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INTEGRATION OF AIR SEPARATION UNIT WITH H₂ SEPARATION MEMBRANE REACTOR IN COAL-BASED POWER PLANT

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ABSTRACT

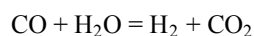
A novel process configuration consisting of integrating the air separation unit with a H₂ separation membrane reactor (HSMR) in a coal gasification based coproduction facility with near zero emissions is described. The plant utilizes an air separation unit operating at elevated pressure to produce an Intermediate Pressure (IP) N₂ stream in addition to the O₂ required by the coal gasifier. The syngas produced by the gasifier after cleanup is supplied to the membrane reactor which produces H₂ by shifting the carbon monoxide while simultaneously separating the H₂. The IP N₂ is used as sweep gas to assist in the separation of the H₂ diffusing across the membrane walls by decreasing the partial pressure of the H₂ on the permeate side. The total pressure of gases on the permeate side may thus be increased such that the H₂ / N₂ mixture may be fed directly to the gas turbines at the required pressure without requiring cooling and compression of the H₂. An added advantage is that the total pressure differential across the membrane wall is reduced. The N₂ in the fuel gas functions both as a thermal diluent for reducing the formation of nitrogen oxides and as additional motive fluid for expansion in the turbine. The carbon dioxide rich gas (non-permeate) leaving the membrane reactor after catalytic oxidation of the residual combustibles constitutes the carbon capture stream which may be further compressed and pipelined for CO₂ sequestration. High purity H₂ may be coproduced for export from a portion of the H₂-N₂ stream leaving the HSMR utilizing a Pressure Swing Adsorption (PSA) unit. The techno-economic advantages of such a coproduction facility are addressed.

INTRODUCTION

Due to the projected increases in fossil fuel usage world wide, emissions of CO₂ to the atmosphere are expected to increase by about 60% over the 1990 level by 2015. CO₂ is the primary constituent in the earth's atmosphere that contributes to the greenhouse effect. The greenhouse effect is the entrapment of heat by the earth's atmosphere by gases such as CO₂; the sun's radiation falling on the earth's surface is re-radiated as infrared heat which is absorbed by the greenhouse gases. It should be noted that the CO₂ generated from a given fuel per unit of power produced is inversely proportional to the thermal efficiency of a power plant for a given fuel conversion emphasizing the importance of plant thermal efficiency. In addition to CO₂, pollutants such as oxides of sulfur, oxides of nitrogen, CO, unburned hydrocarbons and Hg are introduced into the atmosphere when traditional power generation technologies relying on combustion are used. The amount of pollutants emitted to the atmosphere depends on the degree of pollution abatement measures incorporated. These pollution abatement measures, however, increase the plant operating and capital costs.

The FutureGen plant concept [Der, 2003] is aimed at reducing the environmental impacts of fossil fuel usage by generating electric power with near zero emissions while co-producing a carbon free fuel (H₂). This paper compares the thermal performance, environmental signature and economics of an advanced FutureGen type plant consisting of the HSMR coupled with Warm Gas Cleanup with a state-of-the-art Integrated Gasification Combined Cycle (IGCC) based FutureGen type plant defined as the Base Case. Both Warm Gas Cleanup and HSMRs are currently under development; the HSMRs utilizing high temperature membranes and containing

a suitable catalyst to promote the water gas shift reaction:



Two major type of membranes being developed are the microporous inorganic membranes [Judkins and Bischoff, 2004], and Pd-Cu alloy membranes [Enick et. al., 2003] which are especially suitable for producing a high purity H_2 stream but tend to be significantly more expensive. The projected cost for the microporous inorganic membranes is $\sim \$1,000/\text{m}^2$ [Longanbach et. al., 2002] while that of the Pd-Cu alloy membranes is $\sim \$3,000/\text{m}^2$ [Chiesa, Kreutz and Lozza, 2005]. Another advantage of the microporous membrane is that the H_2 flux is directly proportional to the partial pressure gradient across the membrane wall and not the square root difference as is the case of the Pd membranes [Judkins and Bischoff, 2004].

BASE CASE DEFINITION

The scheme for controlling the carbon emissions for the state-of-the-art IGCC plant consists of the following steps: (1) shift the raw syngas leaving the particulate scrubber utilizing a sour shift catalyst after preheating to the required temperature and (2) desulfurize the syngas while removing the CO_2 in a Selexol® Acid Gas Removal Unit after recovering the heat from the raw syngas [Rao and Stobbs, 2003].

Metal carbonyls that may be present in the raw syngas, such as those of Ni and Fe, deposit as Ni sulfide at elevated temperatures (such as those in the shift reactors) in the presence of a catalyst in the top layers of the first-stage shift reactor catalyst bed. It has been found from operating plant experience that the top 0.5 meters (1 to 2 ft) of the shift catalyst needs to be replaced approximately every two years due to increased pressure drop caused by the sulfide deposition. The impact on the annual operating cost of replacing the top section of the bed at a greater frequency (2 years instead of the normal 3 years) is expected not to have a significant effect on the overall economics of the plant.

