3.1.3 WATER-GAS SHIFT REACTOR

For the conversion of the reformer gas to hydrogen, the first step is to convert most of the carbon monoxide (CO) to hydrogen and carbon dioxide (CO_2) by reacting the CO with water over a bed containing iron-based catalysts, which promote the water-gas shift reaction. This produces the balance of the gross hydrogen product by converting approximately 90 percent of the carbon monoxide to hydrogen and CO_2. The product stream from the reformer contains sufficient amounts of water vapor to meet the necessary water-to-gas ratio at the shift reactor inlet. The CO shift converter consists of four fixed-bed reactors with two reactors in series and two in parallel. Two reactors in series with cooling between the two are required to control the exothermic temperature rise. Two reactors in parallel are required due to the high gas mass flow rate.

Effluent from the second stage is cooled by exchanging heat with incoming feed, by an air cooler, and finally by a water cooler. The exit gas is predominantly hydrogen and CO_2 with some residual CO and methane.

3.1.4 ACID GAS REMOVAL

With conventional production of hydrogen from natural gas, CO_2 is normally not recovered from the syngas stream. The excess steam generated in the boiler is exported offsite. However, this plant utilizes a proprietary amine-based process to remove and recover 99 percent of the CO_2 from the syngas stream. The CO_2 is removed by chemical absorption with a highly selective, hybrid amine. From the shift reactor, gas is passed through an amine tower where it is contacted counter-currently with a circulating stream of lean aqueous amine solution. CO_2 in the feed averages approximately 12 mole % and is removed from the gas stream by the circulating lean amine. The rich amine from the absorber is then sent to a stripper column where the amine is regenerated with a steam reboiler to remove the CO_2 by fractionation. Because of the steam load required to regenerate CO_2, there is no steam export from the plant removing CO_2. Regenerated lean amine is then cooled and sent back to the amine tower. The regenerated CO_2 stream is recovered at 27 psia and 121°F and is sent offsite.

3.1.5 HYDROGEN PURIFICATION

The PSA process is used for hydrogen purification, based on the ability to produce high-purity hydrogen, low amounts of CO and CO_2, and ease of operation. Treated gas from the amine unit is fed directly to the PSA unit where hydrogen is purified up to approximately 99.6 percent. Carbon oxides are limited to 10 ppm in the final hydrogen product. The PSA process is based on the principle of adsorbent beds adsorbing more impurities at high gas-phase partial pressure than at low partial pressure.

The gas stream is passed through adsorption beds at approximately 350 psia, and the impurities are purged from the beds at 2.5 psia. Using a recycle compressor, purge gas is sent back to the gas-fired steam/reformer as supplemental fuel. Purified hydrogen is available as a product at 346 psia. The PSA process operates on a cyclic basis and is controlled by automatic switching valves. Multiple beds are used in order to provide constant product and purge gas flows.

A simplified basic flow sheet of Case 1, Conventional Steam Reforming Process *without* CO_2 Recovery, is shown in Figure 3-1. The overall performance and cost summary for the 150 MMscfd plant is shown in Table 3-2. A simplified basic flow sheet of Case 2, Conventional Steam Reforming Process *with* CO_2 Recovery, is shown in Figure 3-2. The overall performance and cost summary for the 150 MMscfd plant is shown in Table 3-3.

Also included in these comparisons is Case 3, Hydrogen from Partial Oxidation of Natural Gas. This plant, which uses an oxygen-blown gasifier and a hydrogen separation membrane, intuitively will not be economically competitive with other approaches to producing hydrogen. It was not evaluated economically. The high costs of capital and natural gas would result in a rather high cost for hydrogen. A simplified basic flow sheet of Case 3, Partial Oxidation of Natural Gas with 600°C HSD, is shown in Figure 3-3. The overall performance summary for the plant is shown in Table 3-4.

