

Hydrogen Production Facilities Plant Performance and Cost Comparisons

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TABLE OF CONTENTS

<u>Section</u>	<u>Title</u>	<u>Page</u>
	List of Tables	iii
	List of Figures	v
	List of Acronyms and Abbreviations	vi
	EXECUTIVE SUMMARY	ES-1
1	INTRODUCTION	1-1
2	HYDROGEN FUEL FROM COAL PLANTS	2-1
	2.1 Hydrogen Separation Device Process Design	2-1
	2.1.1 Updated Assumptions for HSD Design	2-2
	2.1.2 HSD Design	2-3
	2.2 Hydrogen Fuel Plant – 1402°F (761°C) Membrane	2-3
	2.2.1 Process Description	2-9
	2.2.2 Effective Thermal Efficiency	2-11
	2.3 Hydrogen Fuel Plant – 1112°F (600°C) Membrane	2-12
	2.3.1 Process Description	2-13
	2.3.2 Effective Thermal Efficiency	2-16
	2.4 Hydrogen Fuel Plant – 572°F (300°C) Membrane	2-16
	2.4.1 Process Description	2-17
	2.4.2 Effective Thermal Efficiency	2-21
	2.5 Hydrogen Fuel Plant – 1112°F (600°C) Membrane and 80 Percent Hydrogen Transport	2-21
	2.5.1 Process Description	2-22
	2.5.2 Effective Thermal Efficiency	2-26
	2.6 Cost Estimating	2-26
	2.6.1 Approach to Cost Estimating	2-26
	2.6.2 Production Costs (Operations and Maintenance)	2-28
	2.6.3 Cost Results	2-29
	2.7 Summary and Conclusions	2-33
3	HYDROGEN FROM COAL AND NATURAL GAS-BASED PLANTS	3-1
	3.1 Cases 1, 2, and 3 – Hydrogen from Natural Gas Without and With CO ₂ Recovery	3-1
	3.1.1 Natural Gas Conditioning	3-1

<u>Section</u>	<u>Title</u>	<u>Page</u>
	3.1.2 Natural Gas Reformer/Boiler	3-2
	3.1.3 Water-Gas Shift Reactor	3-3
	3.1.4 Acid Gas Removal	3-3
	3.1.5 Hydrogen Purification	3-3
3.2	Cases 4 and 5 – Hydrogen from Conventional Coal Gasification without or with CO ₂ Removal	3-10
	3.2.1 Gasifier	3-10
	3.2.2 Air Separation Unit	3-10
	3.2.3 Particulate Removal	3-10
	3.2.4 Shift	3-11
	3.2.5 Sulfur Removal/Hydrogen Purification	3-11
3.3	Summary and Conclusions	3-16
4	HYDROGEN FUEL FROM WYODAK COAL/BIOMASS BLEND	4-1
4.1	Introduction	4-1
4.2	Process Description	4-4
	4.2.1 Gasifier	4-5
	4.2.2 Air Separation Unit	4-5
	4.2.3 Hot Gas Cleanup System	4-5
	4.2.4 Sulfuric Acid Plant	4-7
	4.2.5 Hydrogen Separation/Conventional Turbine Expander	4-7
	4.2.6 Effective Thermal Efficiency	4-8
4.3	Cost Estimate	4-8
4.4	Summary and Conclusions	4-11
5	REFERENCES	5-1

LIST OF TABLES

Table No.	<u>Title</u>	<u>Page</u>
ES-1	Performance and Cost Summary Comparisons – Hydrogen Fuel Plants with Alternative HSD Temperatures	ES-2
ES-2	Comparison of Hydrogen Cost from Conventional and Advanced Plant Designs	ES-4
ES-3	Performance and Cost Summary Comparisons – Hydrogen Fuel Plants with Alternative Feedstocks	ES-5
1-1	Consistent Design Parameters (Unless Noted in Text)	1-2
1-2	Consistent Financial Parameters (Unless Noted in Text)	1-2
2-1	Revised Assumptions for Hydrogen Separation Device	2-2
2-2	Hydrogen Separation Device Designs	2-4
2-3	Design Basis for Hydrogen Fuel Production Facility with Conventional Expansion Turbine and Hot Gas Cleanup	2-6
2-4	Performance Summary – Baseline Hydrogen Fuel Plant	2-9
2-5	Auxiliary Power Load – Baseline Hydrogen Fuel Plant	2-9
2-6	Plant Water Balance – Baseline Hydrogen Fuel Plant	2-11
2-7	Design Basis for Baseline Hydrogen Fuel Plant – 1112°F (600°C) Membrane	2-12
2-8	Performance Summary – 1112°F Membrane	2-15
2-9	Auxiliary Power Load – 1112°F Membrane	2-15
2-10	Plant Water Balance – 1112°F Membrane	2-16
2-11	Design Basis for Baseline Hydrogen Fuel Plant – 572°F (300°C) Membrane	2-17
2-12	Performance Summary – 572°F Membrane	2-20
2-13	Auxiliary Power Load – 572°F Membrane	2-20
2-14	Plant Water Balance – 572°F Membrane	2-21
2-15	Design Basis for Baseline Hydrogen Fuel Plant – 1112°F (600°C) Membrane and 80 Percent Hydrogen Transport	2-22
2-16	Performance Summary – 1112°F Membrane and 80% Hydrogen Transport	2-25
2-17	Auxiliary Power Load – 1112°F Membrane and 80% Hydrogen Transport	2-25
2-18	Plant Water Balance – 1112°F Membrane and 80% Hydrogen Transport	2-26
2-19	Estimate Basis/Financial Criteria	2-27
2-20	Capital Investment & Revenue Requirement Summary – Baseline	2-29
2-21	Capital Investment & Revenue Requirement Summary – 600°C	2-30

<u>Table No.</u>	<u>Title</u>	<u>Page</u>
2-22	Capital Investment & Revenue Requirement Summary – 300°C	2-31
2-23	Capital Investment & Revenue Requirement Summary – 600°C with 80% H ₂	2-32
2-24	Performance and Cost Summary Comparisons – Hydrogen Fuel Plants with Alternative HSD Temperatures	2-33
3-1	Composition of Synthetic Gas	3-2
3-2	Performance and Cost Summary – Case 1 – Hydrogen from Natural Gas without CO ₂ Capture	3-5
3-3	Performance and Cost Summary – Case 2 – Hydrogen from Natural Gas with CO ₂ Capture by Amine Process	3-7
3-4	Performance Summary – Case 3 – Natural Gas Partial Oxidation Plant with CO ₂ Capture – 600°C Inorganic Membrane	3-9
3-5	Performance and Cost Summary – Case 4 – Conventional Hydrogen from Coal without CO ₂ Capture	3-12
3-6	Performance and Cost Summary – Case 5 – Conventional Hydrogen from Coal with Maximum CO ₂ Capture	3-14
3-7	Comparison of Hydrogen Cost from Conventional and Advanced Plant Designs	3-16
4-1	Wyodak Coal and Biomass Properties (As Received)	4-1
4-2	Design Basis for Hydrogen Fuel Production Facility with Conventional Expansion Turbine and Hot Gas Cleanup – Biomass/Coal Feed	4-2
4-3	Performance Summary – Wyodak Coal/Biomass Blend	4-4
4-4	Auxiliary Power Load – Wyodak Coal/Biomass Blend	4-4
4-5	Plant Water Balance – Wyodak Coal/Biomass Blend	4-8
4-6	Capital Estimate and Revenue Requirement Summary – \$0.65/MMBtu Feedstock	4-9
4-7	Capital Estimate and Revenue Requirement Summary – \$0.50/MMBtu Feedstock	4-10
4-8	Performance and Cost Summary Comparisons – Wyodak/Biomass Blend vs. Pittsburgh No. 8	4-11
4-9	Performance and Cost Summary Comparisons – 1999 Wyodak Substitution for Pittsburgh No. 8	4-12

LIST OF FIGURES

Figure No.	<u>Title</u>	<u>Page</u>
2-1	Hydrogen Separation Device Concept	2-5
2-2	Block Flow Diagram – Baseline Hydrogen Fuel Plant – 1402°F (761°C) Hydrogen Separation Device	2-7
2-3	Process Flow Diagram – Hydrogen Fuel Plant – Hot Gas Desulfurization and Conventional Turbine	2-8
2-4	Block Flow Diagram – Baseline Hydrogen Fuel Plant – 1112°F (600°C) Hydrogen Separation Device	2-13
2-5	Process Flow Diagram – Hydrogen Fuel Plant – 600°C Hydrogen Separation Membrane	2-14
2-6	Block Flow Diagram – Baseline Hydrogen Fuel Plant – 572°F (300°C) HSD	2-18
2-7	Process Flow Diagram – Hydrogen Fuel Plant – 572°F (300°C) Membrane	2-19
2-8	Block Flow Diagram – Baseline Hydrogen Fuel Plant – 1112°F (600°C) HSD at 80 Percent Hydrogen Transport	2-23
2-9	Process Flow Diagram – Hydrogen Fuel Plant – 600°C Hydrogen Separation Membrane and 80 Percent Hydrogen Transport	2-24
3-1	Block Flow Diagram Case 1 – Steam Reforming Natural Gas	3-5
3-2	Block Flow Diagram Case 2 – Steam Reforming Natural Gas with CO ₂ Removal	3-7
3-3	Block Flow Diagram Case 3 – Partial Oxidation Natural Gas with 600°C HSD	3-9
3-4	Block Flow Diagram Case 4 – Conventional Hydrogen Plant without CO ₂ Removal	3-12
3-5	Block Flow Diagram Case 5 – Conventional Hydrogen Plant with CO ₂ Removal	3-14
4-1	Block Flow Diagram – Hydrogen Plant with Wyodak Coal/Biomass Fuel	4-3
4-2	Process Flow Diagram – Hydrogen Fuel Plant with 90% Wyodak Coal/ 10% Biomass – 600°C Hydrogen Separation Membrane	4-6

LIST OF ACRONYMS AND ABBREVIATIONS

AGR	acid gas removal
AR	as received
ASU	air separation unit
ATS	advanced turbine system
C&PS	Office of Coal and Power Systems
¢/kscf	cents per thousand standard cubic feet
CCT	clean coal technology
CH ₄	methane
CO	carbon monoxide
CO ₂	carbon dioxide
DOE	U.S. Department of Energy
ETE	effective thermal efficiency
ETTP	Eastern Tennessee Technology Park
FGD	flue gas desulfurization
FY	fiscal year
H ₂ SO ₄	sulfuric acid
HHV	higher heating value
HRSG	heat recovery steam generator
HSD	hydrogen separation device
ID	inside diameter
IGCC	integrated gasification combined cycle
lb/h	pound per hour
MMBtu	million British thermal units
MW	megawatt
N/A	not applicable
NO _x	oxides of nitrogen
O&M	operations and maintenance
OD	outside diameter
ORNL	Oak Ridge National Laboratory
PFBC	pressurized fluidized-bed combustion
ppm	parts per millions
PSA	pressure swing adsorption
psia	pounds per square inch absolute
psig	pounds per square inch gage
R&D	research and development
scfd	standard cubic feet per day
SCR	selective catalytic reduction
SF	separation factor
SNCR	selective non-catalytic reduction
SO ₂	sulfur dioxide
SO ₃	sulfur trioxide
TBD	to be determined
TCR	total capital requirement
tpd	tons per day

EXECUTIVE SUMMARY

In support of the U.S. Department of Energy (DOE) Advanced Research Program, conceptual systems and cost analyses were developed by the Parsons Corporation for coal processing plants to produce hydrogen while recovering carbon dioxide (CO₂) for offsite processing or sequestration. These plants had been referred to as “decarbonized fuel plants,” but are now called “hydrogen fuel plants.” The scope of work for this analysis entailed the following:

- Identifying alternative processes and technologies utilized for production of hydrogen from coal.
- Reviewing the technical and economic characteristics of developmental materials and technologies for separating hydrogen and oxygen from gas mixtures.
- Conceptualizing process plant designs that utilize developing technologies and materials, resulting in costs of product and CO₂ sequestration significantly lower than with conventional approaches.
- Comparing the costs of a hydrogen fuel plant with plants designed to produce hydrogen from coal utilizing conventional technology.
- Performing sensitivity analyses on the baseline conceptual hydrogen fuel plants to determine the effect of modifying plant design on cost of product.
- Presenting data and results on this study at periodic conferences and workshops.

Introduction

An alternative plant was conceived for producing hydrogen from coal utilizing a hydrogen separation device (HSD) being developed by Oak Ridge National Laboratory (ORNL). The HSD is based on a high-temperature membrane separation concept that can be designed to selectively separate hydrogen from other gases. By utilizing the HSD, it should be possible to separate hydrogen from CO₂ passively and economically.

This report is a compilation of a series of letter reports issued between 1999 and 2001 to document the activity and results from this investigation. It includes the following:

- An establishment of a baseline plant design for hydrogen production based on the ORNL membrane concept,
- A comparison of this design to the conventional methods of producing hydrogen from natural gas and coal, and
- An evaluation of the HSD based on gasifying a mixture of Wyodak coal and biomass.

Hydrogen Fuel from Coal Plants

Through mid-1999, designs and cost estimates for fuel plants utilizing the inorganic membrane were based on information derived from a 1997 conversation with Oak Ridge National Laboratory. The reporting and presentation of work associated with the membranes stimulated significant levels of interest in membrane applications, both within the DOE and in private industry. The primary report from this activity was a letter report prepared in June 1999. Nearly two years had passed since the initial information exchange, which led to a meeting held at Eastern Tennessee Technology Park (ETTP) in Oak Ridge, November 1999 to review the status of the properties and characteristics of the inorganic membrane for hydrogen transport. As a result of data gained from the meeting, assumptions applied to the membrane, which could have an impact on the baseline plant designs and on future membrane applications, were updated.

Utilizing the revised assumptions for the HSD, updated plant concepts were prepared for HSD operation at 572°F (300°C) and 1112°F (600°C). For comparisons, the initial plant operating at 1402°F (761°C) is also presented. A plant with HSD performance reduced from 95 to 80 percent hydrogen transport was also evaluated to show the impact of not reaching the HSD goal of 95 percent separation. Table ES-1 summarizes and compares the performance and economics of the four plants.

Table ES-1
Performance and Cost Summary Comparisons
Hydrogen Fuel Plants with Alternative HSD Temperatures

	1402°F Membrane (761°C)	1112°F Membrane (600°C) Baseline Case	572°F Membrane (300°C)	1112°F Membrane with 80% Hydrogen Transport
HSD Exit Temperature	1402°F (761°C)	1112°F (600°C)	572°F (300°C)	1112°F (600°C)
Coal Feed	221,631 lb/h	221,631 lb/h	221,631 lb/h	221,631 lb/h
Oxygen Feed (95%)	231,218 lb/h	224,519 lb/h	218,657 lb/h	287,917 lb/h
Hydrogen Product Stream	35,205 lb/h	35,903 lb/h	36,565 lb/h	28,562 lb/h
CO ₂ Product Stream	581,657 lb/h	582,566 lb/h	585,598 lb/h	583,220 lb/h
Sulfuric Acid Product	19,482 lb/h	19,482 lb/h	19,482 lb/h	19,482 lb/h
Gross Power Production	94 MW	84 MW	71 MW	131 MW
Auxiliary Power Requirement	76 MW	77 MW	76 MW	83 MW
Net Power Production	18 MW	7 MW	(6 MW)	48 MW
Effective Thermal Efficiency, HHV	80.2%	80.4%	80.3%	69.2%
Capital Cost, \$1,000 (Year 2000)	\$368,448	\$359,791	\$356,797	\$385,650
Hydrogen Product Cost, \$/MMBtu	\$5.11	\$5.06	\$5.10	\$6.02

The lower temperature favors hydrogen recovery but reduces the efficiency of the steam cycle. The 1112°F (600°C) plant was selected as the baseline design since this temperature is the operational goal of the membranes; in addition, this concept maintained a high hydrogen recovery while minimizing costs.

These designs were based on goals that have been set by membrane developers but not yet experimentally demonstrated. These goals include:

- Hydrogen Flux – The hydrogen flux was based on the R&D goal of 0.1 std cc/minute/cm²/cm Hg P_{H₂} differential.
- Separation Factor – The separation determines the hydrogen purity and is high for hydrogen, increasing with higher temperatures. Even at 300°C the separation factor would be above 200.
- Operating Pressure and Temperature – It was assumed that a 950 psi pressure differential can be contained by the inorganic membrane. The operational goal for the membranes is currently 600°C, and a vessel design could be prepared today to operate with confidence up to 300°C.
- CO Shift Properties – It was assumed that the shift reaction on the membrane surface goes to equilibrium without catalyst.

The 80 percent hydrogen transport case reduces the amount of hydrogen recovered but increases the amount of power produced in the topping cycle. The cost of hydrogen increases from the baseline case, but proportionally less than the reduction in hydrogen recovered.

Based on consistent financial parameters and technical parameters taken from the goals of the membrane developers, hydrogen can be produced ranging from \$5.06 to \$5.11 per million Btu including CO₂ capture. With 80 percent hydrogen transport, the cost increases to \$6.02 per million Btu including CO₂ capture.

Hydrogen from Natural Gas and Coal-Based Plants

The previous work resulted in a baseline plant for production of hydrogen from coal utilizing the ORNL-developed inorganic membrane for separation of hydrogen from syngas. The purpose was to compare hydrogen cost from conventional methods, with and without CO₂ recovery, against the baseline hydrogen fuel plant. Table ES-2 summarizes and compares the performance and economics of the conventional hydrogen plants with the hydrogen fuel plants.

Table ES-2
Comparison of Hydrogen Cost from Conventional and Advanced Plant Designs

	Case 1 Hydrogen from Natural Gas without CO₂ Capture	Case 2 Hydrogen from Natural Gas with CO₂ Capture by Amine Process	Case 4 Conventional Hydrogen from Coal without CO₂ Capture	Case 5 Conventional Hydrogen from Coal with Maximum CO₂ Capture	Baseline Case Advanced Hydrogen Plant with CO₂ 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, MMBtu (MMscfd)	2,868 MMBtu (65.5 MMscfd)	2,640 MMBtu (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 CO₂ removal, would be competitive with hydrogen produced from both natural gas- and coal-based conventional technologies even without CO₂ removal. With only 80 percent hydrogen transport, hydrogen production would still be competitive with conventional coal-based technology.

Hydrogen Fuel from Wyodak Coal/Biomass Blend

The purpose of this study was to compare the economics of producing hydrogen from a Wyodak/biomass blend against producing hydrogen from bituminous coal for plants that have

the same dry coal feedrate. Table ES-3 is a summary comparison of the performance and cost results. The costs of hydrogen from both feedstocks are approximately equal. This is due to a balance of capital charges, fuel costs, and byproduct credits.

Table ES-3
Performance and Cost Summary Comparisons
Hydrogen Fuel Plants with Alternative Feedstocks

	90% Wyodak 10% Biomass	Baseline Case Pittsburgh No. 8 600°C Membrane
Coal Feed	283,833 lb/h	221,631 lb/h
Biomass Feed	31,537 lb/h	N/A
Oxygen Feed (95%) to Gasifier	186,650 lb/h	165,818 lb/h
Oxygen Feed to Retentate Combustor	25,300 lb/h	58,701 lb/h
Water to Prepare Feed Slurry	114,009 lb/h	94,025 lb/h
Hydrogen Product Stream	33,337 lb/h	35,903 lb/h
CO ₂ Product Stream	575,923 lb/h	582,566 lb/h
Sulfuric Acid Product	5,057 lb/h	19,482 lb/h
Gross Power Production		
Turbine Expander	55 MW	84 MW
Steam Turbine	28 MW	N/A
Auxiliary Power Requirement	(69 MW)	(77 MW)
Net Power Production	14 MW	7 MW
Net Plant Water Makeup	100,979 lb/h	198,150 lb/h
Effective Thermal Efficiency, HHV	79.8%	80.4%
Capital Cost, \$1,000	\$365,662	\$359,791
Hydrogen Product Cost, \$/MMBtu	\$5.22 (\$0.65 feedstock) \$5.04 (\$0.50 feedstock)	\$5.06

The amount of hydrogen produced from the Wyodak/biomass blend is lowered by about 7 percent, primarily due to the higher level of CO₂ produced in the gasifier. This resulted in a lowered amount of reactive syngas (H₂ and CO) available for hydrogen production.

