Turb Generator System 71.1 (1) 209.3
HRSG/Boiler Plt/BFW/DW/Cond 23.2 (1) 5.3
Steam Turb Generator System 18.1 (1) 97.4
CW Pumps & CT Fans 14.9
Flue Gas CO2 Recovery
BOP 99.0

Total Installed Cost (TIC) 821.1
Home Office @ 5.5 % of TIC 45.0
Contingency @ 15.0 % of TIC 123.2

Total Plant Cost (TPC) 989.3

1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.

H2 Prod HHV Btu/SCF = 324

GTG/10

INCOMES & EXPENSES: \$MM/Yr

AR Coal @ \$29.4/ST 60.4
Fluxant Feed 0.0
Pwr Import, \$85.4/MWH 0.5
Raw Water Import 0.0
Cat & Chem Consumptn 14.0
Royalties 0.0
Mainten Labor & Mat 29.7
Admin & Labor Salary 8.8
Overheads & Benefits 3.1
Insurances 9.9
Local Taxes 9.9
Power Sale, \$85.4/MWH 71.1
Sulfur Sale 1.4
Ash & Slag Disposal 2.4
Waste Water Disposal 0.0
CO2 Sequestn Credit 0.0
Bank Loan Repayment 90.2
Investment Recovery 22.6

Hydrogen Sales 178.7
Pre Income Tax Total 251.2
Fed Income Tax (FIT) 0.0
Stat Income Tax (SIT) 0.0
Post Income Tax Total 251.2

CAPITAL RECOVERY BASIS:

Total Plant Capital Cost, \$MM 989.3
Initial Operating Capital, \$MM 34.6
Land Cost @ \$0,000/Acre, \$MM 0.0
Total Initial Investment, \$MM 1,023.9
Construction Period, Years 0
Plant Life, Years 25
Annual Onstream Factor, % 80
Owner's Investment, % Initial 20
Owner's ROI, % per Year 10
Bank Loan Interest Rate, %/Yr 10

Hydrogen, MMSCFD 206.5
Power Export, MWe 119
Sulfur, STPD 155
Slag & Ash, STPD Dry 806
Waste Water, GPM 1,979
Recovered CO2, STPD 13,133

CONSUMABLES:

As Rec'd Coal, STPD 7,034
Fluxant, STPD Dry 0
Power Import, MW 0
Raw Water, GPM 4,335
95.0% O2, STPD 5,538

PROCESS DATA:

Shift: Type Overall Bypass 0.0 % Fd 2-Stg Sour
1st Stg Bypass 50.0 % Fd Stm/CO Ratio: 0.59
1st Stg Fd 5.00
1st Stg Prod 77.99
2nd Stg Fd 4.41

AGR: Type CO2 Removal Sweet Gas CO2 PP 4.9 psia
SRU: Type Claus AG Fd H2S Conc 45 vol% Dry

H2 Rec: Type ITN Membra
FD Bypass 0.0 %
Min H2 PP Dif 120.3 psi
Net H2 Recov 69.4 %

Power Plt: Type GE 7FA CC
EGT 1062 deg F
FG H2 Conc 67 vol% Dry
Stack Temp 260 deg F
Stack CO2 Emission 2164 STPD
% C Sequestered 85 %

HHV, Btu/# 10,900
LHV, Btu/# 10,295

Total Wt% 100.00

COAL PROPERTIES:

Type Illin #6
Analysis: C 59.87 H 4.27 N 1.11 S 2.21 O 7.07
Chlorine 0.11 Ash 10.86 Moisture 14.50

Power Balance, MW Produced Consumed

Without Diluent 3.4 299.4
After Sat w Dil 4.2 299.4
With Diluent 46.8 185.0
Target Max w Dil 28.1

Cap Cost Summary \$MM(2001) Trains # of FG LHV, Btu/SCF:

Coal Handling/Feeding 67.6 (2) 3.4 299.4
Gasification 134.4 (2) 4.2 299.4
Spare Gasifiers (Total 1 Spare) 19.2 (1) 185.0
Air Separation 107.6 (2) 46.8 185.0
Shift & LT Gas Cooling Incl ASU 48.2 (2) 28.1
Sulfur Recovery 111.1 (2) 18.5
H2 Recovery (243382 Ft2 Memb) 30.8 (1) 0.0
H2 Compression (1000 PSIG) 13.3 (1) 26.5
CO2 Dehydr'n & Comp (2000 PSIG) 52.1 (1) 38.5
Fuel Gas Compression (345 PSIG) 0.4 (1) 0.0
N2 Compression (290 PSIG) 0.0 (1) 0.0
Gas Turb Generator System 71.1 (1) 209.3
HRSG/Boiler Plt/BFW/DW/Cond 23.2 (1) 5.3
Steam Turb Generator System 18.1 (1) 97.4
CW Pumps & CT Fans 14.9
Flue Gas CO2 Recovery
BOP 99.0

Total Installed Cost (TIC) 821.1
Home Office @ 5.5 % of TIC 45.0
Contingency @ 15.0 % of TIC 123.2

Total Plant Cost (TPC) 989.3

1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.

H2 Prod HHV Btu/SCF = 324

GTG/10

INCOMES & EXPENSES: \$MM/Yr

AR Coal @ \$29.4/ST 60.4
Fluxant Feed 0.0
Pwr Import, \$85.4/MWH 0.5
Raw Water Import 0.0
Cat & Chem Consumptn 14.0
Royalties 0.0
Mainten Labor & Mat 29.7
Admin & Labor Salary 8.8
Overheads & Benefits 3.1
Insurances 9.9
Local Taxes 9.9
Power Sale, \$85.4/MWH 71.1
Sulfur Sale 1.4
Ash & Slag Disposal 2.4
Waste Water Disposal 0.0
CO2 Sequestn Credit 0.0
Bank Loan Repayment 90.2
Investment Recovery 22.6

Hydrogen Sales 178.7
Pre Income Tax Total 251.2
Fed Income Tax (FIT) 0.0
Stat Income Tax (SIT) 0.0
Post Income Tax Total 251.2

CAPITAL RECOVERY BASIS:

Total Plant Capital Cost, \$MM 989.3
Initial Operating Capital, \$MM 34.6
Land Cost @ \$0,000/Acre, \$MM 0.0
Total Initial Investment, \$MM 1,023.9
Construction Period, Years 0
Plant Life, Years 25
Annual Onstream Factor, % 80
Owner's Investment, % Initial 20
Owner's ROI, % per Year 10
Bank Loan Interest Rate, %/Yr 10

Hydrogen, MMSCFD 206.5
Power Export, MWe 119
Sulfur, STPD 155
Slag & Ash, STPD Dry 806
Waste Water, GPM 1,979
Recovered CO2, STPD 13,133

CONSUMABLES:

As Rec'd Coal, STPD 7,034
Fluxant, STPD Dry 0
Power Import, MW 0
Raw Water, GPM 4,335
95.0% O2, STPD 5,538

PROCESS DATA:

Shift: Type Overall Bypass 0.0 % Fd 2-Stg Sour
1st Stg Bypass 50.0 % Fd Stm/CO Ratio: 0.59
1st Stg Fd 5.00
1st Stg Prod 77.99
2nd Stg Fd 4.41

AGR: Type CO2 Removal Sweet Gas CO2 PP 4.9 psia
SRU: Type Claus AG Fd H2S Conc 45 vol% Dry

H2 Rec: Type ITN Membra
FD Bypass 0.0 %
Min H2 PP Dif 120.3 psi
Net H2 Recov 69.4 %

Power Plt: Type GE 7FA CC
EGT 1062 deg F
FG H2 Conc 67 vol% Dry
Stack Temp 260 deg F
Stack CO2 Emission 2164 STPD
% C Sequestered 85 %

HHV, Btu/# 10,900
LHV, Btu/# 10,295

Total Wt% 100.00

COAL PROPERTIES:

Type Illin #6
Analysis: C 59.87 H 4.27 N 1.11 S 2.21 O 7.07
Chlorine 0.11 Ash 10.86 Moisture 14.50

Power Balance, MW Produced Consumed

Without Diluent 3.4 299.4
After Sat w Dil 4.2 299.4
With Diluent 46.8 185.0
Target Max w Dil 28.1

Cap Cost Summary \$MM(2001) Trains # of FG LHV, Btu/SCF:

Coal Handling/Feeding 67.6 (2) 3.4 299.4
Gasification 134.4 (2) 4.2 299.4
Spare Gasifiers (Total 1 Spare) 19.2 (1) 185.0
Air Separation 107.6 (2) 46.8 185.0
Shift & LT Gas Cooling Incl ASU 48.2 (2) 28.1
Sulfur Recovery 111.1 (2) 18.5
H2 Recovery (243382 Ft2 Memb) 30.8 (1) 0.0
H2 Compression (1000 PSIG) 13.3 (1) 26.5
CO2 Dehydr'n & Comp (2000 PSIG) 52.1 (1) 38.5
Fuel Gas Compression (345 PSIG) 0.4 (1) 0.0
N2 Compression (290 PSIG) 0.0 (1) 0.0
Gas Turb Generator System 71.1 (1) 209.3
HRSG/Boiler Plt/BFW/DW/Cond 23.2 (1) 5.3
Steam Turb Generator System 18.1 (1) 97.4
CW Pumps & CT Fans 14.9
Flue Gas CO2 Recovery
BOP 99.0

Total Installed Cost (TIC) 821.1
Home Office @ 5.5 % of TIC 45.0
Contingency @ 15.0 % of TIC 123.2

Total Plant Cost (TPC) 989.3

1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.

H2 Prod HHV Btu/SCF = 324

GTG/10

Run ID: 2S7I2G09 = ITN Membrane H2 Recov w 0% Rd Bypass, 2-Std Sour Shift, 5# resid CO2 PP, 0 MPH HRSG Firing, w Stm as NOx Diluent.

INCOMES & EXPENSES:		INVESTMENT & CAPITAL RECOVERY:		Economic Analysis Basis:	
Expenses	Incomes				
\$MM/Yr	\$MM/Yr				
Coal Feed	60.4	Total Plant Capital Cost, \$MM	989.3	As Rec'd Coal Cost, \$/ST	29.4
Fluxant Feed	0.0	Initial Operating Capital, \$MM	34.6	Fluxant Cost, \$/ST Dry	10
Power Import	0.0	Land Cost @ \$0,000/Acre, \$MM	0.0	Import Power Cost, \$/MWH	85.351
Raw Water Import	0.5			Raw Water Cost, \$/1000 Gal	0.25
Cat & Chem Consumptn	14.0	Total Initial Investment, \$MM	1,023.9	Hydrogen Worth, \$/1000 SCF	To be Calc
Royalties	0	Owner Initial Investment, \$MM	204.8	Export Power Worth, \$/MWH	85.351
Mainten Labor & Mat	29.7	Bank Loan, \$MM	819.1	Sulfur Worth, \$/ST	30
Admin & Labor Salary	8.8			Slag & Ash Worth, \$/ST Dry	-10
Overheads & Benefits	3.1	Construction Period, Years	0	Waste H2O Worth, \$/1000 Gal	0
Insurances	9.9	Plant Life, Years	25	Recovered CO2 Worth, \$/ST	0
Local Taxes	9.9			CO2 Emission Penalty, \$/ST	0
Power Sales	71.1	Bank Loan Interest Rate, % per Year	10	Annual Maint Labor+Mat, % TPC	0
Sulfur Sales	1.4	Bank Loan Repayment Period, Years	25	Employees, lxAdmin + 4xOper	3
Ash & Slag Disposal	2.4	Annual Bank Loan Repayment, \$MM/Yr	90.2	Avg Salaries, \$/Year	125
Waste Water Disposal	0.0	Equival Depreciatn & Interest Expenses:		OVBDS & Benefits, % Labor	70,000
CO2 Sequestn Credit	0.0	Bank Loan S/L Depreciation, \$MM/Yr	32.8	Construction Period, Years	35
CO2 Emission Penalty	0.0	Bank Loan Interest, \$MM/Year	57.5	Land Cost, \$/Fc2	0
Bank Loan Repayment	90.2			Initial Oper Fund, Months	0
Investment Recovery	22.6	Owner's Investment ROI, % per Year	10	Plant Life, Years	3
		Owner's Invest Repayment Period, Yrs	25	Annual Onstream Factor, %	25
Hydrogen Sales	178.7	Annual Capital Recovery, \$MM/Yr	22.6	Owner's Investment, % TPC	80
		Equival Depreciatn & Interest Expenses:		Owner Invest Repay Period, Yrs	20
Total Pre Income Taxes	251.2	Equival S/L Depreciatn, \$MM/Yr	8.2	Owner's ROI, % per Year	25
Fed Income Tax (FIT)	0.0	Equival Loan Interest, \$MM/Yr	14.4	Bank Loan Interest Rate, %/Yr	10
Stat Income Tax (SIT)	0.0			Insurances, % TPC	1
		Overall Ht Bal, MM(LHV)/Hr:	In Out	FIT, % Net Income	30
Total Post Income Tax	251.2			SIT, % Net Income less FIT	10
				Local (Property) Taxes, % TPC	1
				Royalties:	
				Gasification, STPD MF Coal	0
				Sour Shift, \$/MMSCFD Feed	0
				AGR, \$/MMSCFD Acid Gas	0
				Sulfur Plt, \$/STPD Sulfur	0
				TGTU, \$/STPD Sulfur	0
				PSA, \$/MMSCFD H2	0
				Catalysts & Chemicals, % TIC:	1.7
				Memb Perf Fact, x 20ml/min/cm2:	2.5