Figure 3-1
Block Flow Diagram Case 1
Steam Reforming Natural Gas

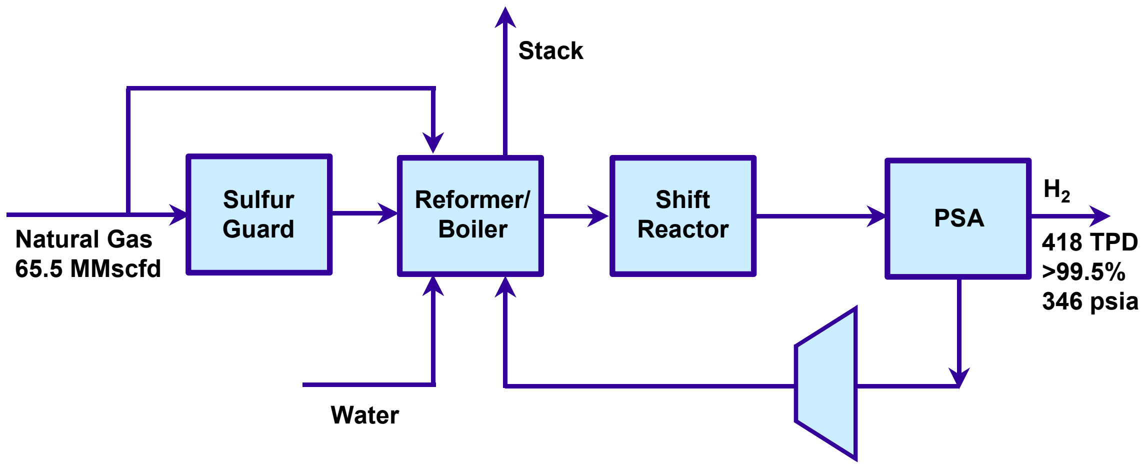

Table 3-2
Performance and Cost Summary
Case 1 – Hydrogen from Natural Gas without CO_2 Capture

Plant Size, tons H_2/day (MMscfd) @ 346 psia	417.8 (150)
Coal Feed (dry basis)	N/A
Natural Gas Feed, MMBtuh (MMscfd)	2,868 (65.5)
Fuel Cost, $/MMBtu	$3.15
Plant Availability	90%
Cold Gas Efficiency	74.2%
Equivalent Thermal Efficiency, HHV	83.9%
Steam Export?	220,000 lb/h
CO_2 Recovered, tpd (percent)	N/A
Net Power	(6 MW)
Total Plant Cost $1,000, Year 2000	$130,998
Cost of Hydrogen, $/MMBtu (c/kscf)	$5.54 (180)

Table 3-2 (Cont'd)
Performance and Cost Summary (Case 1)

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY

TITLE/DEFINITION			
Case:	NG H_2 Plant w/o CO_2 Capture		
Plant Size:	417.8 H_2 TPD	HeatRate:	0 (Btu/kWh)
Primary/Secondary Fuel(type):	Natural Gas	Cost:	3.15 ($/MMBtu)
Design/Construction:	2 (years)	BookLife:	20 (years)
TPC(Plant Cost) Year:	2000 (Jan.)	TPI Year:	2005 (Jan.)
Capacity Factor:	90 (%)		

CAPITAL INVESTMENT		$x1000	$x1000/$H_2$TPD
Process Capital & Facilities		130,998	313.5
Engineering(incl.C.M.,H.O.& Fee)			
Process Contingency			
Project Contingency			
TOTAL PLANT COST(TPC)		$130,998	313.5
TOTAL CASH EXPENDED	$130,998		
AFDC	$6,368		
TOTAL PLANT INVESTMENT(TPI)		$137,366	328.8
Royalty Allowance			
Preproduction Costs		5,017	12.0
Inventory Capital		657	1.6
Initial Catalyst & Chemicals(w/equip.)			
Land Cost			
TOTAL CAPITAL REQUIREMENT(TCR)		$143,040	342.3

OPERATING & MAINTENANCE COSTS (2000 Dollars)	$x1000	000/H_2TPD
Operating Labor	1,489	3.6
Maintenance Labor	943	2.3
Maintenance Material	1,415	3.4
Administrative & Support Labor	608	1.5
TOTAL OPERATION & MAINTENANCE	$4,455	10.7
FIXED O & M		9.60
VARIABLE O & M		1.07

CONSUMABLE OPERATING COSTS,less Fuel (2000 Dollars)	$x1000	$/T H_2-yr
Water	31	0.23
Chemicals	1,678	12.22
Other Consumables		
Waste Disposal		
TOTAL CONSUMABLE OPERATING COSTS	$1,709	12.45
BY-PRODUCT CREDITS (2000 Dollars)	($8,396)	-61.17
FUEL COST (2000 Dollars)	$71,226	518.92

PRODUCTION COST SUMMARY	1st Year (2005 $)	$/T H_2-yr	Levelized (10th.Year $)	$/T H_2-yr
Fixed O & M	9.6/kW-yr	9.60	9.6/kW-yr	9.60
Variable O & M		1.07		1.07
Consumables		12.45		12.45
By-product Credit		-61.17		-61.17
Fuel		552.72		569.55
TOTAL PRODUCTION COST		514.66	47.9281	531.49
LEVELIZED CARRYING CHARGES(Capital)				145.90
LEVELIZED (10th.Year) BUSBAR COST OF POWER				677.39
Equivalent $/MMBtu				5.54

Figure 3-2
Block Flow Diagram Case 2
Steam Reforming Natural Gas with CO_2 Removal