Total plant costs are roughly equal, resulting from a combination of increased and decreased equipment requirements. The cost adjustments to the hydrogen plant due to the changeover to the Wyodak/biomass blend are reflected in increased feedstock handling, increased oxygen plant size due to the higher water content (and associated increase in CO₂ content), and the need for a steam turbine that produces 28 MW from excess low-pressure steam. The capital costs were lower in sulfur control areas because of the low-sulfur feedstock, resulting in only 61 tpd sulfuric acid production from the blend versus 234 tpd from bituminous coal. This resulted in a lowering of byproduct credits.

1. INTRODUCTION

The Office of Planning and Environmental Analysis within the Office of Coal and Power Systems (C&PS), renamed Fuel and Power Systems for FY 2002, is responsible for evaluating the reasonableness of C&PS strategic goals, and views the evaluation of innovative systems in fossil energy power generation and liquid fuels production as key elements in that assessment. The Advanced Research Program within the C&PS supports basic research and the development of innovative systems in fossil energy power generation and liquid fuels production. Several research targets have been identified, including low-cost O₂ separation and high-temperature H₂ separation. In support of this program, conceptual systems and cost analyses were developed by the Parsons Corporation for a coal processing plant to produce hydrogen while recovering carbon dioxide (CO₂) for offsite processing or sequestration. This had been referred to as a “decarbonized fuel plant” and is now referred to as a “hydrogen fuel plant.” The scope of work for this analysis entailed the following:

- Identifying alternative processes and technologies utilized for production of hydrogen from coal.
- Reviewing the technical and economic characteristics of developmental materials and technologies for separating hydrogen and oxygen from gas mixtures.
- Conceptualizing process plant designs that utilize developing technologies and materials, resulting in costs of product and CO₂ sequestration significantly lower than with conventional approaches.
- Comparing the costs of a hydrogen fuel plant with plants designed to produce hydrogen from coal utilizing conventional technology.
- Performing sensitivity analyses on the baseline conceptual hydrogen fuel plants to determine the effect of modifying plant design on cost of product.
- Presenting data and results on this study at periodic conferences and workshops.

With an increased interest in greenhouse gas sequestration and production of hydrogen from coal, conceptual designs and resulting economic analyses of syngas and hydrogen plants utilizing conventional technologies were also developed. Throughout the program, certain plant design and economic parameters remained constant to ensure normalized comparisons. These parameters are shown in Table 1-1 and Table 1-2. The conventional approaches included processes such as coal gasification, shift conversion, acid gas removal, and pressure swing adsorption to produce hydrogen. The results of previous studies indicated that the economics of producing syngas and hydrogen from coal by conventional methods is not presently cost competitive.^{1,2,3}

Table 1-1
Consistent Design Parameters (Unless Noted in Text)

Coal	Pittsburgh No. 8
Gasifier Coal Feed	221,631 lb/h as received
Gasifier	E-Gas (Destec two-stage entrained) oxygen-blown
Hydrogen Product	High purity, 346 psia
Sulfur Recovery	Sulfuric acid
CO ₂ Recovery	Low pressure

Table 1-2
Consistent Financial Parameters (Unless Noted in Text)

Cost Basis	Year 2000	
Capacity Factor	Coal-based – 80% Natural gas-based – 90%	
Delivered Cost of: Natural Gas Coal	3.15 \$/MMBtu 1.00 \$/MMBtu	
Project Book Life	20 Years	
Capital:	% of Total	Cost (%)
Common Equity	20	16.5
Debt	80	6.3
Weighted Cost of Capital: (after tax)	6.4%	

An alternative plant was conceived for producing hydrogen from coal utilizing a hydrogen separation device (HSD). The HSD is based on a high-temperature membrane separation concept being developed by Oak Ridge National Laboratory (ORNL)⁴ that can be designed to selectively separate hydrogen from other gases. By utilizing the HSD, it should be possible to separate hydrogen from CO₂ passively and economically.

This report is a compilation of a series of letter reports issued between 1999 and 2001 to document the activity and results from this investigation. Section 2 of this report establishes the baseline plant design for hydrogen production based on the ORNL membrane concept. Section 3 compares these designs to the conventional method of producing hydrogen from natural gas and coal. Section 4 evaluates the HSD based on gasifying a mixture of Wyodak coal and biomass.

2. HYDROGEN FUEL FROM COAL PLANTS

Through mid-1999, designs and cost estimates for hydrogen fuel plants utilizing the inorganic membrane were based on information derived from a 1997 conversation with Oak Ridge National Laboratory.⁴ The reporting and presentation of work associated with the membranes stimulated significant levels of interest in membrane applications, both within the DOE and in private industry. The primary documentation from this activity was a letter report prepared in June 1999.⁵ Nearly two years had passed since the initial information exchange, which led to a meeting held at Eastern Tennessee Technology Park (ETTP) in Oak Ridge in November 1999 to review the status of the properties and characteristics of the inorganic membrane for hydrogen transport. As a result of data gained from the meeting, assumptions applied to the membrane that could have an impact on the baseline plant designs and on future membrane applications were updated.

2.1 HYDROGEN SEPARATION DEVICE PROCESS DESIGN

The HSD is a high-temperature membrane device in a shell and tube configuration, with the high-pressure side being on the inside of the inorganic membrane tubes. The inorganic membrane is designed to have pore sizes of controlled diameters, and it can be made of Al_2O_3 or other ceramic materials. According to ORNL,⁴ the confidential manufacturing process is sufficiently flexible to accommodate a variety of gas compositions and design requirements. The resultant membrane material is analogous to a packed bed through which interstitial pores can be controlled to less than 5 angstroms, while acting like a molecular sieve (that is, it excludes larger molecules).

The separation factor (SF) for hydrogen is high, increasing with higher temperatures. The definition of SF is the rate at which hydrogen passes through, relative to the balance of molecules passing through. For example, the purity of hydrogen resulting from an SF of 1,000 would be calculated as follows:

$$\text{Purity} = (1.0 - 1/1000) \times 100 = 99.90\%$$

The balance is made up of the other gases in the initial mixture.

Pressurized syngas, to which steam has been added, enters the tube side of the HSD, which is assumed to have gas contact catalytic properties that promote the water-gas shift reaction. Possibly this will be achieved by lining the inner tube surfaces with catalytic material. However, the hydrogen-deficient surface, resulting from hydrogen migrating through the membrane, will also promote water-gas shift reaction.

It was assumed that, as hydrogen is extracted from the gas stream through the HSD membrane, gas composition at the catalytic surface will become hydrogen deficient and, with excess steam, equilibrium will be shifted to convert available CO to CO_2 and hydrogen. The hydrogen will then migrate to the HSD membrane surface and be transported across. Eventually the CO will reach equilibrium with the steam at the system temperature.

The HSD transports hydrogen across the membrane in proportion to the relative hydrogen partial pressure differentials, where P_1 equals the upstream hydrogen pressure, and P_2 equals the product hydrogen pressure. The initial HSD was designed to operate at an equilibrium temperature of 761°C (1402°F) and at 950 psia. At 950 psia upstream (assuming 42 percent hydrogen), 95 percent of the hydrogen will be separated, with a downstream pressure of 20 psia, according to the following relationship:

$$\begin{aligned} \text{H}_2 \text{ transport} &= (1 - P_2/P_1) \times 100 = \% \text{ transport} \\ &= (1 - 20/(950 \times 0.42)) \times 100 = 95\% \text{ transport} \end{aligned}$$

Fuel value remaining in the separated gas (or retentate) will be about 5 percent of the original feed gas. At that temperature, the hydrogen purity will be better than 99.5 percent. The hydrogen stream leaving the HSD at 20 psia and 1402°F passes through a heat recovery steam generator (HRSG) and a compressor, which reduces hydrogen stream temperature to 117°F and compresses the hydrogen to 346 psia. The steam raised in the HRSG is added to the steam that is injected into the raw gas coming from the gasifier to promote the shift reaction.

2.1.1 UPDATED ASSUMPTIONS FOR HSD DESIGN

The HSD design was modified, based on revised assumptions. This discussion incorporates the latest thinking on inorganic membranes from ETTP at Oak Ridge. The design basis for the revised HSD is the result of conversations with the membrane developers at ETTP. At that meeting, the characteristics of the membrane were identified, and changes were indicated according to Table 2-1. Note that many of the assumptions remain the same. Most significant is the reduced operating temperature.

Table 2-1
Revised Assumptions for Hydrogen Separation Device

	Original Assumptions	Revised Assumptions
Separation Factor	>200	~200
Hydrogen Transport Flux	0.1 cc/min/cm ² /cmHgPH ²	0.1 cc/min/cm ² /cmHgPH ²
CO Shift Reaction	On surface of membrane without catalyst	On surface of membrane without catalyst
HSD Tube Size	60 mm OD	0.625 inch OD 0.50 inch ID
Gas Exit Temperature	1402°F (761°C)	1112°F (600°C) 572°F (300°C)
Membrane Cost	\$100/ft ²	\$100/ft ²
Number of Tubes per 8-Foot Vessel Diameter	4,096	11,800
Vessel Pressurization	Shell side	Tube side

The security classification of inorganic membranes is now described in an exclusive “Classification Guide for Inorganic Membranes.” This document was prepared within the last two years. The security level is confidential, essentially requiring a “need to know” for the

manufacturing procedure. The membranes themselves are unclassified. ETTP is now able to talk “freely” about the performance and characteristics of the membranes, but refrains from describing the manufacturing process.

At the meeting, results of the baseline hydrogen fuel plants were discussed, and it was agreed that the HSD can be instrumental in lowering the cost of hydrogen from coal. Parsons’ assumptions were reviewed, which led to discussion of the membrane status and ETTP’s thoughts on membrane performance. ETTP gave Parsons a paper presented at the Pittsburgh Coal Conference by Douglas Fain⁶. In it, the transport mechanism is discussed, along with several comments on the Parsons paper from May 1998.⁷

This led to discussing Parsons’ large membrane vessel. ETTP views the large vessel with many internal tubes as conventional technology. A picture of an ORNL gaseous diffusion vessel was shown, and its dimensions were comparable to the Parsons conceptual vessel. ETTP commented that the pressurization should be from the tube side and the tubes should be smaller diameter, about 10 to 15 mm. Internal pressurization would avoid the cost of a high-pressure shell.

The following salient points were discussed regarding design assumptions for the HSD:

- Separation Factor – The SFs provided by ORNL in 1997 were based on binary gas mixtures. SFs are still based on binary mixtures. At 300°C, the hydrogen SF relative to all gases is ~150.
- Hydrogen Flux through the Membrane – Parsons’ initial assumption was that the gas flux was purely proportional to the hydrogen partial pressure differential. ETTP stated that temperature increase aids hydrogen flow in a non-Knudsen manner. The flux can increase with thinner membranes, but the risk of membrane defect increases.
- Operating Pressure and Temperature – ETTP believes that a 1000 psi pressure differential can be contained by the inorganic membrane. The furnace temperature that is used for testing can reach 650°C, and the operational goal for the membranes is currently 600°C. ETTP uses glass seals to join the membrane tubes to the end support tubes, and these are satisfactory up to 300°C. As a result, a vessel design could be prepared to operate with confidence up to 300°C.
- Sulfur and Steam Tolerance – In reply to being asked if steam or H₂S has any effect on the membrane structure, the ETTP response was negative.
- CO Shift Properties and Requirement for Sweep Gas – Testing is planned, but there are no empirical data available. Parsons will continue to assume shift reaction on the membrane surface without catalyst.
- Membrane Costs – Parsons’ assumptions of \$100/ft² are still “in the ball park.”

Table 2-1 summarizes the revised assumptions for the inorganic membrane and the HSD.

2.1.2 HSD DESIGN

A review of literature for designing vessels for gas separation membranes showed that two approaches can be taken regarding the gas flow to the membrane surface area.⁸ The

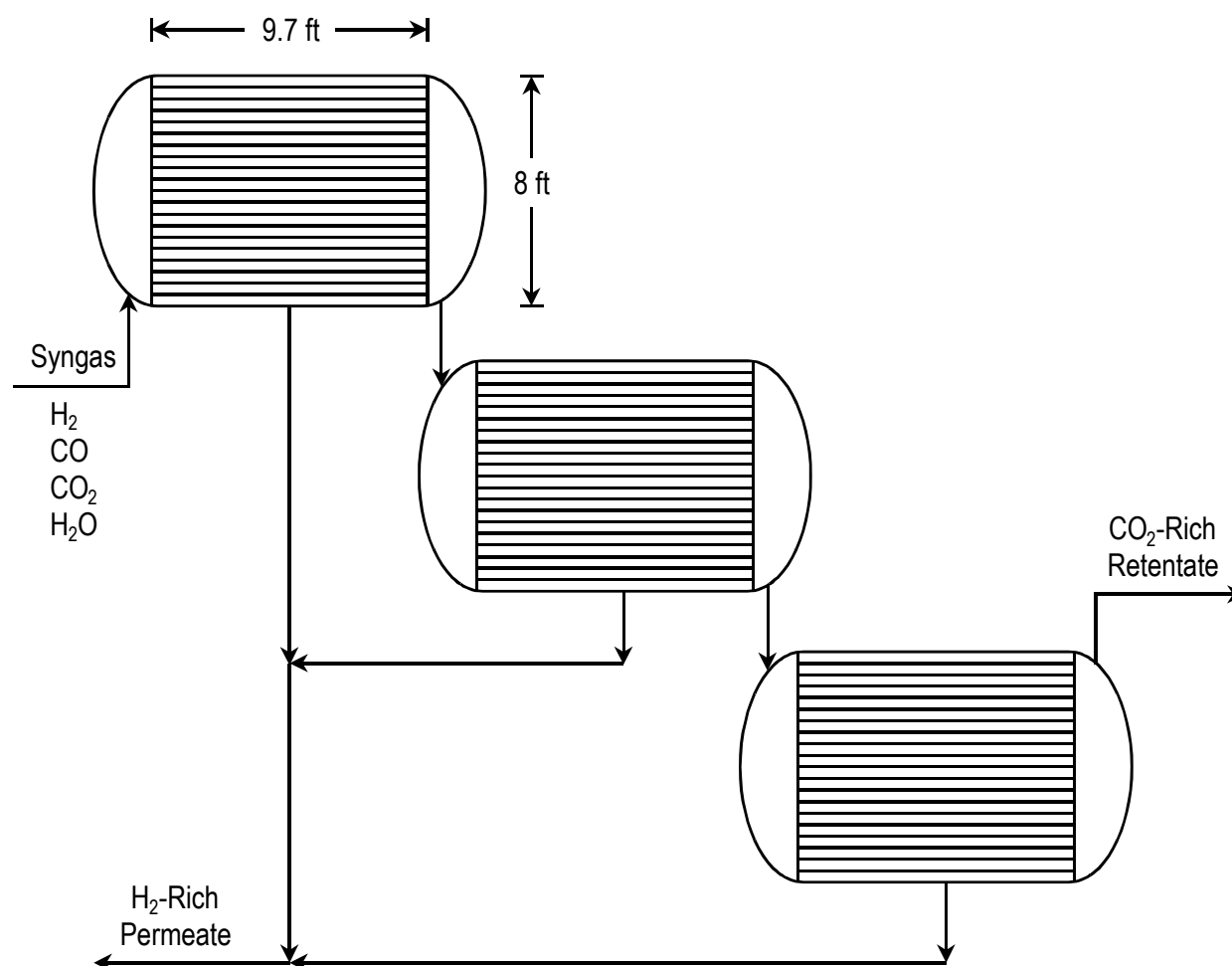
characterization of membrane flux in the laboratory utilizes a long retention time. It appears that an inlet gas flow is used, which results in a retention time of one minute at the membrane. Conversely, concerns were raised as the design of a full-size vessel for integrated gasification combined cycle (IGCC) applications were discussed. The primary concern was the existence of a laminar boundary layer at the membrane surface, which would reduce membrane flux. The final design used a flow velocity of 10 feet per second, and a retention time of 1 second to achieve turbulent flow. The primary difference between the referenced design and the HSD design is their use of catalyst to promote the shift reaction. Their design included shift catalyst in the initial section of membrane tubing, which also created a turbulent region.

The HSD design retains the previous concept to promote the shift reaction by product extraction at the membrane surface. The scenario is based on the gas proceeding along the membrane surface in turbulent flow. Hydrogen product partial pressure is both maintained and extracted at the membrane surface. CO continues to react with steam until the CO-steam equilibrium is reached. The remaining gas then passes from the membrane without further reaction. To ensure the shift reaction going to completion, the membrane path was increased 25 percent above theoretical. Conceptually, the HSD resembles the sketch in Figure 2-1. The basis for the HSD design is shown in Table 2-2.

Table 2-2
Hydrogen Separation Device Designs

	Initial Design 761°C HSD	Revised Design 600°C HSD	Revised Design 300°C HSD
Hydrogen Production	35,205 lb/h 1402°F	35,903 lb/h 1112°F	36,564 lb/h 571°F
Syngas Inlet Conditions	684,000 lb/h 1000 psia, 956°F 12,228 acfm	684,000 lb/h 1000 psia, 605°F 10,382 acfm	684,000 lb/h 1000 psia, 404°F 6,771 acfm
Minimum Membrane Area	35,205 ft ²	35,903 ft ²	36,564 ft ²
Minimum Membrane Area Increased by ~25% to Reach Design	45,000 ft ²	45,000 ft ²	45,000 ft ²
Vessel Diameter	8 ft ID	8 ft ID	8 ft ID
Tube Dimensions	0.625 inch OD 0.50 inch ID	0.625 inch OD 0.50 inch ID	0.625 inch OD 0.50 inch ID
Tubes per Vessel	11,800	11,800	11,800
Preliminary Tube Length	29 ft	29 ft	29 ft
Gas Velocity through Tubes	12.7 ft/sec	10.8 ft/sec	7.0 ft/sec
Gas Retention Time	2.3 sec	2.7 sec	4.1 sec
Reynolds Number	~19,000	~22,500	~28,000
Number of Vessels and Configuration of Tube Bundle	3 vessels 8 x 9.7 ft	3 vessels 8 x 9.7 ft	3 vessels 8 x 9.7 ft
Vessel Flow Arrangement	Series	Series	Series

Figure 2-1
Hydrogen Separation Device Concept



2.2 HYDROGEN FUEL PLANT -- 1402°F (761°C) MEMBRANE

This hydrogen fuel production facility conceptual plant design was the initial plant used to evaluate the conversion of coal to synthesis gas, and achieve essentially total separation of hydrogen from the CO_2 . This concept utilizes hot gas desulfurization and particulate removal upstream of the HSD along with a modern non-ATS (conventional) gas turbine in the CO_2 -rich stream. Table 2-3 provides the design basis established for the plant.

This concept reduces the temperature of the fuel gas stream from the gasifier to 1100°F before the gas is desulfurized and filtered in a transport reactor desulfurizer and a ceramic candle filter. This eliminates the need for a downstream flue gas desulfurization (FGD) unit.

