CAPITAL AND OPERATING COST OF HYDROGEN PRODUCTION FROM COAL GASIFICATION

Final Report

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PARSONS

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LIST OF ACRONYMS AND ABBREVIATIONS

acfm	actual cubic feet per minute
AGR	acid gas removal
Al_2O_3	aluminum oxide
ASU	air separation unit
BOP	balance of plant
Btu	British thermal unit
°C	degree Centigrade
cfm	cubic feet per minute
CO	carbon monoxide
CO_2	carbon dioxide
COS	carbonyl sulfide
DEA	diethanolamine
DGA	diglycolamine
DIPA	diisopropanolamine
DOE	U.S. Department of Energy
°F	degrees Fahrenheit
Fe_2O_3	ferric (iron) oxide
gpm	gallons per minute
h	hour
H_2O	water
H_2S	hydrogen sulfide
H_2SO_4	sulfuric acid
HHV	higher heating value
HP	high pressure
HRSG	heat recovery steam generator
HT	high temperature
IGCC	integrated gasification combined cycle
IP	intermediate pressure
IRR	internal rate of return
K_2O	potassium oxide
kW_e	kilowatt hour
lb	pound
LP	low pressure
LT	low temperature
MAF	moisture and ash free
Mcf	thousand cubic feet
MDEA	methyl-diethanolamine
MEA	monoethanolamine
MgO	magnesium oxide
MM	million
MW_e	megawatt electric

Na_2O	sodium oxide
NREL	National Renewable Energy Laboratory
O&M	operation and maintenance
O_2	oxygen
OH	overhead
OJ	operating job
P_2O_5	phosphorous pentoxide
ppm	parts per million
ppmv	parts per million by volume
PRB	Powder River Basin
PSA	pressure swing adsorption
psia	pounds per square inch absolute
psig	pounds per square inch gauge
rpm	revolutions per minute
scf	standard cubic feet
scfd	standard cubic feet per day
scfm	standard cubic feet per minute
SiO_2	silica
SO_2	sulfur dioxide
SO_3	sulfur trioxide
SS	stainless steel
TCR	total capital requirement
TG	turbine generator
TiO_2	titanium dioxide
TM	trademark
TPC	total plant cost
tpd	tons per day
tph	tons per hour
vol %	volume percent

EXECUTIVE SUMMARY

BACKGROUND

In 1999, both Parsons Infrastructure & Technology Group (Parsons) and the National Renewable Energy Laboratory (NREL) prepared conceptual plant designs and cost estimates for producing hydrogen from coal gasification. Parsons' approach to producing hydrogen focused on integrating high-temperature ceramic membranes with coal gasification to both shift and separate hydrogen from the syngas.[1] Parsons also prepared a base case design for hydrogen from coal gasification utilizing conventional technology. The NREL approach to plant design focused on advanced and conventional technology for hydrogen production with high-temperature gas cleanup, shift, and PSA purification, augmented with various concepts to sequester CO_2 and increase hydrogen production.[2] These concepts consisted of a base case design for production of hydrogen from coal gasification accompanied by CO_2 sequestration in coal seams, reforming extracted methane, and producing power from extracted coal seam methane. The base case cost for producing hydrogen from coal gasification was reported by Parsons to be $5.57/MMBtu, while NREL reported the base case cost for hydrogen from coal gasification to be $18.97/MMBtu.

The primary differences in the cost of hydrogen from the Parsons and NREL plants can be realized from the Total Plant Investment (TPI). The TPI for the NREL plant per unit of hydrogen production is 2.3 times that of the Parsons plant.

Due to the wide differences in reported costs for capital and the need to provide a baseline cost for hydrogen production, NETL has tasked Parsons to review its prior plant design and cost estimate for producing hydrogen from coal gasification utilizing commercial technology. The key benefit of utilizing commercial technology is the obtaining of credible cost estimates for the plant, with a minimum of process contingency. The results of this effort are intended to prepare a basis from which to utilize individualized financial parameters in the U.S. Department of Energy (DOE) Integrated Gasification Combined Cycle (IGCC) Cost Estimating Model to arrive at a selling price for hydrogen.

Focus of the plant design will be from a common thermal gasifier throughput. Two coals will be reviewed, Pittsburgh No. 8 and PRB Wyodak. Hydrogen costs from these coals will be prepared to quantify the differing plant characteristics associated with bituminous coal or sub-bituminous coal.

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[1] "Decarbonized Fuel Production Facilities/Base Case Comparisons," Letter Report, U.S. DOE, June 1999.

[2] Spath, Pamela and Amos, Wade, "Technoeconomic Analysis of Hydrogen Production from Low-Btu Western Coal Augmented with CO_2 Sequestration and Coalbed Methane Recovery Including Delivered Hydrogen Costs," NREL, September 1999.

INTRODUCTION

The objective of this task was to prepare capital and operating cost data to be used to arrive at a plant gate cost for hydrogen produced from coal gasification. The two coals used in this study are Pittsburgh No. 8 bituminous and Wyodak Powder River Basin (PRB) sub-bituminous. Hydrogen cost was determined by first preparing two plant designs for hydrogen production, based on currently available process technology, and meeting current permitting regulations for environmental compliance. These baseline plants will not capture CO_2.

To arrive at a cost estimate for hydrogen, the design included commercially available process technology obtained from verifiable sources. The plants utilized commercially available technology including a Wabash River-scale Destec (E-Gas™) gasifier, conventional gas cooling, commercial shift conversion and acid gas cleanup, commercial sulfuric acid technology, and commercial pressure swing adsorption (PSA). The E-Gas™ gasifier is the gasifier of choice for this study since it has been operated on both bituminous and sub-bituminous coals. Figure ES-1 is the block flow diagram for the plant.

Based on financial assumptions typically used by Parsons for IGCC, the cost of hydrogen was estimated for Pittsburgh No. 8 coal and for Wyodak PRB coal at the plant gate. The results of these two cases were imported into the DOE IGCC financial model. Using the financial model, sensitivities of the effect of financial parameters can easily be determined. When different financial parameters are defined, the impact can be quantified.

Figure ES-1
Block Flow Diagram
Conventional Hydrogen Plant

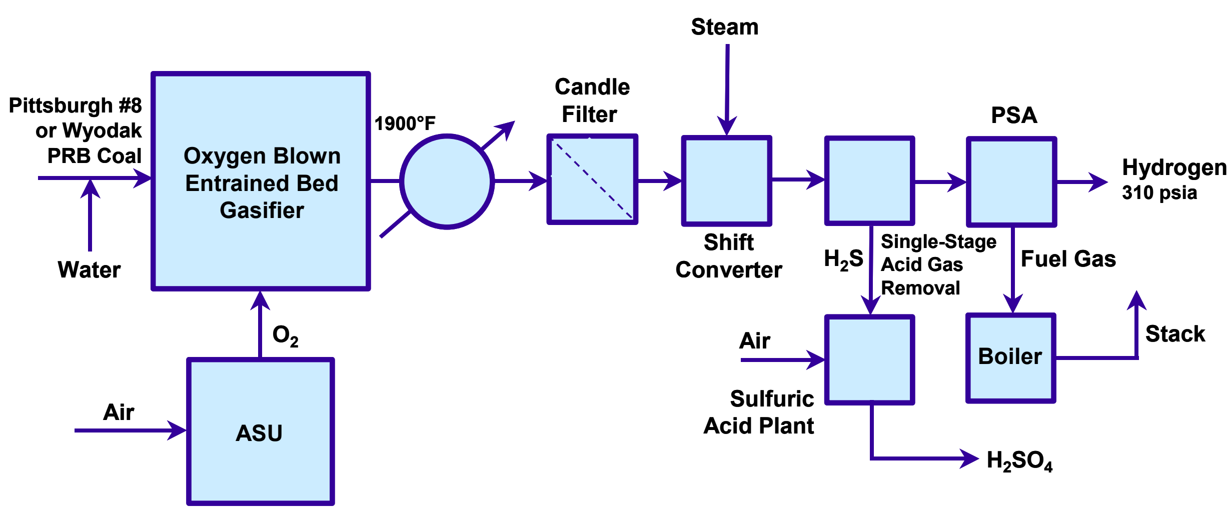


Table ES-1 lists the plant design criteria and site conditions.

Table ES-1
Design Criteria for Conventional Hydrogen Production Plant

Hydrogen Production Plant Parameter	Hydrogen Production Plant Design Basis
Ambient Conditions	14.7 psia, 60°F, river water access
Coal Feed	Pittsburgh No. 8/PRB Wyodak
Gasifier	Oxygen-blown E-Gas™ with second stage adjusted for 1900°F output
Coal Feed Rate	2,500 tpd dry basis
Hot Gas Temperature	~1900°F
Gasifier Outlet Pressure	450 psia
Gas Quench/Cooling	625°F
Metallic Candle Filter	Following quench/cooling
CO-Shift	Single-stage high-temperature, sulfur-tolerant
Desulfurization	Proprietary amine
Sulfur Recovery	Sulfuric acid byproduct
CO_2 Recovery	None
Hydrogen Purification	Pressure swing adsorption (PSA)
PSA Retinate Gas	Fired in auxiliary boiler
CO_2 Product Pressure	N/A
Hydrogen Utilization	315 psia at plant gate
Auxiliary Power Block	Steam turbine generator
Plant Size	Maximum hydrogen production from 2,500 tpd dry coal feed
Plant Capacity Factor	90 percent

<u>Process Selection</u>

Gasifier. The E-Gas™ gasifier is selected for these plants because of the wide differences in the coals to be compared. The E-Gas™ two-stage design has resulted in successful operation on both bituminous and sub-bituminous coals. By comparison, the Texaco gasifier with its single-stage entrained slurry feed reaches operational limitations with high-moisture coals, e.g., sub-bituminous and lignite.