Hg present in the coal typically volatilizes within the gasifier and leaves the gasifier along with the raw gas. Compounds of Hg can cause damage to organs. Sulfided activated carbon has been used to remove mercury (and arsenic) from coal derived syngas at the Tennessee Eastman gasification plant [Rutkowski, Klett and Maxwell, 2002]. Calgon offers this type of activated carbon for removal of mercury, reducing its concentration to as low as 0.01 to 0.1 $\mu\text{g}/\text{Nm}^3$ Hg in the syngas depending on the operating temperature and moisture content. Hg is captured predominantly as a sulfide, but some of it is captured in its elemental form. The spent carbon has to be disposed of as a hazardous waste although attempts are being made to recover elemental Hg as a saleable product assuming

a market does exist for it. Hg capture by sulfided carbon beds is unaffected by pressure of the syngas. The capture efficiency is reduced, however, as the operating temperature is increased and as the relative humidity of the syngas is increased.

Any metal (Ni and Fe) carbonyls that may remain in the syngas after passing through the shift reactors would be captured by the sulfided activated carbon bed.

DESIGN BASIS

The coal utilized in this study consists of the bituminous Illinois No. 6 coal. Site ambient conditions correspond to ISO conditions while the plant heat rejection is accomplished utilizing mechanical draft cooling towers. It is assumed that fresh makeup water is available. The gas turbines consist of "F class" technology. These gas turbines that employ an air cooled turbine are commercially offered for syngas applications. The firing (rotor inlet) temperature of these gas turbines in syngas applications is approximately 1260°C (2300°F). A condensing reheat steam turbine is fed with superheated steam at 106 bar, 538°C (1530 psia, 1000°F). Reheat steam conditions at the steam turbine inlet are 26.5 bar, 538°C (384 psia, 1000°F) while the condenser is operated at 0.0508 bar (0.737 psia). The CO_2 is exported at a pressure of 13,800 kPa (2,015 psia) while the H_2 is exported at a pressure of 2,400 kPa (350 psia). The minimum overall carbon capture as defined below is 90%:

Overall Carbon Capture = Carbon Captured from Syngas / Carbon Entering with Coal.

Producing a sour CO_2 stream for sequestration, i.e., bulk acid gas removal while letting the sulfur compounds leave the plant with the CO_2 is not considered in this study. Reduction in both the plant heat rate and cost may be realized but the pipeline will be classified as being in sour service resulting in an increase in its cost while the acceptability of the sour CO_2 by the end user and the safety associated with the high H_2S content would be issues.

Twenty percent of the syngas is diverted to the H_2 recovery unit in the Selexol® based case while 80% percent of this H_2 is recovered for export. This same H_2 export amount on a per tonne of coal feed basis is utilized for the HSMR based case.

The plant simulations were developed utilizing the Advanced Power Systems Analysis Tool [Rao and Samuelsen, 2002].

BASE CASE PLANT DESCRIPTION

An overall block flow sketch for this Selexol® based FutureGen type plant is presented in Figure 1

while the major stream data is presented in Table 1. A slurry fed high pressure entrained bed gasifier such as that offered by General Electric is utilized. The gasification process is a proven, non-catalytic partial oxidation process, in which hydrocarbonaceous materials in the coal react with a controlled amount of O_2 under high pressure and temperature conditions. The Air Separation Unit (ASU) operating at elevated pressure supplies 95% purity O_2 to gasifiers and produces N_2 at an intermediate pressure for injection into the gas turbines as a thermal diluent for NO_x control. By operating the ASU at an elevated pressure, the temperature at which the cryogenic separation of air occurs is increased leading to a more efficient IGCC plant as long as the O_2 and N_2 streams produced during the air separation may be utilized at or above the pressure at which these streams leave the cryogenic unit.

The gasifiers operate at a nominal pressure of 7,350 kPa (1066 psia). The coal is wet ground in rod mills to form a slurry and introduced into the gasifiers. The gasifiers partially oxidize the coal with the O_2 to generate a hot raw gas (syngas), slag and char. The raw syngas leaving the gasification zone is cooled by direct contact with water (Total Quench Design). This method of cooling the gas eliminates the need for syngas coolers that tend to be expensive; the quenching operation introduces water vapor into the syngas which is required for reacting with the CO present in the syngas to form H_2 and CO_2 within the shift reactors. The gas leaving the quench section of the gasifier is wet scrubbed to remove any entrained solids. The other contaminants such as soluble alkali salts, hydrogen halides and a small fraction of the NH_3 are also removed. The contaminated water (black water) is treated to remove the fine slag, and the remaining water (gray water) is treated before discharge.