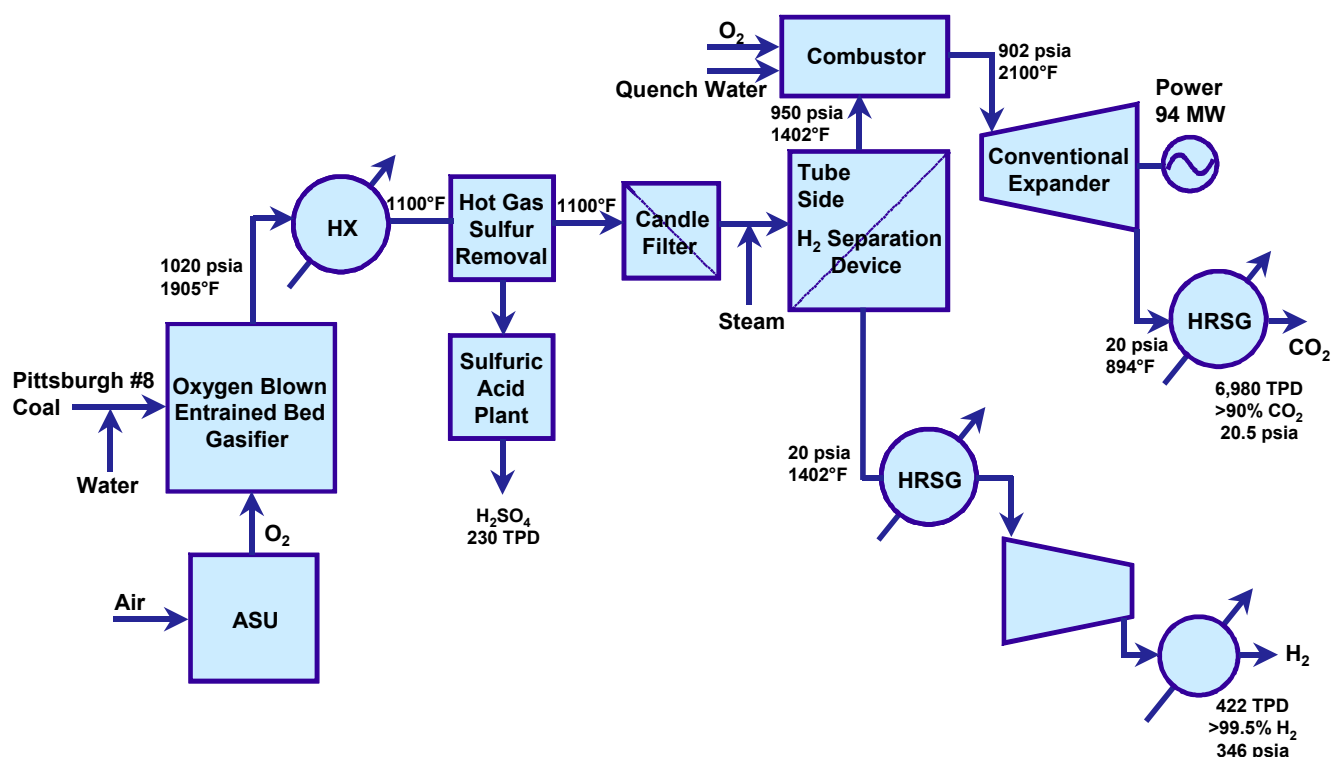
Table 2-3
Design Basis for Hydrogen Fuel Production Facility
with Conventional Expansion Turbine and Hot Gas Cleanup

Hydrogen Fuel Production Facility Parameter	Hydrogen Fuel Production Facility Plant Design Basis
Coal Feed	Pittsburgh No. 8, <10% ash
Limestone Sorbent	None
Gasifier	Oxygen-blown Destec with second stage adjusted for 1905°F output
Hot Gas Temperature	1905°F
Gasifier Outlet Pressure	1000 psia
Ambient Conditions	14.7 psia, 60°F
Hot Gas Desulfurization	Yes, 1100°F
Sulfur Recovery	Sulfuric acid
Ceramic Candle Filter	Before HSD
Hydrogen Separation	H ₂ separation device Shell and tube configuration 95% separation 99.5% pure H ₂ Zero sulfur 20 psia hydrogen compressed to 346 psia
Separated Gas	CO shifted to 1402°F equilibrium 5% of fuel value in gas 950 psia
Separated Gas Utilization	Combustion with oxygen Steam injection conventional turbine expander
CO ₂ Product Pressure	19.4 psia
Hydrogen Utilization	346 psia offsite
Auxiliary Power Block	Conventional turbine expander
Plant Size	Maximum H ₂ production from 2,500 tpd dry gasifier Excess power sold offsite

A block flow diagram of the plant is shown on Figure 2-2. The flows and state points on the attached process flow diagram (Figure 2-3) result from the heat and material balance for the plant. Key process components included in the plant are a Destec high-pressure slurry-feed gasifier, the ORNL HSD, and the transport reactor for desulfurization.

The selected processes exhibit some unique features that result in a simplification of plant design and may contribute to lowering of capital cost. 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. The hot raw gas is cleaned of larger particulates in a cyclone and then is cooled in a firetube boiler to 1100°F. A hot gas cleanup system consisting of a transport reactor desulfurizer and a ceramic candle filter removes sulfur and particulates from the fuel gas stream. Sulfur is recovered as sulfuric acid.

Figure 2-2
Block Flow Diagram
Baseline Hydrogen Fuel Plant
1402°F (761°C) Hydrogen Separation Device



A considerable amount of steam is added, ensuring adequate water content for the high-temperature shift reaction to occur. The gas enters the HSD at 956°F and leaves the HSD at 1402°F as a result of the exothermic shift reaction. The hydrogen produced from the HSD is 99.5 percent pure. It goes through a HRSG and then is compressed to 346 psia.

The CO₂-rich gas leaving the HSD at 950 psia contains about 5 percent of the fuel value of the inlet syngas stream. This gas goes to the gas turbine combustor with which oxygen is injected to convert CO and hydrogen to CO₂ and H₂O, respectively. Water is also injected into the combustor to moderate the temperature to 2100°F. The hot gas is expanded to 20 psia and 894°F through the conventional gas turbine expander to produce 94 MW electric power. The gas is cooled in a HRSG, and steam produced is combined with other steam produced from cooling the hydrogen for process applications. There is no power produced from steam. The CO₂ product is cooled to 100°F, dewatered, and sent offsite.

Page 2-8

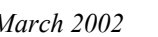


Table 2-4 presents the performance summary for the plant, and Table 2-5 identifies the plant power requirements.

Table 2-4
Performance Summary

Coal Feed	221,631 lb/h
Oxygen Feed (95%)	231,218 lb/h
Hydrogen Product Stream	35,205 lb/h
CO ₂ Product Stream	581,657 lb/h
Sulfuric Acid Product	19,482 lb/h
Gross Power Production	94 MW
Auxiliary Power Requirement	76 MW
Net Power Production	18 MW
Effective Thermal Efficiency (ETE), HHV	80.2%

Table 2-5
Auxiliary Power Load, kW

Gasifier Auxiliary Oxygen Compressor	10,300
Combustor Oxygen Compressor	4,000
ASU Air Compressor	30,900
Gasifier Slurry Pumps	190
Water Spray Pump	350
Boiler Feedwater Pumps	640
Coal Handling	210
Slag Handling	530
Regenerator Air Blower	2,960
Gas Turbine Auxiliary	400
Hydrogen Compressor	25,180
Miscellaneous Balance of Plant	750
Total Auxiliary Load	76,410 kW

Following are more detailed descriptions of the key process elements.

2.2.1 PROCESS DESCRIPTION

2.2.1.1 GASIFIER

The high-pressure system for producing hydrogen has resulted in utilizing two Destec gasifier trains, each having a capacity of 1,250 tpd coal (dry basis).⁵ The Destec high-pressure entrained flow gasifier consists of two stages to gasify a coal-water slurry feed with oxygen. The slurry is prepared by fine grinding coal to about 200 mesh and mixing with water to achieve a ratio of 65 percent solids and 35 percent water, including the moisture content of the coal. The gasifier can operate at any pressure up to the capability of the oxygen compressor. By operating in two

stages, it is possible to adjust the flow split between stages to achieve a desired outlet temperature of the product gas. A typical operating temperature for the Destec gasifier is 1900°F. This temperature is reached 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 gasifiers at 1905°F goes through an internal cyclone that separates entrained particles from the gas for recycle to the gasifiers, a firetube boiler to cool the gas to 1100°F, and a hot gas cleanup system. Steam is then injected into the gas stream, promoting the shift reaction, which will occur downstream in the HSD.

2.2.1.2 AIR SEPARATION UNIT

Oxygen supply for this plant is provided through a conventional cryogenic air separation unit (ASU). The air separation plant is designed to produce a nominal output of 3,000 tons/day of 95 percent pure O₂. 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.

2.2.1.3 HOT GAS CLEANUP SYSTEM

The transport reactor desulfurizer consists of a riser tube, a disengager, and a standpipe for both the absorber section and regeneration section. Sorbent from the absorber passes through the regenerator riser, disengages, and transfers back to the absorber through the standpipe. Regeneration is conducted with neat air to minimize heat release and limit temperature. The regeneration heat has negligible effect on the sorbent temperature in the absorber. The regeneration off-gas containing predominantly SO₂ is sent to the sulfuric acid plant. Elutriated particles are disengaged from the gas by high-efficiency cyclones at the top of the absorber. A final ceramic candle filter is located downstream.

2.2.1.4 SULFURIC ACID PLANT

Key to the double-absorption contact sulfuric acid plant process is use of an intermediate absorber in the four-pass converter developed by Monsanto. The reaction from SO₂ to SO₃ is an exothermic reversible reaction. Using a vanadium catalyst, a contact plant takes advantage of both rate and equilibrium considerations by first allowing the gases to enter over a part of the catalyst at about 800°F, and then allowing the temperature to increase adiabatically as the reaction proceeds. The reaction essentially stops when about 60 to 70 percent of the SO₂ has been converted, at a temperature in the vicinity of 1100°F. The gas is cooled in a waste heat boiler and passed through subsequent stages until the temperature of the gases passing over the last portion of catalyst does not exceed 800°F. The gases leaving the converter, having passed through two or three layers of catalyst, are cooled and passed through an intermediate absorber tower where some of the SO₃ is removed with 98 percent H₂SO₄. The gases leaving this tower are then reheated, and flow through the remaining layers of catalyst in the converter. The gases are then cooled and pass through the final absorber tower before discharge to the atmosphere. In

this manner, more than 99.7 percent of the SO₂ is converted into SO₃ and subsequently into product sulfuric acid.

CO₂-Rich Separated Gas Stream/Conventional Turbine Expander

The gas, which is separated from the hydrogen, leaves the HSD at 950 psia and 1407°F, and has a fuel value of about 15 Btu/scf. A conventional expansion turbine is utilized to extract the energy from the gas stream by producing power and steam. The gas stream is fired with oxygen in the combustor, resulting in conversion of CO and hydrogen to CO₂ and water vapor, respectively. Water is injected into the combustor to moderate the stream temperature to 2100°F, making it suitable for expansion through the turbine expander. The turbine expander reduces the gas pressure to 20 psia and its temperature to 894°F, while generating 94 MW power. In-plant power requirements and transformer losses amount to 76 MW, resulting in export power sales of 18 MW. The gas then passes through a HRSG where it is cooled to 250°F, while raising steam for in-plant process use. The CO₂ product is cooled to 100°F, dried, and sent offsite. Table 2-6 identifies the overall water balance for the plant.

**Table 2-6
Plant Water Balance**

Water Source	
Makeup Water	193,426 lb/h
Recycled from Stack Condenser	288,382 lb/h
Water Consumption Point	
Boiler Feed	320,084 lb/h
Gasifier Coal Slurry Preparation	94,025 lb/h
Combustor Quench	64,118 lb/h
Sulfuric Acid Water	3,581 lb/h

2.2.2 EFFECTIVE THERMAL EFFICIENCY (ETE)

For comparative purposes and to arrive at a figure of merit for the plant design, an ETE was derived for the plant performance based on HHV thermal value of hydrogen produced and offsite power sales, divided by the fuel input to the plant. The formula is:

$$\text{ETE} = \frac{(\text{Hydrogen Heating Value} + \text{Electrical Btu Equivalent})}{\text{Fuel Heating Value (HHV)}}$$

$$\text{ETE} = \frac{35,205 \text{ lb H}_2/\text{h} \times 61,095 \text{ Btu/lb} + 18,000 \text{ kW} \times 3,414 \text{ Btu/kWh}}{221,631 \text{ lb coal/h} \times 12,450 \text{ Btu/lb}}$$

$$\text{ETE} = 80.2\%$$

2.3 HYDROGEN FUEL PLANT – 1112°F (600°C) MEMBRANE

Utilizing the revised assumptions for the HSD, updated plant concepts were prepared for HSD operation at 1112°F (600°C). The initial plant design was modified to determine the impact of the revised assumptions and of changing the operating temperature of the HSD to 1112°F (600°C) equilibrium. This concept utilizes the same hot gas desulfurization and particulate removal upstream of the HSD along with a modern non-ATS (conventional) gas turbine in the CO₂-rich stream. Table 2-7 provides the design basis established for the plant.

Table 2-7
Design Basis for Baseline Hydrogen Fuel Plant
1112°F (600°C) Membrane

Hydrogen Fuel Production Facility Parameter	Hydrogen Fuel Production Facility Plant Design Basis
Coal Feed	Pittsburgh No. 8, <10% ash
Limestone Sorbent	None
Gasifier	Oxygen-blown Destec with second stage adjusted for 1905°F output
Hot Gas Temperature	1905°F
Gasifier Outlet Pressure	1000 psia
Ambient Conditions	14.7 psia, 60°F
Hot Gas Desulfurization	Yes, 1100°F
Sulfur Recovery	Sulfuric acid
Ceramic Candle Filter	Before HSD
Hydrogen Separation	H ₂ separation device Shell and tube configuration 95% separation 99.5% pure H ₂ Zero sulfur 20 psia hydrogen compressed to 346 psia
Separated Gas	CO shifted to 1112°F equilibrium 5% of fuel value in gas 950 psia
Separated Gas Utilization	Combustion with oxygen Steam injection conventional turbine expander
CO ₂ Product Pressure	19.4 psia
Hydrogen Utilization	346 psia offsite
Auxiliary Power Block	Conventional turbine expander
Plant Size	Maximum H ₂ production from 2,500 tpd dry gasifier Excess power sold offsite

2.3.1 PROCESS DESCRIPTION

A block flow diagram of the plant is shown on Figure 2-4. The flows and state points on the attached process flow diagram (Figure 2-5) result from the heat and material balance for the plant. Key process components included in the plant are a Destec high-pressure slurry-feed gasifier, the ETTP HSD, and the transport reactor for desulfurization. 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. The hot raw gas is cleaned of larger particulates in a cyclone and then is cooled in a firetube boiler to 1100°F. A hot gas cleanup system consisting of a transport reactor desulfurizer and a ceramic candle filter removes sulfur and particulates from the fuel gas stream. Sulfur is recovered as sulfuric acid.

For this case, both steam and water are added to cool the syngas while ensuring adequate water content for the high-temperature shift reaction to occur at the lower HSD inlet temperature. The gas enters the HSD at 605°F and leaves the HSD at 1112°F as a result of the exothermic shift reaction. The hydrogen produced from the HSD is 99.5 percent pure. It goes through a HRSG and then is compressed to 346 psia.

Figure 2-4
Block Flow Diagram
Baseline Hydrogen Fuel Plant
1112°F (600°C) Hydrogen Separation Device

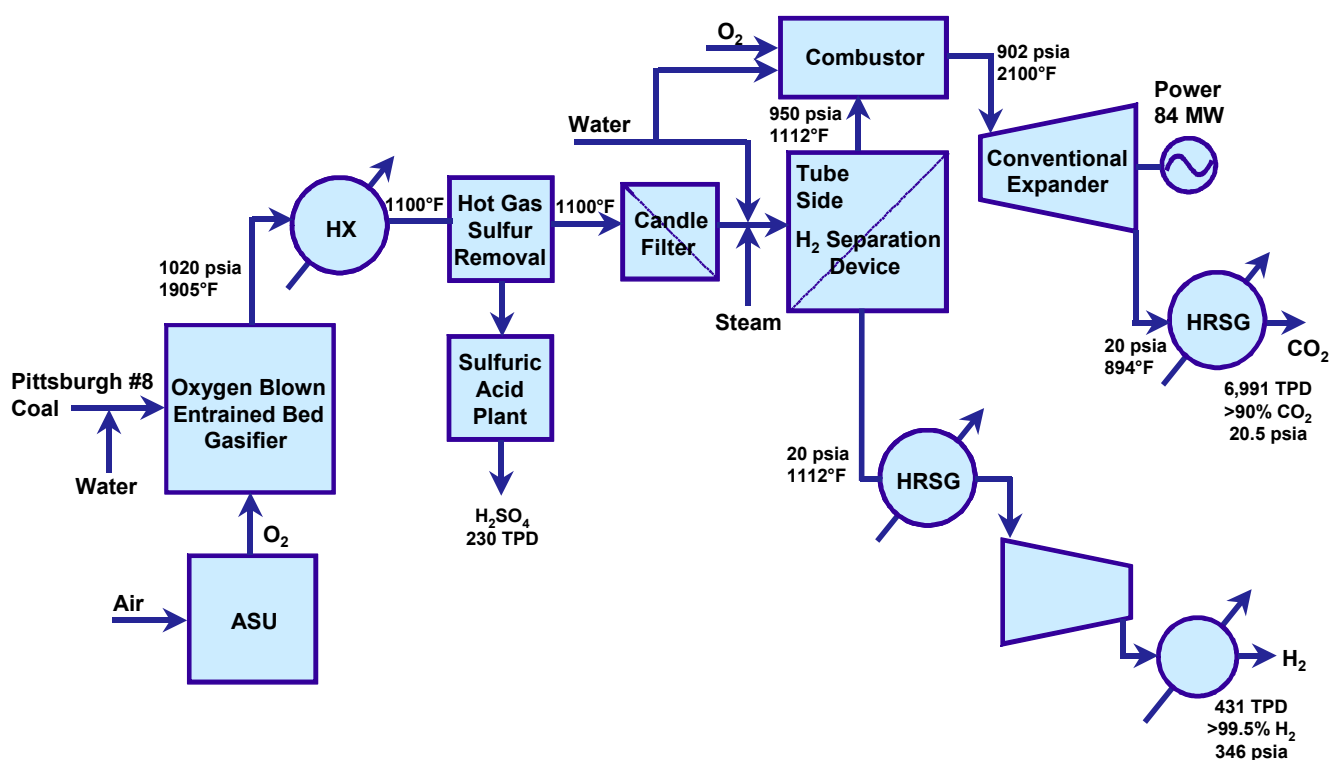
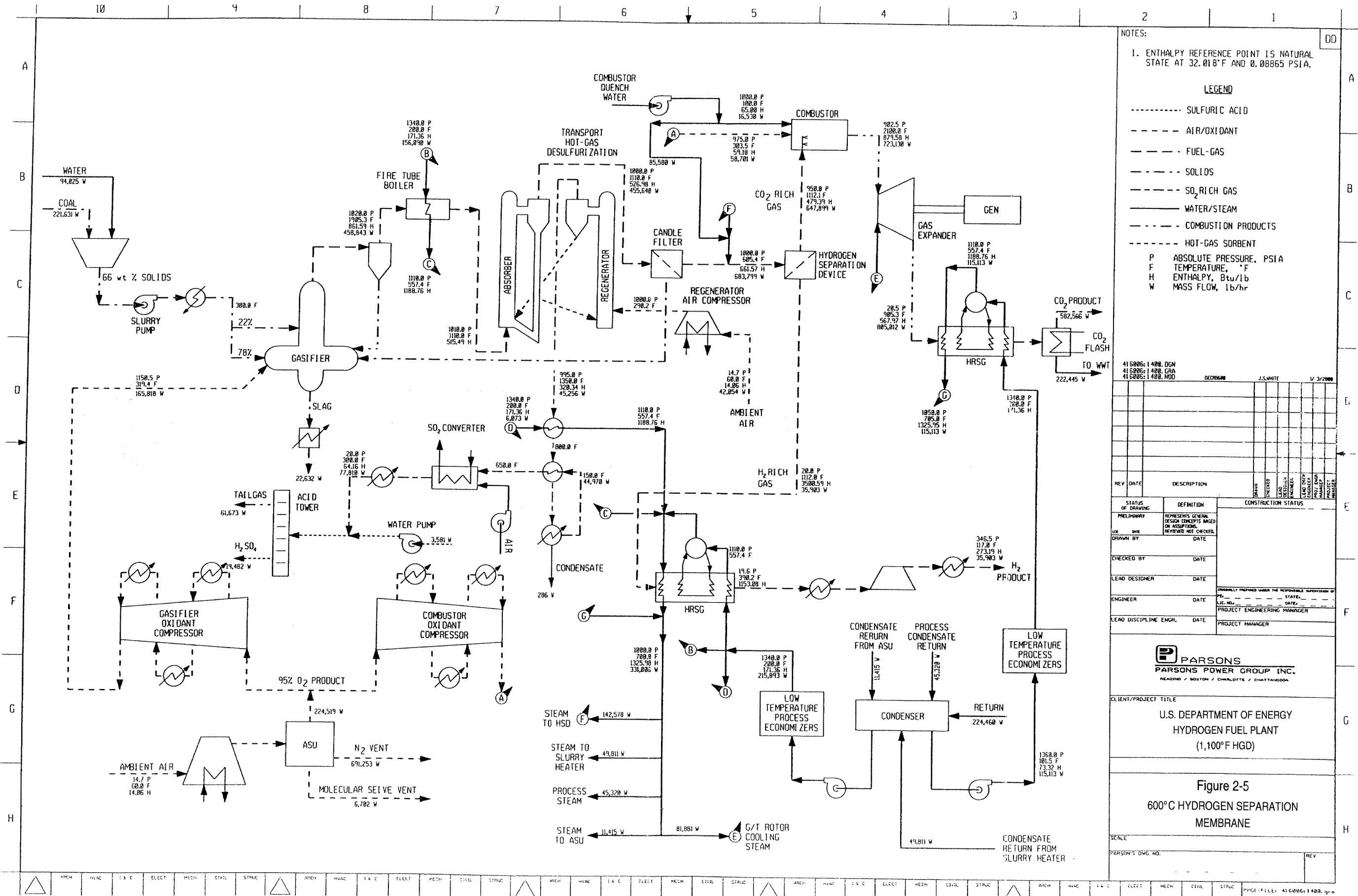