Shift Reactor Catalyst. For this plant design the CO converter was located upstream of the acid gas removal (AGR) unit. The CO shift catalyst selected for these plants is the Haldor-Topsoe SSK Sulfur Tolerant CO Conversion Catalyst. The plant will utilize a single-stage high-temperature shift, resulting in a CO conversion of greater than 80 percent. The SSK catalyst also promotes COS hydrolysis, thereby resulting in an acid gas consisting of all H_2S.

Acid Gas Removal. The traditional approach to acid gas removal is with regenerable amines. Other methods include removal of H_2S with membranes systems or with molecular sieves. Regenerable amines are by far the most popular means of removal of acid gas from all types of gaseous streams. Therefore, the AGR process selected for these plants is a proprietary amine with an H_2S concentrator on the regenerated acid gas. The gas from the AGR process, concentrated in H_2S, will be used as a feed for a Monsanto H_2S-fired sulfuric acid plant.

Hydrogen Purification. The three main processes for hydrogen purification are the pressure swing adsorption, the selective permeation process using polymer membranes, and the cryogenic separation process. Each of these processes is based on a different separation principle, and the process characteristics differ significantly. The PSA system was selected based on the ability to produce high purity (99.9 percent) hydrogen, low amounts of CO and CO_2, ease of operation, and a single system.

PITTSBURGH NO. 8 COAL

This section is dedicated to the design and cost estimate for a hydrogen plant fed with Pittsburgh No. 8 bituminous coal. This coal is characterized having high volatility, low ash and moisture content, and high as-received heating value. The high sulfur content results in a significant value-added from the sulfuric acid byproduct.

Heat and Material Balance

The heat and material balance for the IGCC plant is based on the maximum hydrogen production from 2,500 tons per day of dry coal. Ambient operating conditions are indicated in the plant design basis. The pressurized entrained flow E-Gas™ two-stage gasifier uses a coal/water slurry and oxygen to produce a medium heating value fuel gas. The syngas produced in the gasifier first stage at about 2450°F (1343°C) is quenched to 1900°F (1038°C) by reacting with slurry injected into the second stage. The syngas passes through a fire tube boiler syngas cooler and leaves at 1300°F (704°C). A second gas cooler in series cools the gas further to 645°F (341°C). High-pressure saturated steam is generated in the syngas coolers and is joined with the main steam supply.

The gas goes through a series of additional gas coolers and cleanup processes including a scrubber. Slag captured by the syngas scrubber is recovered in a slag recovery unit.

The syngas stream from the syngas scrubber enters the high-temperature shift converter, which contains a bed of sulfided shift catalyst. The shift reaction converts over 80 percent of the CO to hydrogen and CO_2 and hydrolyzes COS to H_2S. Following the shift converter, the cooled gas stream passes through a proprietary amine acid gas removal process, which removes H_2S and some of the CO_2. The clean gas stream then passes through the PSA for final purification of the hydrogen. Regeneration gas from the PSA contains fuel value, and is fed to the heat recovery steam generator (HRSG). Regeneration gas from the AGR plant is fed to a sulfuric acid plant.

The cryogenic oxygen plant supplies 99 percent purity oxygen to the gasifiers at the rated pressure. A dedicated air compressor provides air supply for the oxygen plants.

The steam cycle is based on maximizing heat recovery from the gasifier cooler and HRSG, as well as utilizing steam generation opportunities in the shift process.

Overall performance for the entire plant is summarized in Table ES-2, which includes auxiliary power requirements. The net plant output power, after plant auxiliary power requirements are deducted, is nominally 38 MW_e. The overall plant thermal effective efficiency (thermal value of hydrogen and power produced) is 62.3 percent, on an HHV basis.

Table ES-2
Performance Summary
Hydrogen Production from Pittsburgh No. 8 Coal

Plant Size, tons H_2/day (MMscfd) @ 346 psia	312.6 (112.2)
Coal Feed (dry basis)	2,500 tpd
Plant Availability	90%
Cold Gas Efficiency	57.7%
Equivalent Thermal Efficiency, HHV	62.3%
Gross Power Production	78.5 MW
Auxiliary Power	40.9 MW
Net Power	37.6 MW

Capital Cost

The total plant cost for the plant producing 313 tons of hydrogen per day from Pittsburgh No. 8 coal is $376.1 million in 2001 dollars. The capital cost summary is included in Table ES-3.

Consumables

Shift Catalyst:

- Change-out every 3 years
- 0.0045 pound of catalyst per 1,000 standard cubic feet of hydrogen
- 250 tons initial charge
- 85 tons per year annual cost

Proprietary Amine:

- 12 pounds per hour
- 100,000 pounds per year

Table ES-3
Capital Cost Summary – Hydrogen Production from Pittsburgh No. 8 Coal

Client: DEPARTMENT OF ENERGY — **Report Date:** 03-Jun-2002 01:19 PM
Project: NETL H_2 Production Facility

TOTAL PLANT COST SUMMARY

Case: Bituminous H_2 Plant w/o CO_2 Capture
Plant Size: 312.6 H_2 TPD — **Estimate Type:** Conceptual — Cost Base (Dec) 2001 ($X1000 & $X1000/TPD)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $	TOTAL PLANT COST $/TPD
1	COAL & SORBENT HANDLING	5,354	1,099	3,955	277		**$10,686**	855		1,154	**$12,694**	41
2	COAL & SORBENT PREP & FEED	8,987	1,307	4,905	343		**$15,542**	1,733		1,728	**$19,003**	61
3	FEEDWATER & MISC. BOP SYSTEMS	4,506	1,339	2,679	187		**$8,711**	697		941	**$10,348**	33
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries (E	51,616		21,976	1,538		**$75,130**	9,016		8,415	**$92,560**	296
4.2	Syngas Cooling	w/4.1		w/ 4.1	w/ 4.1			w/ 4.1		w/ 4.1		
4.3	ASU/Oxidant Compression	29,284		w/equip.			**$29,284**	2,343		3,163	**$34,789**	111
4.4-4.9	Other Gasification Equipment	6,881	5,691	4,811	337		**$17,720**	1,447		1,917	**$21,084**	67
	SUBTOTAL 4	*87,781*	*5,691*	*26,787*	*1,875*		***$122,134***	*12,805*		*13,494*	***$148,433***	*475*
5	HYDROGEN SEPARATION/GAS CLEAN	59,135	4,205	21,176	1,482		**$85,999**	10,111		9,611	**$105,721**	338
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Expander Turbine/Generator											
6.2-6.9	Combustion Turbine Accessories											
	SUBTOTAL 6											
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	4,533		551	39		**$5,123**	410		553	**$6,086**	19
7.2-7.9	HRSG Accessories, Ductwork and Stack	561	209	335	23		**$1,129**	90		122	**$1,341**	4
	SUBTOTAL 7	*5,094*	*209*	*886*	*62*		***$6,251***	*500*		*675*	***$7,427***	*24*
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	7,612		1,061	74		**$8,747**	700		945	**$10,392**	33
8.2-8.9	Turbine Plant Auxiliaries and Steam Pip	3,470	106	1,611	113		**$5,300**	424		572	**$6,296**	20
	SUBTOTAL 8	*11,082*	*106*	*2,672*	*187*		***$14,047***	*1,124*		*1,517*	***$16,688***	*53*
9	COOLING WATER SYSTEM	2,375	1,163	1,901	133		**$5,572**	446		602	**$6,619**	21
10	ASH/SPENT SORBENT HANDLING SYS	5,147	653	2,196	154		**$8,150**	888		904	**$9,941**	32
11	ACCESSORY ELECTRIC PLANT	5,029	2,077	4,075	285		**$11,467**	917		1,238	**$13,623**	44
12	INSTRUMENTATION & CONTROL	5,292	1,257	3,903	273		**$10,725**	858		1,158	**$12,741**	41
13	IMPROVEMENTS TO SITE	1,566	900	2,593	182		**$5,241**	419		566	**$6,226**	20
14	BUILDINGS & STRUCTURES		2,663	2,711	190		**$5,564**	445		601	**$6,610**	21
	TOTAL COST	**$201,347**	**$22,671**	**$80,439**	**$5,631**		**$310,088**	**$31,798**		**$34,189**	**$376,074**	**1203**

PSA Sorbent:

- Periodic change-out with scheduled maintenance

SO_2 Conversion Catalyst:

- Periodic change-out with scheduled maintenance

Byproduct Credits

The production of 229 tons of sulfuric acid per day is taken as a byproduct credit at $75 per ton.

WYODAK PRB COAL

This section is dedicated to the design and cost estimate for a hydrogen plant fed with Wyodak PRB sub-bituminous coal. This coal is characterized having low volatility, high ash and moisture content, and a lower as-received heating value. The low sulfur content results in a lesser value-added from the sulfuric acid byproduct.