PROCESS ECONOMIC EVALUATION SUMMARY:

Breakeven H2 Price, \$/1000 SCF	2.96
Breakeven H2 Price, \$/MMBtu(HHV)	9.15
TPC Payback Periods, Years	8.8
% Return On Invest (Excl Land)	11.02

CASE 3

\$26/ST Credit for Recovered CO₂

28712G09 = ITN Membrane H2 Recov w 0% Fd Bypass, 2-Stdg Sour Shift, 5# resid CO2 PP, 0 MPH HRSG Firing, w Stm as NOx Diluent. (\$26/ST recov CO2 credit & \$0/ST CO2

OVERALL (H2 + MWe) EFFICIENCIES: PROCESS ECONOMIC EVALUATION SUMMARY:

1.	2.	3.	4.	5.	6.	7.	8.	OVERALL (H2 + MWe) EFFICIENCIES:	PROCESS ECONOMIC EVALUATION SUMMARY:
Syngas from Destec Gasifiers Shift	Quenched Syngas to Acid Gas Removal	Shifted Syngas to Acid Gas Removal	Treated Syngas to H2 Recovery	Saturated Tail Gas to H2 Plant	Hydrogen Product to Compress'n	GT Nox Diluent			
685	685	616	616	643	0	0			
0	0	0	0	0	0	0			
8,141	8,141	25,786	469	570	0	0			
220	220	198	198	206	0	0			
15,803	15,803	32,682	32,682	8,946	0	0			
22,118	22,118	1,447	1,447	1,399	1	0			
428	428	385	0	0	0	0			
21	21	19	0	0	0	0			
1,844	1,844	1,660	1,660	1,605	1	0			
0	0	0	0	0	0	0			
12,158	12,978	128	0	1,177	0	8,994			
61,417	62,237	62,921	37,071	14,546	22,669	8,994			
Temperature, deg F	1,850	324	95	862	91	550			
Pressure, psig	595	521	386	371	340	290			

Stream Flows, MPH:

N2	O2	CO2	Ar	H2	CO	H2S	COS	CH4	C2H6	Steam	Total	Temperature, deg F	Pressure, psig
685	0	8,141	220	15,803	22,118	428	21	1,844	0	12,158	61,417	1,850	595
685	0	8,141	220	15,803	22,118	428	21	1,844	0	12,158	61,417	1,850	595
616	0	25,786	198	32,682	1,447	385	19	1,660	128	12,978	62,237	324	521
616	0	469	198	32,682	1,447	385	19	1,660	128	12,978	62,237	324	521
643	0	570	206	8,946	1,399	0	0	1,605	0	1,177	37,071	95	386
0	0	0	0	0	1	0	0	0	0	0	22,669	862	371
0	0	0	0	0	0	0	0	0	0	0	91	862	371
0	0	0	0	0	0	0	0	0	0	0	80	340	371
0	0	0	0	0	0	0	0	0	0	0	22,669	862	371
0	0	0	0	0	0	0	0	0	0	0	91	862	371
0	0	0	0	0	0	0	0	0	0	0	80	340	371

PROCESS DATA: 48.9
Sulfur Sale 1.4
Ash & Slag Disposal 2.4
Waste Water Disposal 0.0
CO2 Sequestn Credit 99.9
CO2 Emission Penalty 0.0
Bank Loan Repayment 90.2
Investment Recovery 22.6

Hydrogen Sales 101.1
Pre Income Tax Total 251.2
Fed Income Tax (FIT) (0.0)
Stat Income Tax (SIT) (0.0)
Post Income Tax Total 251.2

Capital Recovery Basis: 251.2
Total Plant Capital Cost, \$MM 989.3
Initial Operating Capital, \$MM 34.6
Land Cost @ \$0,000/Acre, \$MM 0.0
Total Initial Investment, \$MM 1,023.9
Construction Period, Years 0
Plant Life, Years 25
Annual Onstream Factor, % 80
Owner's Investment, % Initial 20
Owner's ROI, % per Year 10
Bank Loan Interest Rate, %/Yr 10

1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat. H2 Prod HHV Btu/SCF = 324 GTG/10

Process Design Program : IGH2CC9B R.Chu

03-Dec-2004

Run ID: 2S712G09 = ITN Membrane H2 Recov w 0% Pd Bypass, 2-Stg Sour Shift, 5# resid CO2 PP, 0 MPH HRSG Firing, w Stm as NOx Diluent.

INCOMES & EXPENSES:		Expenses Incomes		Economic Analysis Basis:	
	\$MM/Yr	\$MM/Yr	\$MM/Yr		
Coal Feed	60.4			As Rec'd Coal Cost, \$/ST	29.4
Fluxant Feed	0.0			Fluxant Cost, \$/ST Dry	10
Power Import	0.0			Import Power Cost, \$/MMH	58.679
Raw Water Import	0.5			Raw Water Cost, \$/1000 Gal	0.25
Cat & Chem Consump	14.0				
Royalties	0			Hydrogen Worth, \$/1000 SCF	To be Calc
Mainten Labor & Mat	29.7			Export Power Worth, \$/MMH	58.679
Admin & Labor Salary	8.8			Sulfur Worth, \$/ST	30
Overheads & Benefits	3.1			Slag & Ash Worth, \$/ST Dry	-10
Insurances	9.9			Waste H2O Worth, \$/1000 Gal	0
Local Taxes	9.9			Recovered CO2 Worth, \$/ST	26.038
Power Sales	48.9			CO2 Emission Penalty, \$/ST	0
Sulfur Sales	1.4			Annual Maint Labor+Mat, % TPC	3
Ash & Slag Disposal	2.4			Employees, 1xAdmin + 4xOper	125
Waste Water Disposal	0.0			Avg Salaries, \$/Year	70,000
CO2 Sequestn Credit	0.0			OVHDS & Benefits, % Labor	35
CO2 Emission Penalty	0.0			Construction Period, Years	0
Bank Loan Repayment	90.2			Land Cost, \$/Ft2	0
Investment Recovery	22.6			Initial Oper Fund, Months	3
Hydrogen Sales		101.1		Plant Life, Years	25
Total Pre Income Taxes	251.2	251.2		Annual Onstream Factor, %	80
Fed Income Tax (FIT)	(0.0)			Owner's Investment, % TPC	20
Stat Income Tax (SIT)	(0.0)			Owner Invest Repay Period, Yrs	25
Total Post Income Tax	251.2	251.2		Owner's ROI, % per Year	10
				Bank Loan Interest Rate, %/Yr	10
				Insurances, % TPC	1
				FIT, % Net Income	30
				SIT, % Net Income less FIT	10
				Local (Property) Taxes, % TPC	1
				Royalties:	
				Gasification, STPD MF Coal	0
				Sour Shift, \$/MMSCFD Feed	0
				AGR, \$/MMSCFD Acid Gas	0
				Sulfur plt, \$/STPD Sulfur	0
				TGTU, \$/STPD Sulfur	0
				PSA, \$/MMSCFD H2	0
				Catalysts & Chemicals, % TIC:	1.7
				Memb Perf Fact, x 20ml/min/cm2:	2.5

PROCESS ECONOMIC EVALUATION SUMMARY:

Breakeven H2 Price, \$/1000 SCF	1.68
Breakeven H2 Price, \$/MMBtu (HHV)	5.18
TPC Payback Periods, Years	8.8
% Return On Invest (Excl Land)	11.02

CASE 4

No Credit For Recovered CO₂

15710G9Z = ITN Membrane H2 Recov w 0% Fd Bypass, None Shift, 56# resid CO2 PP, 0 MPH HRSG Firing, w Stm as Nox Diluent. (\$0/ST recov CO2 credit & \$0/ST CO2 emissi

1.	2.	3.	4.	5.	6.	7.	8.
<hr/>							
PROCESS ECONOMIC EVALUATION SUMMARY:							
<hr/>							
OVERALL (H2 + MWe) EFFICIENCIES:							
<hr/>							

Syngas	Quenched	Shifted	Treated Saturated Hydrogen	GT	HHV Basis	36.0 % HHV	Breakeven H ₂ Price, \$/1000 SCF
							5.05

[illegible][illegible]

342	308	316	0	Hydrogen, MMSCFD	46.7
=====				Hydrogen, MMSCFD	46.7

[illegible][illegible][illegible][illegible][illegible]

Power Import, MW	0	0	0	0	Admin & Labor Salary	8.8
Dev. Rate, \$/kW	5,426	5,605	5,426	5,426		
	6,029	6,029	6,029	6,029		

30,645	30,645	27,580	27,373	22,292	5,126	0	95.0% O2, STPD	2,719	Insurances	6.6
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[illegible]

Activity	1.1	1.2	1.3	1.4
Surplus Sale	200			
Ash & Slag Disposal				
Destec				
Gasifier				

Air separation	Power Balance, MW	Cap Cost	0.0
Conventional			0.0
Waste Water Disposal			0.0
CO2 Sequenstrn Credit			0.0

[illegible]

	Investment Recovery	1st Stg Bypass Not Appl.	Investment Recovery
ing/Feeding	33.8	(1)	stm/cv. Bypass.
Without Diluent	195.4	After Sat w/o Dil	195.4
With Diluent	195.4	After Sat w/ Dil	195.4
Investment Recovery	15.0		

[illegible]

(1)	23.0		Pre Income Tax Total	161.3
53.1	23.0		1st Stg Prod 0.55	161.3
million				
total per				
unit				

Gas Cooling	4.1	(1)	0.0	Fed Income Tax (FIT)	(0.0)
COAL PROPERTIES:					
				Fed Income Tax (FIT)	(0.0)

0.0	0.1	ack: type	med temp	stat income tax (sri)	(0.0)
20.5	(1)	Type	CO2 Removal	0.1 %	(0.0)
		Illin #6	CO2 Removal	0.1 %	(0.0)

	AS REC'd	Sweet Gas	CO ₂	PP	56.0 psia	Post Income Tax Total	161.3
Analysis:	(1)	Sulfuric Acid	SRU: Type	CO ₂	PP	56.0 psia	161.3
	(1)	Sulfuric Acid	SRU: Type	CO ₂	PP	56.0 psia	161.3

C	39.0	26.3	93	vol% Dry	CAPITAL RECOVERY BASIS:
H	0.0	0.0	59.87	AG Fd H2S Conc	
H	0.5	0.0	4.27	AG Fd H2S Conc	

	N	H2 Rec:	Type	ITN Membra	Total plant Capital Cost,	\$MM
0.0	0.0	1.11	Ed Brnacc	0 %	Total plant Capital Cost,	660.0
(1) Billion (290 PSIG)	(1)	2.21	Ed Brnacc	0 %	Total plant Capital Cost,	660.0
generator System	71-1	(1)				

		H2	Min	H2	pp	Dif	Land Cost @ \$0.000/Acre, \$MM
Plt/BFW/DN/Cond	22.2	(1)	7.07	6.11	3.5	psi	0.0
Generator System	10.4	(1)	52.4	0.1	3.5	psi	0.0

CT Fans	27.6		
CT Investments	52.5		
CT Receivables	52.5		
Ash	16.4	10.86	
Total Initial Investment, \$MM			681.6

[illegible]

	Total	261.9	109.6	100.00
lled Cost (TIC)	543.4			
Stock Temp 216 deg F				
Annual Onstream Factor, %				
Owner's Investment, % Initial				
80				
20				
20				

[illegible][illegible][illegible]

Run ID: 1S710G9Z = ITN Membrane H2 Recov w 0% Fd Bypass, None Shift, 56# resid CO2 PP, 0 MPH HRSG Firing, w Stm as NOx Diluent.