The scrubbed gas enters the shift unit where most of the CO present in the gas is reacted with water vapor contained in the gas (no steam addition is required) to produce H_2 and CO_2 . Heat generated by the exothermic shift reaction is recovered by the generation of High Pressure (HP) and Medium Pressure (MP) steam. The shifted gas leaving the shift unit is further cooled in a series of heat exchangers consisting of Low Pressure (LP) steam generation, syngas humidifier circulating water heater, vacuum condensate heater and a trim cooler against cooling water. The cooled syngas is superheated by about 11 C (20 F) to avoid pore condensation and then fed to a sulfided activated carbon bed which removes the Hg. The effluent gas is then cooled against cooling water and fed into the Selexol® Acid Gas Removal Unit. The high temperature condensate separated from the gas is recycled to the scrubber while the colder condensate is fed to a sour water stripper. The sour gases stripped

off from the water are routed to the Claus Sulfur Recovery Unit.

The Selexol® process scheme consists of a four column design and utilizes solvent refrigeration to produce the clean syngas, acid gas which is supplied to the Sulfur Recovery Unit and CO_2 streams at three different pressures. About 50% of the CO_2 is produced at a pressure of 1,034 kPa (150 psia), about 40% is at a pressure of 345 kPa (50 psia) and the remainder at a pressure of 115 kPa (16.7 psia) and provided to the CO_2 compression system. The CO_2 is compressed to the mixture supercritical pressure, dehydrated and is then pumped to the pipeline pressure of 13,800 kPa (2,015 psia). The tail gas leaving the Claus unit which contains CO_2 and residual sulfur compounds as well as elemental sulfur vapor, is hydrogenated to form H_2S from the sulfur species. The hydrogenated tail gas is then recycled back to the Selexol® unit.

A portion of the clean syngas leaving the Selexol® unit with ultra low sulfur content is treated in the PSA unit to produce high purity H_2 for export. The PSA produces a product H_2 stream at pressure (about 2,400 kPa or 350 psia) and a tail gas stream which consists of the remaining fuel gas components mostly CO, CO_2 , H_2 , and N_2 . The remainder of the clean syngas along with the compressed PSA tail gas consisting mostly of H_2 with some CO and inerts, is humidified in a counter-current column by direct contact with hot water such that the humidified gas has a heating value of 7,870 kJ/Nm³ (200 BTU/SCF) on a LHV basis. The heat required by the humidifier is recovered from the low temperature heat contained in the shifted gas (downstream of the shift unit in low temperature gas cooling).

The humid gas is preheated and then fed to the gas turbines consisting of two units. IP N_2 from the ASU after preheating is also fed to the gas turbines but through separate nozzles within the combustors. The introduction of the N_2 and the moisture (via the syngas) reduces the formation of NO_x within the combustor of the gas turbine while the humidification of the syngas allows for efficient recovery of the low temperature heat from the plant.

The exhaust gas from the two gas turbines is fed to triple pressure heat recovery steam generators (HRSGs) which provide the superheated steam to a single condensing steam turbine. A Selective Catalytic Reduction (SCR) unit may be included in each of the HRSGs to further reduce the NO_x emissions. Note that the ultra low sulfur content of the fuel gas to the gas turbines makes it feasible to utilize the SCR without being constrained by the formation of ammonium salts. The plant includes the necessary general facilities such as a cooling water system, instrument air, flare, etc.

HSMR CASE PLANT DESCRIPTION

An overall block flow sketch for this HSMR based FutureGen type plant is presented in Figure 2 while the major stream data is presented in Table 2. The plant consists of the following major process units:

- ASU which is similar to the previous Selexol based case except that the amount of air supplied by the gas turbines is lower while the flow rate of N_2 provided at pressure by the ASU is higher (less air extraction is required to fully load the gas turbines, while more N_2 is utilized as the diluent in the gas turbine fuel since syngas humidification is not utilized)
- Gasification Unit except that each of the gasifiers is equipped with a radiant syngas cooler
- Warm Gas Cleanup Unit (eliminates the generation of low temperature heat and consequently syngas humidification)
- Sulfuric Acid Unit to convert the SO_2 produced by the sulfur capture unit during the regeneration step
- HSMR unit utilizing IP N_2 supplied by the ASU as sweep gas while producing the fuel gas for the gas turbines and feed gas for a PSA unit
- PSA unit recovers high purity H_2 for export while the tail gas is compressed and combined with fuel gas from the HSMR and supplied to the gas turbines
- CO_2 Compression Unit
- Power Block (similar to the previous Selexol based case except that the amount of air extracted from the gas turbines is lower) while the gas turbines combust a fuel gas that is a mixture of H_2 and N_2 (and small concentrations of the other components present in the clean syngas that may leak to the permeate side within the HSMR).

The necessary general facilities as described in the Selexol based case are also included. The following provides highlights of the plant that are unique to this case.