Figure 2-5



The CO₂-rich gas leaving the HSD at 950 psia contains about 5 percent of the fuel value of the inlet syngas stream. This gas goes to the gas turbine combustor with which oxygen is injected to convert CO and hydrogen to CO₂ and H₂O, respectively. Both steam and water are injected into the combustor to moderate the temperature to 2100°F. The hot gas is expanded to 20 psia and 905°F through the conventional gas turbine expander to produce 84 MW electric power. The gas is cooled in a HRSG, and the steam produced is combined with other steam produced from cooling the hydrogen for process applications. There is no power produced from steam. The CO₂ product is cooled to 100°F, dried, and sent offsite.

Table 2-8 presents the performance summary for the plant, and Table 2-9 identifies the plant power requirements. Table 2-10 is the plant water balance.

Table 2-8
Performance Summary

Coal Feed	221,631 lb/h
Oxygen Feed (95%)	224,519 lb/h
Hydrogen Product Stream	35,903 lb/h
CO ₂ Product Stream	582,566 lb/h
Sulfuric Acid Product	19,482 lb/h
Gross Power Production	84 MW
Auxiliary Power Requirement	77 MW
Net Power Production	7 MW
Effective Thermal Efficiency (ETE), HHV	80.4%

Table 2-9
Auxiliary Power Load, kW

Gasifier Auxiliary Oxygen Compressor	10,300
Combustor Oxygen Compressor	4,000
ASU Air Compressor	30,900
Gasifier Slurry Pumps	190
Water Spray Pump	400
Boiler Feedwater Pumps	510
Coal Handling	210
Slag Handling	530
Regenerator Air Blower	2,960
Gas Turbine Auxiliary	400
Hydrogen Compressor	25,690
Miscellaneous Balance of Plant	750
Total Auxiliary Load	76,840 kW

Table 2-10
Plant Water Balance

Water Source	
Makeup Water	198,150 lb/h
Recycled from Stack Condenser	222,445 lb/h
Water Consumption Point	
Boiler Feed	224,460 lb/h
Gasifier Coal Slurry Preparation	94,025 lb/h
Shift Water	85,580 lb/h
Quench	16,530 lb/h

2.3.2 EFFECTIVE THERMAL EFFICIENCY

For comparative purposes and to arrive at a figure of merit for the plant design, an ETE was derived for the plant performance based on HHV thermal value of hydrogen produced and offsite power sales, divided by the fuel input to the plant. The formula is:

$$\text{ETE} = \frac{(\text{Hydrogen Heating Value} + \text{Electrical Btu Equivalent})}{\text{Fuel Heating Value (HHV)}}$$

$$\text{ETE} = \frac{35,903 \text{ lb H}_2/\text{h} \times 61,095 \text{ Btu/lb} + 7,060 \text{ kW} \times 3,414 \text{ Btu/kWh}}{221,631 \text{ lb coal/h} \times 12,450 \text{ Btu/lb}}$$

$$\text{ETE} = 80.4\%$$

2.4 HYDROGEN FUEL PLANT – 572°F (300°C) MEMBRANE

Utilizing the revised assumptions for the HSD, updated plant concepts were prepared for HSD operation at 572°F (300°C) to determine the impact of lowering the operating temperature of the HSD. The plant design basis is shown in Table 2-11.

Table 2-11
Design Basis for Baseline Hydrogen Fuel Plant
572°F (300°C) Membrane

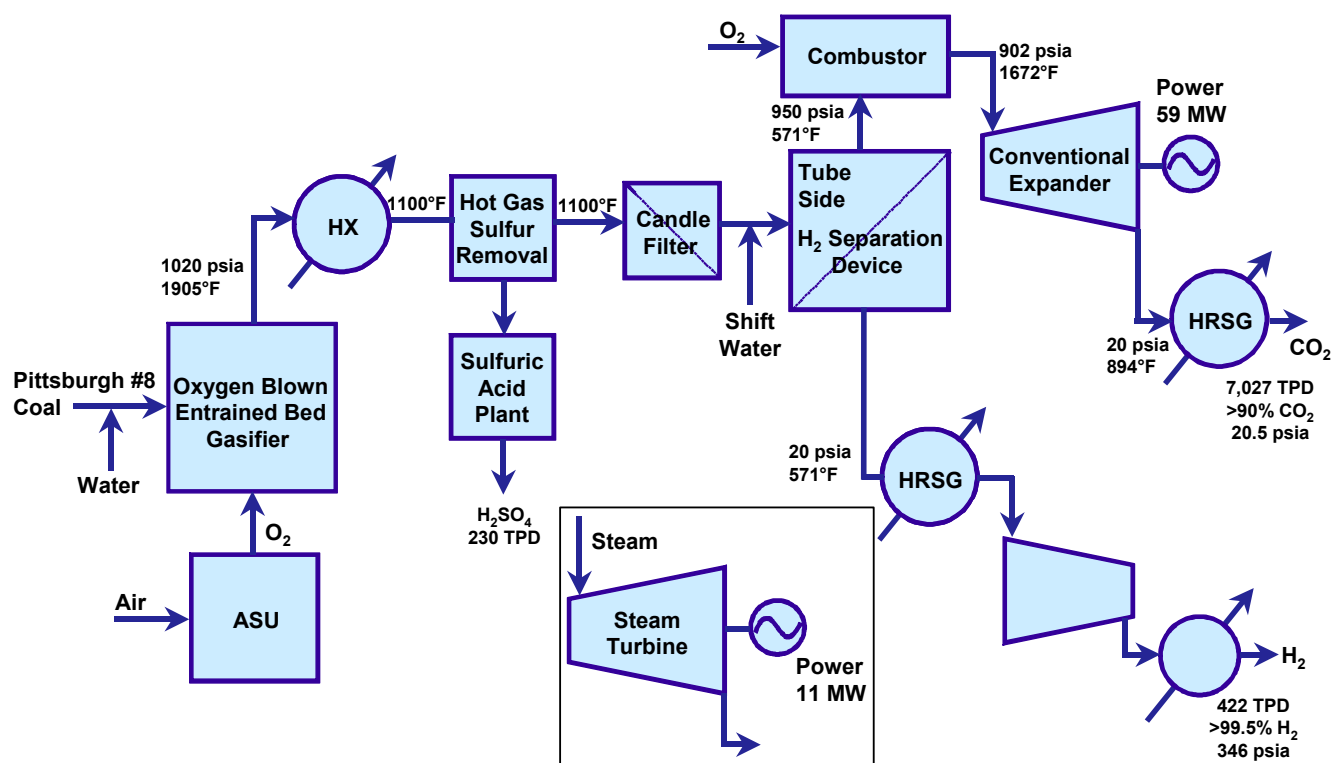
Hydrogen Fuel Production Facility Parameter	Hydrogen Fuel Production Facility Plant Design Basis
Coal Feed	Pittsburgh No. 8, <10% ash
Limestone Sorbent	None
Gasifier	Oxygen-blown Destec with second stage adjusted for 1905°F output
Hot Gas Temperature	1905°F
Gasifier Outlet Pressure	1000 psia
Ambient Conditions	14.7 psia, 60°F
Hot Gas Desulfurization	Yes, 1100°F
Sulfur Recovery	Sulfuric acid
Ceramic Candle Filter	Before HSD
Hydrogen Separation	H ₂ separation device Shell and tube configuration 95% separation 99.5% pure H ₂ Zero sulfur 20 psia hydrogen compressed to 346 psia
Separated Gas	CO shifted to 572°F equilibrium 5% of fuel value in gas 950 psia
Separated Gas Utilization	Combustion with oxygen Steam injection conventional turbine expander
CO ₂ Product Pressure	19.4 psia
Hydrogen Utilization	346 psia offsite
Auxiliary Power Block	Conventional turbine expander
Plant Size	Maximum H ₂ production from 2,500 tpd dry gasifier

2.4.1 PROCESS DESCRIPTION

A block flow diagram of the plant is shown on Figure 2-6. The flows and state points on the attached process flow diagram (Figure 2-7) result from the heat and material balance for the plant.

For this case, only water is added to cool the syngas, ensuring adequate water content for the high-temperature shift reaction to occur at the lower HSD inlet temperature. The gas enters the HSD at 404°F and leaves the HSD at 571°F as a result of the exothermic shift reaction. The hydrogen produced from the HSD is 99.5 percent pure. It goes through a HRSG and then is compressed to 346 psia.

Figure 2-6
Block Flow Diagram
Baseline Hydrogen Fuel Plant
572°F (300°C) Hydrogen Separation Device



The CO₂-rich gas leaving the HSD at 950 psia contains about 5 percent of the fuel value of the inlet syngas stream. This gas goes to the gas turbine combustor with which oxygen is injected to convert CO and hydrogen to CO₂ and H₂O, respectively. Since gas exits the combustor at 1672°F, it is not necessary to add steam or water to moderate the temperature. The hot gas is expanded to 20 psia and 782°F through the conventional gas turbine expander to produce 59 MW electric power. The gas is cooled in a HRSG, and the steam produced is combined with other steam produced from cooling the hydrogen for process applications. This plant has excess steam, which is used to produce 11 MW power. The CO₂ product is cooled to 100°F, dried, and sent offsite.

Figure 2-7

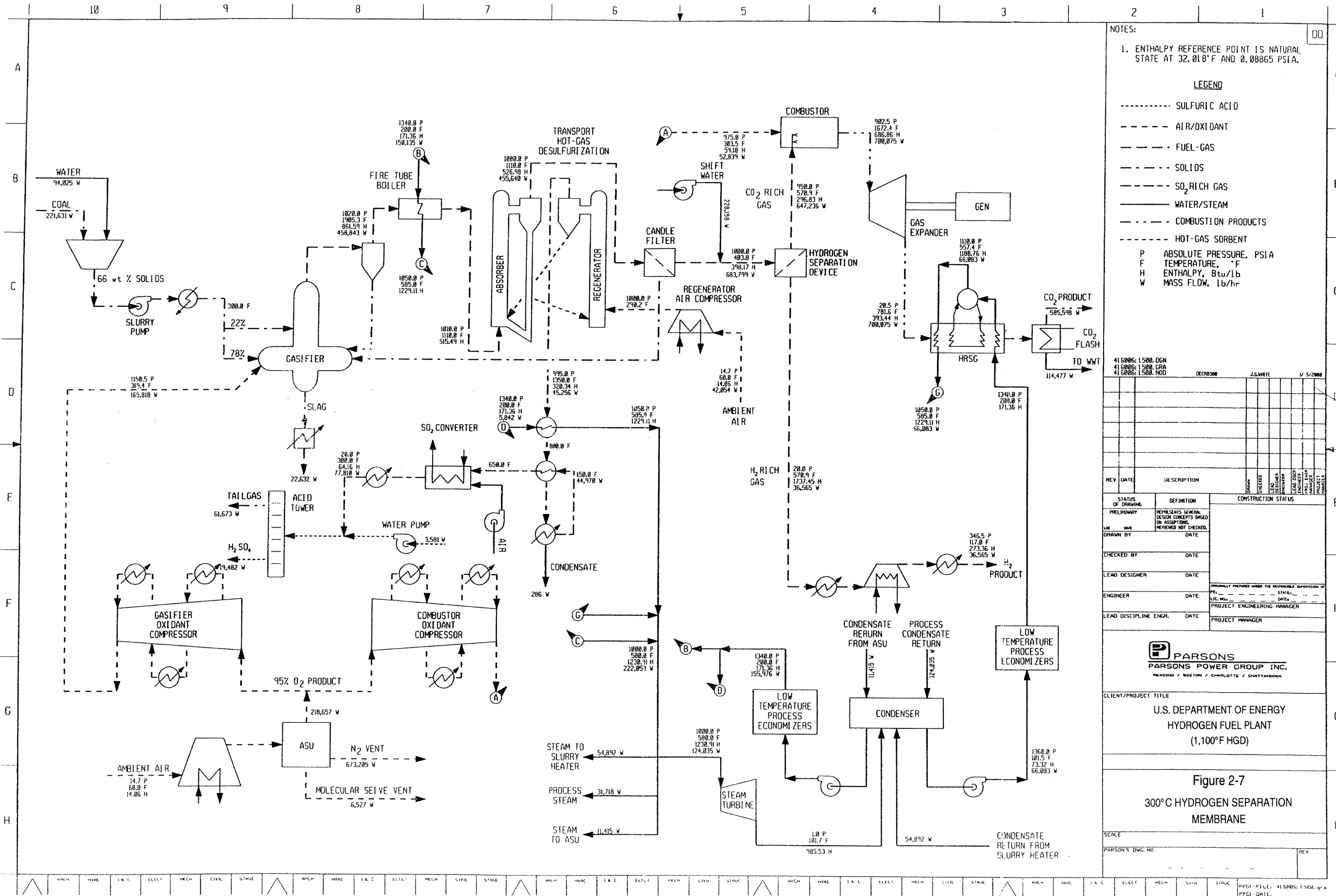


Table 2-12 presents the performance summary for the plant, and Table 2-13 identifies the plant power requirements. Table 2-14 shows the plant water balance.

Table 2-12
Performance Summary

Coal Feed	221,631 lb/h
Oxygen Feed (95%)	218,657 lb/h
Hydrogen Product Stream	36,565 lb/h
CO ₂ Product Stream	585,598 lb/h
Sulfuric Acid Product	19,482 lb/h
Gas Turbine Gross Power	59 MW
Steam Turbine Gross Power	11 MW
Auxiliary Power Requirement	76 MW
Net Power Production	(6 MW)
Effective Thermal Efficiency (ETE), HHV	80.3%

Table 2-13
Auxiliary Power Load, kW

Gasifier Auxiliary Oxygen Compressor	10,300
Combustor Oxygen Compressor	3,200
ASU Air Compressor	29,220
Gasifier Slurry Pumps	190
Water Spray Pump	400
Boiler Feedwater Pumps	340
Coal Handling	210
Slag Handling	530
Regenerator Air Blower	2,960
Gas Turbine Auxiliary	400
Steam Turbine Auxiliary	400
Cooling Tower	890
Hydrogen Compressor	26,180
Miscellaneous Balance of Plant	750
Total Auxiliary Load	75,970 kW

Table 2-14
Plant Water Balance

Water Source	
Makeup Water	nil
Recycled from Stack Condenser	114,477 lb/h
Water Consumption Point	
Gasifier Coal Slurry Preparation	94,025 lb/h
Sulfuric Acid Water	3,581 lb/h
Shift Water	16,530 lb/h

2.4.2 EFFECTIVE THERMAL EFFICIENCY

For comparative purposes and to arrive at a figure of merit for the plant design, an ETE was derived for the plant performance based on HHV thermal value of hydrogen produced and offsite power sales, divided by the fuel input to the plant. The formula is:

$$\text{ETE} = \frac{(\text{Hydrogen Heating Value} + \text{Electrical Btu Equivalent})}{\text{Fuel Heating Value (HHV)}}$$

$$\text{ETE} = \frac{36,565 \text{ lb H}_2/\text{h} \times 61,095 \text{ Btu/lb} - 5,510 \text{ kW} \times 3,414 \text{ Btu/kWh}}{221,631 \text{ lb coal/h} \times 12,450 \text{ Btu/lb}}$$

$$\text{ETE} = 80.3\%$$

2.5 HYDROGEN FUEL PLANT – 1112°F (600°C) MEMBRANE AND 80 PERCENT HYDROGEN TRANSPORT

The baseline plant design having 600°C HSD equilibrium was modified to determine the impact of less than optimum performance from the membrane. It was assumed that only 80 percent of the syngas was recovered as hydrogen, with the rest being utilized in the topping cycle to generate electricity. The plant design basis is shown in Table 2-15.