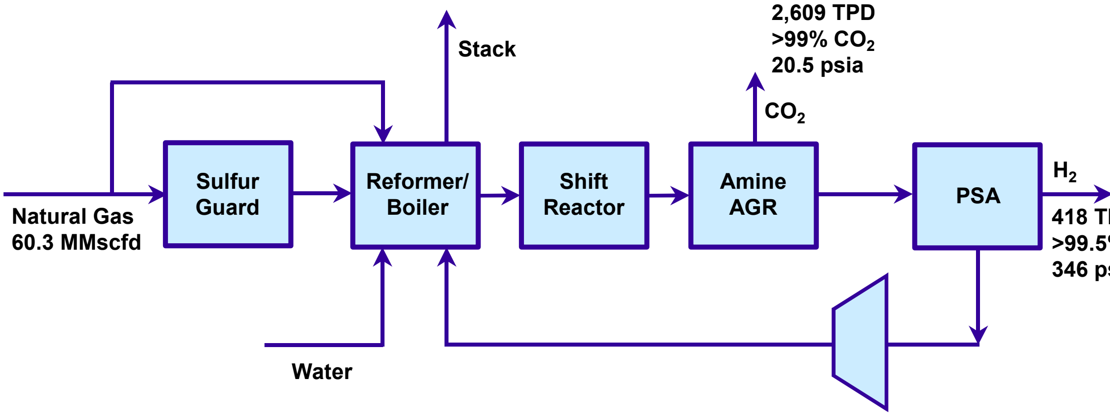

Table 3-3
Performance and Cost Summary
Case 2 – Hydrogen from Natural Gas with CO_2 Capture by Amine Process

Plant Size, tons H_2/day (MMscfd) @ 346 psia	417.8 (150)
Coal Feed (dry basis)	N/A
Natural Gas Feed, MMBtuh (MMscfd)	2,640 (60.3)
Fuel Cost, $/MMBtu	$3.15
Plant Availability	90%
Cold Gas Efficiency	80.6%
Equivalent Thermal Efficiency, HHV	78.6%
Steam Export?	No
CO_2 Recovered, tpd (percent)	2,609 (71%)
Net Power	(15 MW)
Total Plant Cost, $1,000, Year 2000	$142,370
Cost of Hydrogen, $/MMBtu (c/kscf)	$5.93 (192)

Table 3-3 (Cont'd)
Performance and Cost Summary (Case 2)

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY

TITLE/DEFINITION				
Case:	NG H_2 Plant w/CO_2 Capture			
Plant Size:	417.8 H_2 TPD		HeatRate:	0 (Btu/kWh)
Primary/Secondary Fuel(type):	Natural Gas		Cost:	3.15 ($/MMBtu)
Design/Construction:	2 (years)		BookLife:	20 (years)
TPC(Plant Cost) Year:	2000 (Jan.)		TPI Year:	2005 (Jan.)
Capacity Factor:	90 (%)			
CAPITAL INVESTMENT			**$x1000**	**$x1000/$H_2$TPD**
Process Capital & Facilities			142,370	340.7
Engineering(incl.C.M.,H.O.& Fee)				
Process Contingency				
Project Contingency				
TOTAL PLANT COST(TPC)			$142,370	340.7
TOTAL CASH EXPENDED		$142,370		
AFDC		$6,921		
TOTAL PLANT INVESTMENT(TPI)			$149,291	357.3
Royalty Allowance				
Preproduction Costs			5,342	12.8
Inventory Capital			714	1.7
Initial Catalyst & Chemicals(w/equip.)				
Land Cost				
TOTAL CAPITAL REQUIREMENT(TCR)			$155,346	371.8
OPERATING & MAINTENANCE COSTS (2000 Dollars)			**$x1000**	**000/H_2TPD**
Operating Labor			1,489	3.6
Maintenance Labor			1,025	2.5
Maintenance Material			1,538	3.7
Administrative & Support Labor			629	1.5
TOTAL OPERATION & MAINTENANCE			$4,680	11.2
FIXED O & M				10.08
VARIABLE O & M				1.12
CONSUMABLE OPERATING COSTS,less Fuel (2000 Dollars)			**$x1000**	**$/T H_2-yr**
Water			31	0.23
Chemicals			4,161	30.32
Other Consumables				
Waste Disposal				
TOTAL CONSUMABLE OPERATING COSTS			$4,192	30.54
BY-PRODUCT CREDITS (2000 Dollars)				
FUEL COST (2000 Dollars)			$65,563	477.67

PRODUCTION COST SUMMARY	**1st Year (2005 $)**	**$/T H_2-yr**	**Levelized (10th.Year $)**	**$/T H_2-yr**
Fixed O & M	10.1/kW-yr	10.08	10.1/kW-yr	10.08
Variable O & M		1.12		1.12
Consumables		30.54		30.54
By-product Credit				
Fuel		508.78		524.27
TOTAL PRODUCTION COST		550.53	52.05135	566.02
LEVELIZED CARRYING CHARGES(Capital)				158.45
LEVELIZED (10th.Year) BUSBAR COST OF POWER				724.47
Equivalent $/MMBtu				5.93

Figure 3-3
Block Flow Diagram Case 3
Partial Oxidation Natural Gas with 600°C HSD