The plant configuration developed consists of high pressure O_2 blown entrained bed gasifiers using radiant coolers followed by spray cooling to $333^\circ C$ (the sprays will have to be properly designed to avoid splattering of liquid water on any surfaces; atomization of the water should be beneficial in this regard). The O_2 is supplied by an elevated pressure cryogenic ASU which also produces IP N_2 for utilization in the HSMR. The shifting reaction is driven within the HSMR by constantly separating the H_2 as it is formed through the membrane wall. Thus, lesser amount of water vapor is required in the syngas prior to shifting as compared to the previous Selexol® based case (Stream 5 in Tables 3 and 4)

making the radiant syngas coolers (which produce HP steam) suitable [Chiesa, Kreutz and Lozza, 2005]. The gasifier type (high-pressure entrained-bed) and its operating conditions as well as the gas turbine model are kept the same as in the previous case in order to better understand the advantages of the HSMR integrated with the ASU and warm gas cleanup.

The gas is then processed in the Warm Gas Cleanup unit before it is supplied to the HSMR to produce the decarbonized fuel gas. The syngas leaving the radiant syngas cooler at $760^\circ C$ is cooled to $333^\circ C$ by partially quenching the gas with water. It then goes to a barrier filter where over 99.99% of the particulates entrained in the syngas are removed. Next the syngas is treated in a nahcolite bed, which removes chlorides as well as the other halides. This is followed by another barrier filter after which it is treated in a transport desulfurizer containing ZnO . The ZnO is converted to ZnS which is then regenerated using air extracted from the gas turbine. During the regeneration operation, the sulfur is released as SO_2 from which the saleable product H_2SO_4 is made.

Warm gas Hg removal processes are being developed and one such process is that being developed by ADA technologies (funded by the EPA and the DOE) that operates around 300° to $400^\circ C$ [Butz 2003] and uses a fixed bed reactor containing an Amended Silicates™ sorbent where the Hg is chemisorbed from the syngas. An added advantage of Warm Gas Cleanup is that it operates outside the temperature window where Ni and Fe carbonyls form.

These above described Warm Gas Cleanup technologies have been tested at pilot scale levels but have not been commercially demonstrated while the HSMR technology itself is in much earlier stages of development. It is expected that the Warm Gas Cleanup technologies will be available for commercial offering by the time the HSMR technology is made available for large-scale applications.

The cleaned gas is next pre-shifted before being supplied to the HSMR consisting of the microporous inorganic membrane. Pre-shifting reduces the shifting load on the HSMR [De Lorenzo, Kreutz and Chiesa and Williams, 2005]. As depicted in Figure 3, IP N_2 produced by the ASU is used as the sweep gas in the HSMR (flowing on the permeate side counter-currently to the feed syngas flowing on the other side of the membrane wall) to assist in the separation of the H_2 diffusing across the membrane wall. The corresponding H_2 partial pressure on the permeate side varies from 0 to a maximum of 1216 kPa (176.4 psia). The H_2 - N_2 mixture leaving the HSMR is supplied to the gas turbines as fuel gas. The IP N_2 as

sweep gas assists in the separation of the H₂ by decreasing the partial pressure of the H₂ on the permeate side. The total pressure of gases on the permeate side may thus be increased such that the H₂-N₂ mixture can be fed directly to the gas turbines at the required pressure without requiring costly cooling, compression and preheating. An added advantage is that the total pressure differential across the membrane wall is reduced. The N₂ in the fuel gas functions both as a thermal diluent for NO_x control and as additional motive fluid for expansion in the turbine.

Microporous membranes typically have lower selectivity than the Pd membranes for H₂ separation and small fractions of the other gas components present in the syngas are expected to be also transferred across the membrane depending on the operating temperature and the membrane pore size. Since the purity requirement for the export H₂ needs to be of "PSA quality," a fraction of the H₂-N₂ mixture is sent to the PSA unit for recovering high purity H₂ for export while the tail gas leaving the PSA unit containing the N₂ along with un-recovered H₂ is compressed and then combined with the remainder of the fuel gas (H₂-N₂ mixture) and supplied to the gas turbines. As in the previous Selexol case, the high purity H₂ leaving the PSA unit is at about 2,400 kPa (essentially at the same pressure as the inlet gas to the PSA) and no compression of this stream is required before it is supplied to the pipeline.

The CO₂ rich non-permeate gas from the HSMR contains residual amounts of combustibles which are oxidized in a catalytic combustor using O₂, cooled while recovering the heat and then compressed to the export pressure required for pipelining the CO₂ for sequestration. Note that the CO₂ rich stream leaving the HSMR is at high pressure (4,100kPa) and so the compression power is minimized.