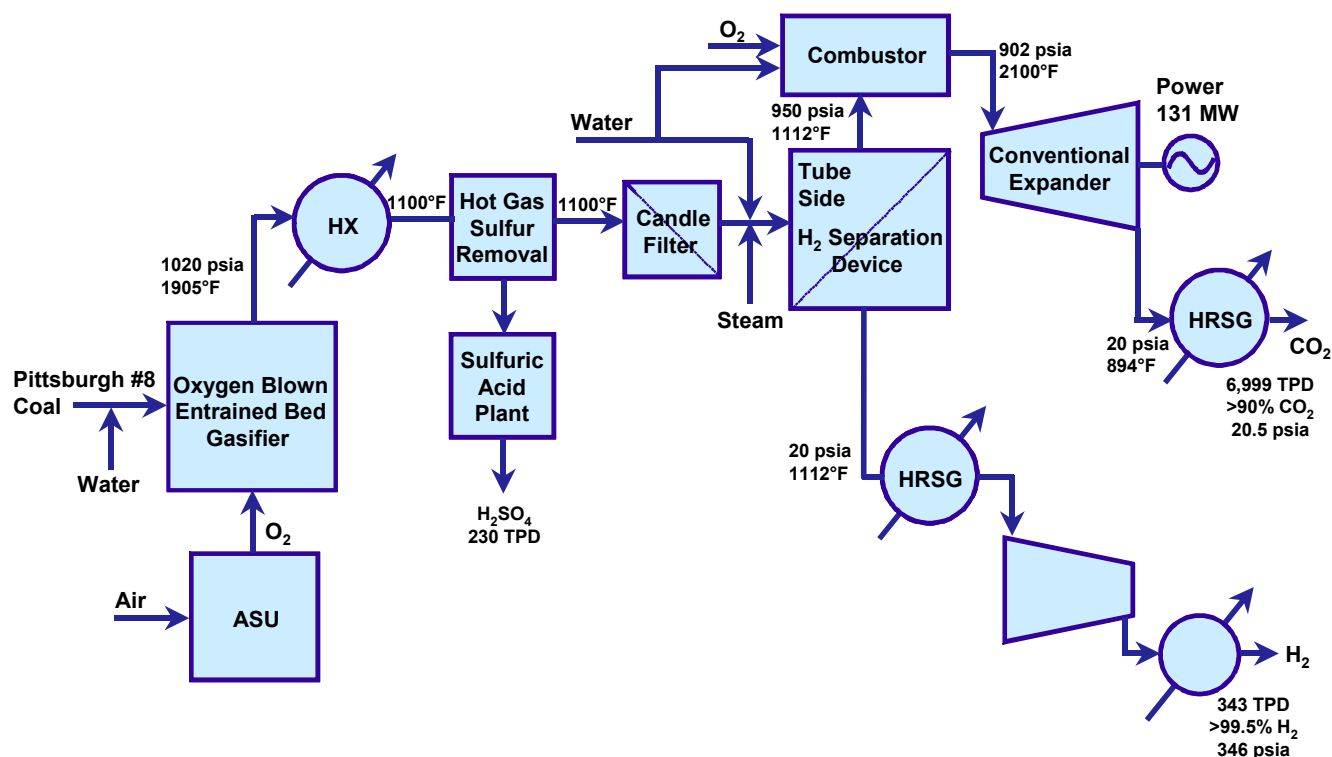
Table 2-15
Design Basis for Baseline Hydrogen Fuel Plant
1112°F (600°C) Membrane and 80 Percent Hydrogen Transport

Hydrogen Fuel Production Facility Parameter	Hydrogen Fuel Production Facility Plant Design Basis
Coal Feed	Pittsburgh No. 8, <10% ash
Limestone Sorbent	None
Gasifier	Oxygen-blown Destec with second stage adjusted for 1905°F output
Hot Gas Temperature	1905°F
Gasifier Outlet Pressure	1000 psia
Ambient Conditions	14.7 psia, 60°F
Hot Gas Desulfurization	Yes, 1100°F
Sulfur Recovery	Sulfuric acid
Ceramic Candle Filter	Before HSD
Hydrogen Separation	H ₂ separation device Shell and tube configuration 95% separation x 0.8 99.5% pure H ₂ Zero sulfur 20 psia hydrogen compressed to 346 psia
Separated Gas	CO shifted to 572°F equilibrium ~ 1.2 x 5% of fuel value in gas 950 psia
Separated Gas Utilization	Combustion with oxygen Steam injection conventional turbine expander
CO ₂ Product Pressure	19.4 psia
Hydrogen Utilization	346 psia offsite
Auxiliary Power Block	Conventional turbine expander
Plant Size	Maximum H ₂ production from 2,500 tpd dry gasifier

2.5.1 PROCESS DESCRIPTION

A block flow diagram of the plant is shown on Figure 2-8. The flows and state points on the attached process flow diagram (Figure 2-9) result from the heat and material balance for the plant. Key process components included in the plant are a Destec high-pressure slurry-feed gasifier, the ETTP HSD, and the transport reactor for desulfurization. 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. The hot raw gas is cleaned of larger particulates in a cyclone and then is cooled in a firetube boiler to 1100°F. A hot gas cleanup system consisting of a transport reactor desulfurizer and a ceramic candle filter removes sulfur and particulates from the fuel gas stream. Sulfur is recovered as sulfuric acid.

Figure 2-8
Block Flow Diagram
Baseline Hydrogen Fuel Plant
1112°F (600°C) Hydrogen Separation Device at 80 Percent Hydrogen Transport



For this case, both steam and water are added to cool the syngas while ensuring adequate water content for the high-temperature shift reaction to occur at the lower HSD inlet temperature. The gas enters the HSD at 605°F and leaves the HSD at 1112°F as a result of the exothermic shift reaction. For modeling purposes, 20 percent of the syngas bypasses the HSD, resulting in reduced hydrogen flow. Hydrogen produced is still 99.5 percent pure. It goes through a HRSG and then is compressed to 346 psia.

The CO₂-rich gas leaving the HSD at 950 psia contains more than the previous 5 percent of the fuel value of the inlet syngas stream. This gas goes to the gas turbine combustor with which oxygen is injected to convert CO and hydrogen to CO₂ and H₂O, respectively. A larger oxygen plant is needed to fire this gas. Both steam and water are injected into the combustor to moderate the temperature to 2100°F. The hot gas is expanded to 20 psia and 905°F through the conventional gas turbine expander to produce 84 MW electric power. The gas is cooled in a HRSG, and steam produced is combined with other steam produced from cooling the hydrogen for process applications. There is no power produced from steam. The CO₂ product is cooled to 100°F, dried, and sent offsite.

Figure 2-9

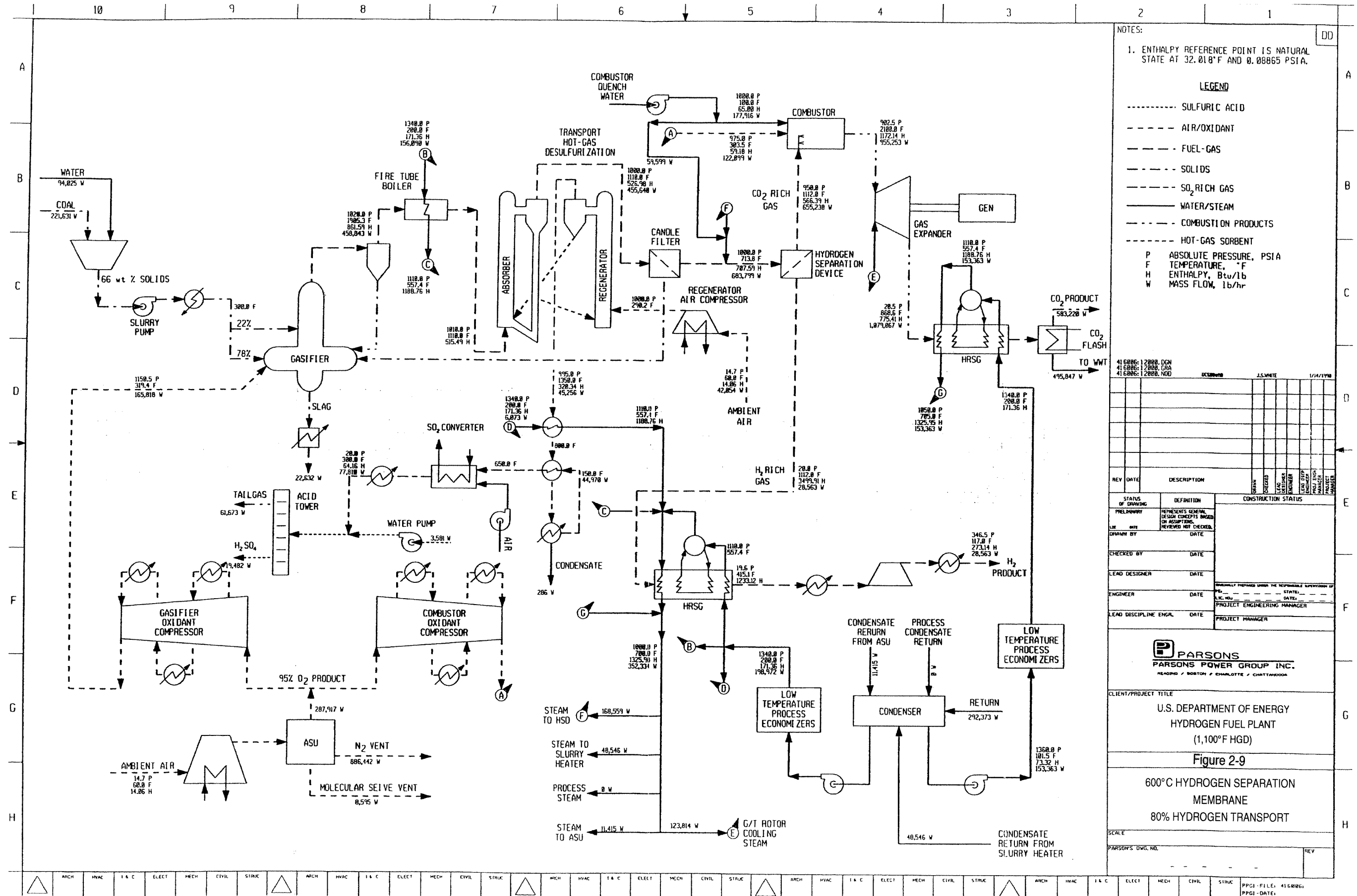


Table 2-16 presents the performance summary for the plant, and Table 2-17 identifies the plant power requirements. Table 2-18 is the plant water balance.

Table 2-16
Performance Summary

Coal Feed	221,631 lb/h
Oxygen Feed (95%)	287,917 lb/h
Hydrogen Product Stream	28,563 lb/h
CO ₂ Product Stream	583,220 lb/h
Sulfuric Acid Product	19,482 lb/h
Gross Power Production	131 MW
Auxiliary Power Requirement	83 MW
Net Power Production	48 MW
Effective Thermal Efficiency (ETE), HHV	69.2%

Table 2-17
Auxiliary Power Load, kW

Gasifier Auxiliary Oxygen Compressor	10,300
Combustor Oxygen Compressor	7,340
ASU Air Compressor	38,480
Gasifier Slurry Pumps	190
Water Spray Pump	400
Boiler Feedwater Pumps	550
Coal Handling	210
Slag Handling	530
Regenerator Air Blower	2,960
Gas Turbine Auxiliary	400
Hydrogen Compressor	20,440
Miscellaneous Balance of Plant	750
Total Auxiliary Load	82,550 kW

Table 2-18
Plant Water Balance

Water Source	
Makeup Water	63,734 lb/h
Recycled from Stack Condenser	495,847 lb/h
Water Consumption Point	
Boiler Feed	224,460 lb/h
Gasifier Coal Slurry Preparation	94,025 lb/h
Shift Water	59,599 lb/h
Combustion Quench	177,916 lb/h
Sulfuric Acid Water	3,581 lb/h

2.5.2 EFFECTIVE THERMAL EFFICIENCY

For comparative purposes and to arrive at a figure of merit for the plant design, an ETE was derived for the plant performance based on HHV thermal value of hydrogen produced and offsite power sales, divided by the fuel input to the plant. The formula is:

$$\text{ETE} = \frac{(\text{Hydrogen Heating Value} + \text{Electrical Btu Equivalent})}{\text{Fuel Heating Value (HHV)}}$$

$$\text{ETE} = \frac{28,563 \text{ lb H}_2/\text{h} \times 61,095 \text{ Btu/lb} + 47,950 \text{ kW} \times 3,414 \text{ Btu/kWh}}{221,631 \text{ lb coal/h} \times 12,450 \text{ Btu/lb}}$$

$$\text{ETE} = 69.2\%$$

2.6 COST ESTIMATING

For this economic analysis, the capital and operating costs for the four plants being evaluated have been upgraded to year 2000 dollars. Coal cost has been retained at \$1.00 per MMBtu.

2.6.1 APPROACH TO COST ESTIMATING

Economics in this report are stated primarily in terms of levelized cost of product, \$/short ton (\$/ton), or \$/MMBtu. The cost of product is developed from the identified financial parameters in Table 2-19, which are common to all the cost estimates in this report, and:

- Total capital requirement of the plant (TCR).
- Fixed operating and maintenance cost (fixed O&M).
- Non-fuel variable operating and maintenance costs (variable O&M).
- Consumables and byproducts costs and credits.
- Fuel costs.

Table 2-19

ESTIMATE BASIS/FINANCIAL CRITERIA for REVENUE REQUIREMENT CALCULATIONS			
GENERAL DATA/CHARACTERISTICS			
Case Title:	Hydrogen Fuel Facility		
Unit Size:/Plant Size:	422.5 H ₂ TPD	422.5 H ₂ TPD	
Location:	Middletown, USA		
Fuel:	Pittsburgh		
Levelized Capacity Factor / Preproduction (equivalent months)	80 %	1 month	
Capital Cost Year Dollars (Reference Year)	2000 (January)		
Delivered Cost of Coal	1.00\$/MBtu	\$/MBtu	
Design/Construction Period:	3 years		
Plant Startup Date (1 st Year Dollars):	2000 (January)		
Land Area/Unit Cost:	150 acres	\$1,500/acre	
FINANCIAL CRITERIA			
Project Book Life:	20 years		
Book Salvage Value:	%		
Project Tax Rate:	20 year		
Tax Depreciation Method:	Accel. based on ACRS Class		
Property Tax Rate:	% per		
Insurance Tax Rate:	% per		
Federal Income Tax Rate:	34.0%		
State Income Tax Rate:	6.0%		
Investment Tax Credit/% Eligible:	%	%	
Economic Basis:	Over Book Constant Dollars		
Capital Structure:	<u>% of Total</u>	<u>Cost (%)</u>	
Common Equity	20	16.5	
Preferred Stock			
Debt	80	6.3	
Weighted Cost of Capital: (after tax)		6.4 %	
	<u>Over Book Life</u>	<u>2000 to</u>	
Escalation Rates:	General	% per year	% per year
	Primary Fuel	-1.1% per year	-0.6% per year
	Secondary Fuel	0.6% per year	1.27% per year

The IGCC cost model used in the June 1999 Letter Report⁵ was the basis for developing the bulk of the balance-of-plant cost portion of the estimate. Use of this model assured consistency in the evaluation of balance-of-plant costs. As before, the capital cost for the gasifiers, gas cleanup including CO₂ removal, and the gas turbine was based on recent studies conducted by Parsons. Destec gasifier pricing was adjusted to reflect the impact of using a total quench in the second stage rather than a firetube boiler, followed by a ceramic candle filter. Balance-of-plant process system costs were estimated from cost curves developed by Parsons based in large part on the results of completed construction projects.

Costs for the HSD were developed independently of the cost model, based on several major assumptions listed below:

- The H₂ ceramic molecular sieve membrane requirement was calculated utilizing the membrane coefficient R&D goal confirmed by ETTP:

$$0.1 \text{ std cc/minute/cm}^2/\text{cm Hg P}_{\text{H}_2} \text{ differential}$$

Using this coefficient and a hydrogen pressure of 905 cm Hg differential pressure, the English coefficient on hydrogen weight basis becomes:

$$1.0 \text{ lb H}_2/\text{h/ft}^2$$

This coefficient is convenient due to the heat and material balance being expressed in lb/h.

- The cost of the ceramic molecular sieve material was based on a unit cost of \$100/ft². ORNL indicated that commercially available filters cost about \$300/ft², and they project the hydrogen membrane to be one-third of that cost.
- The shell and tube configuration can be conceived as being similar in design to shell and tube heat exchangers, except that the heat exchange surfaces are replaced by the ceramic molecular sieve.
- The cost base for the ceramic candle filter was the Westinghouse design used in pressurized fluidized-bed combustion (PFBC) hot gas cleanup applications. For the HSD, the cost of the shell and internals was applied, excluding the ceramic candles. The cost of the ceramic candles was replaced by the cost of the ceramic molecular sieve. On the basis of the typical 36,000 pounds of hydrogen per hour, 45,000 square feet of inorganic membrane are required for the nominal plant. Referring back to Table 2-2, 45,000 square feet of membrane is the design requirement for each plant. It was determined that the membranes could be contained in three vessels with a tube bundle configuration of 0.625-inch-diameter tubes by 9.7 feet long. Each 8-foot-diameter vessel contains 11,800 tubes.

2.6.2 PRODUCTION COSTS (OPERATION AND MAINTENANCE)

The production costs for the plant consist of several broad categories of cost elements. These cost elements include operating labor, maintenance material and labor, administrative and support labor, consumables (water and water treating chemicals, solid waste disposal costs, byproducts such as power sales, and fuel costs). Note that production costs do not include capital charges and should not be confused with cost of product.

2.6.3 COST RESULTS

The results of the cost estimating activity are summarized in Table 2-20 through Table 2-23.

Table 2-20

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY			
TITLE/DEFINITION			
Case:	Hydrogen Fuel Facility w/Hot Gas Desulfurization		
Plant Size:	422.5 H ₂ TPD	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/H₂TPD
Process Capital & Facilities		296,792	702.5
Engineering(incI.C.M.,H.O.& Fee)		28,890	68.4
Process Contingency		9,270	21.9
Project Contingency		33,495	79.3
TOTAL PLANT COST(TPC)		\$368,448	872.1
TOTAL CASH EXPENDED	\$368,448		
AFDC	\$23,625		
TOTAL PLANT INVESTMENT(TPI)		\$392,073	928.1
Royalty Allowance			
Preproduction Costs		9,466	22.4
Inventory Capital		3,067	7.3
Initial Catalyst & Chemicals(w/ equip.)			
Land Cost		150	0.4
TOTAL CAPITAL REQUIREMENT(TCR)		\$404,755	958.1
		\$x1000	\$x1000/H₂TPD
OPERATING & MAINTENANCE COSTS (2000 Dollars)			
Operating Labor		3,574	8.5
Maintenance Labor		2,733	6.5
Maintenance Material		4,099	9.7
Administrative & Support Labor		1,577	3.7
TOTAL OPERATION & MAINTENANCE		\$11,982	28.4
FIXED O & M			22.69
VARIABLE O & M			5.67
		\$x1000	\$/T H₂-yr
CONSUMABLE OPERATING COSTS,less Fuel (2000 Dollars)			
Water		89	0.72
Chemicals		761	6.17
Other Consumables			
Waste Disposal		793	6.43
TOTAL CONSUMABLE OPERATING COSTS		\$1,643	13.32
BY-PRODUCT CREDITS (2000 Dollars)		(\$8,739)	-70.84
FUEL COST (2000 Dollars)		\$19,337	156.76
PRODUCTION COST SUMMARY			
	1st Year (2005 \$)	Levelized (Over Book Life \$)	
	\$/T H₂-yr	\$/T H₂-yr	
Fixed O & M	77.71	77.71	
Variable O & M	5.67	5.67	
Consumables	13.32	13.32	
By-product Credit/ Penalty	-70.84	-70.84	
Fuel	152.11	138.75	
TOTAL PRODUCTION COST	177.97	164.61	
LEVELIZED CARRYING CHARGES(Capital)		459.36	
LEVELIZED(Over Book Life)COST/Ton of H ₂		623.97	
Equivalent \$/ MMBtu		5.11	

Table 2-21

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY				
TITLE/DEFINITION				
Case:	Hydrogen Fuel Facility w/600C Hot Gas Desulfurization			
Plant Size:	430.6 H ₂ TPD	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.TPD	
Process Capital & Facilities		290,035	673.2	
Engineering(incl.C.M.,H.O.& Fee)		28,255	65.6	
Process Contingency		8,792	20.4	
Project Contingency		32,708	75.9	
TOTAL PLANT COST(TPC)		\$359,791	835.1	
TOTAL CASH EXPENDED	\$359,791			
AFDC	\$23,070			
TOTAL PLANT INVESTMENT(TPI)		\$382,861	888.6	
Royalty Allowance				
Preproduction Costs		9,266	21.5	
Inventory Capital		3,038	7.1	
Initial Catalyst & Chemicals(w/equip.)				
Land Cost		150	0.3	
TOTAL CAPITAL REQUIREMENT(TCR)		\$395,315	917.6	
		\$x1000	\$x1000/H.TPD	
OPERATING & MAINTENANCE COSTS (2000 Dollars)				
Operating Labor		3,574	8.3	
Maintenance Labor		2,663	6.2	
Maintenance Material		3,994	9.3	
Administrative & Support Labor		1,559	3.6	
TOTAL OPERATION & MAINTENANCE		\$11,790	27.4	
FIXED O & M			21.89	
VARIABLE O & M			5.47	
CONSUMABLE OPERATING COSTS,less Fuel (2000 Dollars)		\$x1000	\$/T H ₂ -yr	
Water		96	0.77	
Chemicals		765	6.08	
Other Consumables				
Waste Disposal		793	6.30	
TOTAL CONSUMABLE OPERATING COSTS		\$1,654	13.15	
BY-PRODUCT CREDITS (2000 Dollars)		(\$6,504)	-51.70	
FUEL COST (2000 Dollars)		\$19,337	153.71	
PRODUCTION COST SUMMARY		1st Year (2005 \$)	Levelized (Over Book Life \$)	
		\$/T H ₂ -yr	\$/T H ₂ -yr	
Fixed O & M		74.97	74.97	0.6135793
Variable O & M		5.47	5.47	0.0447913
Consumables		13.15	13.15	0.107596
By-product Credit/Penalty		-51.70	-51.70	-0.423092
Fuel		149.15	136.05	1.113442
TOTAL PRODUCTION COST		191.05	177.95	1.4563165
LEVELIZED CARRYING CHARGES(Capital)			439.92	3.6003157
LEVELIZED(Over Book Life)COST/Ton of H ₂			617.87	
Equivalent \$/MMBtu			5.06	