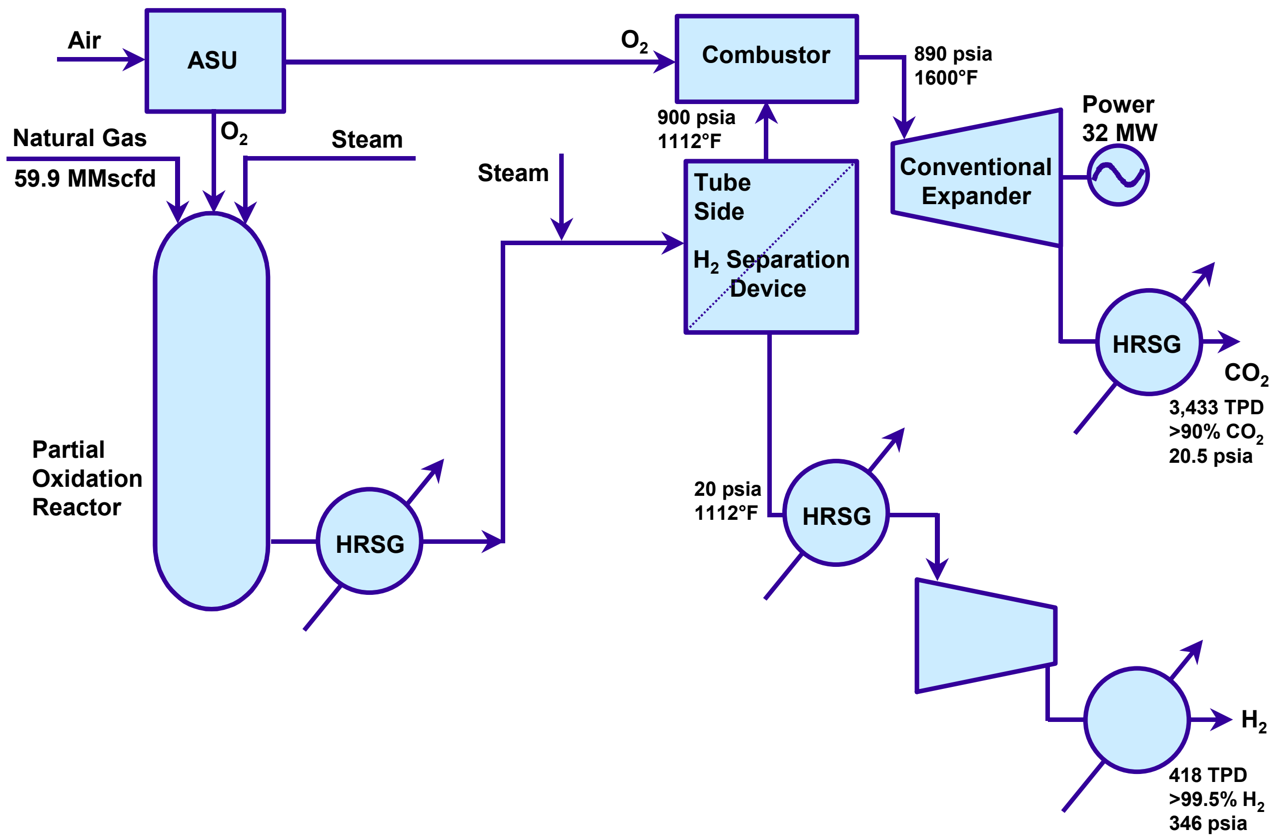

Table 3-4
Performance Summary
Case 3 – Natural Gas Partial Oxidation Plant with CO_2 Capture
600°C Inorganic Membrane

Plant Size, tons H_2/day (MMscfd) @ 346 psia	417.8 (150)
Coal Feed (dry basis)	N/A
Natural Gas Feed, MMBtuh (MMscfd)	2,618 (59.9)
Fuel Cost, $/MMBtu	$3.15
Plant Availability	90%
Cold Gas Efficiency	81.2%
Equivalent Thermal Efficiency, HHV	87.4%
Steam Export?	220,000 lb/h
CO_2 Recovered, tpd (percent)	3,433 (94%)
Net Power	(27 MW)

3.2 CASES 4 AND 5 – HYDROGEN FROM COAL GASIFICATION WITHOUT OR WITH CO_2 REMOVAL

A fuel production facility conceptual plant design was prepared to evaluate the conversion of coal to hydrogen utilizing conventional gas stream cleanup and processing.

The Destec gasifier and coal handling equipment are identical to those in the previous hydrogen plants. The high-pressure syngas produced in the gasifier is quenched to 1905°F as a result of adjustments in the second stage of the gasifier, and utilizes a firetube heat exchanger to cool the gas further to 625°F. The gas is cleaned of particles with a ceramic candle filter and shifted utilizing a sulfur-tolerant catalyst. The gas can be cleaned of CO_2 and sulfur in a double-stage Selexol unit. H_2S from the acid gas removal process is used to manufacture sulfuric acid byproduct. Hydrogen is purified in a PSA unit, and the PSA tail gas is fired in a heat recovery steam generator (HRSG). For the CO_2 removal case, the PSA tail gas is fired in the HRSG with oxygen, resulting in a concentrated CO_2 stream in the stack for recovery. Excess steam produced from hot gas cooling and the HRSG is used to produce power for in-plant use and the balance for sale.