Heat and Material Balance

The heat and material balance for the IGCC plant is based on the maximum hydrogen production from 2,500 tons per day of dry coal. Ambient operating conditions are indicated in the plant design basis. The pressurized entrained flow E-Gas™ two-stage gasifier uses a coal/water slurry and oxygen to produce a medium heating value fuel gas. The syngas produced in the gasifier first stage at about 2450°F (1343°C) is quenched to 1900°F (1038°C) by reacting with slurry injected into the second stage. The syngas passes through a fire tube boiler syngas cooler and leaves at 1300°F (704°C). A second gas cooler in series cools the gas further to 645°F (341°C). High-pressure saturated steam is generated in the syngas coolers and is joined with the main steam supply.

The gas goes through a series of additional gas coolers and cleanup processes including a scrubber. Slag captured by the syngas scrubber is recovered in a slag recovery unit.

The syngas stream from the syngas scrubber enters the high-temperature shift converter, which contains a bed of sulfided shift catalyst. The shift reaction converts over 80 percent of the CO to hydrogen and CO_2 and hydrolyzes COS to H_2S. Following the shift converter, the cooled gas stream passes through a proprietary amine acid gas removal process, which removes H_2S and some of the CO_2. The clean gas stream then passes through the PSA for final purification of the hydrogen. Regeneration gas from the PSA contains fuel value, and is fed to the HRSG. Regeneration gas from the AGR plant is fed to a sulfuric acid plant.

The cryogenic oxygen plant supplies 99 percent pure oxygen to the gasifiers at the rated pressure. A dedicated air compressor provides air supply for the oxygen plants.

The steam cycle is based on maximizing heat recovery from the gasifier cooler and HRSG, as well as utilizing steam generation opportunities in the shift process.

Overall performance for the entire plant is summarized in Table ES-4, which includes auxiliary power requirements. The net plant output power, after plant auxiliary power requirements are deducted, is nominally 42 MW_e. The overall plant thermal effective efficiency (thermal value of hydrogen and power produced) is 59.7 percent, on an HHV basis.

Table ES-4
Performance Summary
Hydrogen Production from Wyodak Coal

Plant Size, tons H_2/day (MMscfd) @ 346 psia	259.2 (93.1)
Coal Feed (dry basis)	2,500 tpd
Plant Availability	90%
Cold Gas Efficiency	54.2%
Equivalent Thermal Efficiency, HHV	59.7%
Gross Power Production	81.5 MW
Auxiliary Power	39.6 MW
Net Power	41.9 MW

Capital Cost

The total plant cost for the plant producing 313 tons of hydrogen per day from Wyodak PRB coal is $364.6 million in 2001 dollars. The capital cost summary is included in Table ES-5.

Table ES-5
Capital Cost Summary – Hydrogen Production from Wyodak Coal

Client: DEPARTMENT OF ENERGY
Project: NETL H_2 Production Facility
Report Date: 03-Jun-2002 03:16 PM

TOTAL PLANT COST SUMMARY

Case: Sub-Bituminous H_2 Plant w/o CO_2 Capture
Plant Size: 259.2 H_2 TPD **Estimate Type:** Conceptual Cost Base (Dec) 2001 ($X1000 & $X1000/TPD)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $	TOTAL PLANT COST $/TPD
1	COAL & SORBENT HANDLING	6,242	1,282	4,611	323		**$12,457**	997		1,345	**$14,799**	57
2	COAL & SORBENT PREP & FEED	10,581	1,539	5,776	404		**$18,299**	2,040		2,034	**$22,373**	86
3	FEEDWATER & MISC. BOP SYSTEMS	4,259	1,253	2,561	179		**$8,253**	660		891	**$9,804**	38
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries (E	54,709		23,270	1,629		**$79,607**	9,553		8,916	**$98,076**	378
4.2	Syngas Cooling	w/4.1		w/ 4.1	w/ 4.1			w/ 4.1		w/ 4.1		
4.3	ASU/Oxidant Compression	29,814		w/equip.			**$29,814**	2,385		3,220	**$35,419**	137
4.4-4.9	Other Gasification Equipment	7,665	5,742	5,067	355		**$18,829**	1,538		2,037	**$22,404**	86
	SUBTOTAL 4	*92,188*	*5,742*	*28,336*	*1,984*		*$128,250*	*13,476*		*14,173*	*$155,899*	*602*
5	HYDROGEN SEPARATION/GAS CLEAN	46,572	3,701	16,976	1,188		**$68,437**	8,005		7,644	**$84,086**	324
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Expander Turbine/Generator											
6.2-6.9	Combustion Turbine Accessories											
	SUBTOTAL 6											
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	4,533		551	39		**$5,123**	410		553	**$6,086**	23
7.2-7.9	HRSG Accessories, Ductwork and Stack	496	185	296	21		**$997**	80		108	**$1,185**	5
	SUBTOTAL 7	*5,029*	*185*	*847*	*59*		*$6,120*	*490*		*661*	*$7,271*	*28*
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	7,843		1,094	77		**$9,013**	721		973	**$10,707**	41
8.2-8.9	Turbine Plant Auxiliaries and Steam Pip	3,567	109	1,656	116		**$5,447**	436		588	**$6,471**	25
	SUBTOTAL 8	*11,409*	*109*	*2,749*	*192*		*$14,460*	*1,157*		*1,562*	*$17,178*	*66*
9	COOLING WATER SYSTEM	2,438	1,193	1,951	137		**$5,719**	458		618	**$6,795**	26
10	ASH/SPENT SORBENT HANDLING SYS	4,322	559	1,845	129		**$6,855**	745		760	**$8,360**	32
11	ACCESSORY ELECTRIC PLANT	4,974	2,052	4,028	282		**$11,335**	907		1,224	**$13,466**	52
12	INSTRUMENTATION & CONTROL	5,045	1,199	3,720	260		**$10,224**	818		1,104	**$12,146**	47
13	IMPROVEMENTS TO SITE	1,465	842	2,425	170		**$4,902**	392		529	**$5,824**	22
14	BUILDINGS & STRUCTURES		2,630	2,702	189		**$5,521**	442		596	**$6,559**	25
	TOTAL COST	**$194,524**	**$22,285**	**$78,527**	**$5,497**		**$300,832**	**$30,586**		**$33,142**	**$364,560**	**1407**

Consumables

Shift Catalyst:

- Change-out every 3 years
- 0.0045 pound of catalyst per 1,000 standard cubic feet of hydrogen
- 210 tons initial charge
- 70 tons per year annual cost

Proprietary Amine:

- 12 pounds per hour
- 100,000 pounds per year

PSA Sorbent:

- Periodic change-out with scheduled maintenance

SO_2 Conversion Catalyst:

- Periodic change-out with scheduled maintenance

Byproduct Credits

The production of 61 tons of sulfuric acid per day is taken as a byproduct credit at $75 per ton.

BASIS OF COST OF HYDROGEN COMPARISONS FOR VARIOUS FINANCIAL ASSUMPTIONS

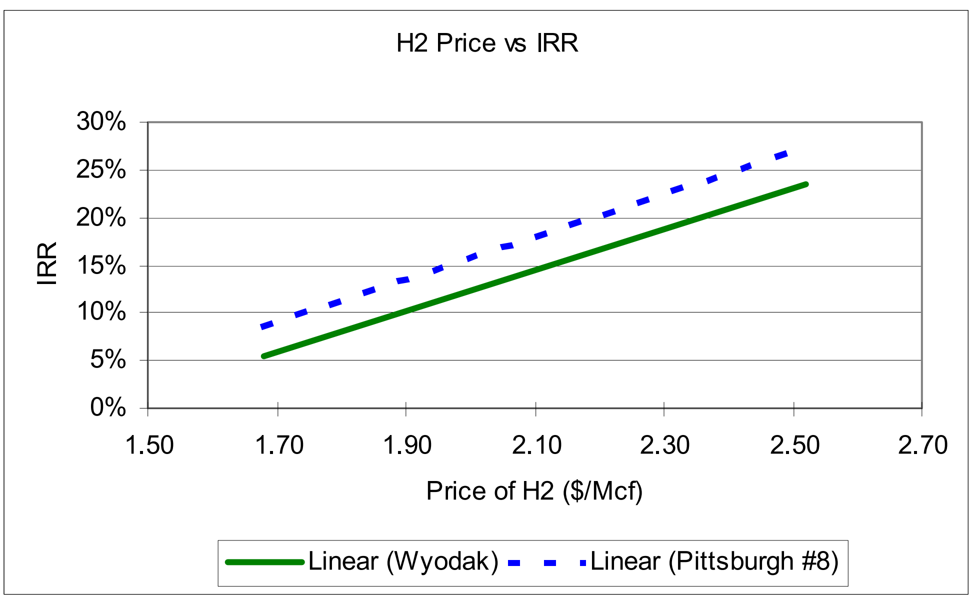
Based on financial assumptions typically used by Parsons for IGCC (see Table ES-6), the cost of hydrogen was estimated to be $6.01/MMBtu ($2.06/Mcf) for Pittsburgh No. 8 and $6.44/MMBtu ($2.20/Mcf) for Wyodak PRB coal at plant gate.

The results of these two cases were imputed into the DOE IGCC financial model. Using the financial model, sensitivities of the effect of financial parameters can easily be determined. When different financial parameters are defined, the impact can be quantified. Figure ES-2 shows one such variation, the internal rate of return (IRR) versus the cost of hydrogen.