RESULTS AND DISCUSSIONS

Performance summaries for the two cases are presented in Table 3. Both plants consume the same amount of coal (5,660 tonne/d or 6,240 ST/D on an "as received" basis) while exporting 1.42 x 10⁶ nM³/d (53.1 MM SCFD) of H₂ (which is equivalent to about 12% of the coal bound energy on an HHV basis). The Selexol® based FutureGen type plant produces 496 MW of net electric power while the HSMRbased FutureGen type plant produces 537 MW of net electric power which is as much as 8% higher than the Selexol® case. The overall plant efficacy¹

¹ Efficacy = (Energy contained in Exported H₂ on HHV basis + Net Electric Power) / (Energy Contained in Coal on HHV basis)

[McGurl, 2005] for the four column Selexol® based case is 39.8% while that for the HSMRbased case is 42.1% (it should be noted that the efficacy difference between the two cases is dampened by the inclusion of the HHV energy of the exported H₂). An overall thermal efficiency may also be calculated by adding to the net electrical energy generated by the plants, the electric equivalent of the energy contained in the exported H₂ utilizing an efficiency of 60% on an LHV basis. The corresponding overall plant thermal efficiencies are then 33.9% and 36.2% for the Selexol® and the HSMRbased cases, a more that 6% decrease in heat rate for the HSMR case. The degree of CO₂ capture for the Selexol® based cases is 91% while that for the HSMR based case is expected to be greater than 95% (<100% however, since small fractions of the carbon species in the syngas diffuse across the membrane walls). The capture and pressurization of the CO₂ has a significant penalty on the plant thermal efficiency especially for the Selexol® based case; the heat rate and the cost of electricity are typically increased by more than 20%.

Table 3: Performance Summaries

Case	Selexol Based	HSMR Based
Coal Feed Rate (as Received), MT / D	5,660	
MWt (HHV)	1,778	
Gas Turbine Power, MW	420	420
Steam Turbine Power, MW	249	286
Total Gross Power, MW	669	706
Internal Power Consumption, MW	173	169
Net Electric Power, MW	496	537
H ₂ Exported, MWt (HHV)	211	
% of Coal HHV	11.86	
1000 Nm ³ / D	1426	
% of Coal Carbon Captured	91	>95
Efficacy, % HHV	39.8	42.1
Efficiency, % HHV	33.9	36.2

Next, rough order of magnitude installed plant cost estimates are developed for the two cases. The cost of the Selexol® unit is estimated using ICARUS, an ASPEN Suite product. The geometry of HSMR unit is assumed to be similar to that of a tubular reactor (tubes packed with granular shift catalyst) or a shell and tube heat exchanger with the feed syngas flowing inside the tubes constructed out of the microporous inorganic membrane material supported by a porous metal structure while the sweep gas along with the permeate flow on the shell side. The cost of this unit is developed by adding the fabrication and other material costs of \$1,500/m² [De Lorenzo, Kreutz and Chiesa and Williams, 2005] to the cost of the microporous inorganic membrane material (as stated earlier, projected at ~ \$1,000/m²).

Such an inorganic membrane is under development at the Oakridge National Laboratory and consists of composite structures of a macroporous support tube (pore diameter > 50 nm, usually 2 to 5 μm) coated with one or more membrane layers. The layers are about 2 μm thick and for H_2 separation, the separative layer pore size is typically < 1 nm [Judkins, and Bischoff, 2004]. A permeance of 1.1×10^{-6} kg $\text{H}_2/\text{s}/\text{m}^2/\text{kPa}$ [Longanbach et. al., 2002] is utilized in estimating the size of the HSMR. The HSMR is modeled on a thermodynamic basis by dividing the HSMR in the axial direction into a series of alternating shift reactors and membrane units. The impact of having small amounts of CO , CO_2 and CH_4 in the " H_2 - N_2 mixture" leaving the HSMR decreases the amount of carbon capture somewhat (the simulations were performed assuming that only H_2 is transferred across the microporous membrane). It is important to operate the HSMR at high temperatures while the pore diameter of the membrane should be minimized to increase the H_2 selectivity. The permeance of lighter molecules such as H_2 increases more rapidly with temperature than do the permeances of the heavier molecules such as CO and CO_2 . For example, at a temperature of 600K, the permeance of H_2 is almost 10 times that of CO_2 while below 400K, the two permeances are similar based data shown in plots by Judkins and Bischoff [Judkins and Bischoff, 2004]. Decrease in pore diameter has a significant effect on selectivity; for example the H_2 - CO_2 separation factor increases exponentially below a pore diameter of 0.6 nm using the Hard Sphere Transport Model [Judkins and Bischoff, 2004].

The estimated total installed cost for the Selexol based plant is \$949 million and that for the HSMR based case is \$1,013 million which is about 7% higher while the amount of power generated is almost 8% higher than that of the Selexol® based case (see Table 4).

An effective cost of electricity is calculated by adding to the net electrical energy generated by the plants, the electric equivalent of the energy contained in the exported H_2 utilizing the efficiency of 60% on an LHV basis. The 10th year effective levelized cost of electricity utilizing the Electric Power Research Institute's methodologies [Applegren and Vejtasa, 1982] and the basis summarized in Table 5 result in the following: the HSMR based case has a cost of electricity of \$47.81/MWh and is about 1.5% lower than that of the Selexol based case which is at \$48.56/MWh. It should be noted that the advantage of the HSMR based case will be higher at higher coal prices since the HSMR based case is significantly more efficient for electric power generation.