INCOMES & EXPENSES:		INVESTMENT & CAPITAL RECOVERY:		Economic Analysis Basis:	
\$MM/Yr	Incomes	\$MM	\$MM/Yr		
Coal Feed	30.2	Total Plant Capital Cost, \$MM	660.0	As Rec'd Coal Cost, \$/ST	29.4
Fluxant Feed	0.0	Initial Operating Capital, \$MM	21.5	Fluxant Cost, \$/ST Dry	10
Power Import	0.0	Land Cost @ \$0,000/Acre, \$MM	0.0	Import Power Cost, \$/MWH	85.351
Raw Water Import	0.8			Raw Water Cost, \$/1000 Gal	0.25
Cat & Chem Consump	9.2	Total Initial Investment, \$MM	681.6	Hydrogen Worth, \$/1000 SCF	To be Calc
Royalties	0	Owner Initial Investment, \$MM	136.3	Export Power Worth, \$/MWH	85.351
Mainten Labor & Mat	19.8	Bank Loan, \$MM	545.3	H2SO4 Worth, \$/ST S Eq	60
Admin & Labor Salary	8.8	Construction Period, Years	0	Slag & Ash Worth, \$/ST Dry	-10
Overheads & Benefits	3.1	Plant Life, Years	25	Recovered CO2 Worth, \$/ST	0
Insurances	6.6	Bank Loan Interest Rate, % per Year	10	CO2 Emission Penalty, \$/ST	0
Local Taxes	6.6	Bank Loan Repayment Period, Yrs	25	Annual Maint Labor+Mat, % TPC	0
Power Sales		Annual Bank Loan Repayment, \$MM/Yr	25	Employees, 1xAdmin + 4xOper	3
Sulfur Sales	91.1	Equival Depreciatn & Interest Expenses:	60.1	Avg Salaries, \$/Year	125
Ash & Slag Disposal	1.2	Bank Loan S/L Depreciation, \$MM/Yr	21.8	OYHDS & Benefits, % Labor	35
Waste Water Disposal	0.0	Bank Loan Interest, \$MM/Year	38.3	Construction Period, Years	0
CO2 Sequestn Credit	0.0	Owner's Investment ROI, % per Year	10	Land Cost, \$/Ft2	0
CO2 Emission Penalty	0.0	Owner's Invest Repayment Period, Yrs	25	Initial Oper Fund, Months	3
Bank Loan Repayment	60.1	Annual Capital Recovery, \$MM/Yr	15.0	Plant Life, Years	25
Investment Recovery	15.0	Equival Depreciatn & Interest Expenses:		Annual Onstream Factor, %	80
Hydrogen Sales	68.8	Equival S/L Depreciatn, \$MM/Yr	5.5	Owner's Investment, % TPC	20
Total Pre Income Taxes	161.3	Equival Loan Interest, \$MM/Yr	9.6	Owner Invest Repay Period, Yrs	25
Fed Income Tax (FIT)	(0.0)	Overall Ht Bal, MM(LHV)/Hr:		Owner's ROI, % per Year	10
Stat Income Tax (SIT)	(0.0)	In	Out	Bank Loan Interest Rate, %/Yr	10
Total Post Income Tax	161.3			Insurances, % TPC	1
				FIT, % Net Income	30
				SIT, % Net Income less FIT	10
				Local (Property) Taxes, % TPC	1
				Royalties:	
				Gasification, STPD MF Coal	0
				Sour Shift, \$/MMSCFD Feed	0
				AGR, \$/MMSCFD Acid Gas	0
				Sulfur Plt, \$/STPD Sulfur	0
				TGTU, \$/STPD Sulfur	0
				PSA, \$/MMSCFD H2	0
				Catalysts & Chemicals, % TIC:	1.7
				Memb Perf Fact, x 20ml/min/cm2:	2.5

PROCESS ECONOMIC EVALUATION SUMMARY:

Breakeven H2 Price, \$/1000 SCF	5.05
Brkeven H2 Price, \$/MMBtu(HHV)	15.58
TPC Payback Periods, Years	8.8
% Return On Invest (Excl Land)	11.02

CASE 4

\$26/ST Credit for Recovered CO₂

1S710G9Z = ITN Membrane H2 Recov w 0% Pd Bypass, None Shift, 56# resid CO2 PP, 0 MPH HRSG Firing, w Stm as NOx Diluent. (\$26/ST recov CO2 credit & \$0/ST CO2 emiss

PROCESS ECONOMIC EVALUATION SUMMARY:

Syngas Quenched Shifted		Treated Saturated Hydrogen	GT	HHV Basis		35.0 % HHV	Breakeven H2 Price, \$/1000 SCF	3.47
from Syngas to Syngas		Syngas Tail Gas Product	Nox	LHV Basis		34.9 % LHV	Brkeven H2 Price, \$/MMBtu (HHV)	10.71
Destec to Acid Gas to H2		to Power	Diluent	(HHV w S)		36.8 % HHV		
Gasifiers Shift Removal		Recovery Plant	Compres'n	(LHV w S)		35.7 % LHV	TPC Payback Periods, Years	8.8
				PRODUCTS:			% Return On Invest (Excl Land)	11.02
Stream Flows, MPH:								

Stream Flows, MPH:

[illegible]

Temperature, deg F
Pressure, psia

1,850	800	800	995	91	550	Local Taxes	6.6
465	415	405	340	5	290	Power Sale, \$58.7/MWH	62.6
						Sulfur Sale	1.4
						=====	
						PROCESS DATA:	
						=====	

	Cap Cost	Power Balance, MW
1	0.00	0.00
2	0.00	0.00
3	0.00	0.00
4	0.00	0.00
5	0.00	0.00
6	0.00	0.00
7	0.00	0.00
8	0.00	0.00
9	0.00	0.00
10	0.00	0.00
11	0.00	0.00
12	0.00	0.00
13	0.00	0.00
14	0.00	0.00
15	0.00	0.00
16	0.00	0.00
17	0.00	0.00
18	0.00	0.00
19	0.00	0.00
20	0.00	0.00
21	0.00	0.00
22	0.00	0.00
23	0.00	0.00
24	0.00	0.00
25	0.00	0.00
26	0.00	0.00
27	0.00	0.00
28	0.00	0.00
29	0.00	0.00
30	0.00	0.00
31	0.00	0.00
32	0.00	0.00
33	0.00	0.00
34	0.00	0.00
35	0.00	0.00
36	0.00	0.00
37	0.00	0.00
38	0.00	0.00
39	0.00	0.00
40	0.00	0.00
41	0.00	0.00
42	0.00	0.00
43	0.00	0.00
44	0.00	0.00
45	0.00	0.00
46	0.00	0.00
47	0.00	0.00
48	0.00	0.00
49	0.00	0.00
50	0.00	0.00
51	0.00	0.00
52	0.00	0.00
53	0.00	0.00
54	0.00	0.00
55	0.00	0.00
56	0.00	0.00
57	0.00	0.00
58	0.00	0.00
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67	0.00	0.00
68	0.00	0.00
69	0.00	0.00
70	0.00	0.00
71	0.00	0.00
72	0.00	0.00
73	0.00	0.00
74	0.00	0.00
75	0.00	0.00
76	0.00	0.00
77	0.00	0.00
78	0.00	0.00
79	0.00	0.00
80	0.00	0.00
81	0.00	0.00
82	0.00	0.00
83	0.00	0.00
84	0.00	0.00
85	0.00	0.00
86	0.00	0.00
87	0.00	0.00
88	0.00	0.00
89	0.00	0.00
90	0.00	0.00
91	0.00	0.00
92	0.00	0.00
93	0.00	0.00
94	0.00	0.00
95	0.00	0.00
96	0.00	0.00
97	0.00	0.00
98	0.00	0.00
99	0.00	0.00
100	0.00	0.00

[illegible]

COAL PROPERTIES:

Item	Value	Unit	Notes
Shift & LT Gas Cooling	4.1	(1)	
Acid Gas Removal & SWS	9.4	(1)	
Sulfur Recovery	20.5	(1)	
H2 Recovery (88487 Ft2 Membr)	14.0	(1)	
H2 Compression (1000 PSIG)	7.6	(1)	
COAL PROPERTIES:			
=====			
Type	Illin #6		
AGR: Type	CO2 Removal		
	0.1 %		
Sweet Gas CO2 PP	56.0 psia		
SRU: Type	Sulfuric Acid		
Analysis:			
	10.6		
Fed Income Tax (FIT)			(0.0)
Stat Income Tax (SIT)			(0.0)
Post Income Tax Total			161.3
			161.3

AG Fd H₂S Conc 93 vol%Dry CAPITAL RECOVERY BASIS:

[illegible]

Total Plant Cost (TPC)

850 F 1501 # SHHP Stm, Surf Cond @ 24.4 inHg & 116 F appr, 71 F CW, wo FG Sat.	H2 Prod HHV Btu/SCF = 324
660.0 Total Plant Cost (TPC)	GTG/6

Run ID: 1S710G9Z = ITN Membrane H2 Recov w 0% Fd Bypass, None Shift, 56# resid CO2 PP, 0 MPH HRSG Firing, w Stm as NOx Diluent.