Table ES-6
Financial Parameters

Levelized capacity factor	90%
Design/construction period	4 years
Plant startup date	January 2005
Land area/Unit cost	100 acres @ 41,500/acre
Project book life	20 years
Project tax life	20 years
Tax depreciation method	Accelerated based on ACRS class
Property tax rate	1.0% per year
Insurance tax rate	1.0% per year
Federal income tax rate	34.0%
State income tax rate	4.2%
Capital structure	
Common equity	20% @ 16.50% annum
Debt	80% @ 6.30% annum
Weighted cost of capital (after tax)	6.49%
Sulfur credit	$75/ton
Power sales	$30.00/MWh

Figure ES-2
Sensitivity of IRR to Hydrogen Costs



1. INTRODUCTION

1.1 BACKGROUND

In 1999, both Parsons Infrastructure & Technology Group (Parsons) and the National Renewable Energy Laboratory (NREL) prepared conceptual plant designs and cost estimates for producing hydrogen from coal gasification. Parsons' approach to producing hydrogen focused on integrating high-temperature ceramic membranes with coal gasification to both shift and separate hydrogen from the syngas.[3] Parsons also prepared a base case design for hydrogen from coal gasification utilizing conventional technology. This included a Wabash River-scale Destec gasifier, conventional gas cooling, commercial acid gas cleanup, commercial sulfuric acid technology, and commercial pressure swing adsorption.

The NREL approach to plant design focused on advanced and conventional technology for hydrogen production with high-temperature gas cleanup, shift, and pressure swing adsorption (PSA) purification, augmented with various concepts to sequester carbon dioxide (CO_2) and increase hydrogen production.[4] These concepts consisted of a base case design for production of hydrogen from coal gasification accompanied by CO_2 sequestration in coal seams, reforming extracted methane, and producing power from extracted coal seam methane. The base case cost for producing hydrogen from coal gasification was reported by Parsons to be $5.57/MMBtu, while NREL reported the base case cost for hydrogen from coal gasification to be $18.97/MMBtu.

1.1.1 Plant Configurations

Comparing the two base case plants, there are numerous differences. One difference between the plant designs had to do with the selection of coal. Parsons used Pittsburgh No. 8, while NREL used Wyodak PRB (Powder River Basin). The plants have the Destec gasifier in common, followed by quench and cooling with a fire-tube boiler heat exchanger. The single Destec gasifiers handled similar mass throughputs, and hydrogen production per thermal input is similar. However, the Wyodak coal resulted in less hydrogen production due to its lower heating value. The gas from both plants is cleaned of particulate with a metallic or ceramic filter at about 600°F.

To remove H_2S from the gas prior to the carbon monoxide (CO) shift reaction, NREL uses an advanced DOE hot gas desulfurization process. The sulfur dioxide (SO_2) released from regeneration of the sorbent is sent to a sulfuric acid plant. The desulfurized gas is sent to the high- and low-temperature shift reactors.

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[3] "Decarbonized Fuel Production Facilities/Base Case Comparisons," Letter Report, U.S. DOE, June 1999.

[4] Spath, Pamela and Amos, Wade, "Technoeconomic Analysis of Hydrogen Production from Low-Btu Western Coal Augmented with CO_2 Sequestration and Coalbed Methane Recovery Including Delivered Hydrogen Costs," NREL, September 1999.

By using a sulfur-tolerant catalyst, as in the Parsons plant, the CO shift reaction can occur directly following the syngas cooler and filter. The hydrogen sulfide (H_2S) and most of the remaining CO_2 following the shift can be removed at low temperature. Haldor Topsoe and others[5] have indicated that they offer sulfur-tolerant shift catalysts for which the presence of H_2S is actually beneficial to maintaining catalyst activity. The Parsons plant design uses a staged Selexol acid gas removal (AGR) process following the sour shift. This results in separate CO_2 and H_2S streams. The CO_2 is recovered at low pressure, and the H_2S is fed to a sulfuric acid plant.

Parsons uses the combined sources of steam from the plant cooling for generation of power with steam turbines, and the low-pressure steam is used for regeneration of the AGR process. To achieve maximum CO_2 recovery, the retinate gas from the PSA can be fired in a heat recovery steam generator (HRSG) with oxygen to produce steam for power and a clean recoverable CO_2 stream in the stack.

Table 1-1 lists the differences and common features in the two designs.

Table 1-1
Comparison of Parsons and NREL Plants

	Differing		**Common**
	Parsons	**NREL**	
Coal	Pittsburgh No. 8	PRB Wyodak	
Gasifier			Destec
Gas Cleanup	Cold	Hot	
ASU			Conventional cryogenic
Acid Gas Removal	Selexol	Hot zinc titanate	
Sulfur Product			Sulfuric acid
Water Gas Shift	Sour gas high temperature	Clean gas, high and low temperature	
Hydrogen Purification			PSA
Excess Steam	Used to make power	Shipped off-site	
PSA Off-Gas	Fired in HRSG with oxygen to maximize CO_2	Treated as CO_2 and used for coal bed methane	
Captured CO_2	Pure stream sent off-site	Coal bed methane	

[5] Rasmussen, H.W. and Houken, J., "Topsoe Hydrogen Plant Catalysts with Focus on Industrial Experience and Solutions to Operational Problems," Haldor Topsoe Refining Seminar, San Antonio, Texas, September 17-19, 1997.

1.1.2 Summary of Plant Differences

As shown in Table 1-2, the overall equivalent efficiency of the NREL plant differs from the Parsons plant due to the value of steam and PSA off-gas being credited in the numerator as a byproduct. Parsons used available energy streams to produce electricity, resulting in a lower equivalent efficiency.

Table 1-2
Performance Summary

Parameter	Parsons	NREL	Significant Differences
Coal Feed	221,631 lb/h	249,764 lb/h	
Thermal Input	2,759 MMBtu/h	2,155 MMBtu/h	
Oxygen Feed (95%)	178,860 lb/h	NOT REPORTED	
Hydrogen Product	26,487 lb/h	17,645 lb/h	
Sulfuric Acid Byproduct	19,100 lb/h	2,598 lb/h	Parsons has a larger H_2SO_4 plant due to higher sulfur in coal
Gross Power Production	77 MW	NOT REPORTED	
Auxiliary Power Requirement	41 MW	NOT REPORTED	
Net Power Production	36 MW	12 MW	Parsons produces more power
Equivalent Plant Efficiency, HHV	63.1%	83.0%	NREL takes credit for fuel value of PSA off-gas

1.1.3 Financial Assumptions and Cost Data

The differences in financial assumptions as shown in Table 1-3 should be recognized as being unique to the project for which the respective studies were conducted. Similar assumptions would lead to similar results.

Table 1-3
Financial Assumptions

Parameter	Parsons	NREL
Capacity Factor	95%	90%
Excess Steam Usage	Make power	Sell off-site
Coal Cost	$1.00/MMBtu	$0.82/MMBtu
Electricity Revenue	$30.00/MWh	$50.00/MWh
Debt/Equity	80/20	0/100
Cost of Debt	6.3%	N/A
Return on Equity	16.5%	15%
Book Life	30 years	20 years

1.1.4 Capital and Operating Costs

The primary differences in the cost of hydrogen from the Parsons and NREL plants can be realized from the Total Plant Investment (TPI) as shown in Table 1-4. The TPI for the NREL plant per unit of hydrogen production is 2.3 times that of the Parsons plant.

Table 1-4
Capital and Operating Cost Comparison

Parameter	Parsons	NREL	Significant Differences
Total Plant Cost	$374 million ('97)	NOT REPORTED	
Total Plant Investment	$398 million ('97)	$612 million ('95)	Significantly higher for comparable-sized plant
Cost/lbH_2/h	$15,026/lb/h	$34,684/lb/h	2.3x higher
Cost/scfd	$3.49/scfd	$8.05/scfd	2.3x higher
Annual Coal Cost	$22.960 million	$13.93 million	
Annual O&M	$12.836 million	17 million	
Annual Credits	($14.963 million)	($5.289 million)	
Steam Credit	None	NOT REPORTED	

1.1.5 The Need for Reconciliation of Cost Estimates

Due to the wide differences in reported costs for capital and the need to provide a baseline cost for hydrogen production, NETL has tasked Parsons to review its prior plant design and cost estimate for producing hydrogen from coal gasification. To arrive at a cost estimate for hydrogen based on reliable information that is acceptable throughout DOE, a design will be prepared using commercially available process technology, supplied by vendor sources that can provide quotations based on direct experience.

The key benefit of utilizing commercial technology is the obtaining of credible cost estimates for the plant, with a minimum of process contingency. The results of this effort are intended to prepare a basis from which to utilize individualized financial parameters in the DOE IGCC Cost Estimating Model to arrive at a selling price for hydrogen.