Following are more detailed descriptions of the key process elements:

3.2.1 GASIFIER

For this application, to produce lower pressure syngas, a single-train Destec gasifier of the Wabash River configuration is utilized. The net temperature for gas leaving the gasifier is 1900°F by using a 78/22 flow split between the first and second stages of the gasifier. Slag produced in the high-temperature gasifier reaction flows to the bottom of the first stage where it falls into a water bath and is cooled and shattered to become an inert frit.

Gas leaving the gasifier at 1905°F goes through an internal cyclone that separates entrained particles from the gas for recycle to the gasifier, followed by a fire-tube boiler to reduce gas temperature to 625°F. Following the cooler, the remaining particulates are removed from the gas with a ceramic candle filter and are returned to the gasifier.

3.2.2 AIR SEPARATION UNIT

Oxygen supply for this plant is also provided through a conventional cryogenic air separation unit (ASU). The air separation plant is designed to produce a nominal output of 2,100 tons/day of 95 percent pure O_2. The high-pressure plant is designed with two 50 percent capacity production trains, with liquefaction and liquid oxygen storage providing an 8-hour backup supply of oxygen.

3.2.3 PARTICULATE REMOVAL

The particulate removal device is a ceramic candle configuration operating at the relatively low temperature of 625°F. The vessel and candle array is similar to the Westinghouse configuration used at the Piñon Pine clean coal technology (CCT) demonstration plant. A single-train particulate removal vessel is adequate for each gasifier train.

3.2.4 SHIFT

After leaving the particulate control unit, steam is injected into the gas stream, and the CO in the syngas is shifted to hydrogen and CO_2 in the shift converter utilizing sulfur-tolerant shift catalysts. Heat is removed from the gas stream following the shift, the gases are cooled, water is condensed, and the gas stream is sent to the sulfur removal unit.

3.2.5 SULFUR REMOVAL/HYDROGEN PURIFICATION

In order to remove H_2S and CO_2 separately from the hydrogen product stream, a double-stage Selexol unit was selected. This process removes H_2S from the cooled syngas and then removes CO_2 from the desulfurized syngas. The acid gas removal (AGR) process utilizes a physical sorbent and several design features to effectively remove and recover H_2S and CO_2 from the syngas stream. Syngas leaves the shift converter reactor at 857°F and is cooled to 105°F prior to entering the absorber tower at 353 psia. The product hydrogen stream exits the absorber at 338 psia and is sent to a PSA unit to purify the hydrogen. The product hydrogen leaves the PSA unit at 310 psia, and the PSA tail gas is sent to the fired HRSG. For the CO_2 removal case, the PSA tail gas is fired in the HRSG with oxygen, resulting in a concentrated CO_2 stream in the stack for recovery.

The conventional hydrogen from coal plant described in the June 1999[5] letter report included provisions for recovering CO_2. The amount of CO_2 recovered, relative to the total amount that could be produced from the coal carbon, was about 75 percent. This was a result of having some CO remaining in the syngas following the shift reactors. Upon separating the hydrogen from the syngas in the PSA, the PSA off-gas was fired in a HRSG with air, and the CO_2 in the flue gas would be emitted to the atmosphere. To put the product costs of the conventional plant on an equal basis with other plants, process adjustments were made to maximize the amount of CO_2 captured. This was accomplished by firing the PSA retentate with oxygen in the HRSG, resulting in a stack gas containing only CO_2 and water vapor. The CO_2 is then cooled and recovered.

The Selexol unit consists of two absorbers: the first absorbs H_2S from the cooled syngas, providing a desulfurized syngas, and the second absorbs CO_2 from the desulfurized syngas. The two absorbers are integrated, with solvent flowing between them. A low-pressure H_2S stream is sent to the sulfuric acid plant and a low-pressure CO_2 stream is sent offsite for sequestration.

A simplified basic flow sheet of Case 4, Conventional Hydrogen from Coal *without* CO_2 Recovery, is shown in Figure 3-4. The overall performance and cost summary for the plant is shown in Table 3-5. A simplified basic flow sheet of Case 5, Conventional Hydrogen from Coal *with Maximum* CO_2 Recovery, is shown in Figure 3-5. The overall performance and cost summary for the plant is shown in Table 3-6.