Table 4 shows that the major cost component in the case of the HSMR based plant is the gasification section due to the radiant syngas coolers. A

significant reduction in plant cost may be realized by eliminating these coolers, i.e., by utilizing the same quench design as employed in the Selexol case. Some efficiency will be sacrificed and the trade-off between loss of efficiency and savings in capital cost is the subject of a future study.

Table 4: Plant Cost Summaries

Plant Section	Total Plant Cost (\$1,000, Year 2005 Basis)	
	Selexol Based	HSMR Based
Air Separation Unit	105,101	105,101
Coal Receiving, Grinding & Slurry Prep	29,238	29,238
Gasification	168,693	245,524
Low Temperature Gas Cooling, Shift & Humidification	61,348	-
Warm Gas Cleanup & Pre-Shift	-	98,001
Acid Gas Removal	74,397	-
Sulfur Recovery & Tail Gas Recycle	30,017	-
HSMR	-	63,105
PSA & Tail Gas Compression	7,457	10,156
Power Block	191,882	199,494
CO_2 Pressurization & Dehydration	30,526	12,525
General Facilities	250,138	249,595
TOTAL PLANT COST	948,797	1,012,739

Table 5: Basis for Economic Analysis

Coal Price (Illinois No. 6)	\$1.07/GJ (\$1.13/MMBtu)
Project Book and Tax Life	20 Years
Escalation Rates	
Coal Price	0.42%/y
Total Plant Cost	3%/y
General	3%/y
Property Taxes	0.1% of Escalated Plant Cost
Insurance	0.7% of Escalated Plant Cost
Federal + State Income Tax	38%
Project Financing	
Common Equity	65%
Debt	35%
Maximum Annual Capacity Factor	85%

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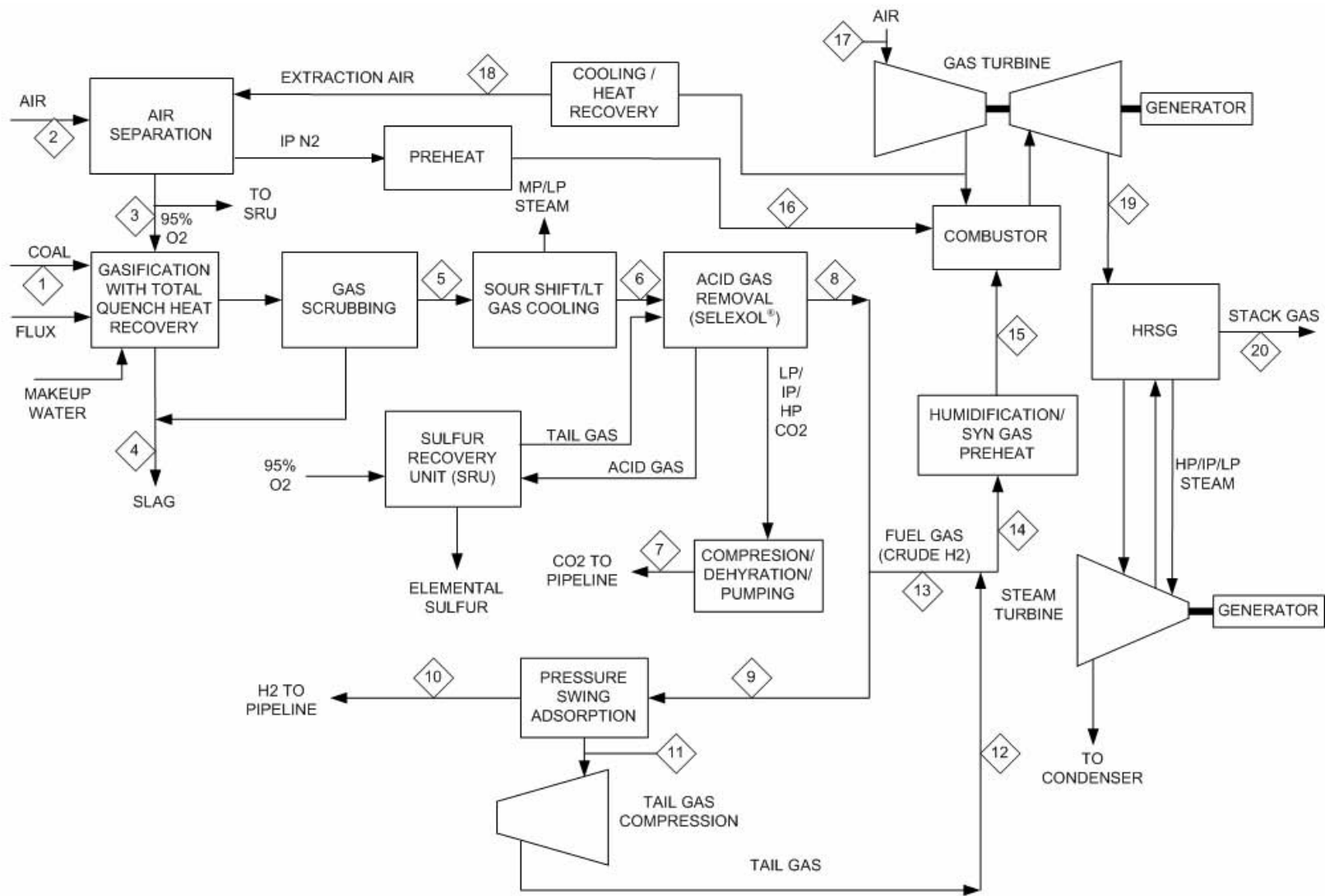