INCOMES & EXPENSES:		INVESTMENT & CAPITAL RECOVERY:		Economic Analysis Basis:	
Expenses	Incomes				
\$MM/Yr	\$MM/Yr				
Coal Feed	30.2	Total Plant Capital Cost, \$MM	660.0	As Rec'd Coal Cost, \$/ST	29.4
Fluxant Feed	0.0	Initial Operating Capital, \$MM	21.5	Fluxant Cost, \$/ST Dry	10
Power Import	0.0	Land Cost @ \$0,000/Acre, \$MM	0.0	Import Power Cost, \$/MWH	58.679
Raw Water Import	0.8			Raw Water Cost, \$/1000 Gal	0.25
Cat & Chem Consumptn	9.2	Total Initial Investment, \$MM	681.6	Hydrogen Worth, \$/1000 SCF	To be Calc
Royalties	0	Owner Initial Investment, \$MM	136.3	Export Power Worth, \$/MWH	58.6794099
Mainten Labor & Mat	19.8	Bank Loan, \$MM	545.3	H2SO4 Worth, \$/ST S Eq	60
Admin & Labor Salary	8.8	Construction Period, Years	0	Slag & Ash Worth, \$/ST Dry	-10
Overheads & Benefits	3.1	Plant Life, Years	25	Waste H2O Worth, \$/1000 Gal	0
Insurances	6.6	Bank Loan Interest Rate, % per Year	10	Recovered CO2 Worth, \$/ST	26.038
Local Taxes	6.6	Bank Loan Repayment Period, Years	25	CO2 Emission Penalty, \$/ST	0
Power Sales	62.6	Annual Bank Loan Repayment, \$MM/Yr	60.1	Annual Maint Labor+Mat, % TPC	3
Sulfur Sales	1.4	Equivalent Depreciatn & Interest Expenses:		Employees, 1xAdmin + 4xOper	125
Ash & Slag Disposal	1.2	Bank Loan S/L Depreciation, \$MM/Yr	21.8	Avg Salaries, \$/Year	70,000
Waste Water Disposal	0.0	Bank Loan Interest, \$MM/Year	38.3	OYHDS & Benefits, % Labor	35
CO2 Sequestn Credit	0.0	Owner's Investment ROI, % per Year	10	Construction Period, Years	0
CO2 Emission Penalty	0.0	Owner's Invest Repayment Period, Yrs	25	Land Cost, \$/Ft2	0
Bank Loan Repayment	60.1	Annual Capital Recovery, \$MM/Yr	15.0	Initial Oper Fund, Months	3
Investment Recovery	15.0	Equivalent Depreciatn & Interest Expenses:		Plant Life, Years	25
		Equival Loan Interest, \$MM/Yr	5.5	Annual Onstream Factor, %	80
Hydrogen Sales	47.3		9.6	Owner's Investment, % TPC	20
Total Pre Income Taxes	161.3	Overall Ht Bal, MM(LHV)/Hr:		Owner Invest Repay Period, Yrs	25
Fed Income Tax (FIT)	(0.0)	In		Owner's ROI, % per Year	10
Stat Income Tax (SIT)	(0.0)	Out		Bank Loan Interest Rate, %/Yr	10
Total Post Income Tax	161.3			Insurances, % TPC	1
				FIT, % Net Income	30
				SIT, % Net Income less FIT	10
				Local (Property) Taxes, % TPC	1
				Royalties:	
				Gasification, STPD MF Coal	0
				Sour Shift, \$/MMSCFD Feed	0
				AGR, \$/MMSCFD Acid Gas	0
				Sulfur Plt, \$/STPD Sulfur	0
				TGTU, \$/STPD Sulfur	0
				PSA, \$/MMSCFD H2	0
				Catalysts & Chemicals, % TIC:	1.7
				Memb Perf Fact, x 20ml/min/cm2:	2.5

PROCESS ECONOMIC EVALUATION SUMMARY:

Breakeven H2 Price, \$/1000 SCF	3.47
Brkeven H2 Price, \$/MMBtu(HHV)	10.71
TPC Payback Periods, Years	8.8
% Return On Invest (Excl Land)	11.02

CASE 5

No Credit For Recovered CO₂

1XBP2G19 = PSA H2 Recov w 0% Fd Bypass, 2-Stg Sour Shift, 5# resid CO2 PP, w Boiler Plt. (\$0/ST recov CO2 credit & \$0/ST CO2 emission tax)

OVERALL (H2 + MWe) EFFICIENCIES: PROCESS ECONOMIC EVALUATION SUMMARY:									
1.	2.	3.	4.	5.	6.	7.	8.		
Syngas from Destec Gasifiers Shift	Quenched Syngas Shift	Shifted Syngas to Acid Gas Removal	Treated Syngas to H2 Recovery	Saturated Tail Gas to Power	Hydrogen Product to Compress'n	GT Nox Diluent		HHV Basis	54.7 % HHV
								LHV Basis	48.6 % LHV
								(HHV w S)	55.5 % HHV
								(LHV w S)	49.5 % LHV
PRODUCTS: (LHV w S)									
=====									
Stream Flows, MPH:								Hydrogen, MMSCFD	133.9
N2	81	81	73	73	71	1	0	Power Export, MWe	
O2	0	0	0	0	0	0	0	Sulfur, STPD	77
CO2	4,070	4,070	12,884	272	272	0	0	Slag & Ash, STPD Dry	403
Ar	26	26	23	23	23	0	0	Waste Water, GPM	1,014
H2	7,897	7,897	16,328	16,328	1,633	0	0	Recovered CO2, STPD	6,542
CO	11,050	11,050	724	724	724	0	0	CONSUMABLES:	
H2S	214	214	192	0	0	0	0	As Rec'd Coal, STPD	3,517
COS	11	11	10	0	0	0	0	Fluxant, STPD Dry	0
CH4	922	922	830	830	830	0	0	Power Import, MW	18
C2H6	0	0	0	0	0	0	0	Raw Water, GPM	2,157
Steam	6,084	6,580	87	0	0	0	0	99.5% O2, STPD	2,519
Total	30,354	30,850	31,151	18,250	3,554	14,697	0	PROCESS DATA:	
Temperature, deg F	1,850	317	100	95	176	95	550	Gasifier	
Pressure, psig	535	461	326	311	66	291	20	Air Separation	
=====									
Coal Handling/Feeding	33.8		(1)	1.7				Shift: Type	
Gasification	67.2		(1)	2.1				Overall Bypass	0.0 % Fd
Spare Gasifiers (Total 1 Spare)	19.2		(1)					1st Stg Bypass	50.0 % Fd
Air Separation	52.0		(1)					Stm/CO Ratio:	
O2 Compression (635 PSIG)				25.7				Syngas Feed	0.60
Shift & LT Gas Cooling	19.6		(1)	12.4				1st Stg Fd	5.00
Acid Gas Removal & SWS	55.4		(1)	8.5				1st Stg Prod	77.66
Sulfur Recovery	6.3		(1)	0.3				2nd Stg Fd	4.42
H2 Recovery	30.4		(1)	0.1				COAL PROPERTIES:	
H2 Compression (1000 PSIG)	4.0		(1)	8.6				Type	Illin #6
CO2 Dehydr'n & Comp (2000 PSIG)	32.0		(1)	19.2				Analysis:	As Rec'd
Fuel Gas Compression (50 PSIG)	3.2		(1)	2.6				C	59.87
N2 Compression (20 PSIG)	0.0		(1)	0.0				H	4.27
Gas Turb Generator System	0.0		(1)	0.0				N	1.11
HRSG/Boiler Plt/BFW/DM/Cond	14.5		(1)	2.1				S	2.21
Stream Turb Generator System	14.4		(1)	74.5				O	7.07
CW Pumps & CT Fans	10.1		(1)	3.9				Chlorine	0.11
Flue Gas CO2 Recovery	62.2		(1)	4.1				Ash	10.86
BOP								Moisture	14.50
Total Installed Cost (TIC)	424.1		Total	74.5				Total Wt%	100.00
Home Office @ 5.2 % of TIC	22.0			92.1				HHV, Btu/#	10,900
Contingency @ 15.0 % of TIC	63.6							LHV, Btu/#	10,295
Total Plant Cost (TPC)	509.7							% C Sequestered	85 %
								1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.	
								H2 Prod HHV Btu/SCF = 324	BLR/4

Run ID: 1XP2G19 = PSA H2 Recov w 0% Fd Bypass, 2-Std Sour Shift, 5# resid CO2 PP, w Boiler Pit.

INCOMES & EXPENSES:		Expenses		Incomes		INVESTMENT & CAPITAL RECOVERY:		Economic Analysis Basis:	
	\$MM/Yr	\$MM/Yr	\$MM/Yr		\$MM/Yr				
Coal Feed	30.2			Total Plant Capital Cost, \$MM	509.7	As Rec'd Coal Cost, \$/ST			29.4
Fluxant Feed	0.0			Initial Operating Capital, \$MM	20.3	Fluxant Cost, \$/ST Dry			10
Power Import	4.9			Land Cost @ \$0,000/Acre, \$MM	0.0	Import Power Cost, \$/MWH			40
Raw Water Import	0.2					Raw Water Cost, \$/1000 Gal			0.25
Cat & Chem Consumptn	7.2			Total Initial Investment, \$MM	530.0	Hydrogen Worth, \$/1000 SCF			To be Calc
Royalties	0			Owner Initial Investment, \$MM	106.0	Export Power Worth, \$/MWH			40
Mainten Labor & Mat	15.3			Bank Loan, \$MM	424.0	Sulfur Worth, \$/ST			30
Admin & Labor Salary	8.8			Construction Period, Years		Slag & Ash Worth, \$/ST Dry			-10
Overheads & Benefits	3.1			Plant Life, Years	25	Waste H2O Worth, \$/1000 Gal			0
Insurances	5.1			Bank Loan Interest Rate, % per Year	10	Recovered CO2 Worth, \$/ST			0
Local Taxes	5.1			Bank Loan Repayment Period, Years	25	CO2 Emission Penalty, \$/ST			0
Power Sales		0.0		Annual Bank Loan Repayment, \$MM/Yr	46.7	Annual Maint Labor+Mat, % TPC			3
Sulfur Sales		0.7		Equival Depreciatn & Interest Expenses:		Employees, 1xAdmin + 4xOper			125
Ash & Slag Disposal	1.2			Bank Loan S/L Depreciation, \$MM/Yr	17.0	Avg Salaries, \$/Year			70,000
Waste Water Disposal	0.0			Bank Loan Interest, \$MM/Year	29.7	OVHDS & Benefits, % Labor			35
CO2 Sequestn Credit	0.0			Owner's Investment ROI, % per Year	10	Construction Period, Years			0
CO2 Emission Penalty	0.0			Owner's Invest Repayment Period, Yrs	25	Land Cost, \$/Ft2			0
Bank Loan Repayment	46.7			Annual Capital Recovery, \$MM/Yr	11.7	Initial Oper Fund, Months			3
Investment Recovery	11.7			Equival Depreciatn & Interest Expenses:		Plant Life, Years			25
				Equival S/L Depreciatn, \$MM/Yr	4.2	Annual Onstream Factor, %			80
				Equival Loan Interest, \$MM/Yr	7.4	Owner's Investment, % TPC			20
						Owner Invest Repay Period, Yrs			25
						Owner's ROI, % per Year			10
						Bank Loan Interest Rate, %/Yr			10
Hydrogen Sales		138.8		Overall Ht Bal, MM(LHV)/Hr:		Insurances, % TPC			1
						FIT, % Net Income			30
Total Pre Income Taxes	139.4	139.4				SIT, % Net Income less FIT			10
Fed Income Tax (FIT)	0.0					Local (Property) Taxes, % TPC			1
Stat Income Tax (SIT)	0.0					Royalties:			
Total Post Income Taxe	139.4	139.4				Gasification, STPD MF Coal			0
						Sour Shift, \$/MMSCFD Feed			0
						AGR, \$/MMSCFD Acid Gas			0
						Sulfur Plt, \$/STPD Sulfur			0
						TGTU, \$/STPD Sulfur			0
						PSA, \$/MMSCFD H2			0
						Catalysts & Chemicals, % TIC:			1.7
						Memb Perf Fact, x 20ml/min/cm2:			2.5

PROCESS ECONOMIC EVALUATION SUMMARY:

Breakeven H2 Price, \$/1000 SCF	3.55
Breakeven H2 Price, \$/MMBtu(HHV)	10.96
TPC Payback Periods, Years	8.7
% Return On Invest (Excl Land)	11.02

CASE 5

\$26/ST Credit for Recovered CO₂

1XBP2G19 = PSA H2 Recov w 0% Fd Bypass, 2-Stg Sour Shift, 5# resid CO2 PP, w Boiler Plt. (\$26/ST recov CO2 credit & \$0/ST CO2 emission tax)