Figure 3-4
Block Flow Diagram Case 4
Conventional Hydrogen Plant without CO_2 Removal

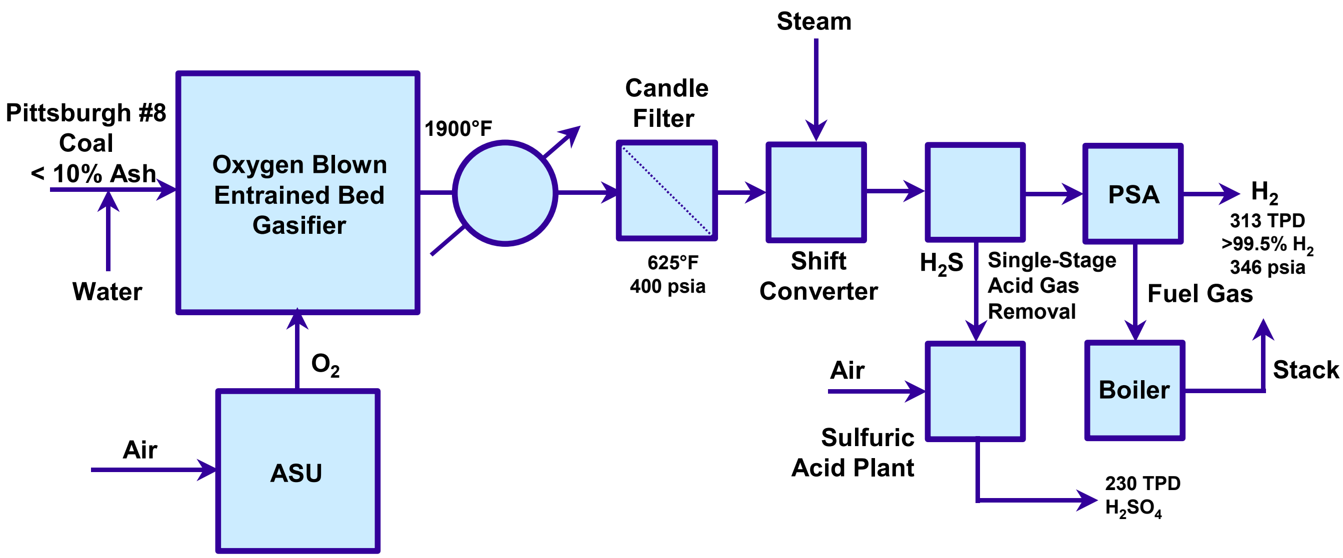

Table 3-5
Performance and Cost Summary
Case 4 – Conventional Hydrogen from Coal without CO_2 Capture

Plant Size, tons H_2/day (MMscfd) @ 346 psia	312.6 (112)
Coal Feed (dry basis)	2,500 tpd
Natural Gas Feed, MMBtuh (MMscfd)	N/A
Fuel Cost, $/MMBtu	$1.00
Plant Availability	80%
Cold Gas Efficiency	57.7%
Equivalent Thermal Efficiency, HHV	62.3%
Steam Export?	No
CO_2 Recovered, tpd (percent)	N/A
Net Power	38 MW
Total Plant Cost, $1,000, Year 2000	$321,824
Cost of Hydrogen, $/MMBtu (c/kscf)	$5.71 (186)

Table 3-5 (Cont'd)
Performance and Cost Summary (Case 4)

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY				
TITLE/DEFINITION				
Case:	Conventional H_2 Plant w/o CO_2 Capture			
Plant Size:	312.6 TPD-Syngas	HeatRate:		(Btu/kWh)
Primary/Secondary Fuel(type):	Pitts. #8	Cost:		1.00 ($/MMBtu)
Design/Construction:	2.5 (years)	BookLife:		20 (years)
TPC(Plant Cost) Year:	2000 (Jan.)	TPI Year:		2005 (Jan.)
Capacity Factor:	80 (%)			
CAPITAL INVESTMENT			$x1000	$x1000/$H_2$TPD
Process Capital & Facilities			264,449	846.0
Engineering(incl.C.M.,H.O.& Fee)			25,558	81.8
Process Contingency			2,560	8.2
Project Contingency			29,257	93.6
TOTAL PLANT COST(TPC)			$321,824	1029.5
TOTAL CASH EXPENDED		$321,824		
AFDC		$20,635		
TOTAL PLANT INVESTMENT(TPI)			$342,459	1095.5
Royalty Allowance				
Preproduction Costs			8,167	26.1
Inventory Capital			2,862	9.2
Initial Catalyst & Chemicals(w/equip.)				
Land Cost			150	0.5
TOTAL CAPITAL REQUIREMENT(TCR)			$353,637	1131.3
OPERATING & MAINTENANCE COSTS (2000 Dollars)			$x1000	$x1000/$H_2$TPD
Operating Labor			3,276	10.5
Maintenance Labor			2,121	6.8
Maintenance Material			3,182	10.2
Administrative & Support Labor			1,349	4.3
TOTAL OPERATION & MAINTENANCE			$9,928	31.8
FIXED O & M				25.41
VARIABLE O & M				6.35
CONSUMABLE OPERATING COSTS,less Fuel (2000 Dollars)			$x1000	$/T H_2-yr
Water			154	1.68
Chemicals			637	6.98
Other Consumables				
Waste Disposal			793	8.69
TOTAL CONSUMABLE OPERATING COSTS			$1,584	17.35
BY-PRODUCT CREDITS (2000 Dollars)			($13,021)	-142.65
FUEL COST (2000 Dollars)			$19,337	211.85