Table 2-22

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY			
TITLE/DEFINITION			
Case:	Hydrogen Fuel Facility w/300C Hot Gas Desulfurization		
Plant Size:	438.8 H ₂ TPD	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/H₂ TPD	
Process Capital & Facilities	288,192	656.8	
Engineering(incl.C.M.,H.O.& Fee)	27,974	63.8	
Process Contingency	8,195	18.7	
Project Contingency	32,436	73.9	
TOTAL PLANT COST(TPC)	\$356,797	813.2	
TOTAL CASH EXPENDED	\$356,797		
AFDC	\$22,878		
TOTAL PLANT INVESTMENT(TPI)	\$379,675	865.3	
Royalty Allowance			
Preproduction Costs	9,053	20.6	
Inventory Capital	3,024	6.9	
Initial Catalyst & Chemicals(w/ equip.)			
Land Cost	150	0.3	
TOTAL CAPITAL REQUIREMENT(TCR)	\$391,903	893.2	
	\$x1000	\$x1000/H₂ TPD	
OPERATING & MAINTENANCE COSTS (2000 Dollars)			
Operating Labor	3,574	8.1	
Maintenance Labor	2,608	5.9	
Maintenance Material	3,912	8.9	
Administrative & Support Labor	1,545	3.5	
TOTAL OPERATION & MAINTENANCE	\$11,639	26.5	
FIXED O & M		21.22	
VARIABLE O & M		5.31	
CONSUMABLE OPERATING COSTS,less Fuel (2000 Dollars)	\$x1000	\$/T H₂-yr	
Water	195	1.52	
Chemicals	808	6.31	
Other Consumables			
Waste Disposal	793	6.19	
TOTAL CONSUMABLE OPERATING COSTS	\$1,797	14.02	
BY-PRODUCT CREDITS (2000 Dollars)	(\$3,861)	-30.14	
FUEL COST (2000 Dollars)	\$19,337	150.93	
PRODUCTION COST SUMMARY	1st Year (2005 \$)	Levelized (Over Book Life \$)	
	\$/T H₂-yr	\$/T H₂-yr	
Fixed O & M	72.68	72.68	
Variable O & M	5.31	5.31	
Consumables	14.02	14.02	
By-product Credit/ Penalty	-30.14	-30.14	
Fuel	146.46	133.59	
TOTAL PRODUCTION COST	208.33	195.46	
LEVELIZED CARRYING CHARGES(Capital)		428.24	
LEVELIZED(Over Book Life)COST/Ton of H₂		623.70	
Equivalent \$/ MMBtu		5.10	

Table 2-23

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY			
TITLE/DEFINITION			
Case:	Hydrogen Fuel Facility w/600C (80% H ₂) Hot Gas Desulfurization		
Plant Size:	342.8 H ₂ TPD	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/H₂ TPD
Process Capital & Facilities		311,678	909.3
Engineering(incl.C.M.,H.O.& Fee)		30,006	87.5
Process Contingency		8,907	26.0
Project Contingency		35,059	102.3
TOTAL PLANT COST(TPC)		\$385,650	1125.1
TOTAL CASH EXPENDED	\$385,650		
AFDC	\$24,728		
TOTAL PLANT INVESTMENT(TPI)		\$410,378	1197.3
Royalty Allowance			
Preproduction Costs		9,724	28.4
Inventory Capital		3,132	9.1
Initial Catalyst & Chemicals(w/ equip.)			
Land Cost		150	0.4
TOTAL CAPITAL REQUIREMENT(TCR)		\$423,384	1235.2
		\$x1000	\$x1000/H₂ TPD
OPERATING & MAINTENANCE COSTS (2000 Dollars)			
Operating Labor		3,574	10.4
Maintenance Labor		2,855	8.3
Maintenance Material		4,283	12.5
Administrative & Support Labor		1,607	4.7
TOTAL OPERATION & MAINTENANCE		\$12,319	35.9
FIXED O & M			28.75
VARIABLE O & M			7.19
CONSUMABLE OPERATING COSTS,less Fuel (2000 Dollars)			
		\$x1000	\$/T H₂-yr
Water		172	1.72
Chemicals		798	7.97
Other Consumables			
Waste Disposal		793	7.92
TOTAL CONSUMABLE OPERATING COSTS		\$1,763	17.61
BY-PRODUCT CREDITS (2000 Dollars)		(\$15,100)	-150.88
FUEL COST (2000 Dollars)		\$19,337	193.21
PRODUCTION COST SUMMARY			
	1st Year (2005 \$)	Levelized (Over Book Life \$)	
	\$/T H₂-yr	\$/T H₂-yr	
Fixed O & M	98.47	98.47	
Variable O & M	7.19	7.19	
Consumables	17.61	17.61	
By-product Credit/ Penalty	-150.88	-150.88	
Fuel	187.48	171.01	
TOTAL PRODUCTION COST	159.87	143.40	
LEVELIZED CARRYING CHARGES(Capital)		592.24	
LEVELIZED(Over Book Life)COST/ Ton of H ₂		735.64	
Equivalent \$/ MMBtu		6.02	

2.7 SUMMARY AND CONCLUSIONS

Utilizing the revised assumptions for the HSD, updated plant concepts were prepared for HSD operation at 572°F (300°C) and 1112°F (600°C). For comparisons, the initial plant operating at 1402°F (761°C) is also presented. A plant with HSD performance reduced from 95 to 80 percent hydrogen transport was also evaluated to show the impact of not reaching the HSD goal of 95 percent separation. Table 2-24 summarizes and compares the performance and economics of the four plants.

Table 2-24
Performance and Cost Summary Comparisons
Hydrogen Fuel Plants with Alternative HSD Temperatures

	1402°F Membrane (761°C)	1112°F Membrane (600°C) Baseline Case	572°F Membrane (300°C)	1112°F Membrane with 80% Hydrogen Transport
HSD Exit Temperature	1402°F (761°C)	1112°F (600°C)	572°F (300°C)	1112°F (600°C)
Coal Feed	221,631 lb/h	221,631 lb/h	221,631 lb/h	221,631 lb/h
Oxygen Feed (95%)	231,218 lb/h	224,519 lb/h	218,657 lb/h	287,917 lb/h
Hydrogen Product Stream	35,205 lb/h	35,903 lb/h	36,564 lb/h	28,562 lb/h
CO ₂ Product Stream	581,657 lb/h	582,566 lb/h	585,598 lb/h	583,220 lb/h
Sulfuric Acid Product	19,482 lb/h	19,482 lb/h	19,482 lb/h	19,482 lb/h
Gross Power Production	94 MW	84 MW	71 MW	131 MW
Auxiliary Power Requirement	76 MW	77 MW	76 MW	83 MW
Net Power Production	18 MW	7 MW	(6 MW)	48 MW
Effective Thermal Efficiency, HHV	80.2%	80.4%	80.3%	69.2%
Capital Cost, \$1,000 (Year 2000)	\$368,448	\$359,791	\$356,797	\$385,650
Hydrogen Product Cost, \$/MMBtu	\$5.11	\$5.06	\$5.10	\$6.02

The lower temperature favors hydrogen recovery but reduces the efficiency of the steam cycle. The 1112°F (600°C) plant was selected as the baseline design since this temperature is the operational goal of the membranes and also this concept maintained a high hydrogen recovery while minimizing costs.

These designs were based on goals that have been set by membrane developers but not yet experimentally demonstrated. These goals include:

- Hydrogen Flux – The hydrogen flux was based on the R&D goal of 0.1 std cc/minute/cm²/cm Hg P_{H₂} differential.
- Separation Factor – The separation determines the hydrogen purity and is high for hydrogen, increasing with higher temperatures. Even at 300°C the separation factor would be above 200.
- Operating Pressure and Temperature – It was assumed that a 950 psi pressure differential can be contained by the inorganic membrane. The operational goal for the membranes is

currently 600°C, and a vessel design could be prepared today to operate with confidence up to 300°C.

- CO Shift Properties – It was assumed that the shift reaction on the membrane surface goes to equilibrium without catalyst.

The 80 percent hydrogen transport case reduces the amount of hydrogen recovered but increases the amount of power produced in the topping cycle. The cost of hydrogen increases from the baseline case, but proportionally less than the reduction in hydrogen recovered.

3. HYDROGEN FROM COAL AND NATURAL GAS-BASED PLANTS

Throughout 1999 and 2000, conceptual systems and cost analyses were developed by Parsons for a coal processing plant to produce hydrogen while recovering carbon dioxide (CO₂) for offsite processing or sequestration. This has been referred to as a hydrogen fuel plant.

This work has been reported in several venues including the June 1999 letter report⁵, and U.S. Department of Energy (DOE) sponsored conferences.^{9,10,11,12,13} This work has resulted in a baseline plant for production of hydrogen from coal utilizing the ORNL-developed inorganic membrane for separation of hydrogen from syngas.

The purpose of this section is to compare hydrogen cost from conventional methods, with and without CO₂ recovery, against the baseline hydrogen fuel plant.

3.1 CASES 1, 2, AND 3 – HYDROGEN FROM NATURAL GAS WITHOUT AND WITH CO₂ RECOVERY

Cases 1 and 2 are based on steam reforming. Also included in these comparisons is Case 3, which uses an oxygen-blown gasifier and a hydrogen separation membrane. Intuitively it will not be economically competitive with other approaches to producing hydrogen; thus it was not evaluated economically.

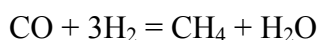
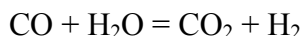
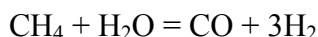
Steam reforming of hydrocarbons continues to be the most efficient, economical, and widely used process for production of hydrogen and hydrogen/carbon monoxide mixtures. The process involves a catalytic conversion of the hydrocarbon and steam to hydrogen and carbon oxides. Since the process works only with light hydrocarbons that can be vaporized completely without carbon formation, the feedstocks used range from methane (natural gas) to naphtha to No. 2 fuel oil.

3.1.1 NATURAL GAS CONDITIONING

Natural gas is fed to the plant from the pipeline at a pressure of 450 psia. To protect the catalysts in the hydrogen plant, the natural gas must be desulfurized before being fed to the reformer. The gas is generally sulfur-free, but odorizers with mercaptans must be cleaned from the gas to prevent contamination of the reformer catalyst. This is accomplished with a zinc oxide polishing bed.

3.1.2 NATURAL GAS REFORMER/BOILER

The desulfurized natural gas feedstock is mixed with process steam to be reacted over a nickel-based catalyst contained inside a system of high alloy steel tubes. The following reactions for methane take place in the reformer:



The reforming reaction is strongly endothermic, with energy supplied by firing the reformer on the outside of the catalyst tubes with recycled syngas from the hydrogen purification process. The metallurgy of the tubes usually limits the reaction temperature to 1400-1700°F. The flue gas path of the fired reformer is integrated with additional boiler surfaces to produce about 700,000 lb/hour steam. Of this, about 450,000 lb/hour is superheated to 450 psia and 750°F, to be added to the incoming natural gas. Additional steam from the boiler is either shipped offsite or used within the plant for regeneration of CO₂ from the acid gas removal process.

The CO-shift and methanation reactions quickly reach equilibrium at all points in the catalyst bed. High steam-to-carbon ratio, low pressure, and high temperature favor the equilibrium composition of the reformed gas. The process generally employs a steam-to-carbon ratio of 3 to 5 at a process temperature of around 1500°F and pressures up to 500 psig to convert more than 70 percent of hydrocarbon to oxides of carbon at the outlet of the reformer so as to ensure a minimum concentration of CH₄ in the product gas. After the reformer, the process gas mixture of CO and H₂ passes through a heat recovery step and is fed into a water-gas shift reactor to produce additional H₂.

The typical composition of the synthesis gas at 450 psia leaving a steam-methane reformer is shown in Table 3-1:

Table 3-1
Composition of Synthetic Gas

Component	Volume %
CH ₄	8
CO	7
CO ₂	6
H ₂	44
H ₂ O	35
Total	100

The reformer burner uses a low-NO_x design to limit NO_x emissions to 20 ppm, very low for a gas-fired boiler. This consists of burning predominantly pressure swing adsorption (PSA) purge gas with air at ambient temperature. Neither selective non-catalytic reduction (SNCR) nor selective catalytic reduction (SCR) for NO_x reduction is used with this plant design.

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 (CO₂) 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 CO₂. 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 CO₂ with some residual CO and methane.

3.1.4 ACID GAS REMOVAL

With conventional production of hydrogen from natural gas, CO₂ 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 CO₂ from the syngas stream. The CO₂ 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. CO₂ 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 CO₂ by fractionation. Because of the steam load required to regenerate CO₂, there is no steam export from the plant removing CO₂. Regenerated lean amine is then cooled and sent back to the amine tower. The regenerated CO₂ 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 CO₂, 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* CO₂ 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* CO₂ 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

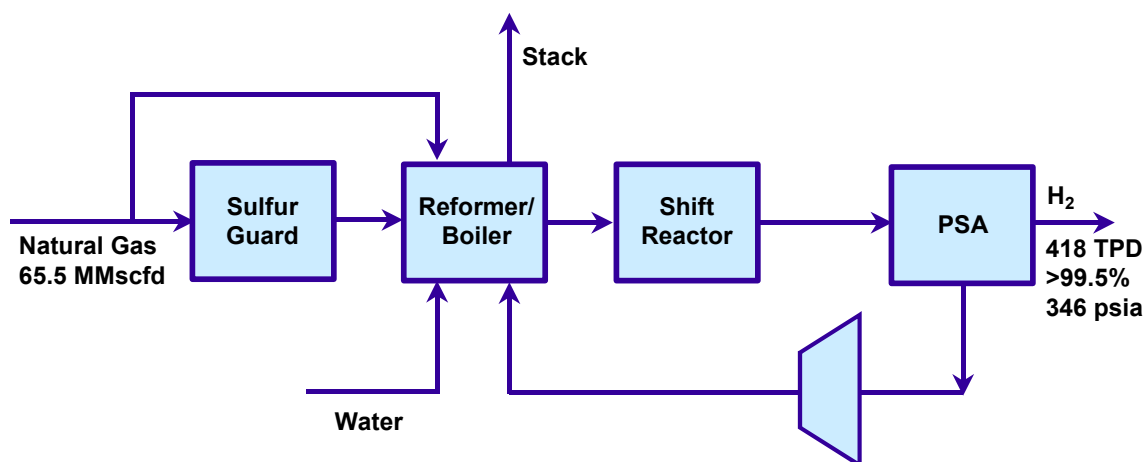


Table 3-2
Performance and Cost Summary
Case 1 – Hydrogen from Natural Gas without CO₂ Capture

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/H₂TPD
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/H₂TPD
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 H₂-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 \$)	Levelized (10th Year \$)	
	\$/T H₂-yr	\$/T H₂-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 CO₂ Removal

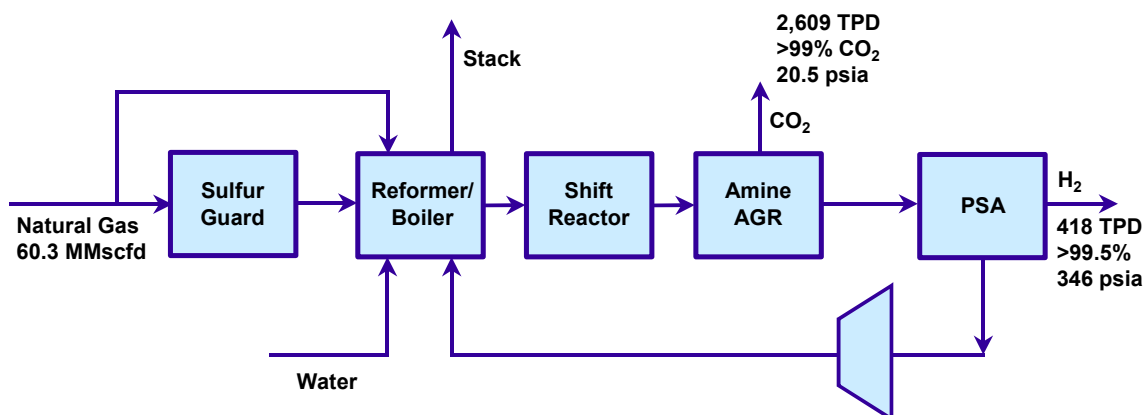


Table 3-3
Performance and Cost Summary
Case 2 – Hydrogen from Natural Gas with CO₂ Capture by Amine Process

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/H₂TPD	
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/H₂TPD	
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 H₂-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 \$)	Levelized (10th Year \$)	
	\$/T H₂-yr	\$/T H₂-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	524.27	
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

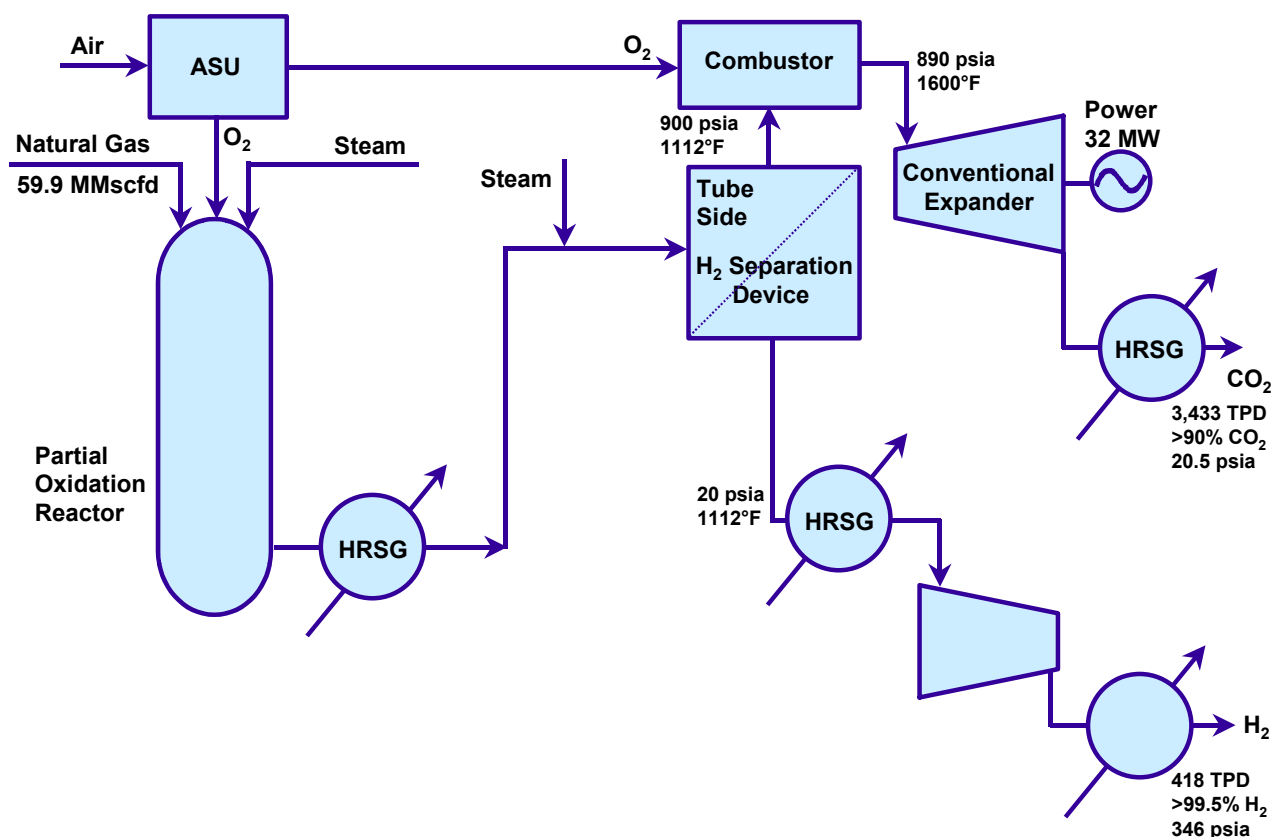


Table 3-4
Performance Summary
Case 3 – Natural Gas Partial Oxidation Plant with CO₂ Capture
600°C Inorganic Membrane

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 CO₂ 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 CO₂ and sulfur in a double-stage Selexol unit. H₂S 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 CO₂ removal case, the PSA tail gas is fired in the HRSG with oxygen, resulting in a concentrated CO₂ 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 O₂. 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 CO₂ 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 H₂S and CO₂ separately from the hydrogen product stream, a double-stage Selexol unit was selected. This process removes H₂S from the cooled syngas and then removes CO₂ from the desulfurized syngas. The acid gas removal (AGR) process utilizes a physical sorbent and several design features to effectively remove and recover H₂S and CO₂ 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 CO₂ removal case, the PSA tail gas is fired in the HRSG with oxygen, resulting in a concentrated CO₂ stream in the stack for recovery.