OVERALL (H2 + MWe) EFFICIENCIES: PROCESS ECONOMIC EVALUATION SUMMARY:									
1.	2.	3.	4.	5.	6.	7.	8.		
Syngas from Destec Gasifiers Shift	Quenched Syngas to Acid Gas Removal	Shifted Syngas to H2 to Power	Treated Syngas to H2 to Power	Saturated Syngas to H2 to Power	Hydrogen Product to Diluent	GT NOX			
81	81	73	73	71	1	0			
0	0	0	0	0	0	0			
4,070	4,070	12,884	272	272	0	0			
26	26	23	23	23	0	0			
7,897	7,897	16,328	16,328	1,633	14,695	0			
11,050	11,050	724	724	724	0	0			
214	214	192	0	0	0	0			
11	11	10	0	0	0	0			
922	922	830	830	830	0	0			
0	0	0	0	0	0	0			
6,084	6,580	87	0	0	0	0			
30,354	30,850	31,151	18,250	3,554	14,697	0			
Temperature, deg F	1,850	317	95	176	95	550			
Pressure, psig	535	461	326	66	291	20			
Stream Flows, MPH:									
N2	81	73	73	71	1	0			
O2	0	0	0	0	0	0			
CO2	4,070	12,884	272	272	0	0			
Ar	26	23	23	23	0	0			
H2	7,897	16,328	16,328	1,633	14,695	0			
CO	11,050	724	724	724	0	0			
H2S	214	192	0	0	0	0			
COS	11	10	0	0	0	0			
CH4	922	830	830	830	0	0			
C2H6	0	0	0	0	0	0			
Steam	6,084	6,580	87	0	0	0			
Total	30,354	30,850	31,151	18,250	3,554	14,697	0		
PROCESS DATA:									
Temperature, deg F	1,850	317	95	176	95	550			
Pressure, psig	535	461	326	66	291	20			
Cap Cost Summary (\$MM(2001))									
Coal Handling/Feeding	33.8	(1)	1.7	12.4	1.7	403.2			
Gasification	67.2	(1)	2.1	8.5	2.1	403.2			
Spare Gasifiers (Total 1 Spare)	19.2	(1)	---	0.3	---	403.2			
Air Separation	52.0	(1)	25.7	12.4	25.7	185.0			
O2 Compression (635 PSIG)	Incl ASU	---	---	---	---	---			
Shift & LT Gas Cooling	19.6	(1)	0.7	0.7	0.7	---			
Acid Gas Removal & SWS	55.4	(1)	8.5	8.5	8.5	---			
Sulfur Recovery	6.3	(1)	0.3	0.3	0.3	---			
H2 Recovery	30.4	(1)	0.1	0.1	0.1	---			
H2 Compression (1000 PSIG)	4.0	(1)	19.2	19.2	19.2	---			
CO2 Dehydr'n & Comp (2000 PSIG)	32.0	(1)	8.6	8.6	8.6	---			
Fuel Gas Compression (50 PSIG)	3.2	(1)	0.0	0.0	0.0	---			
N2 Compression (20 PSIG)	0.0	(1)	0.0	0.0	0.0	---			
Gas Turb Generator System	0.0	(1)	0.0	0.0	0.0	---			
HRSG/Boiler Plt/BFW/DW/Cond	14.5	(1)	2.1	2.1	2.1	---			
Steam Turb Generator System	14.4	(1)	74.5	74.5	74.5	---			
CW Pumps & CT Fans	10.1	(1)	3.9	3.9	3.9	---			
Flue Gas CO2 Recovery	---	---	---	---	---	---			
BOP	62.2	---	4.1	4.1	4.1	---			
Total Installed Cost (TIC)	424.1	Total	74.5	92.1	92.1	---			
Home Office @ 5.2 % of TIC	22.0								
Contingency @ 15.0 % of TIC	63.6								
Total Plant Cost (TPC)	509.7								
1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.									
H2 Prod HHV Btu/SCF = 324 BLR/4									

OVERALL (H2 + MWe) EFFICIENCIES: PROCESS ECONOMIC EVALUATION SUMMARY:									
1.	2.	3.	4.	5.	6.	7.	8.		
Syngas from Destec Gasifiers Shift	Quenched Syngas to Acid Gas Removal	Shifted Syngas to H2 to Power	Treated Syngas to H2 to Power	Saturated Syngas to H2 to Power	Hydrogen Product to Diluent	GT NOX			
81	81	73	73	71	1	0			
0	0	0	0	0	0	0			
4,070	4,070	12,884	272	272	0	0			
26	26	23	23	23	0	0			
7,897	7,897	16,328	16,328	1,633	14,695	0			
11,050	11,050	724	724	724	0	0			
214	214	192	0	0	0	0			
11	11	10	0	0	0	0			
922	922	830	830	830	0	0			
0	0	0	0	0	0	0			
6,084	6,580	87	0	0	0	0			
30,354	30,850	31,151	18,250	3,554	14,697	0			
Temperature, deg F	1,850	317	95	176	95	550			
Pressure, psig	535	461	326	66	291	20			
Cap Cost Summary (\$MM(2001))									
Coal Handling/Feeding	33.8	(1)	1.7	12.4	1.7	403.2			
Gasification	67.2	(1)	2.1	8.5	2.1	403.2			
Spare Gasifiers (Total 1 Spare)	19.2	(1)	---	0.3	---	403.2			
Air Separation	52.0	(1)	25.7	12.4	25.7	185.0			
O2 Compression (635 PSIG)	Incl ASU	---	---	---	---	---			
Shift & LT Gas Cooling	19.6	(1)	0.7	0.7	0.7	---			
Acid Gas Removal & SWS	55.4	(1)	8.5	8.5	8.5	---			
Sulfur Recovery	6.3	(1)	0.3	0.3	0.3	---			
H2 Recovery	30.4	(1)	0.1	0.1	0.1	---			
H2 Compression (1000 PSIG)	4.0	(1)	19.2	19.2	19.2	---			
CO2 Dehydr'n & Comp (2000 PSIG)	32.0	(1)	8.6	8.6	8.6	---			
Fuel Gas Compression (50 PSIG)	3.2	(1)	0.0	0.0	0.0	---			
N2 Compression (20 PSIG)	0.0	(1)	0.0	0.0	0.0	---			
Gas Turb Generator System	0.0	(1)	0.0	0.0	0.0	---			
HRSG/Boiler Plt/BFW/DW/Cond	14.5	(1)	2.1	2.1	2.1	---			
Steam Turb Generator System	14.4	(1)	74.5	74.5	74.5	---			
CW Pumps & CT Fans	10.1	(1)	3.9	3.9	3.9	---			
Flue Gas CO2 Recovery	---	---	---	---	---	---			
BOP	62.2	---	4.1	4.1	4.1	---			
Total Installed Cost (TIC)	424.1	Total	74.5	92.1	92.1	---			
Home Office @ 5.2 % of TIC	22.0								
Contingency @ 15.0 % of TIC	63.6								
Total Plant Cost (TPC)	509.7								
1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.									
H2 Prod HHV Btu/SCF = 324 BLR/4									

OVERALL (H2 + MWe) EFFICIENCIES: PROCESS ECONOMIC EVALUATION SUMMARY:									
1.	2.	3.	4.	5.	6.	7.	8.		
Syngas from Destec Gasifiers Shift	Quenched Syngas to Acid Gas Removal	Shifted Syngas to H2 to Power	Treated Syngas to H2 to Power	Saturated Syngas to H2 to Power	Hydrogen Product to Diluent	GT NOX			
81	81	73	73	71	1	0			
0	0	0	0	0	0	0			
4,070	4,070	12,884	272	272	0	0			
26	26	23	23	23	0	0			
7,897	7,897	16,328	16,328	1,633	14,695	0			
11,050	11,050	724	724	724	0	0			
214	214	192	0	0	0	0			
11	11	10	0	0	0	0			
922	922	830	830	830	0	0			
0	0	0	0	0	0	0			
6,084	6,580	87	0	0	0	0			
30,354	30,850	31,151	18,250	3,554	14,697	0			
Temperature, deg F	1,850	317	95	176	95	550			
Pressure, psig	535	461	326	66	291	20			
Cap Cost Summary (\$MM(2001))									
Coal Handling/Feeding	33.8	(1)	1.7	12.4	1.7	403.2			
Gasification	67.2	(1)	2.1	8.5	2.1	403.2			
Spare Gasifiers (Total 1 Spare)	19.2	(1)	---	0.3	---	403.2			
Air Separation	52.0	(1)	25.7	12.4	25.7	185.0			
O2 Compression (635 PSIG)	Incl ASU	---	---	---	---	---			
Shift & LT Gas Cooling	19.6	(1)	0.7	0.7	0.7	---			
Acid Gas Removal & SWS	55.4	(1)	8.5	8.5	8.5	---			
Sulfur Recovery	6.3	(1)	0.3	0.3	0.3	---			
H2 Recovery	30.4	(1)	0.1	0.1	0.1	---			
H2 Compression (1000 PSIG)	4.0	(1)	19.2	19.2	19.2	---			
CO2 Dehydr'n & Comp (2000 PSIG)	32.0	(1)	8.6	8.6	8.6	---			
Fuel Gas Compression (50 PSIG)	3.2	(1)	0.0	0.0	0.0	---			
N2 Compression (20 PSIG)	0.0	(1)	0.0	0.0	0.0	---			
Gas Turb Generator System	0.0	(1)	0.0	0.0	0.0	---			
HRSG/Boiler Plt/BFW/DW/Cond	14.5	(1)	2.1	2.1	2.1	---			
Steam Turb Generator System	14.4	(1)	74.5	74.5	74.5	---			
CW Pumps & CT Fans	10.1	(1)	3.9	3.9	3.9	---			
Flue Gas CO2 Recovery	---	---	---	---	---	---			
BOP	62.2	---	4.1	4.1	4.1	---			
Total Installed Cost (TIC)	424.1	Total	74.5	92.1	92.1	---			
Home Office @ 5.2 % of TIC	22.0								
Contingency @ 15.0 % of TIC	63.6								
Total Plant Cost (TPC)	509.7								
1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.									
H2 Prod HHV Btu/SCF = 324 BLR/4									

OVERALL (H2 + MWe) EFFICIENCIES: PROCESS ECONOMIC EVALUATION SUMMARY:									
1.	2.	3.	4.	5.	6.	7.	8.		
Syngas from Destec Gasifiers Shift	Quenched Syngas to Acid Gas Removal	Shifted Syngas to H2 to Power	Treated Syngas to H2 to Power	Saturated Syngas to H2 to Power	Hydrogen Product to Diluent	GT NOX			
81	81	73	73	71	1	0			
0	0	0	0	0	0	0			
4,070	4,070	12,884	272	272	0	0			
26	26	23	23	23	0	0			
7,897	7,897	16,328	16,328	1,633	14,695	0			
11,050	11,050	724	724	724	0	0			
214	214	192	0	0	0	0			
11	11	10	0	0	0	0			
922	922	830	830	830	0	0			
0	0	0	0	0	0	0			
6,084	6,580	87	0	0	0	0			
30,354	30,850	31,151	18,250	3,554	14,697	0			
Temperature, deg F	1,850	317	95	176	95	550			
Pressure, psig	535	461	326	66	291	20			
Cap Cost Summary (\$MM(2001))									
Coal Handling/Feeding	33.8	(1)	1.7	12.4	1.7	403.2			
Gasification	67.2	(1)	2.1	8.5	2.1	403.2			
Spare Gasifiers (Total 1 Spare)	19.2	(1)	---	0.3	---	403.2			
Air Separation	52.0	(1)	25.7	12.4	25.7	185.0			
O2 Compression (635 PSIG)	Incl ASU	---	---	---	---	---			
Shift & LT Gas Cooling	19.6	(1)	0.7	0.7	0.7	---			
Acid Gas Removal & SWS	55.4	(1)	8.5	8.5	8.5	---			
Sulfur Recovery	6.3	(1)	0.3	0.3	0.3	---			
H2 Recovery	30.4	(1)	0.1	0.1	0.1	---			
H2 Compression (1000 PSIG)	4.0	(1)	19.2	19.2	19.2	---			
CO2 Dehydr'n & Comp (2000 PSIG)	32.0	(1)	8.6	8.6	8.6	---			
Fuel Gas Compression (50 PSIG)	3.2	(1)	0.0	0.0	0.0	---			
N2 Compression (20 PSIG)	0.0	(1)	0.0	0.0	0.0	---			
Gas Turb Generator System	0.0	(1)	0.0	0.0	0.0	---			
HRSG/Boiler Plt/BFW/DW/Cond	14.5	(1)	2.1	2.1	2.1	---			
Steam Turb Generator System	14.4	(1)	74.5	74.5	74.5	---			
CW Pumps & CT Fans	10.1	(1)	3.9	3.9	3.9	---			
Flue Gas CO2 Recovery	---	---	---	---	---	---			
BOP	62.2	---	4.1	4.1	4.1	---			
Total Installed Cost (TIC)	424.1	Total	74.5	92.1	92.1	---			
Home Office @ 5.2 % of TIC	22.0								
Contingency @ 15.0 % of TIC	63.6								
Total Plant Cost (TPC)	509.7								
1000 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.									
H2 Prod HHV Btu/SCF = 324 BLR/4									

OVERALL (H2 + MWe) EFF									
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Run ID: 1XBP2G19 = PSA H2 Recov w 0% Pd Bypass, 2-Std Sour Shift, 5# resid CO2 PP, w Boiler Plt.