PRODUCTION COST SUMMARY	1st Year (2005 $) $/T H_2-yr	Levelized (Over Book Life $) $/T H_2-yr
Fixed O & M	87.01	87.01
Variable O & M	6.35	6.35
Consumables	17.35	17.35
By-product Credit/Penalty	-142.65	-142.65
Fuel	205.57	187.51
TOTAL PRODUCTION COST	173.63	155.57
LEVELIZED CARRYING CHARGES(Capital)		542.39
LEVELIZED(Over Book Life)COST/Ton of Syngas		697.97
Equivalent $/MMBtu		5.71

Figure 3-5
Block Flow Diagram Case 5
Conventional Hydrogen Plant with CO_2 Removal

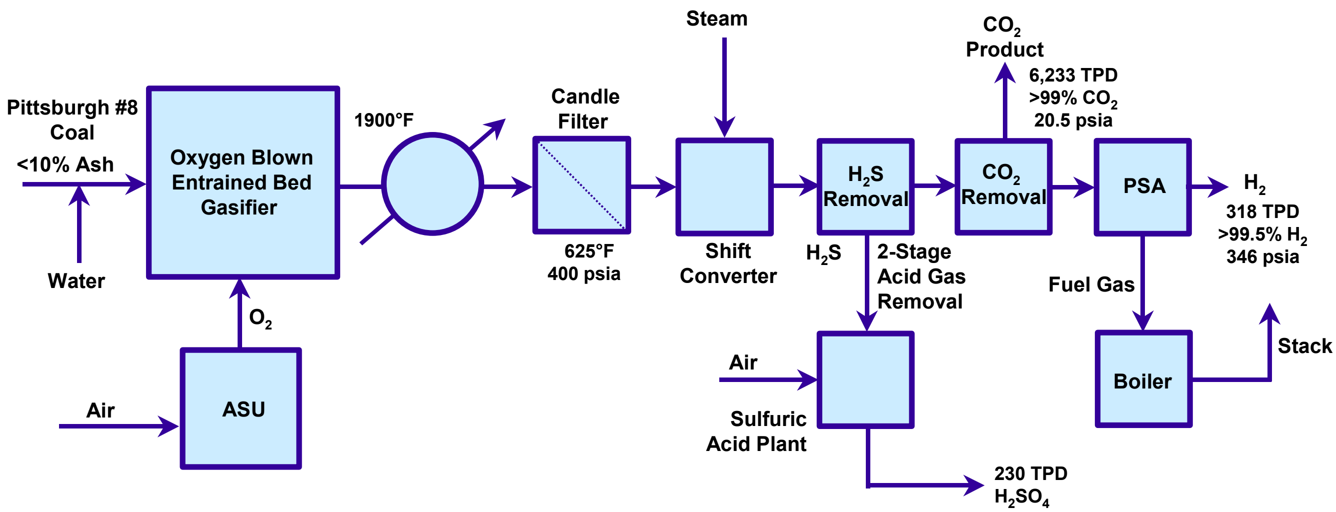

Table 3-6
Performance and Cost Summary
Case 5 – Conventional Hydrogen from Coal with Maximum CO_2 Capture

Plant Size, tons H_2/day (MMscfd) @ 346 psia	317.8 (114)
Coal Feed (dry basis)	2,500 tpd
Natural Gas Feed, MMBtuh (MMscfd)	N/A
Fuel Cost, $/MMBtu	$1.00
Plant Availability	80%
Cold Gas Efficiency	58.6%
Equivalent Thermal Efficiency, HHV	60.1%
Steam Export?	No
CO_2 Recovered, tpd (percent)	6,233 (92%)
Net Power	12 MW
Total Plant Cost, $1,000, Year 2000	$374,906
Cost of Hydrogen, $/MMBtu (c/kscf)	$6.91 (225)

Table 3-6 (Cont'd)
Performance and Cost Summary (Case 5)

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY

TITLE/DEFINITION

Case:	Conventional H_2 Plant w/ Max.CO Removal			
Plant Size:	317.8 TPD-Synga	HeatRate:		(Btu/kWh)
Primary/Secondary Fuel(type):	Pitts. #8	Cost:	1.00	($/MMBtu)
Design/Construction:	2.5 (years)	BookLife:	20	(years)
TPC(Plant Cost) Year:	2000 (Jan.)	TPI Year:	2005	(Jan.)
Capacity Factor:	80 (%)			

CAPITAL INVESTMENT		$x1000	$x1000/$H_2$TPD
Process Capital & Facilities		307,610	967.8
Engineering(incl.C.M.,H.O.& Fee)		30,653	96.4
Process Contingency		2,560	8.1
Project Contingency		34,082	107.2
TOTAL PLANT COST(TPC)		$374,906	1179.5
TOTAL CASH EXPENDED	$374,906		
AFDC	$24,039		
TOTAL PLANT INVESTMENT(TPI)		$398,945	1255.2
Royalty Allowance			
Preproduction Costs		9,522	30.0
Inventory Capital		3,006	9.5
Initial Catalyst & Chemicals(w/equip.)			
Land Cost		150	0.5
TOTAL CAPITAL REQUIREMENT(TCR)		$411,622	1295.0