Figure 1: Overall Block Flow Diagram - Near Zero Emission / H₂ Coproduction with Selexol®

Table 1: Stream Data - Near Zero Emission / H₂ Coproduction with Selexol®

Stream No.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
H2					15.46	56.57	0.92	93.3	93.3	96.9	80.1	80.12	93.3	92.5	74.5					
H2O		0.99			57.28	0.06		0.01	0.01		0.03	0.03	0.01	0.01	19.5		0.99	0.3	15.3	15.3
N2		77.34	1.76		0.39	0.62	0.01	1.04	1.04	1.5			1.04	0.98	0.79	98.8	77.3	77.9	72.9	72.9
NH3					0.11															
O2		20.74	95													0.7	20.7	20.9	10.2	10.2
CO					20.8	1.41	0.11	2.29	2.29		9.85	9.85	2.29	2.71	2.18					
CO2					5.09	40.01	98.92	2.25	2.25		9.65	9.65	2.25	2.65	2.14				0.68	0.68
H2S + COS					0.39	0.62														
CH4					0.03	0.05		0.08	0.08		0.35	0.35	0.08	0.1	0.08					
Ar		0.93	3.24		0.41	0.65	0.03	1.09	1.09	1.57			1.09	1.03	0.83	0.5	0.93	0.94	0.91	0.91
Solids, kg/s	65.52			7.19																
Total, kg/s	65.52	177.5	53.15	19.34	251.6	163.2	139.8	21.4	4.28	2.29	1.99	1.99	17.1	19.1	37.3	124	797	45.9	913	913
Temperature, °C	15	15	115	50	243	27	45	16	16	16	12	144	16	24	288	288	15	34	583	160
Pressure, kPa	101	101	9317	101	6723	6266	13890	2960	2962	2717	148	2951	2960	2950	2860	3089	101	1820	105	101