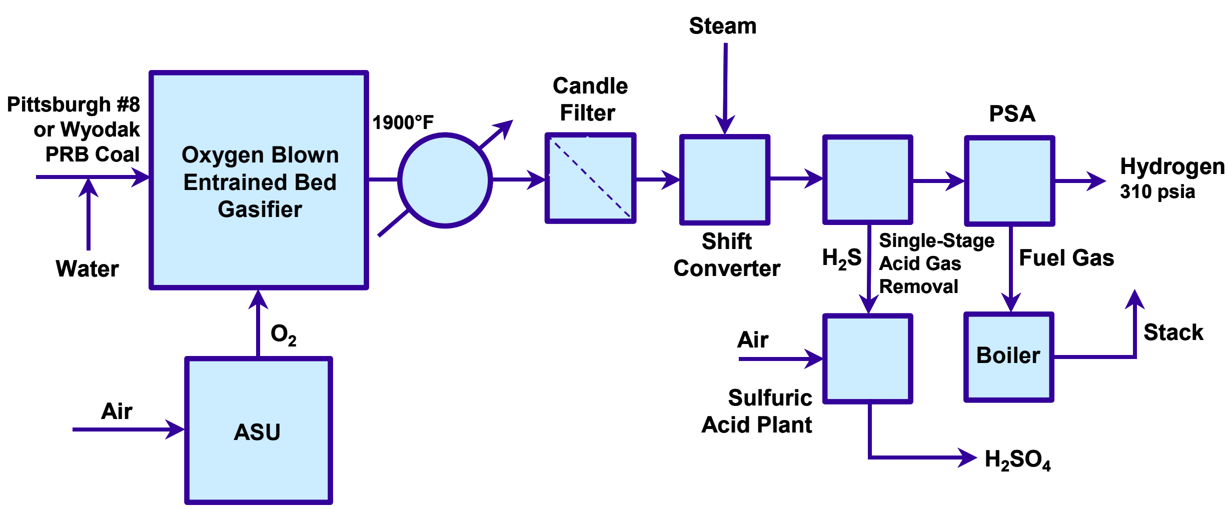
Focus of the plant design will be from a common thermal gasifier throughput. Two coals will be reviewed, Pittsburgh No. 8 and Wyodak PRB. Hydrogen costs from these coals will be prepared to quantify the differing plant characteristics associated with bituminous coal or sub-bituminous coal. Plant design areas common to each coal will be defined, but may be of different size due to coal selection.

1.2 TASK OBJECTIVE

The objective of this task was to prepare capital and operating cost data to be used to arrive at a plant gate cost for hydrogen produced from coal gasification. The two coals used in this study are Pittsburgh No. 8 bituminous and Wyodak PRB sub-bituminous. Hydrogen cost was determined by first preparing two plant designs for hydrogen production, based on currently available process technology, and meeting current permitting regulations for environmental compliance. These baseline plants will not capture CO_2.

To arrive at a cost estimate for hydrogen, the design included commercially available process technology obtained from verifiable sources. The plants utilized commercially available technology including a Wabash River-scale Destec (E-Gas™) gasifier, conventional gas cooling, commercial shift conversion and acid gas cleanup, commercial sulfuric acid technology, and commercial PSA. The E-Gas™ gasifier is the gasifier of choice for this study since it has been operated on both bituminous and sub-bituminous coals. Figure 1-1 is the block flow diagram for the plant.

Figure 1-1
Block Flow Diagram
Conventional Hydrogen Plant



The E-Gas™ gasifier is used to partially react a coal/water slurry with oxygen at high pressure. Gas exiting the gasifier is cooled in a fire-tube boiler to 625°F and cleaned of particulate matter. Particulates are recycled to the gasifier. Steam is added to the raw syngas, which passes through a reactor containing high-temperature sulfur-tolerant CO shift catalyst for conversion of the CO and steam to hydrogen and CO_2. The syngas containing predominantly hydrogen and CO_2 is cooled to less than 105°F and enters the AGR process. H_2S is removed and recovered for conversion to sulfuric acid. The remaining syngas goes through a PSA process to produce pure hydrogen at pressure. The PSA offgas is fired in an auxiliary boiler.

The plant design and cost estimates are addressed in two separate sections: Pittsburgh No. 8 bituminous coal and Wyodak PRB sub-bituminous coal. Sections 1.4 and 1.5 include the rationale for process selection and the basis for determining installed costs of the major process areas, respectively. Each section of this report contains a heat and material balance, process description with process flow diagram and stream composition tables, and a list of major equipment. Each section also presents the capital and operating costs and a calculated cost of hydrogen, based on preliminary economic assumptions. These hydrogen values are used in the final section, wherein the U.S. DOE integrated gasification combined cycle (IGCC) financial model is used to calculate the internal rate of return (IRR) for both coal cases.

1.3 PLANT DESIGN BASIS

1.3.1 Plant Capacity and Availability

The overall availability of the operating plant will be 90 percent. This is a high factor for single train gasification, and will result in two gasifier trains, operating at 50 percent capacity with the capability to ramp up to 100 percent. The balance of plant will be single train, operating at 100 percent capacity, based on commercial process operating experience as verified by equipment vendors.

Product Specifications:

- Sulfur as 98 percent pure H_2SO_4
- Hydrogen: 99 percent pure, 300 psig

Coal Properties:

- Pittsburgh No. 8; see Table 1-5.
- Wyodak PRB; see Table 1-6.

Table 1-5
Coal Analysis – Pittsburgh No. 8

Ultimate Analysis			
Constituent	**Air Dry, %**	**Dry, %**	**As Received, %**
Carbon	71.88	73.79	69.36
Hydrogen	4.97	4.81	5.18
Nitrogen	1.26	1.29	1.22
Sulfur	2.99	3.07	2.89
Ash	10.30	10.57	9.94
Oxygen	8.60	6.47	11.41
Total	100.00	100.00	100.00
Proximate			
		Dry Basis, %	**As Received, %**
Moisture		--	6.00
Ash		10.57	9.94
Volatile Matter		38.20	35.91
Fixed Carbon		51.23	48.15
Total		100.00	100.00
Sulfur		3.07	2.89
Btu Content		13,244	12,450
Moisture and Ash Free (MAF), Btu		14,810	
Ash Analysis, %			
Silica, SiO_2		48.1	
Aluminum Oxide, Al_2O_3		22.3	
Iron Oxide, Fe_2O_3		24.2	
Titanium Dioxide, TiO_2		1.3	
Calcium Oxide, CaO		1.3	
Magnesium Oxide, MgO		0.6	
Sodium Oxide, Na_2O		0.3	
Potassium Oxide, K_2O		1.5	
Sulfur Trioxide, SO_3		0.8	
Phosphorous Pentoxide, P_2O_5		0.1	
Total		100	
Ash Fusion Temperature			
		Reducing Atmosphere, °F	**Oxidizing Atmosphere, °F**
Initial Deformation		2015	2570
Spherical		2135	2614
Hemispherical		2225	2628
Fluid		2450	2685

Table 1-6
Wyodak Coal Properties

Proximate	As Received Basis	Dry Basis
Moisture	26.6	0
Volatile Matter	33.2	45.23
Fixed Carbon	34.4	46.87
Ash	5.8	7.90
Ultimate		
Sulfur	0.6	0.82
Hydrogen	6.5	4.82
Carbon	50.0	68.12
Nitrogen	0.9	1.23
Oxygen	36.2	17.11
Ash	5.8	7.90
Heating Value, HHV	8,630 Btu/lb	11,757 Btu/lb

Table 1-7 lists the plant design criteria and site conditions.

Table 1-7
Design Criteria for Conventional Hydrogen Production Plant

Hydrogen Production Plant Parameter	Hydrogen Production Plant Design Basis
Ambient Conditions	14.7 psia, 60°F, river water access
Coal Feed	Pittsburgh No. 8/PRB Wyodak
Gasifier	Oxygen-blown E-Gas™ with second stage adjusted for 1900°F output
Coal Feed Rate	2,500 tpd dry basis
Hot Gas Temperature	~1900°F
Gasifier Outlet Pressure	450 psia
Gas Quench/Cooling	625°F
Metallic Candle Filter	Following quench/cooling
CO-Shift	Single-stage high-temperature, sulfur-tolerant
Desulfurization	Proprietary amine
Sulfur Recovery	Sulfuric acid byproduct
CO_2 Recovery	None
Hydrogen Purification	Pressure swing adsorption
PSA Retinate Gas	Fired in auxiliary boiler
CO_2 Product Pressure	N/A
Hydrogen Utilization	315 psia at plant gate
Auxiliary Power Block	Steam turbine generator
Plant Size	Maximum hydrogen production from 2,500 tpd dry coal feed
Plant Capacity Factor	90 percent

1.4 PROCESS SELECTION

1.4.1 Gasifier

The E-Gas™ gasifier is selected for these plants because of the wide differences in the coals to be compared. The E-Gas™ two-stage design has resulted in successful operation on both bituminous and sub-bituminous coals. By comparison, the Texaco gasifier with its single-stage entrained slurry feed reaches operational limitations with high-moisture coals, e.g., sub-bituminous and lignite.

1.4.2 Shift Reactor Catalyst

For the conversion of the gasifier product to hydrogen, the first step is to convert most of the CO to hydrogen and CO_2 by reacting the CO with water over a bed of catalysts. This produces approximately 45 percent of the gross hydrogen product and converts more than 80 percent of the carbon monoxide to hydrogen and CO_2. The CO shift converter can be located either upstream of the AGR or immediately downstream. If the CO converter is located downstream of the AGR, then the metallurgy of the unit will be less stringent, but additional equipment must be added to the process. Products from the gasifier will be steam-injected to reach sufficient amounts of water vapor to meet the necessary water to gas criteria at the reactor inlet. If the CO converter is located downstream of the AGR, then the gasifier product would first have to be cooled and the free water separated and treated. Additional steam would have to be generated and reinjected into the CO converter feed to meet the required water-to-gas ratio. If the CO converter is located upstream of the AGR step, no additional equipment is required. Therefore, for this plant design the CO converter was located upstream of the AGR unit.

The CO shift catalyst selected for these plants is the Haldor-Topsoe SSK Sulfur Tolerant CO Conversion Catalyst. The plant will utilize a single-stage high-temperature shift, resulting in a CO conversion of greater than 80 percent. The SSK catalyst also promotes carbonyl sulfide (COS) hydrolysis, thereby resulting in an acid gas consisting of all H_2S.