INCOMES & EXPENSES:		Expenses Incomes		INVESTMENT & CAPITAL RECOVERY:		Economic Analysis Basis:	
	\$MM/Yr	\$MM/Yr	\$MM/Yr				
Coal Feed	30.2			Total Plant Capital Cost, \$MM	509.7	As Rec'd Coal Cost, \$/ST	29.4
Fluxant Feed	0.0			Initial Operating Capital, \$MM	20.3	Fluxant Cost, \$/ST Dry	10
Power Import	4.9			Land Cost @ \$0,000/Acre, \$MM	0.0	Import Power Cost, \$/MMH	40
Raw Water Import	0.2					Raw Water Cost, \$/1000 Gal	0.25
Cat & Chem Consumptn	7.2			Total Initial Investment, \$MM	530.0	Hydrogen Worth, \$/1000 SCF	To be Calc
Royalties	0			Owner Initial Investment, \$MM	106.0	Export Power Worth, \$/MMH	40
Mainten Labor & Mat	15.3			Bank Loan, \$MM	424.0	Sulfur Worth, \$/ST	30
Admin & Labor Salary	8.8					Slag & Ash Worth, \$/ST Dry	-10
Overheads & Benefits	3.1			Construction Period, Years	0	Waste H2O Worth, \$/1000 Gal	0
Insurances	5.1			Plant Life, Years	25	Recovered CO2 Worth, \$/ST	26.038
Local Taxes	5.1					CO2 Emission Penalty, \$/ST	0
Power Sales		0.0		Bank Loan Interest Rate, % per Year	10	Annual Maint Labor+Mat, % TPC	0
Sulfur Sales		0.7		Bank Loan Repayment Period, Years	25	Employees, 1xAdmin + 4xOper	3
Ash & Slag Disposal	1.2			Annual Bank Loan Repayment, \$MM/Yr	46.7	Avg Salaries, \$/Year	125
Waste Water Disposal	0.0			Equival Depreciatn & Interest Expenses:		OVDHS & Benefits, % Labor	70,000
CO2 Sequestn Credit		49.7		Bank Loan S/L Depreciation, \$MM/Yr	17.0	Construction Period, Years	35
CO2 Emission Penalty	0.0			Bank Loan Interest, \$MM/Year	29.7	Land Cost, \$/Ft2	0
Bank Loan Repayment	46.7					Initial Oper Fund, Months	0
Investment Recovery	11.7			Owner's Investment ROI, % per Year	10	Plant Life, Years	3
				Owner's Invest Repayment Period, Yrs	25	Annual Onstream Factor, %	25
Hydrogen Sales		89.0		Annual Capital Recovery, \$MM/Yr	11.7	Owner's Investment, % TPC	80
				Equival Depreciatn & Interest Expenses:		Owner Invest Repay Period, Yrs	20
Total Pre Income Taxes	139.4	139.4		Equival S/L Depreciatn, \$MM/Yr	4.2	Owner's ROI, % per Year	25
Fed Income Tax (FIT)	0.0			Equival Loan Interest, \$MM/Yr	7.4	Bank Loan Interest Rate, %/Yr	10
Stat Income Tax (SIT)	0.0					Insurances, % TPC	1
				Overall Ht Bal, MM(LHV)/Hr:		FIT, % Net Income	30
Total Post Income Tax	139.4	139.4				SIT, % Net Income less FIT	10
						Local (Property) Taxes, % TPC	1

PROCESS ECONOMIC EVALUATION SUMMARY:

Breakeven H2 Price, \$/1000 SCF	2.28
Breakeven H2 Price, \$/MMBtu (HHV)	7.03
TPC Payback Periods, Years	8.7
% Return On Invest (Excl Land)	11.02

Royalties:

Gasification, STPD MF Coal	0
Sour Shift, \$/MMSCFD Feed	0
AGR, \$/MMSCFD Acid Gas	0
Sulfur Plt, \$/STPD Sulfur	0
TGTU, \$/STPD Sulfur	0
PSA, \$/MMSCFD H2	0

Catalysts & Chemicals, % TIC:	1.7
Memb Perf Fact, x 20ml/min/cm2:	2.5

CASE 6

No Credit For Recovered CO₂

1S612G19 = ITN Membrane H2 Recov w 0% Fd Bypass, 2-Stg Sour Shift, 5# resid CO2 PP, 170 MPH HRSG Firing, w Stm as NOx Diluent. (\$0/ST recov CO2 credit & \$0/ST CO2

OVERALL (H2 + MWe) EFFICIENCIES: PROCESS ECONOMIC EVALUATION SUMMARY:

1.	2.	3.	4.	5.	6.	7.	8.	OVERALL (H2 + MWe) EFFICIENCIES:	PROCESS ECONOMIC EVALUATION SUMMARY:
Syngas from Destec Gasifiers Shift	Quenched Syngas to Acid Gas Removal	Shifted Syngas to H2	Treated Syngas to H2	Saturated Tail Gas to Power	Hydrogen Product to Diluent	GT NOx Diluent			
81	81	73	73	70	0	0			
0	0	0	0	0	0	0			
4,070	4,070	12,884	264	296	0	0			
26	26	23	23	22	0	0			
7,897	7,897	16,327	16,327	1,077	14,706	0			
11,050	11,050	725	725	669	0	0			
214	214	192	0	0	0	0			
922	922	830	830	766	0	0			
0	0	0	0	0	0	0			
6,082	6,555	72	0	514	0	2,765			
30,352	30,825	31,136	18,242	3,416	14,708	2,765			
Temperature, deg F	1,850	318	95	1,066	91	550			
Pressure, psig	545	471	336	285	10	235			

Stream Flows, MPH:

1.	2.	3.	4.	5.	6.	7.	8.	OVERALL (H2 + MWe) EFFICIENCIES:	PROCESS ECONOMIC EVALUATION SUMMARY:
N2	81	73	73	70	0	0			
O2	0	0	0	0	0	0			
CO2	4,070	12,884	264	296	0	0			
Ar	26	23	23	22	0	0			
H2	7,897	16,327	16,327	1,077	14,706	0			
CO	11,050	725	725	669	0	0			
H2S	214	192	0	0	0	0			
COS	11	10	0	0	0	0			
CH4	922	830	830	766	0	0			
C2H6	0	0	0	0	0	0			
Steam	6,082	6,555	72	0	514	0			
Total	30,352	30,825	31,136	18,242	3,416	14,708	2,765		
Temperature, deg F	1,850	318	95	1,066	91	550			
Pressure, psig	545	471	336	285	10	235			

TPC Payback Periods, Years 8.7
% Return On Invest (Excl Land) 11.02

1.	2.	3.	4.	5.	6.	7.	8.	OVERALL (H2 + MWe) EFFICIENCIES:	PROCESS ECONOMIC EVALUATION SUMMARY:
Hydrogen, MMSCFD	134.0								
Power Export, MWe	0								
Sulfur, STPD	77								
Slag & Ash, STPD Dry	403								
Waste Water, GPM	977								
Recovered CO2, STPD	6,546								

CONSUMABLES:

1.	2.	3.	4.	5.	6.	7.	8.	OVERALL (H2 + MWe) EFFICIENCIES:	PROCESS ECONOMIC EVALUATION SUMMARY:
As Rec'd Coal, STPD	3,517								
Fluxant, STPD Dry	0								
Power Import, MW	26								
Raw Water, GPM	2,098								
99.5% O2, STPD	2,637								

PROCESS DATA:

1.	2.	3.	4.	5.	6.	7.	8.	OVERALL (H2 + MWe) EFFICIENCIES:	PROCESS ECONOMIC EVALUATION SUMMARY:
Gasifier	Destec								
Air Separation	Conventional								
Shift: Type	2-Stg Sour								
Overall Bypass	0.0 % Fd								
1st Stg Bypass	50.0 % Fd								
Stm/CO Ratio:									
Syngas Feed	0.59								
1st Stg Fd	5.00								
1st Stg Prod	77.66								
2nd Stg Fd	4.41								

Hydrogen Sales 141.3

Pre Income Tax Total 142.0

Fed Income Tax (FIT) 0.0

Stat Income Tax (SIT) 0.0

Post Income Tax Total 142.0

CAPITAL RECOVERY BASIS:

Total Plant Capital Cost, \$MM 511.0

Initial Operating Capital, \$MM 20.8

Land Cost @ \$0,000/Acre, \$MM 0.0

Total Initial Investment, \$MM 531.8

Plant Life, Years 25

Annual Onstream Factor, % 80

Owner's Investment, % Initial 20

Owner's ROI, % per Year 10

Bank Loan Interest Rate, %/Yr 10

H2 Prod HHV Btu/SCF = 324

GTG/299

950 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.

HHV, Btu/# 10,900

LHV, Btu/# 10,295

Stack CO2 Emission 1099 STPD

% C Sequestered 85 %

Power Plt: Type GE 6B

FG H2 Conc 37 vol% Dry

Stack Temp 260 deg F

1000 deg F

Run ID: 1S612G19 = ITN Membrane H2 Recov w 0% Fd Bypass, 2-Stg Sour Shift, 5# resid CO2 PP, 170 MPH HRSG Firing, w Stm as NOx Diluent.