The conventional hydrogen from coal plant described in the June 1999⁵ letter report included provisions for recovering CO₂. The amount of CO₂ 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 CO₂ 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 CO₂ captured. This was accomplished by firing the PSA retentate with oxygen in the HRSG, resulting in a stack gas containing only CO₂ and water vapor. The CO₂ is then cooled and recovered.

The Selexol unit consists of two absorbers: the first absorbs H₂S from the cooled syngas, providing a desulfurized syngas, and the second absorbs CO₂ from the desulfurized syngas. The two absorbers are integrated, with solvent flowing between them. A low-pressure H₂S stream is sent to the sulfuric acid plant and a low-pressure CO₂ stream is sent offsite for sequestration.

A simplified basic flow sheet of Case 4, Conventional Hydrogen from Coal *without* CO₂ 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* CO₂ 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 CO₂ Removal

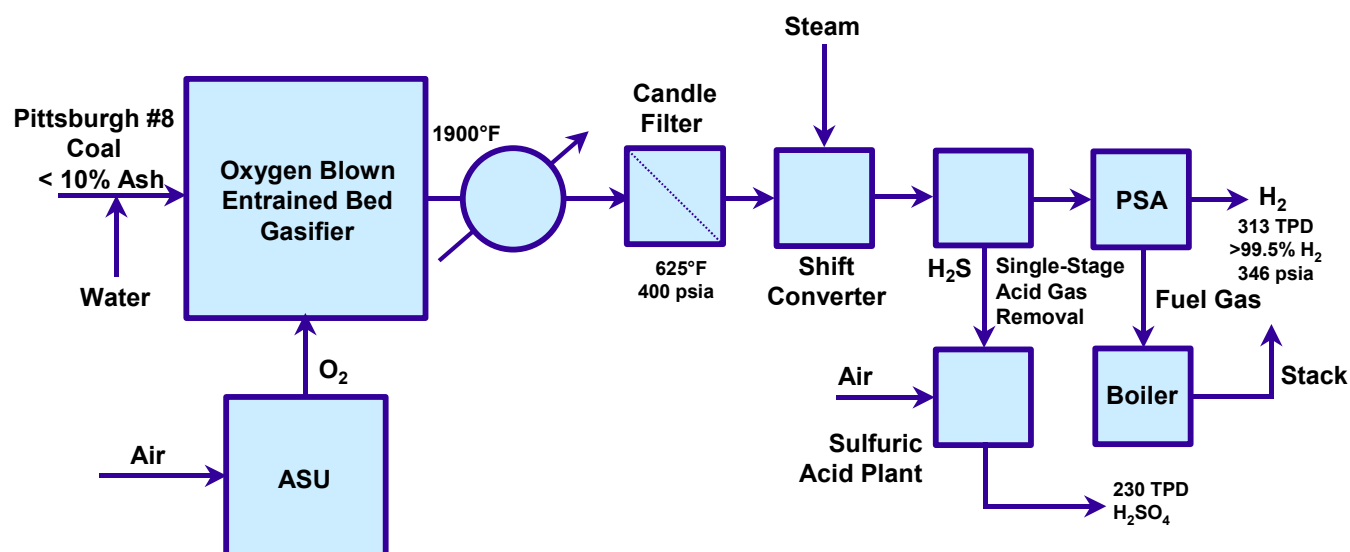


Table 3-5
Performance and Cost Summary
Case 4 – Conventional Hydrogen from Coal without CO₂ Capture

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-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: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: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: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 \$)	Levelized (Over Book Life \$)	
	\$/T H:yr.	\$/T H: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 CO₂ Removal

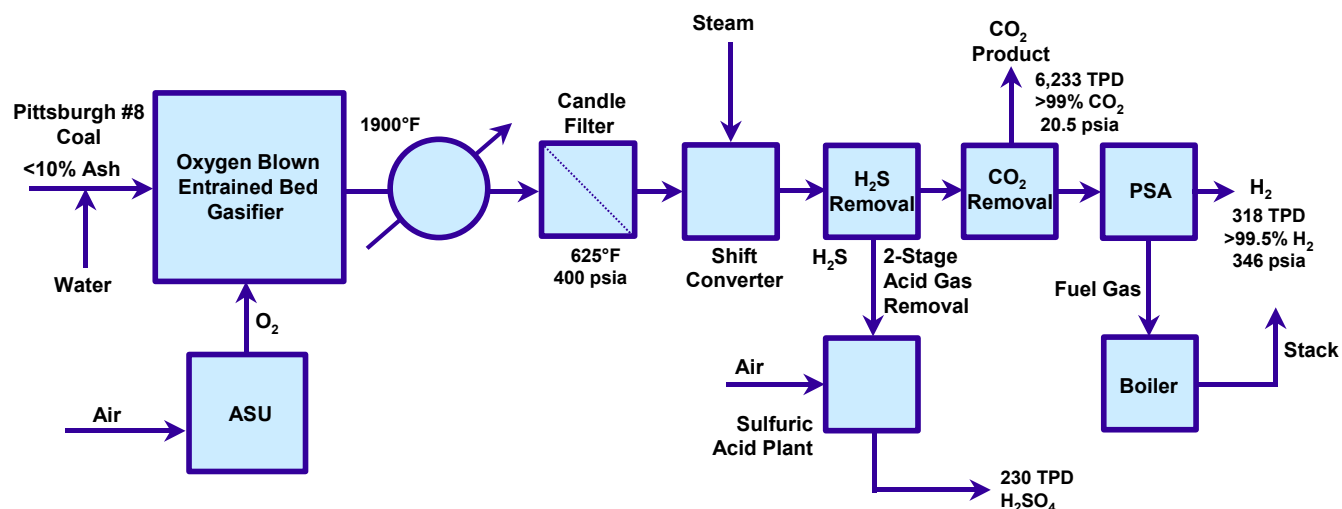


Table 3-6
Performance and Cost Summary
Case 5 – Conventional Hydrogen from Coal with Maximum CO₂ Capture

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/H₂TPD
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
		\$x1000	\$x1000/H₂TPD
OPERATING & MAINTENANCE COSTS (2000 Dollars)			
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 H₂-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
		1st Year (2005 \$)	Levelized (Over Book Life \$)
PRODUCTION COST SUMMARY		\$/T H₂-yr.	\$/T H₂-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 CO₂ Capture	Case 2 Hydrogen from Natural Gas with CO₂ Capture by Amine Process	Case 4 Conventional Hydrogen from Coal without CO₂ Capture	Case 5 Conventional Hydrogen from Coal with Maximum CO₂ Capture	Baseline Case Advanced Hydrogen Plant with CO₂ 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, MMBtu (MMscfd)	2,868 MMBtu (65.5 MMscfd)	2,640 MMBtu (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 CO₂ removal, would be competitive with hydrogen produced from both natural gas- and coal-based conventional technologies even without CO₂ removal. With only 80 percent hydrogen transport, hydrogen production would still be competitive with conventional coal-based technology.

4. HYDROGEN FUEL FROM WYODAK COAL/BIOMASS BLEND

The design and cost assessments of hydrogen fuel plants have been based on Pittsburgh No. 8 bituminous coal since work in this area began. The plants have been normalized by having a throughput of 2,500 tons per day of coal on a dry basis. In response to an inquiry regarding the performance of plants on fuels other than the base coal, a conceptual plant has been designed, based on a fuel consisting of a blend of 90 percent Wyodak subbituminous coal and 10 percent biomass. The results of this assessment are compared to the baseline plant performance and economics from Pittsburgh No. 8 coal.

4.1 INTRODUCTION

The baseline hydrogen fuel plant design (600°C hydrogen separation device [HSD] and hot gas desulfurization) was selected to evaluate the conversion of the coal and biomass mixture to synthesis gas, and achieve essentially total separation of hydrogen from the CO₂. Wyodak subbituminous coal and the biomass analyses for the study were provided by NETL, and are listed in Table 4-1. This concept utilizes hot gas desulfurization and particulate removal upstream of the HSD along with a modern non-ATS (conventional) gas turbine in the CO₂-rich stream. Table 4-2 provides the design basis established for the plant.

Table 4-1
Wyodak Coal and Biomass Properties (As Received)

Proximate	Wyodak Coal	Biomass* (Seward Sawdust)
Moisture	26.6	39.0
Volatile Matter	33.2	49.3
Fixed Carbon	34.4	11.5
Ash	5.8	0.31
Ultimate		
Sulfur	0.6	0.02
Hydrogen	6.5	5.9
Carbon	50.0	49.8
Nitrogen	0.9	0.2
Oxygen	36.2	43.6
Ash	5.8	0.51
Heating Value, HHV	8,630 Btu/lb	5,165 Btu/lb

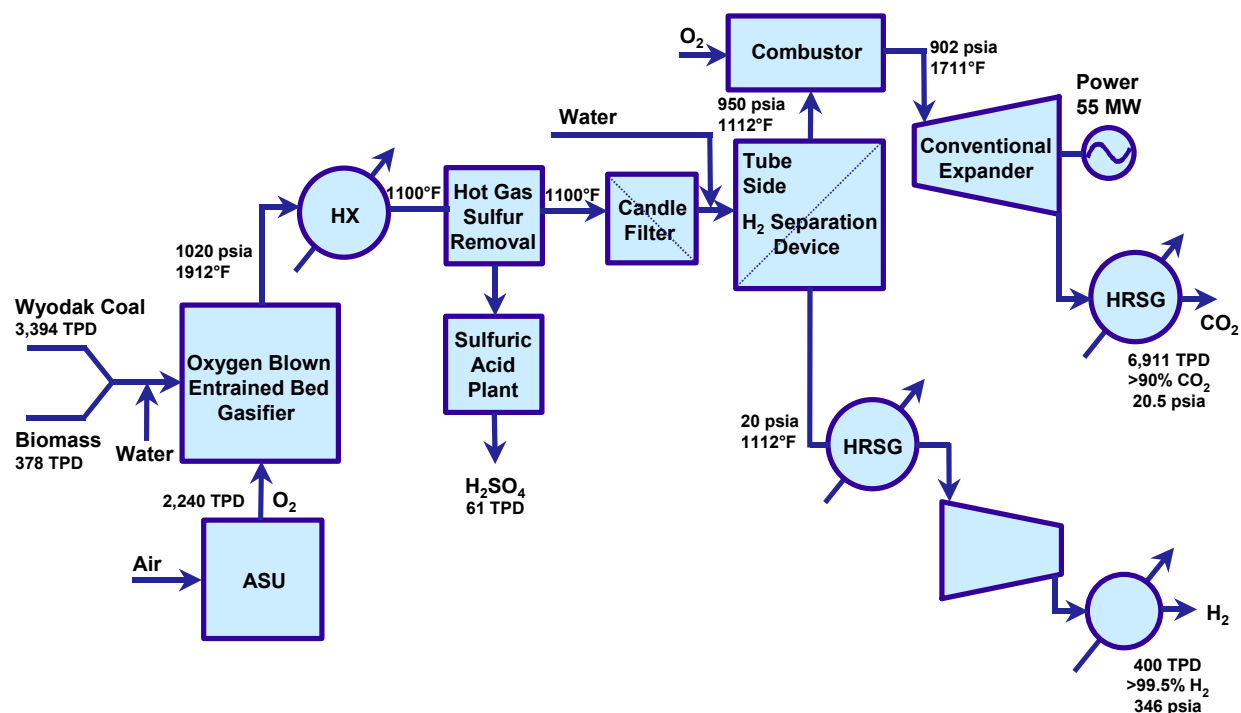
* Biomass is supplied in < ¼-inch size.

Table 4-2
Design Basis for Hydrogen Fuel Production Facility
with Conventional Expansion Turbine and Hot Gas Cleanup
Biomass/Coal Feed

Hydrogen Fuel Production Facility Parameter	Hydrogen Fuel Production Facility Plant Design Basis
Feed	90% Wyodak coal, AR 10% biomass, AR
Gasifier	Oxygen-blown E-Gas with second stage adjusted for 1905°F output
ASU	Cryogenic
Hot Gas Temperature	1905°F
Gasifier Outlet Pressure	1000 psia
Ambient Conditions	14.7 psia, 60°F
Hot Gas Desulfurization	Yes, 1100°F
Sulfur Recovery	Sulfuric acid
Ceramic Candle Filter	Before HSD
Hydrogen Separation	H ₂ separation device Shell and tube configuration 95% separation 99.5% pure H ₂ Zero sulfur 20 psia hydrogen compressed to 346 psia
Separated Gas	CO shifted to 1112°F equilibrium 5% of fuel value in gas 950 psia
Separated Gas Utilization	Combustion with oxygen Conventional turbine expander
CO ₂ Product Pressure	19.4 psia
Hydrogen Utilization	346 psia offsite
Auxiliary Power Block	Conventional turbine expander Steam turbine bottoming cycle
Plant Size	Maximum H ₂ production from 2,500 tpd dry feed basis gasifier Excess power sold offsite

The overall plant concept is shown on Figure 4-1. Key process components included in the plant are an E-Gas high-pressure slurry-feed gasifier, the Eastern Tennessee Technology Park (ETTP) HSD, and the transport reactor for desulfurization. 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. The hot raw gas is cleaned of larger particulates in a cyclone and then is cooled in a firetube boiler to 1100°F. A hot gas cleanup system consisting of a transport reactor desulfurizer and a ceramic candle filter removes sulfur and particulates from the fuel gas stream. Sulfur is recovered as sulfuric acid.

Figure 4-1
Block Flow Diagram
Hydrogen Plant with Wyodak Coal/Biomass Fuel



For this case, water is added to both cool the hot clean syngas and ensure adequate water content for the high-temperature shift reaction to occur at the HSD inlet temperature. The gas enters the HSD at 717°F and leaves the HSD at 1112°F as a result of the exothermic shift reaction. The hydrogen produced from the HSD is 99.5 percent pure. It goes through a HRSG and is then compressed to 346 psia.

The CO₂-rich gas leaving the HSD at 950 psia contains about 5 percent of the fuel value of the inlet syngas stream. This gas goes to the gas turbine combustor with which oxygen is injected to combust CO and hydrogen to CO₂ and H₂O, respectively, resulting in a firing temperature of 1711°F. The hot gas is expanded to 20 psia and 814°F through the conventional gas turbine expander to produce 55 MW electric power. The gas is cooled in a HRSG, and steam is combined with other steam from cooling the hydrogen to produce an additional 28 MW and for process applications. The low-pressure CO₂ product is cooled to 100°F, dried, and sent offsite.

Table 4-3 presents the performance summary for the plant, and Table 4-4 identifies the plant auxiliary power requirements.

Table 4-3
Performance Summary

Wyodak Coal Feed	283,833 lb/h
Biomass Feed	31,537 lb/h
Oxygen Feed (95%) to Gasifier	186,650 lb/h
Oxygen Feed to Retentate Combustor	25,300 lb/h
Water to Prepare Feed Slurry	114,009 lb/h
Hydrogen Product Stream	33,337 lb/h
CO ₂ Product Stream @ 90.2% CO ₂	575,923 lb/h
Sulfuric Acid Product	5,057 lb/h
Gross Power Production	
Turbine Expander	55.4 MW
Steam Turbine	27.7 MW
Auxiliary Power Requirement	(69.2 MW)
Net Power Production	13.9 MW
Effective Thermal Efficiency (ETE), HHV	79.8%

Table 4-4
Auxiliary Power Load, kW

Gasifier Auxiliary Oxygen Compressor	11,560 kW
Combustor Oxygen Compressor	1,520
ASU Air Compressor	28,320
Gasifier Slurry Pumps	220
Sulfuric Acid Plant	90
Water Spray Pump	420
BF Water Pumps	470
Coal/Sawdust Handling	300
Slag Handling	700
Regenerator Air Blower	780
GT/ST Auxiliary	400
Hydrogen Compressor	23,690
Miscellaneous Balance of Plant	750
Total Auxiliary Load	69,220 kW

4.2 PROCESS DESCRIPTION

The baseline hydrogen fuel plant design was modified to accommodate a blend of 90 percent Wyodak coal and 10 percent biomass. The basis for design was retained at a gasifier throughput of 2,500 tons per day *dry* fuel, but due to the higher moisture contents, the coal/biomass blend required increased material handling. However, the syngas and hydrogen production processes had only minor changes. Hydrogen production with the new feedstock is reduced by about

7 percent. The low sulfur content of the fuel also resulted in a reduction in the acid plant size by about 2/3. The flows and state points on Figure 4-2 result from the heat and material balance for the plant. Following are more detailed descriptions of the key process elements.

4.2.1 GASIFIER

The high-pressure system for producing hydrogen has resulted in utilizing two E-Gas gasifier trains, each having a throughput capacity of 1,250 tpd coal (dry basis). The E-Gas high-pressure entrained flow gasifier consists of two stages to gasify the coal-biomass-water slurry feed with oxygen. The slurry was prepared by fine grinding the 90/10 coal biomass (as received) feed fuel to about 200 mesh and mixing with water to achieve a ratio of 66 percent solids and 34 percent water, *including the moisture content of the coal and biomass*. The gasifier can operate at any pressure up to the capability of the oxygen compressor. By operating in two stages, it is possible to adjust the flow split between stages to achieve a desired outlet temperature of the product gas. A typical operating temperature for the E-Gas gasifier is 1900°F. This temperature is reached 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 gasifiers at 1905°F goes through an internal cyclone that separates entrained particles from the gas for recycle to the gasifiers, a firetube boiler to cool the gas to 1100°F, and a hot gas cleanup system. Water is injected into the gas stream to cool and saturate the gas at 717°F, promoting the shift reaction, which will occur downstream in the HSD.