1.4.3 Acid Gas Removal

The traditional approach to acid gas removal is with regenerable amines. Other methods include removal of H_2S with membranes systems or with molecular sieves. Regenerable amines are by far the most popular means of removal of acid gas from all types of gaseous streams.

Acid Gas Removal with Amines. The amine solvents are typically categorized into chemical, physical, and hybrid solvents. Hybrid solvents can be described as weak chemical solvents. The general flow scheme is similar for all of these solvents, and the choice depends on criteria such as AGR requirements, selectivity for H_2S compared to CO_2, organic sulfur removal requirements, regeneration energy requirements, and the presence of heavy hydrocarbons.

Chemical solvents remove CO_2 along with the H_2S. Examples of chemical solvents are monoethanolamine (MEA), diethanolamine (DEA), and diglycolamine (DGA).

Physical solvents are proprietary solvents that are selective toward H_2S and achieve the removal by equilibrium effects due to the more favorable solubility of H_2S. In order to achieve the solubility effect, refrigeration of the solution or compression and recycle is normally required, which increases the capital investment.

Hybrid solvents have removal capabilities between the chemical and physical solvents. The most common hybrid solvent is methy-diethanolamine (MDEA), which is a tertiary amine. In addition to MDEA, other hybrid solvents include diisopropanolamine (DIPA) and specialty amines such as Exxon's FLEXSORB, Union Carbide's UCARSOL HS solvents, and Dow Chemical's GAS/SPEC SS Solvent. FLEXSORB solvent is described as a severely sterically hindered amine and is the most selective for H_2S of any solvent currently marketed.

Acid Gas Removal with Membranes. Cellulose acetate membranes have been used successfully for acid gas pretreatment in gas processing facilities. Membranes are typically used for pretreatment of natural gas streams upstream of an amine unit. However, since the permeation rate for H_2S is similar to that of CO_2, membranes are not suitable for selective removal of H_2S.

Acid Gas Removal with Molecular Sieves. Molecular sieves have a large surface area in addition to highly localized polar charges and can be used for selective removal of H_2S. A type 5A molecular sieve is typically used for this type of application. One problem with the use of molecular sieves for H_2S removal is that the alumina in the molecular sieve catalyzes the formation of COS from H_2S and CO_2.

The basic criteria for selecting the technology were selective removal of H_2S, ease of operation, and a single type of system. Chemical absorption, e.g., MEA, MDEA proprietary amines, operates at lower pressure, and removes both CO_2 and H_2S. Therefore, the AGR process selected for these plants is a proprietary amine with an H_2S concentrator on the regenerated acid gas. The gas from the AGR process, concentrated in H_2S, will be used as a feed for a Monsanto H_2S-fired sulfuric acid plant.

1.4.4 Hydrogen Purification

The three main processes for hydrogen purification are the pressure swing adsorption, the selective permeation process using polymer membranes, and the cryogenic separation process. Each of these processes is based on a different separation principle, and the process characteristics differ significantly.

Pressure Swing Adsorption. The PSA units are based on the capacity of adsorbents to adsorb more impurities at high gas partial pressure than at low gas partial pressure.

These systems are the most commonly used. Two advantages of the PSA process are its ability to remove the very undesirable impurities down to a low level and to produce a very high purity hydrogen product. Typically, hydrogen product purities range from 99 to 99.999 vol %, and removal of CO and CO_2 to less than 10 ppmv is easily achieved. The amount of hydrogen recovered is dependent on inlet pressure, purge gas pressure, level of impurities, and hydrogen concentration. Hydrogen recovery with the feed gas produced for this project should be approximately 85 percent.

Polymer Membranes. Selective permeation through polymer membranes is a relatively recent and rapidly evolving commercial separation development. The process is based on the difference in permeation rates between hydrogen and impurities. Gas phase components must first dissolve into the membrane, then diffuse through it to the permeate side. For a hydrogen recovery membrane system, the very high hydrogen purity is not practical as the recovery of hydrogen falls rapidly as the purity goes up. For example, an increase in hydrogen purity from 95 to 98 percent will result in greater than a 25 percent decrease in hydrogen recovery.

Cryogenic Process. The cryogenic process is a low temperature separation process, which uses the difference in boiling temperatures of the feed components to effect the separation. Hydrogen has a high relative volatility compared to hydrocarbons. However, if the feed contains significant amounts of CO and CO_2 such as the feed in this project, a methane wash column is required. This column is used to wash the impurities from the hydrogen product stream, which is necessary to reduce CO and CO_2 to the low levels required. Also, because of the water in the feed stream, a drying system would have to be added upstream of the cryogenic system. Higher hydrogen recovery at moderate hydrogen purities (95 percent or less) is possible with a cryogenic system; however, very high hydrogen purity is not practical. Because of the type and composition of the feed gas, a cryogenic system is not acceptable.

The PSA system was selected based on the ability to produce high purity (99.9 percent) hydrogen, low amounts of CO and CO_2, ease of operation, and a single system.

1.5 APPROACH TO COST ESTIMATING

1.5.1 Gasifier

The gasifier specified for production of H_2 was the E-Gas™ gasifier. The cost in the evaluation was based on the reported cost of the E-Gas™ gasifier in the IGCC Reference Plant, E-Gas™, Final Report, dated February 2002. The same cost basis was utilized. In this evaluation, compared to the reference cost, two trains of gasifier were utilized to increase availability to 90 percent. The cost for the syngas cooling was adjusted on the basis of the difference in duty. The cost for the low-temperature (LT) heat recovery was evaluated as one common train separate from the equipment associated with each gasifier train.

1.5.2 Acid Gas Removal

The AGR process for this H_2 production plant is a proprietary amine system. The cost basis used in the estimate is an estimate developed by the Parsons process group. The reference price was developed for another application. For this application, the cost was adjusted for the required capacity and inclusion of the H_2S furnace.

1.5.3 Sulfur Recovery and Tail Gas Cleanup

A sulfuric acid plant was specified to handle the gas from the AGR process. For this application, the cost is based on a budgetary quote provided by Monsanto. This furnished price was prepared for a similar IGCC application and adjusted for this evaluation. The plant cost was adjusted for escalation and the change in plant capacity.

1.5.4 Hydrogen Purification

A PSA system was specified for the hydrogen purification process. The cost basis used in the estimate is based on a PSA system estimate developed by the Parsons process group. The reference price was developed for another application. For this application, the cost was adjusted for the significantly larger volumetric capacity (for both the Pittsburgh and Wyodak coals).

1.5.5 Shift Reactors

The shift reactor cost portion of the gas cleanup stream was based on cost information developed by the Parsons process group. The shift reactor portion that was part of another evaluation was utilized for this evaluation. In application for both coals, the cost was adjusted for somewhat smaller capacity of the reactors.

1.5.6 Candle Filters

The candle filters in the gas cleanup train were based on the cost of similar filters from other IGCC applications. The filter costs were originally based on pricing provided by Westinghouse and applied to the IGCC system requirements. These adjustments consisted of selecting the correct number of filter vessels to match the volumetric flow and candle type to match the temperature environment.

1.5.7 Air Separation Unit

The cost of the ASU portion of the gasification system was based on an in-house ASU cost model. This cost model was based on data provided by Air Products. The cost portion of the model was subsequently adjusted to reflect Parsons' experience with competitively furnished costs. The costs in this evaluation were adjusted for parameters such as capacity per day, purity, inlet pressure, and discharge pressure.

1.5.8 Balance of Plant

The costing of the balance of plant that constitutes the complete H_2 production IGCC was based on an in-house IGCC model that has been used to develop the capital costs and economic results for many IGCC applications. Each account within the model is adjusted to reflect the major cost parameter(s) for that component. Costs are adjusted on the basis of heat and mass balance data, equipment list, and plant arrangement drawing data.

2. PITTSBURGH NO. 8 COAL

This section is dedicated to the design and cost estimate for a hydrogen plant fed with Pittsburgh No. 8 bituminous coal. This coal is characterized having high volatility, low ash and moisture content, and high as-received heating value. The high sulfur content results in a significant value-added from the sulfuric acid byproduct.

2.1 HEAT AND MATERIAL BALANCE

The heat and material balance for the IGCC plant is based on the maximum hydrogen production from 2,500 tons per day of dry coal. Ambient operating conditions are indicated in the plant design basis. The pressurized entrained flow E-Gas™ two-stage gasifier uses a coal/water slurry and oxygen to produce a medium heating value fuel gas. The syngas produced in the gasifier first stage at about 2450°F (1343°C) is quenched to 1900°F (1038°C) by reacting with slurry injected into the second stage. The syngas passes through a fire tube boiler syngas cooler and leaves at 1300°F (704°C). A second gas cooler in series cools the gas further to 645°F (341°C). High-pressure saturated steam is generated in the syngas coolers and is joined with the main steam supply. The process flow diagram resulting from the heat and material balance is shown as Figure 2-1.

The gas goes through a series of additional gas coolers and cleanup processes including a scrubber. Slag captured by the syngas scrubber is recovered in a slag recovery unit.