INCOMES & EXPENSES:		Expenses Incomes		INVESTMENT & CAPITAL RECOVERY:		Economic Analysis Basis:	
	\$MM/Yr	\$MM/Yr	\$MM/Yr				
Coal Feed	30.2			Total Plant Capital Cost, \$MM	511.0	As Rec'd Coal Cost, \$/ST	29.4
Fluxant Feed	0.0			Initial Operating Capital, \$MM	20.8	Fluxant Cost, \$/ST Dry	10
Power Import	7.2			Land Cost @ \$0,000/Acre, \$MM	0.0	Import Power Cost, \$/MMH	40
Raw Water Import	0.2					Raw Water Cost, \$/1000 Gal	0.25
Cat & Chem Consumptn	7.2			Total Initial Investment, \$MM	531.8	Hydrogen Worth, \$/1000 SCF	To be Calc
Royalties	0			Owner Initial Investment, \$MM	106.4	Export Power Worth, \$/MMH	40
Mainten Labor & Mat	15.3			Bank Loan, \$MM	425.4	Sulfur Worth, \$/ST	30
Admin & Labor Salary	8.8					Slag & Ash Worth, \$/1000 Gal	-10
Overheads & Benefits	3.1			Construction Period, Years	0	Waste H2O Worth, \$/ST	0
Insurances	5.1			Plant Life, Years	25	Recovered CO2 Worth, \$/ST	0
Local Taxes	5.1					CO2 Emission Penalty, \$/ST	0
Power Sales		0.0		Bank Loan Interest Rate, % per Year	10	Annual Maint Labor+Mat, % TPC	0
Sulfur Sales		0.7		Bank Loan Repayment Period, Years	25	Employees, 1xAdmin + 4xOper	3
Ash & Slag Disposal	1.2			Annual Bank Loan Repayment, \$MM/Yr	46.9	Avg Salaries, \$/Year	125
Waste Water Disposal	0.0			Equival Depreciatn & Interest Expenses:		OVHDS & Benefits, % Labor	35
CO2 Sequestn Credit		0.0		Bank Loan S/L Depreciation, \$MM/Yr	17.0	Construction Period, Years	0
CO2 Emission Penalty	0.0			Bank Loan Interest, \$MM/Year	29.9	Land Cost, \$/Ft2	0
Bank Loan Repayment	46.9					Initial Oper Fund, Months	3
Investment Recovery	11.7			Owner's Investment ROI, % per Year	10	Plant Life, Years	25
				Owner's Invest Repayment Period, Yrs	25	Annual Onstream Factor, %	80
Hydrogen Sales		141.3		Annual Capital Recovery, \$MM/Yr	11.7	Owner's Investment, % TPC	20
Total Pre Income Taxes	142.0	142.0		Equival Depreciatn & Interest Expenses:		Owner Invest Repay Period, Yrs	25
Fed Income Tax (FIT)	0.0			Equival S/L Depreciatn, \$MM/Yr	4.3	Owner's ROI, % per Year	10
Stat Income Tax (SIT)	0.0			Equival Loan Interest, \$MM/Yr	7.5	Bank Loan Interest Rate, %/Yr	10
Total Post Income Taxe	142.0	142.0				Insurances, % TPC	1
				Overall Ht Bal, MM(LHV)/Hr:	In Out	FIT, % Net Income	30
						SIT, % Net Income less FIT	10
						Local (Property) Taxes, % TPC	1
PROCESS ECONOMIC EVALUATION SUMMARY:						Royalties:	
Breakeven H2 Price, \$/1000 SCF	3.61					Gasification, STPD MF Coal	0
Brkeven H2 Price, \$/MMBtu (HHV)	11.14					Sour Shift, \$/MMSCFD Feed	0
						AGR, \$/MMSCFD Acid Gas	0
TPC Payback Periods, Years	8.7					Sulfur Plt, \$/STPD Sulfur	0
% Return On Invest (Excl Land)	11.02					TGTU, \$/STPD Sulfur	0
						PSA, \$/MMSCFD H2	0
						Catalysts & Chemicals, % TIC:	1.7
						Memb Perf Fact, x 20ml/min/cm2:	2.5

CASE 6

\$26/ST Credit for Recovered CO₂

15612G19 = ITN Membrane H2 Recov w 0% Fd Bypass, 2-Stdg Sour Shift, 5# resid CO2 PP, 170 MPH HRSG Firing, w Stm as NOX Diluent. (\$26/ST recov CO2 credit & \$0/ST CO

OVERALL (H2 + MWe) EFFICIENCIES: PROCESS ECONOMIC EVALUATION SUMMARY:

Stream Flows, MPH:	1.	2.	3.	4.	5.	6.	7.	8.	OVERALL (H2 + MWe) EFFICIENCIES:	TPC Payback Periods, Years	% Return On Invest (Excl Land)
N2	81	81	73	73	70	0	0	0	HHV Basis	8.7	11.02
O2	0	0	0	0	0	0	0	0	53.9 % HHV		
CO2	4,070	4,070	12,884	264	296	0	0	0	Brakeven H2 Price, \$/1000 SCF	2.34	
Ar	26	26	23	23	22	0	0	0	47.7 % LHV	7.22	
H2	7,897	7,897	16,327	16,327	1,077	14,706	0	0	Brakeven H2 Price, \$/MMBtu (HHV)		
CO	11,050	11,050	725	725	669	0	0	0	(HHV w S)		
H2S	214	214	192	0	0	0	0	0	54.7 % HHV		
COS	11	11	10	0	0	0	0	0	48.6 % LHV		
CH4	922	922	830	830	766	0	0	0	TPC Payback Periods, Years	8.7	
C2H6	0	0	0	0	0	0	0	0	% Return On Invest (Excl Land)	11.02	
Steam	6,082	6,555	72	0	514	0	2,765	0			
Total	30,352	30,825	31,136	18,242	3,416	14,708	2,765				

INCOMES & EXPENSES:	INCOMES & EXPENSES:	INCOMES & EXPENSES:	INCOMES & EXPENSES:	INCOMES & EXPENSES:	INCOMES & EXPENSES:	INCOMES & EXPENSES:	INCOMES & EXPENSES:	INCOMES & EXPENSES:	INCOMES & EXPENSES:	INCOMES & EXPENSES:	INCOMES & EXPENSES:
AR Coal @ \$29.4/ST	30.2	AR Coal @ \$29.4/ST	30.2	AR Coal @ \$29.4/ST	30.2	AR Coal @ \$29.4/ST	30.2	AR Coal @ \$29.4/ST	30.2	AR Coal @ \$29.4/ST	30.2
Fluxant Feed	0.0	Fluxant Feed	0.0	Fluxant Feed	0.0	Fluxant Feed	0.0	Fluxant Feed	0.0	Fluxant Feed	0.0
Pwr Import, \$40.0/MMH	7.2	Pwr Import, \$40.0/MMH	7.2	Pwr Import, \$40.0/MMH	7.2	Pwr Import, \$40.0/MMH	7.2	Pwr Import, \$40.0/MMH	7.2	Pwr Import, \$40.0/MMH	7.2
Raw Water Import	0.2	Raw Water Import	0.2	Raw Water Import	0.2	Raw Water Import	0.2	Raw Water Import	0.2	Raw Water Import	0.2
Cat & Chem Consumptn	7.2	Cat & Chem Consumptn	7.2	Cat & Chem Consumptn	7.2	Cat & Chem Consumptn	7.2	Cat & Chem Consumptn	7.2	Cat & Chem Consumptn	7.2
Royalties	0.0	Royalties	0.0	Royalties	0.0	Royalties	0.0	Royalties	0.0	Royalties	0.0
Mainten Labor & Mat	15.3	Mainten Labor & Mat	15.3	Mainten Labor & Mat	15.3	Mainten Labor & Mat	15.3	Mainten Labor & Mat	15.3	Mainten Labor & Mat	15.3
Admin & Labor Salary	8.8	Admin & Labor Salary	8.8	Admin & Labor Salary	8.8	Admin & Labor Salary	8.8	Admin & Labor Salary	8.8	Admin & Labor Salary	8.8
Overheads & Benefits	3.1	Overheads & Benefits	3.1	Overheads & Benefits	3.1	Overheads & Benefits	3.1	Overheads & Benefits	3.1	Overheads & Benefits	3.1
Insurances	5.1	Insurances	5.1	Insurances	5.1	Insurances	5.1	Insurances	5.1	Insurances	5.1
Local Taxes	5.1	Local Taxes	5.1	Local Taxes	5.1	Local Taxes	5.1	Local Taxes	5.1	Local Taxes	5.1
Power Sale, \$40.0/MMH	0.0	Power Sale, \$40.0/MMH	0.0	Power Sale, \$40.0/MMH	0.0	Power Sale, \$40.0/MMH	0.0	Power Sale, \$40.0/MMH	0.0	Power Sale, \$40.0/MMH	0.0
Sulfur Sale	0.7	Sulfur Sale	0.7	Sulfur Sale	0.7	Sulfur Sale	0.7	Sulfur Sale	0.7	Sulfur Sale	0.7

EXPENSES	EXPENSES	EXPENSES	EXPENSES	EXPENSES	EXPENSES	EXPENSES	EXPENSES	EXPENSES	EXPENSES	EXPENSES	EXPENSES
Ash & Slag Disposal	1.2	Ash & Slag Disposal	1.2	Ash & Slag Disposal	1.2	Ash & Slag Disposal	1.2	Ash & Slag Disposal	1.2	Ash & Slag Disposal	1.2
Waste Water Disposal	0.0	Waste Water Disposal	0.0	Waste Water Disposal	0.0	Waste Water Disposal	0.0	Waste Water Disposal	0.0	Waste Water Disposal	0.0
CO2 Sequestn Credit	0.0	CO2 Sequestn Credit	0.0	CO2 Sequestn Credit	0.0	CO2 Sequestn Credit	0.0	CO2 Sequestn Credit	0.0	CO2 Sequestn Credit	0.0
CO2 Emission Penalty	46.9	CO2 Emission Penalty	46.9	CO2 Emission Penalty	46.9	CO2 Emission Penalty	46.9	CO2 Emission Penalty	46.9	CO2 Emission Penalty	46.9
Bank Loan Repayment	11.7	Bank Loan Repayment	11.7	Bank Loan Repayment	11.7	Bank Loan Repayment	11.7	Bank Loan Repayment	11.7	Bank Loan Repayment	11.7
Investment Recovery	0.0	Investment Recovery	0.0	Investment Recovery	0.0	Investment Recovery	0.0	Investment Recovery	0.0	Investment Recovery	0.0

CONSUMABLES:	CONSUMABLES:	CONSUMABLES:	CONSUMABLES:	CONSUMABLES:	CONSUMABLES:	CONSUMABLES:	CONSUMABLES:	CONSUMABLES:	CONSUMABLES:	CONSUMABLES:	CONSUMABLES:
As Rec'd Coal,STPD	3,517	As Rec'd Coal,STPD	3,517	As Rec'd Coal,STPD	3,517	As Rec'd Coal,STPD	3,517	As Rec'd Coal,STPD	3,517	As Rec'd Coal,STPD	3,517
Fluxant, STPD Dry	0	Fluxant, STPD Dry	0	Fluxant, STPD Dry	0	Fluxant, STPD Dry	0	Fluxant, STPD Dry	0	Fluxant, STPD Dry	0
Power Import, MW	26	Power Import, MW	26	Power Import, MW	26	Power Import, MW	26	Power Import, MW	26	Power Import, MW	26
Raw Water, GPM	2,098	Raw Water, GPM	2,098	Raw Water, GPM	2,098	Raw Water, GPM	2,098	Raw Water, GPM	2,098	Raw Water, GPM	2,098
99.5% O2, STPD	2,637	99.5% O2, STPD	2,637	99.5% O2, STPD	2,637	99.5% O2, STPD	2,637	99.5% O2, STPD	2,637	99.5% O2, STPD	2,637

PROCESS DATA:	PROCESS DATA:	PROCESS DATA:	PROCESS DATA:	PROCESS DATA:	PROCESS DATA:	PROCESS DATA:	PROCESS DATA:	PROCESS DATA:	PROCESS DATA:	PROCESS DATA:	PROCESS DATA:
Gasifier	Destec	Gasifier	Destec	Gasifier	Destec	Gasifier	Destec	Gasifier	Destec	Gasifier	Destec
Air Separation	Conventional	Air Separation	Conventional	Air Separation	Conventional	Air Separation	Conventional	Air Separation	Conventional	Air Separation	Conventional