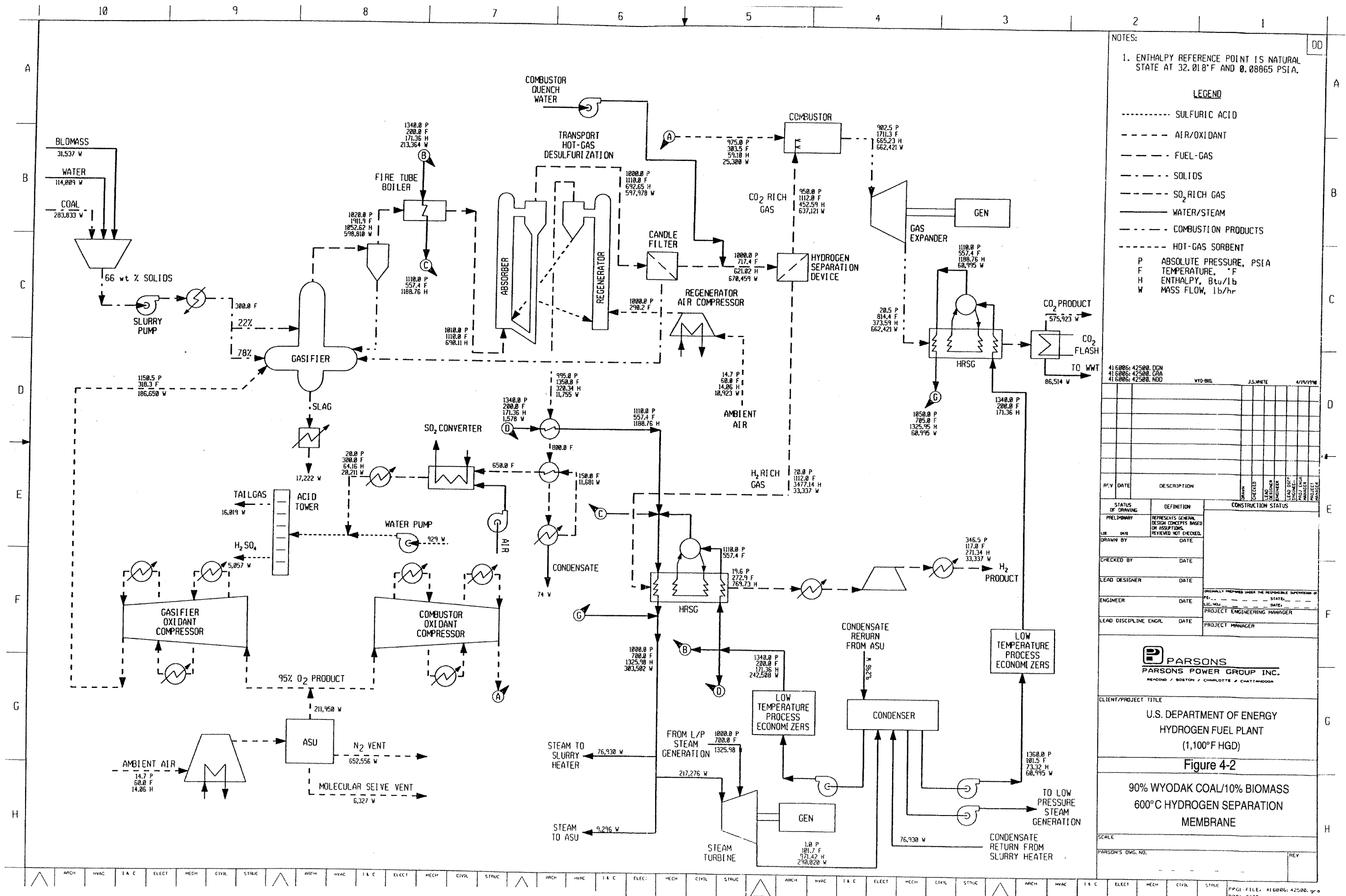
4.2.2 AIR SEPARATION UNIT

Oxygen supply for this plant is provided through a conventional cryogenic ASU. The air separation plant is designed to produce a nominal output of 2,600 tons/day of 95 percent pure O₂. The high-pressure plant is designed with one 100 percent capacity production train, with liquefaction and liquid oxygen storage providing an 8-hour backup supply of oxygen.

4.2.3 HOT GAS CLEANUP SYSTEM

The transport reactor desulfurizer consists of a riser tube, a disengager, and a standpipe for both the absorber section and regeneration section. Sorbent from the absorber passes through the regenerator riser, disengages, and transfers back to the absorber through the standpipe. Regeneration is conducted with neat air to minimize heat release and limit temperature. The regeneration heat has negligible effect on the sorbent temperature in the absorber. The regeneration off-gas containing predominantly SO₂ is sent to the sulfuric acid plant. Elutriated particles are disengaged from the gas by high-efficiency cyclones at the top of the absorber. A final ceramic candle filter is located downstream.

Figure 4-2



4.2.4 SULFURIC ACID PLANT

Key to the double-absorption contact sulfuric acid plant process is use of an intermediate absorber in the four-pass converter developed by Monsanto. The reaction from SO_2 to SO_3 is an exothermic reversible reaction. Using a vanadium catalyst, a contact plant takes advantage of both rate and equilibrium considerations by first allowing the gases to enter over a part of the catalyst at about 800°F, and then allowing the temperature to increase adiabatically as the reaction proceeds. The reaction essentially stops when about 60 to 70 percent of the SO_2 has been converted, at a temperature in the vicinity of 1100°F. The gas is cooled in a waste heat boiler and passed through subsequent stages until the temperature of the gases passing over the last portion of catalyst does not exceed 800°F. The gases leaving the converter, having passed through two or three layers of catalyst, are cooled and passed through an intermediate absorber tower where some of the SO_3 is removed with 98 percent H_2SO_4 . The gases leaving this tower are then reheated, and flow through the remaining layers of catalyst in the converter. The gases are then cooled and pass through the final absorber tower before discharge to the atmosphere. In this manner, more than 99.7 percent of the SO_2 is converted into SO_3 and subsequently into product sulfuric acid.

4.2.5 HYDROGEN SEPARATION/CONVENTIONAL TURBINE EXPANDER

The HSD design retains the previous concept to promote the shift reaction by product extraction at the membrane surface. The scenario is based on the gas proceeding along the membrane surface in turbulent flow. Hydrogen product partial pressure is both maintained and extracted at the membrane surface. CO continues to react with steam until the CO-steam equilibrium is reached. The remaining gas then passes from the membrane without further reaction. To ensure the shift reaction going to completion, the membrane path was increased 25 percent above theoretical.

The hydrogen product diffusing through the HSD is 99.5 percent pure on a weight basis, and is comprised of a stream having 95 percent of the original syngas fuel value. The syngas continues with an exothermic shift to hydrogen and CO_2 on the membrane surface until reaching an equilibrium at 600°C (1112°F).

The retentate gas, which is separated from the hydrogen, leaves the HSD at 950 psia and 1112°F, and has a fuel value of about 15 Btu/scf. A conventional expansion turbine is utilized to extract the energy from the gas stream by producing power and steam. The gas stream is fired with oxygen in the combustor, resulting in conversion of CO and hydrogen to CO_2 , and water vapor, resulting in a turbine inlet temperature of 1711°F. The turbine expander reduces the gas pressure to 20 psia and its temperature to 814°F, while generating 55 MW power. The gas then passes through a HRSG where it is cooled to 250°F, while raising high-pressure steam. This steam is combined with additional steam from cooling the hydrogen product to produce an additional 28 MW. In-plant auxiliary power requirements and transformer losses amount to 69 MW, resulting in export power sales of 14 MW. The CO_2 product is cooled to 100°F, dried, and sent offsite. Table 4-5 identifies the overall water balance for the plant.

Table 4-5
Plant Water Balance

Water Source	
Makeup Water	100,979 lb/h
Recycled from Stack Condenser	86,514 lb/h
Water Consumption Point	
Boiler Blowdown	74 lb/h
Gasifier Coal Slurry Preparation	114,009 lb/h
HSD Inlet Cooler/Saturator	72,481 lb/h
Sulfuric Acid Water	929 lb/h

4.2.6 EFFECTIVE THERMAL EFFICIENCY

For comparative purposes and to arrive at a figure of merit for the plant design, an ETE was derived for the plant performance based on HHV thermal value of hydrogen produced and offsite power sales, divided by the fuel input to the plant. The formula is:

$$\text{ETE} = \frac{(\text{Hydrogen Heating Value} + \text{Electrical Btu Equivalent})}{\text{Fuel Heating Value (HHV)}}$$

$$\text{ETE} = \frac{33,336 \text{ lb H}_2/\text{h} \times 61,095 \text{ Btu/lb} + 13,900 \text{ kW} \times 3,414 \text{ Btu/kWh}}{283,833 \text{ lb coal/h} \times 8,630 \text{ Btu/lb} + 31,537 \text{ lb sawdust/h} \times 5,165 \text{ Btu/lb}}$$

$$\text{ETE} = 79.8\%$$

4.3 COST ESTIMATE

For this economic analysis, the capital and operating costs for the biomass/Wyodak feedstock plant result from a proportional adjustment from the baseline 600°C hydrogen plant which operates on Pittsburgh No. 8 coal. Whereas the cost of Pittsburgh No. 8 coal was \$1.00 per MMBtu, the cost of the Wyodak/biomass blend is assumed to be \$0.65/MMBtu, followed with a sensitivity case of \$0.50/MMBtu. The approach to the cost estimate was the same as before and detailed in Section 2.7. The financial parameters were the same as detailed in Table 1-2 and Table 2-19 except for the type and cost of the coal.

The results of the cost estimating activity are summarized in Table 4-6 and Table 4-7.

Table 4-6
Capital Estimate and Revenue Requirement Summary
\$0.65/MMBtu Feedstock

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY			
TITLE/DEFINITION			
Case:	Hydrogen Fuel Facility (Wy/Bio Blend) w/600C Hot Gas Desulfurization		
Plant Size:	400.0 H ₂ TPD	HeatRate:	(Btu/kWh)
Primary/Secondary Fuel(type):	Wyodak/Bion	Cost:	0.65 (\$/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/H₂TPD	
Process Capital & Facilities	297,459	743.6	
Engineering(incl.C.M.,H.O.& Fee)	28,793	72.0	
Process Contingency	6,169	15.4	
Project Contingency	33,242	83.1	
TOTAL PLANT COST(TPC)	\$365,662	914.1	
TOTAL CASH EXPENDED	\$365,662		
AFDC	\$23,446		
TOTAL PLANT INVESTMENT(TPI)	\$389,109	972.7	
Royalty Allowance			
Preproduction Costs	9,229	23.1	
Inventory Capital	2,304	5.8	
Initial Catalyst & Chemicals(w/equip.)			
Land Cost	150	0.4	
TOTAL CAPITAL REQUIREMENT(TCR)	\$400,792	1001.9	
OPERATING & MAINTENANCE COSTS (2000 Dollars)	\$x1000	\$x1000/H₂TPD	
Operating Labor	3,871	9.7	
Maintenance Labor	2,604	6.5	
Maintenance Material	3,906	9.8	
Administrative & Support Labor	1,619	4.0	
TOTAL OPERATION & MAINTENANCE	\$12,000	30.0	
FIXED O & M		24.00	
VARIABLE O & M		6.00	
CONSUMABLE OPERATING COSTS,less Fuel (2000 Dollars)	\$x1000	\$/T H₂-yr	
Water	88	0.75	
Chemicals	1,066	9.13	
Other Consumables			
Waste Disposal	603	5.17	
TOTAL CONSUMABLE OPERATING COSTS	\$1,757	15.04	
BY-PRODUCT CREDITS (2000 Dollars)	(\$4,251)	-36.39	
FUEL COST (2000 Dollars)	\$11,900	101.87	
PRODUCTION COST SUMMARY	1st Year (2005 \$)	Levelized (Over Book Life \$)	
	\$/T H₂-yr	\$/T H₂-yr	
Fixed O & M	82.18	82.18	0.6725696
Variable O & M	6.00	6.00	0.0490976
Consumables	15.04	15.04	0.1230982
By-product Credit/Penalty	-36.39	-36.39	-0.29785
Fuel	98.85	90.17	0.7379375
TOTAL PRODUCTION COST	165.68	157.00	1.2848532
LEVELIZED CARRYING CHARGES(Capital)		480.35	3.931157
LEVELIZED(Over Book Life)COST/Ton of H₂		637.34	
Equivalent \$/MMBtu		5.22	

Table 4-7
Capital Estimate and Revenue Requirement Summary
\$0.50/MMBtu Feedstock

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY			
TITLE/DEFINITION			
Case:	Hydrogen Fuel Facility (Wy/Bio Blend) w/600C Hot Gas Desulfurization		
Plant Size:	400.0 H ₂ TPD	HeatRate:	(Btu/kWh)
Primary/Secondary Fuel(type):	Wyodak/Biorr	Cost:	0.50 (\$/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/H₂TPD
Process Capital & Facilities		297,459	743.6
Engineering(Incl.C.M.,H.O.& Fee)		28,793	72.0
Process Contingency		6,169	15.4
Project Contingency		33,242	83.1
TOTAL PLANT COST(TPC)		\$365,662	914.1
TOTAL CASH EXPENDED	\$365,662		
AFDC	\$23,446		
TOTAL PLANT INVESTMENT(TPI)		\$389,109	972.7
Royalty Allowance			
Preproduction Costs		9,160	22.9
Inventory Capital		2,026	5.1
Initial Catalyst & Chemicals(w/equip.)			
Land Cost		150	0.4
TOTAL CAPITAL REQUIREMENT(TCR)		\$400,445	1001.0
OPERATING & MAINTENANCE COSTS (2000 Dollars)		\$x1000	\$x1000/H₂TPD
Operating Labor		3,871	9.7
Maintenance Labor		2,604	6.5
Maintenance Material		3,906	9.8
Administrative & Support Labor		1,619	4.0
TOTAL OPERATION & MAINTENANCE		\$12,000	30.0
FIXED O & M			24.00
VARIABLE O & M			6.00
CONSUMABLE OPERATING COSTS,less Fuel (2000 Dollars)		\$x1000	\$/T H₂-yr
Water		88	0.75
Chemicals		1,066	9.13
Other Consumables			
Waste Disposal		603	5.17
TOTAL CONSUMABLE OPERATING COSTS		\$1,757	15.04
BY-PRODUCT CREDITS (2000 Dollars)		(\$4,251)	-36.39
FUEL COST (2000 Dollars)		\$9,154	78.36
PRODUCTION COST SUMMARY	1st Year (2005 \$)	Levelized (Over Book Life \$)	
	\$/T H₂-yr	\$/T H₂-yr	
Fixed O & M	82.18	82.18	0.6725696
Variable O & M	6.00	6.00	0.0490976
Consumables	15.04	15.04	0.1230982
By-product Credit/Penalty	-36.39	-36.39	-0.29785
Fuel	76.04	69.36	0.5676442
TOTAL PRODUCTION COST	142.87	136.19	1.1145599
LEVELIZED CARRYING CHARGES(Capital)		479.93	3.9277537
LEVELIZED(Over Book Life)COST/Ton of H₂		616.12	
Equivalent \$/MMBtu		5.04	

4.4 SUMMARY AND CONCLUSIONS

The purpose of this brief study was to compare the economics of producing hydrogen from a Wyodak/biomass blend against producing hydrogen from bituminous coal in the same sized plant. Table 4-8 is a summary comparison of the performance and cost results. The costs of hydrogen from both feedstocks are approximately equal. This is due to a balance of capital charges, fuel costs, and byproduct credits.

Table 4-8
Performance and Cost Summary Comparisons
Wyodak/Biomass Blend vs. Pittsburgh No. 8

	90% Wyodak 10% Biomass	Baseline Case Pittsburgh No. 8 600°C Membrane
Coal Feed	283,833 lb/h	221,631 lb/h
Biomass Feed	31,537 lb/h	N/A
Oxygen Feed (95%) to Gasifier	186,650 lb/h	165,818 lb/h
Oxygen Feed to Retentate Combustor	25,300 lb/h	58,701 lb/h
Water to Prepare Feed Slurry	114,009 lb/h	94,025 lb/h
Hydrogen Product Stream	33,337 lb/h	35,903 lb/h
CO ₂ Product Stream	575,923 lb/h	582,566 lb/h
Sulfuric Acid Product	5,057 lb/h	19,482 lb/h
Gross Power Production		
Turbine Expander	55 MW	84 MW
Steam Turbine	28 MW	N/A
Auxiliary Power Requirement	(69 MW)	(77 MW)
Net Power Production	14 MW	7 MW
Net Plant Water Makeup	100,979 lb/h	198,150 lb/h
Effective Thermal Efficiency (ETE), HHV	79.8%	80.4%
Capital Cost, \$1,000	\$365,662	\$359,791
Hydrogen Product Cost, \$/MMBtu	\$5.22 (\$0.65 Feedstock) \$5.04 (\$0.50 Feedstock)	\$5.06

Total plant costs are roughly equal, resulting from a combination of increased and decreased equipment requirements. The cost adjustments to the hydrogen plant due to the change over to the Wyodak/biomass blend are reflected in increased feedstock handling, increased oxygen plant size due to the higher water content (and associated increase in CO₂ content), and the need for a steam turbine which produces 28 MW from excess low-pressure steam. The capital costs were lower in sulfur control areas because of the low-sulfur feedstock, resulting in only 61 tpd sulfuric acid production from the blend versus 234 tpd from bituminous coal. This resulted in a lowering of byproduct credits.

The cost of biomass was not explored. Rather, two feedstock costs were used, \$0.65 and \$0.50/MMBtu. The higher cost reflects biomass being equal to Wyodak in delivered cost, the lower reflecting essentially free biomass.

The amount of hydrogen produced from the Wyodak/biomass blend is lowered by about 7 percent, primarily due to the higher level of CO₂ produced in the gasifier. This resulted in a lowered amount of reactive syngas (H₂ and CO) available for hydrogen production.

In April 1999 a version of the base case hydrogen plant was prepared in which Wyodak Coal was substituted for Pittsburgh No. 8 coal. A full description of the comparison is not included in this compilation report because the results are not readily comparable, for the following reasons:

- The HSD operates at 1000°C
- Sulfur is recovered with FGD, rather than sulfuric acid
- Plant Capacity Factor is 95%, Book Life is 30 years
- Costs are in 1997 dollars

A summary of the performance and economic results from the Wyodak substitution are shown in Table 4-9. Less hydrogen is produced, but more power from excess plant steam is produced. The cost of hydrogen from the Wyodak substitution is slightly lower than the Pittsburgh No. 8, primarily because of the lower cost of coal.

Table 4-9
Performance and Cost Summary Comparisons
1999 Wyodak Substitution for Pittsburgh No. 8

	100% Wyodak	100% Pittsburgh No. 8
Coal Feed	283,833 lb/h	221,631 lb/h
Oxygen Feed (95%)	220,986 lb/h	252,369 lb/h
Water to Prepare Feed Slurry	109,249 lb/h	94,025 lb/h
Hydrogen Product Stream	29,221 lb/h	34,004 lb/h
Limestone Sorbent to FGD	10,583 lb/h	25,188 lb/h
CO ₂ Product Stream	538,410 lb/h	603,324 lb/h
Net Plant Water Makeup	99,960 lb/h	188,878 lb/h
Gross Power Production		
ATS Turbine Expander	102 MW	120 MW
Steam Turbine	20 MW	N/A
Auxiliary Power Requirement	(67 MW)	(78 MW)
Net Power Production	55 MW	42 MW
Effective Thermal Efficiency (ETE), HHV	80.4%	79.8%
Capital Cost, \$1,000	\$313,597	\$306,605
Hydrogen Product Cost, \$/MMBtu	\$3.91 (\$0.67/MMBtu Feedstock)	\$4.05 (\$1.00/MMBtu Feedstock)

5. REFERENCES

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