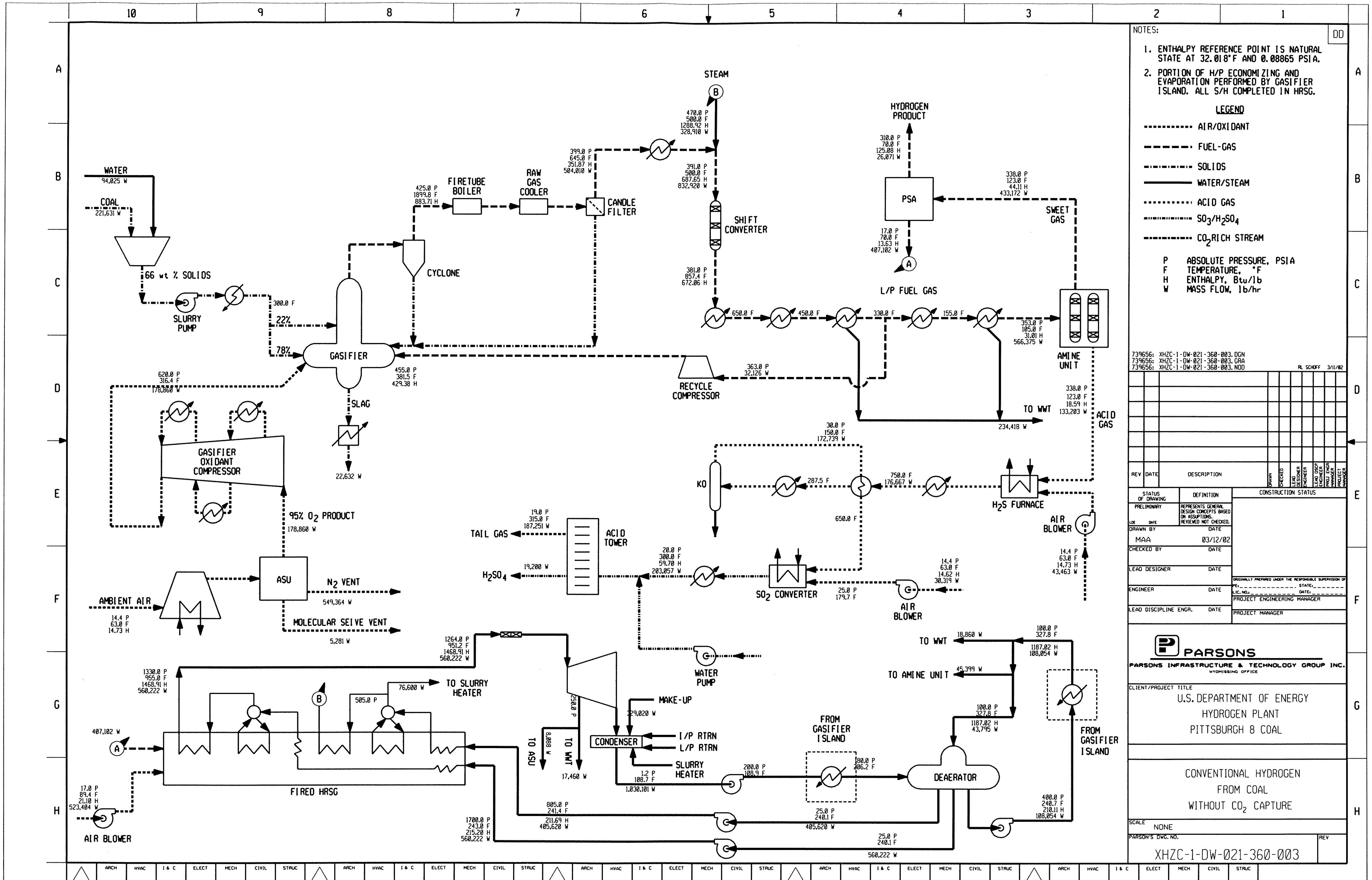
The syngas stream from the syngas scrubber enters the high-temperature shift converter, which contains a bed of sulfided shift catalyst. The shift reaction converts over 80 percent of the CO to hydrogen and CO_2 and hydrolyzes COS to H_2S. Following the shift converter, the cooled gas stream passes through a proprietary amine acid gas removal (AGR) process, which removes H_2S and some of the CO_2. The clean gas stream then passes through the pressure swing adsorption (PSA) for final purification of the hydrogen. Regeneration gas from the PSA contains fuel value, and is fed to the heat recovery steam generator (HRSG). Regeneration gas from the AGR plant is fed to a sulfuric acid plant.

The cryogenic oxygen plant supplies 99 percent pure oxygen to the gasifiers at the rated pressure. A dedicated air compressor provides air supply for the oxygen plants.

The steam cycle is based on maximizing heat recovery from the gasifier cooler and HRSG, as well as utilizing steam generation opportunities in the shift process.

The steam turbine selected to match this cycle is a two-casing, reheat, double-flow (exhaust) machine, exhausting downward to the condenser. The HP and IP turbine sections are contained in one casing, with the LP section in a second casing. The steam turbine drives a 3600 rpm hydrogen-cooled generator. The turbine exhausts to a single-pressure condenser operating at a nominal 1.2 psia at the 100 percent load design point. Two 50 percent capacity, motor-driven pumps are provided for feedwater and condensate.

Figure 2-1


NOTES:
1. ENTHALPY REFERENCE POINT IS NATURAL STATE AT 32.018°F AND 0.08865 PSIA.
2. PORTION OF H/P ECONOMIZING AND EVAPORATION PERFORMED BY GASIFIER ISLAND. ALL S/H COMPLETED IN HRSG.
LEGEND
AIR/OXIDANT
FUEL-GAS
SOLIDS
WATER/STEAM
ACID GAS
SO3/H2SO4
CO2 RICH STREAM
P ABSOLUTE PRESSURE, PSIA
F TEMPERATURE, °F
H ENTHALPY, Btu/lb
W MASS FLOW, lb/hr
WATER
COAL
SLURRY PUMP
GASIFIER
CYCLONE
FIRETUBE BOILER
RAW GAS COOLER
CANDLE FILTER
STEAM
SHIFT CONVERTER
HYDROGEN PRODUCT
PSA
SWEET GAS
L/P FUEL GAS
AMINE UNIT
RECYCLE COMPRESSOR
TO WWT
ACID GAS
SLAG
GASIFIER OXIDANT COMPRESSOR
95% O2 PRODUCT
ASU
N2 VENT
MOLECULAR SEIVE VENT
AMBIENT AIR
TAIL GAS
ACID TOWER
H2SO4
KO
SO2 CONVERTER
H2S FURNACE
AIR BLOWER
WATER PUMP
TO SLURRY HEATER
MAKE-UP
CONDENSER
L/P RTRN
SLURRY HEATER
TO ASU
FIRED HRSG
FROM GASIFIER ISLAND
DEAERATOR
TO AMINE UNIT
PARSONS
PARSONS INFRASTRUCTURE & TECHNOLOGY GROUP INC.
U.S. DEPARTMENT OF ENERGY
HYDROGEN PLANT
PITTSBURGH 8 COAL
CONVENTIONAL HYDROGEN FROM COAL WITHOUT CO2 CAPTURE
SCALE NONE
XHZC-1-DW-021-360-003
DRAWN BY MAA 03/12/02

Overall performance for the entire plant is summarized in Table 2-1, which includes auxiliary power requirements. The net plant output power, after plant auxiliary power requirements are deducted, is nominally 38 MW_e. The overall plant thermal effective efficiency (thermal value of hydrogen and power produced) is 62.3 percent, on an HHV basis.

Table 2-1
Plant Performance Summary
Hydrogen Production from Pittsburgh No. 8 Coal

Plant Output		
Steam Turbine Power	78,460	kW_e
Total	**78,460**	kW_e
Hydrogen Production		
Hydrogen Product	26,049	lb/h
Auxiliary Load		
Gasifier O_2 Compressor	9,470	kW_e
ASU Air Compressor	21,720	kW_e
Gasifier Slurry Pump	190	kW_e
Coal Handling	210	kW_e
Slag Handling	530	kW_e
Amine Unit	300	kW_e
Recycle Compressor	220	kW_e
H_2SO_4 Plant	100	kW_e
H_2S Furnace Air Blower	870	kW_e
Boiler Feedwater Pumps	1,570	kW_e
Steam Turbine Auxiliaries	250	kW_e
Cooling Tower	1,100	kW_e
Circulating Water Pumps	1,760	kW_e
Miscellaneous Balance of Plant	750	kW_e
Condensate Pumps	240	kW_e
Flue Gas Burner Air Blower	1,130	kW_e
Wastewater Treatment	500	kW_e
Total	40,910	kW_e
Plant Performance		
Net Auxiliary Load	40,910	kW_e
Net Plant Power	37,550	kW_e
Net Plant Efficiency (HHV)[1]	62.3%	
Coal Feed Flowrate	221,631	lb/h
Thermal Input[2]	808,673	kW_e
Condenser Duty	570.5	MMBtu/h

1 – Efficiency calculation includes thermal value of hydrogen and power produced.
2 – HHV of as-fed Pittsburgh No. 8 coal is 12,450 Btu/lb.

Figure 2-2 is a block flow diagram for the plant, and is accompanied by Table 2-2, which includes detailed process stream composition and state points.