Shift: Type	Shift: Type	Shift: Type	Shift: Type	Shift: Type	Shift: Type	Shift: Type	Shift: Type	Shift: Type	Shift: Type	Shift: Type	Shift: Type
Overall Bypass	0.0 % Fd	Overall Bypass	0.0 % Fd	Overall Bypass	0.0 % Fd	Overall Bypass	0.0 % Fd	Overall Bypass	0.0 % Fd	Overall Bypass	0.0 % Fd
1st Stg Bypass	50.0 % Fd	1st Stg Bypass	50.0 % Fd	1st Stg Bypass	50.0 % Fd	1st Stg Bypass	50.0 % Fd	1st Stg Bypass	50.0 % Fd	1st Stg Bypass	50.0 % Fd
Stm/CO Ratio:	0.59	Stm/CO Ratio:	0.59	Stm/CO Ratio:	0.59	Stm/CO Ratio:	0.59	Stm/CO Ratio:	0.59	Stm/CO Ratio:	0.59
Syngas Feed	352.9	Syngas Feed	352.9	Syngas Feed	352.9	Syngas Feed	352.9	Syngas Feed	352.9	Syngas Feed	352.9
1st Stg Fd	5.00	1st Stg Fd	5.00	1st Stg Fd	5.00	1st Stg Fd	5.00	1st Stg Fd	5.00	1st Stg Fd	5.00
1st Stg Prod	77.66	1st Stg Prod	77.66	1st Stg Prod	77.66	1st Stg Prod	77.66	1st Stg Prod	77.66	1st Stg Prod	77.66
2nd Stg Fd	4.41	2nd Stg Fd	4.41	2nd Stg Fd	4.41	2nd Stg Fd	4.41	2nd Stg Fd	4.41	2nd Stg Fd	4.41

COAL PROPERTIES:	COAL PROPERTIES:	COAL PROPERTIES:	COAL PROPERTIES:	COAL PROPERTIES:	COAL PROPERTIES:	COAL PROPERTIES:	COAL PROPERTIES:	COAL PROPERTIES:	COAL PROPERTIES:	COAL PROPERTIES:	COAL PROPERTIES:
Type	Illin #6	Type	Illin #6	Type	Illin #6	Type	Illin #6	Type	Illin #6	Type	Illin #6
Analysis:	As Rec'd	Analysis:	As Rec'd	Analysis:	As Rec'd	Analysis:	As Rec'd	Analysis:	As Rec'd	Analysis:	As Rec'd
C	59.87	C	59.87	C	59.87	C	59.87	C	59.87	C	59.87
H	4.27	H	4.27	H	4.27	H	4.27	H	4.27	H	4.27
N	1.11	N	1.11	N	1.11	N	1.11	N	1.11	N	1.11
S	2.21	S	2.21	S	2.21	S	2.21	S	2.21	S	2.21
O	7.07	O	7.07	O	7.07	O	7.07	O	7.07	O	7.07
Chlorine	0.11	Chlorine	0.11	Chlorine	0.11	Chlorine	0.11	Chlorine	0.11	Chlorine	0.11
Ash	10.86	Ash	10.86	Ash	10.86	Ash	10.86	Ash	10.86	Ash	10.86
Moisture	14.50	Moisture	14.50	Moisture	14.50	Moisture	14.50	Moisture	14.50	Moisture	14.50
Total Wt%	100.00	Total Wt%	100.00	Total Wt%	100.00	Total Wt%	100.00	Total Wt%	100.00	Total Wt%	100.00

AGR: Type	AGR: Type	AGR: Type	AGR: Type	AGR: Type	AGR: Type	AGR: Type	AGR: Type	AGR: Type	AGR: Type	AGR: Type	AGR: Type
CO2 Removal	98.0 %	CO2 Removal	98.0 %	CO2 Removal	98.0 %	CO2 Removal	98.0 %	CO2 Removal	98.0 %	CO2 Removal	98.0 %
Sweet Gas CO2 PP	4.9 psia	Sweet Gas CO2 PP	4.9 psia	Sweet Gas CO2 PP	4.9 psia	Sweet Gas CO2 PP	4.9 psia	Sweet Gas CO2 PP	4.9 psia	Sweet Gas CO2 PP	4.9 psia
SRU: Type	Claus	SRU: Type	Claus	SRU: Type	Claus	SRU: Type	Claus	SRU: Type	Claus	SRU: Type	Claus
AG Fd H2S Conc	45 vol% Dry	AG Fd H2S Conc	45 vol% Dry	AG Fd H2S Conc	45 vol% Dry	AG Fd H2S Conc	45 vol% Dry	AG Fd H2S Conc	45 vol% Dry	AG Fd H2S Conc	45 vol% Dry

ITN Membra	ITN Membra	ITN Membra	ITN Membra	ITN Membra	ITN Membra	ITN Membra	ITN Membra	ITN Membra	ITN Membra	ITN Membra	ITN Membra
FD Bypass	0.0 %	FD Bypass	0.0 %	FD Bypass	0.0 %	FD Bypass	0.0 %	FD Bypass	0.0 %	FD Bypass	0.0 %
Min H2 PP Dif	65.0 psi	Min H2 PP Dif	65.0 psi	Min H2 PP Dif	65.0 psi	Min H2 PP Dif	65.0 psi	Min H2 PP Dif	65.0 psi	Min H2 PP Dif	65.0 psi
Net H2 Recov	90.1 %	Net H2 Recov	90.1 %	Net H2 Recov	90.1 %	Net H2 Recov	90.1 %	Net H2 Recov	90.1 %	Net H2 Recov	90.1 %

Power Plt: Type	Power Plt: Type	Power Plt: Type	Power Plt: Type	Power Plt: Type	Power Plt: Type	Power Plt: Type	Power Plt: Type	Power Plt: Type	Power Plt: Type	Power Plt: Type	Power Plt: Type
EGT	1000 deg F	EGT	1000 deg F	EGT	1000 deg F	EGT	1000 deg F	EGT	1000 deg F	EGT	1000 deg F
FG H2 Conc	37 vol% Dry	FG H2 Conc	37 vol% Dry	FG H2 Conc	37 vol% Dry	FG H2 Conc	37 vol% Dry	FG H2 Conc	37 vol% Dry	FG H2 Conc	37 vol% Dry
Stack Temp	260 deg F	Stack Temp	260 deg F	Stack Temp	260 deg F	Stack Temp	260 deg F	Stack Temp	260 deg F	Stack Temp	260 deg F
Stack CO2 Emission	1099 STPD	Stack CO2 Emission	1099 STPD	Stack CO2 Emission	1099 STPD	Stack CO2 Emission	1099 STPD	Stack CO2 Emission	1099 STPD	Stack CO2 Emission	1099 STPD
% C Sequestered	85 %	% C Sequestered	85 %	% C Sequestered	85 %	% C Sequestered	85 %	% C Sequestered	85 %	% C Sequestered	85 %

HHV,Btu/#	HHV,Btu/#	HHV,Btu/#	HHV,Btu/#	HHV,Btu/#	HHV,Btu/#	HHV,Btu/#	HHV,Btu/#	HHV,Btu/#	HHV,Btu/#	HHV,Btu/#	HHV,Btu/#
10,900	10,900	10,900	10,900	10,900	10,900	10,900	10,900	10,900	10,900	10,900	10,900
Stack CO2 Emission	1099 STPD	Stack CO2 Emission	1099 STPD	Stack CO2 Emission	1099 STPD	Stack CO2 Emission	1099 STPD	Stack CO2 Emission	1099 STPD	Stack CO2 Emission	1099 STPD
% C Sequestered	85 %	% C Sequestered	85 %	% C Sequestered	85 %	% C Sequestered	85 %	% C Sequestered	85 %	% C Sequestered	85 %

950 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.	950 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.	950 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.	950 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.	950 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.	950 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.	950 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.	950 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.	950 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.	950 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.	950 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.	950 F 1501 # SHHP Stm, Surf Cond @ 1.5 inHg & 6 F appr, 71 F CW, wo FG Sat.
84.7	84.7	84.7	84.7	84.7	84.7	84.7	84.7	84.7	84.7	84.7	84.7
110.4	110.4	110.4	110.4	110.4	110.4	110.4	110.4	110.4	110.4	110.4	110.4

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Run ID: 1S612G19 = 1TN Membrane H2 Recov w 0% Pd Bypass, 2-Std Sour Shift, 5# resid CO2 PP, 170 MPH HRSG Firing, w Stm as Nox Diluent.

INCOMES & EXPENSES:		Expenses		Incomes		INVESTMENT & CAPITAL RECOVERY:		Economic Analysis Basis:	
	\$MM/Yr	\$MM/Yr	\$MM/Yr		\$MM/Yr				
Coal Feed	30.2			Total Plant Capital Cost, \$MM	511.0	As Rec'd Coal Cost, \$/ST			29.4
Fluxant Feed	0.0			Initial Operating Capital, \$MM	20.8	Fluxant Cost, \$/ST Dry			10
Power Import	7.2			Land Cost @ \$0,000/Acre, \$MM	0.0	Import Power Cost, \$/MWH			40
Raw Water Import	0.2					Raw Water Cost, \$/1000 Gal			0.25
Cat & Chem Consump	7.2			Total Initial Investment, \$MM	531.8				
Royalties	0			Owner Initial Investment, \$MM	106.4	Hydrogen Worth, \$/1000 SCF		To be Calc	
Mainten Labor & Mat	15.3			Bank Loan, \$MM	425.4	Export Power Worth, \$/MWH			40
Admin & Labor Salary	8.8					Sulfur Worth, \$/ST			30
Overheads & Benefits	3.1			Construction Period, Years	0	Slag & Ash Worth, \$/1000 Gal			-10
Insurances	5.1			Plant Life, Years	25	Waste H2O Worth, \$/ST			0
Local Taxes	5.1					Recovered CO2 Worth, \$/ST			26.038
Power Sales		0.0		Bank Loan Interest Rate, % per Year	10	CO2 Emission Penalty, \$/ST			0
Sulfur Sales		0.7		Bank Loan Repayment Period, Years	25	Annual Maint Labor+Mat, % TPC			3
Ash & Slag Disposal	1.2			Annual Bank Loan Repayment, \$MM/Yr	46.9	Employees, 1xAdmin + 4xOper			125
Waste Water Disposal	0.0			Equival Depreciatn & Interest Expenses:		Avg Salaries, \$/Year			70,000
CO2 Sequestn Credit		49.8		Bank Loan S/L Depreciation, \$MM/Yr	17.0	OVRDs & Benefits, % Labor			35
CO2 Emission Penalty	0.0			Bank Loan Interest, \$MM/Year	29.9				
Bank Loan Repayment	46.9					Construction Period, Years			0
Investment Recovery	11.7			Owner's Investment ROI, % per Year	10	Land Cost, \$/Ft2			0
				Owner's Invest Repayment Period, Yrs	25	Initial Oper Fund, Months			3
Hydrogen Sales		91.5		Annual Capital Recovery, \$MM/Yr	11.7	Plant Life, Years			25
				Equival Depreciatn & Interest Expenses:		Annual Onstream Factor, %			80
Total Pre Income Taxes	142.0	142.0		Equival S/L Depreciatn, \$MM/Yr	4.3	Owner's Investment, % TPC			20
				Equival Loan Interest, \$MM/Yr	7.5	Owner Invest Repay Period, Yrs			25
Fed Income Tax (FIT)	(0.0)					Owner's ROI, % per Year			10
Stat Income Tax (SIT)	(0.0)					Bank Loan Interest Rate, %/Yr			10
				Overall Ht Bal, MM(LHV)/Hr:					
Total Post Income Tax	142.0	142.0		In	Out	Insurances, % TPC			1
						FIT, % Net Income			30
						SIT, % Net Income less FIT			10
						Local (Property) Taxes, % TPC			1
						Royalties:			
						Gasification, STPD MF Coal			0
						Sour Shift, \$/MMSCFD Feed			0
						AGR, \$/MMSCFD Acid Gas			0
						Sulfur Plt, \$/STPD Sulfur			0
						TGTU, \$/STPD Sulfur			0
						PSA, \$/MMSCFD H2			0
						Catalysts & Chemicals, % TIC:			1.7
						Memb Perf Fact, x 20ml/min/cm2:			2.5

PROCESS ECONOMIC EVALUATION SUMMARY:

Breakeven H2 Price, \$/1000 SCF	2.34
Breakeven H2 Price, \$/MMBtu(HHV)	7.22
TPC Payback Periods, Years	8.7
% Return On Invest (Excl Land)	11.02