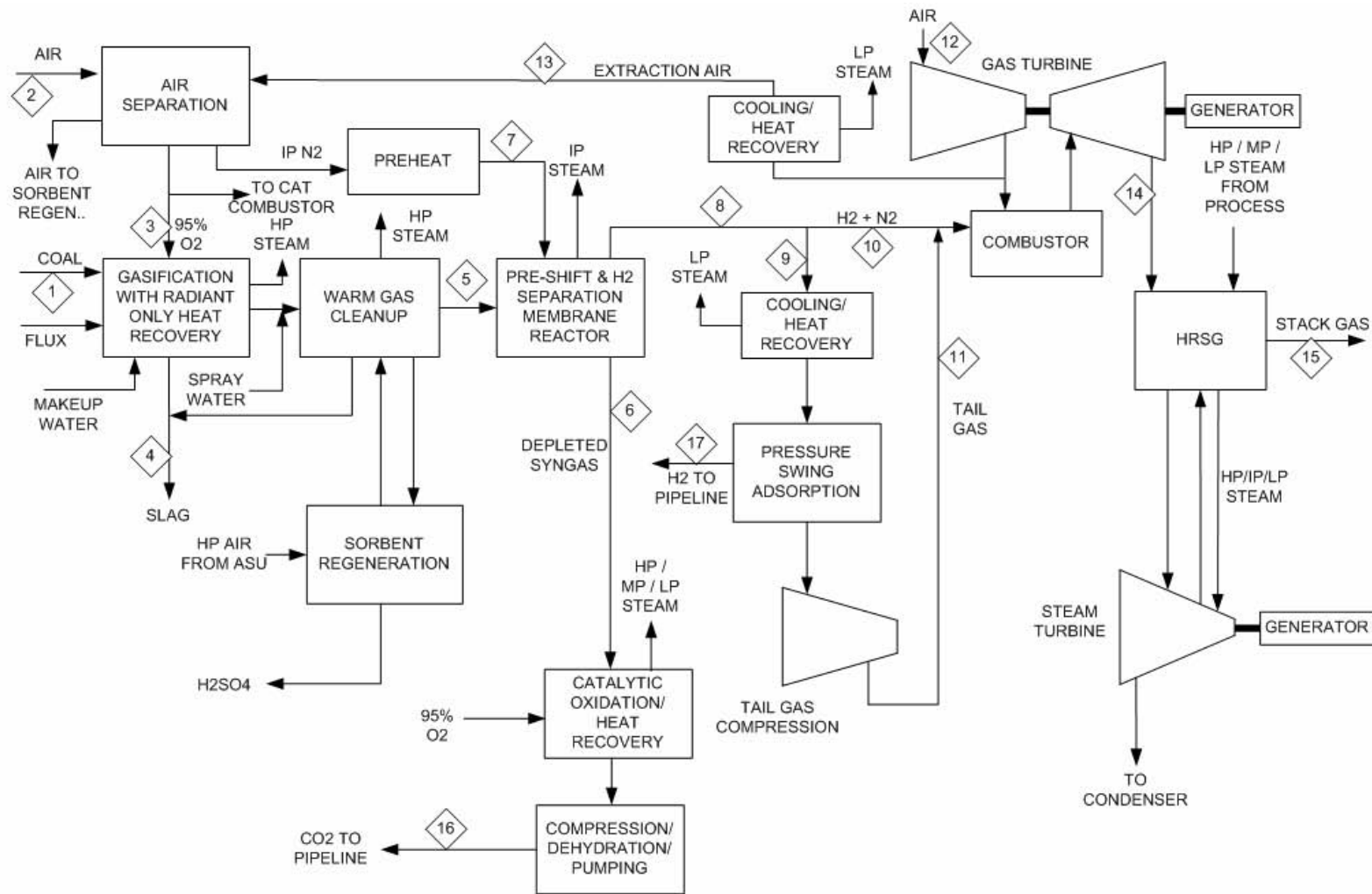


Figure 2: Overall Block Flow Diagram - Near Zero Emission / H₂ Coproduction with H₂ Separation Membrane Reactor

Table 2: Stream Data - Near Zero Emission / H₂ Coproduction with H₂ Separation Membrane Reactor

Stream No.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
H2					20.67	1.68		46.27	46.27	46.27	14.69					0.25	100
H2O		0.99			43.65	30.35						0.99	0.3	12.69	12.69		
N2		77.34	1.76		0.52	0.99	98.8	53.08	53.08	53.08	84.28	77.34	77.88	76.26	76.26	1.47	
O2		20.74	95				0.07	0.38	0.38	0.38	0.6	20.74	20.88	10.25	10.25		
CO					27.79	0.15											0.06
CO2					6.79	65.71											96.65
CH4					0.04	0.08											
Ar		0.93	3.24		0.54	1.04	0.5	0.27	0.27	0.27	0.43	0.93	0.94	0.8	0.8	1.57	
Solids, kg/s	65.52			7.19													
Total, kg/s	65.52	229.1	53.15	19.34	191.04	181.63	152.42	161.84	31.96	129.87	30.48	758.07	19.61	898.72	898.72	153.5	1.49
Temperature, °C	15	15	115	50	309	338	288	297	297	297	149	15	34	582	162	58	27
Pressure, kPa	101	101	9317	101	5459	4667	3089	2628	2628	2628	2951	101	1821	105	101	15200	2386

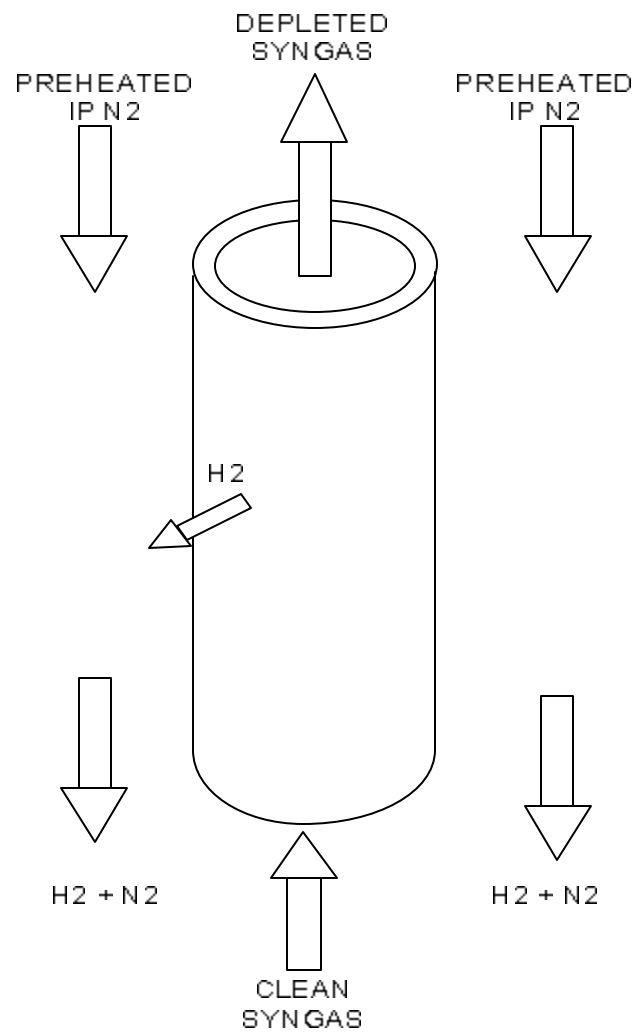


Figure 3: H₂ Production for Gas Turbine - H₂ Separation Membrane Reactor Section with N₂ Sweep Gas