OPERATING & MAINTENANCE COSTS (2000 Dollars)	$x1000	$x1000/$H_2$TPD
Operating Labor	3,276	10.3
Maintenance Labor	2,534	8.0
Maintenance Material	3,801	12.0
Administrative & Support Labor	1,452	4.6
TOTAL OPERATION & MAINTENANCE	$11,064	34.8
FIXED O & M		27.85
VARIABLE O & M		6.96

CONSUMABLE OPERATING COSTS,less Fuel (2000 Dollars)	$x1000	$/T H_2-yr
Water	154	1.66
Chemicals	637	6.87
Other Consumables		
Waste Disposal	793	8.54
TOTAL CONSUMABLE OPERATING COSTS	$1,584	17.07
BY-PRODUCT CREDITS (2000 Dollars)	($7,437)	-80.13
FUEL COST (2000 Dollars)	$19,337	208.35

PRODUCTION COST SUMMARY	1st Year (2005 $) $/T H_2-yr	Levelized (Over Book Life $) $/T H_2-yr	
Fixed O & M	95.37	95.37	0.780471
Variable O & M	6.96	6.96	0.0569744
Consumables	17.07	17.07	0.13968
By-product Credit/Penalty	-80.13	-80.13	-0.655812
Fuel	202.18	184.42	1.5092652
TOTAL PRODUCTION COST	241.44	223.68	1.8305787
LEVELIZED CARRYING CHARGES(Capital)		620.91	5.0815259
LEVELIZED(Over Book Life)COST/Ton of Syngas		844.59	
Equivalent $/MMBtu		6.91	

3.3 SUMMARY AND CONCLUSIONS

Table 3-7 is a summary of the results of comparing hydrogen costs from conventional natural gas and coal sources with the cost of producing hydrogen from coal using advanced membrane technology.

Table 3-7
Comparison of Hydrogen Cost from Conventional and Advanced Plant Designs

	Case 1 Hydrogen from Natural Gas without CO_2 Capture	Case 2 Hydrogen from Natural Gas with CO_2 Capture by Amine Process	Case 4 Conventional Hydrogen from Coal without CO_2 Capture	Case 5 Conventional Hydrogen from Coal with Maximum CO_2 Capture	Baseline Case Advanced Hydrogen Plant with CO_2 Capture 600°C Membrane
Plant Size, tons H_2/day (MMscfd) (Pressure, psia)	417.8 tpd (150 MMscfd) (346)	417.8 tpd (150 MMscfd) (346)	312.6 tpd (112 MMscfd) (346)	317.8 tpd (114 MMscfd) (346)	430.8 tpd (147 MMscfd) (346)
Coal Feed (dry basis)	N/A	N/A	2,500 tpd	2,500 tpd	2,500 tpd
Natural Gas Feed, MMBtuh (MMscfd)	2,868 MMBtuh (65.5 MMscfd)	2,640 MMBtuh (60.3 MMscfd)	N/A	N/A	N/A
Fuel Cost, \$/MMBtu	\$3.15/MMBtu	\$3.15/MMBtu	\$1.00/MMBtu	\$1.00/MMBtu	\$1.00/MMBtu
Plant Availability	90%	90%	80%	80%	80%
Cold Gas Efficiency[1]	74.2%	80.6%	57.7%	58.6%	79.5%
Equivalent Thermal Efficiency, HHV	83.9%	78.6%	62.3%	60.1%	80.4%
Steam Export?	220,000 lb/h	No	No	No	No
CO_2 Recovered, tpd (percent) (Pressure, psia)	N/A	2,609 tpd (71%) (30)	N/A	6,233 tpd (92%) (30)	6,362 tpd (94%) (20)
Net Power	(6 MW)	(15 MW)	38 MW	12 MW	7 MW
Total Plant Cost \$1,000, Year 2000	\$130,998	\$142,370	\$321,824	\$374,906	\$359,791
Cost of Hydrogen, \$/MMBtu (¢/kscf)	\$5.54/MMBtu (180 ¢/kscf)	\$5.93/MMBtu (192 ¢/kscf)	\$5.71/MMBtu (186 ¢/kscf)	\$6.91/MMBtu (225 ¢/kscf)	\$5.06/MMBtu (164 ¢/kscf)

[1] Cold gas efficiency equals HHV of the product gas divided by the HHV of the feed x 100.

Given that the R&D goals can be achieved, hydrogen production from the baseline hydrogen fuel plant, which includes CO_2 removal, would be competitive with hydrogen produced from both natural gas- and coal-based conventional technologies even without CO_2 removal. With only 80 percent hydrogen transport, hydrogen production would still be competitive with conventional coal-based technology.