Figure 2-2
Process Block Flow Diagram
E-Gas™ Gasifier-Based Hydrogen Production Plant – Pittsburgh No. 8 Coal

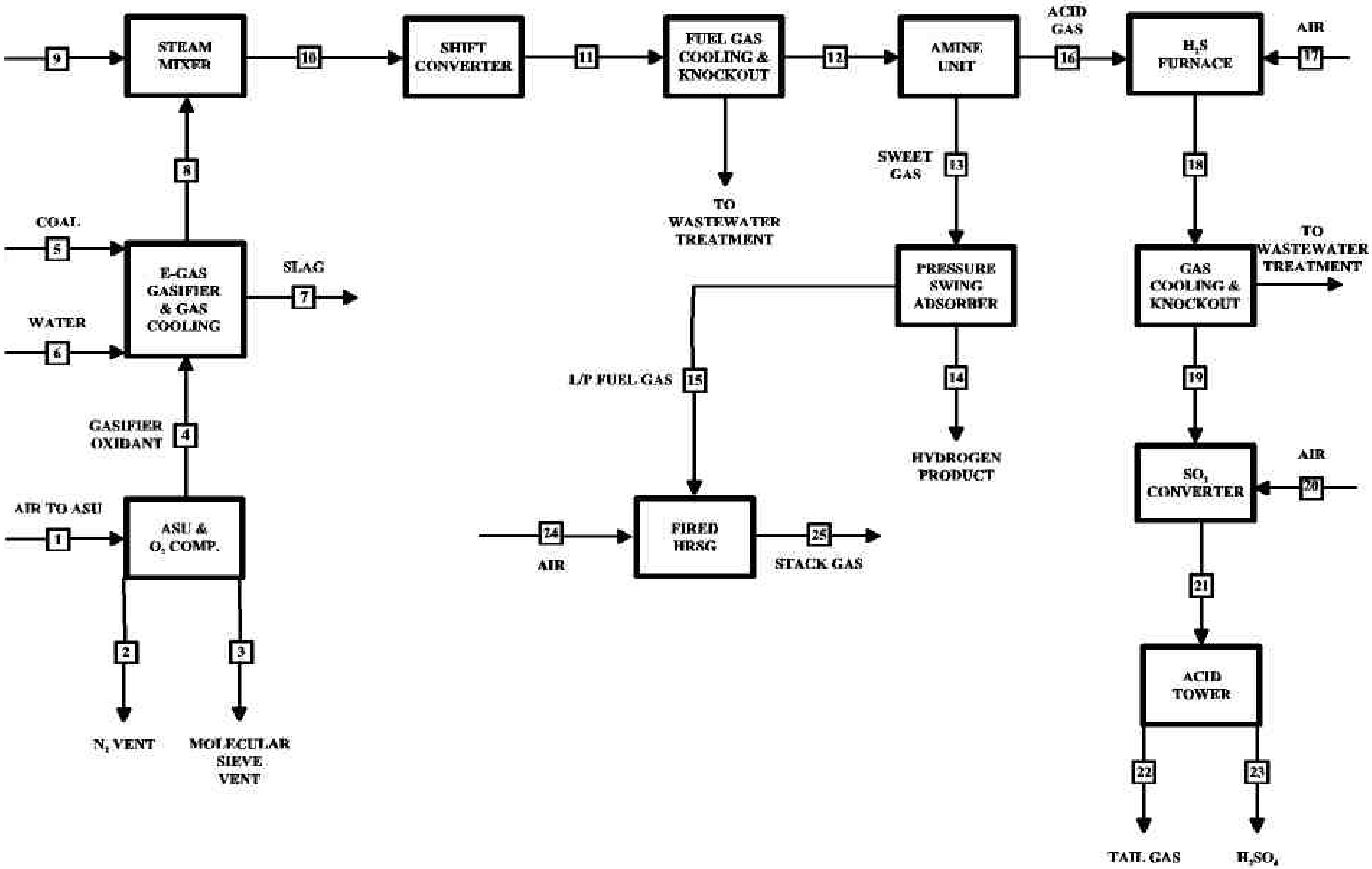


Table 2-2
Process Stream Compositions and State Points – Pittsburgh No. 8 Coal

STREAM NUMBER	1	2	3	4	5	6	7	8	9	10	11	12	13
Vapor - Liquid													
Mole Fraction													
Ar	0.0094	0.0027	0.0000	0.0360	0.0000	0.0000	0.0000	0.0082	0.0000	0.0048	0.0048	0.0069	0.0077
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0042	0.0000	0.0024	0.0024	0.0035	0.0039
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.4195	0.0000	0.2443	0.0627	0.0905	0.1011
CO_2	0.0003	0.0000	0.0266	0.0000	0.0000	0.0000	0.0000	0.0975	0.0000	0.0568	0.2387	0.3435	0.2735
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3320	0.0000	0.1933	0.3750	0.5414	0.6053
H_2O	0.0108	0.0000	0.9734	0.0000	0.0000	1.0000	0.0000	0.1219	1.0000	0.4887	0.3068	0.0027	0.0031
H_2SO_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.7719	0.9973	0.0000	0.0140	0.0000	0.0000	0.0000	0.0057	0.0000	0.0033	0.0033	0.0048	0.0054
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0028	0.0000	0.0016	0.0016	0.0000	0.0000
O_2	0.2076	0.0000	0.0000	0.9500	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
COS (ppmv)	0	0	0	0	0	0	0	388	0	226	0	0	0
H_2S (ppmv)	0	0	0	0	0	0	0	7,772	0	4,526	4,752	6,738	15
SO_2 (ppmv)	0	0	0	0	0	0	0	0	0	0	0	0	0
SO_3 (ppmv)	0	0	0	0	0	0	0	0	0	0	0	0	0
Total V-L Flow (lb_{mol}/hr)	25,425	19,588	282	5,550	0	5,219	0	25,458	18,257	43,716	43,716	29,067	25,990
Total V-L Flow (lb/hr)	733,505	549,364	5,281	178,860	0	94,025	0	504,010	328,910	832,920	832,920	566,376	433,172
Solids Flow													
Coal	0	0	0	0	221,631	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0	0	0
Slag	0	0	0	0	0	0	22,632	0	0	0	0	0	0
Temperature (°F)	63	60	100	316		60		500	500	500	857	105	123
Pressure (psia)	14.4	14.7	87.5	620.0	14.7	14.7	15.0	391.0	470.0	391.0	381.0	353.0	338.0

Table 2-2 (Cont'd)
Process Stream Compositions and State Points – Pittsburgh No. 8 Coal

STREAM NUMBER	14	15	16	17	18	19	20	21	22	23	24	25
Vapor - Liquid												
Mole Fraction												
Ar	0.0000	0.0148	0.0000	0.0094	0.0032	0.0033	0.0094	0.0046	0.0048	0.0000	0.0094	0.0129
CH_4	0.0000	0.0076	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.1953	0.0003	0.0000	0.0002	0.0002	0.0000	0.0002	0.0002	0.0000	0.0000	0.0000
CO_2	0.0009	0.5274	0.9340	0.0003	0.6412	0.6740	0.0003	0.5509	0.5736	0.0000	0.0003	0.3431
H_2	0.9980	0.2396	0.0020	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0000	0.0059	0.0000	0.0108	0.0486	0.0000	0.0108	0.0022	0.0000	0.0103	0.0108	0.1292
H_2SO_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.9897	0.0000	0.0000
N_2	0.0011	0.0093	0.0000	0.7719	0.2594	0.2727	0.7719	0.3783	0.3939	0.0000	0.7719	0.4928
NH_3	0.0000	0.0000	0.0001	0.0000	0.0001	0.0001	0.0000	0.0001	0.0001	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.2076	0.0037	0.0038	0.2076	0.0263	0.0274	0.0000	0.2076	0.0219
COS (ppmv)	0	0	0	0	0	0	0	0	0	0	0	0
H_2S (ppmv)	0	29	63,511	0	0	0	0	0	0	0	0	0
SO_2 (ppmv)	0	0	0	0	43,597	45,826	0	112	117	0	0	14
SO_3 (ppmv)	0	0	0	0	0	0	0	37,340	0	0	0	0
Total V-L Flow (lb_{mol}/hr)	12,531	13,459	3,077	1,507	4,483	4,265	1,051	5,219	5,012	196	18,143	28,672
Total V-L Flow (lb/hr)	26,071	407,102	133,203	43,463	176,667	172,739	30,319	203,057	187,251	19,200	523,404	930,451
Solids Flow												
Coal	0	0	0	0	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0	0
Slag	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	70	70	123	331	750	650	180	300	315	315	89	280
Pressure (psia)	310.0	17.0	338.0	45.0	42.0	29.5	25.0	20.0	19.0	19.0	17.0	14.7

2.2 PROCESS DESCRIPTION

Following are more detailed descriptions of the key process elements:

2.2.1 Gasifier

For this application to produce hydrogen, a dual-train E-Gas™ gasifier of the Wabash River configuration consisting of two 50 percent gasifiers 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 1900°F goes through an internal cyclone that separates entrained particles from the gas for recycle to the gasifier, followed by a fire-tube boiler and a gas cooler to reduce gas temperature to 645°F (341°C). Following the cooler, remaining particulates are removed from the gas with a metallic filter and are returned to the gasifier.

2.2.2 Air Separation Unit

Oxygen supply for this plant is also provided through a conventional cryogenic ASU. The air separation plant is designed to produce a nominal output of 2,150 tons/day of 99 percent pure O_2. The high-pressure plant is designed with one production train, with liquefaction and liquid oxygen storage providing an 8-hour backup supply of oxygen.

2.2.3 Particulate Removal

The particulate removal device is a sintered metal candle configuration, operating at the relatively low temperature of 645°F (341°C). The vessel and candle array is similar to the configuration used at the Wabash River demonstration plant. A particulate removal vessel is provided for each gasifier train.

2.2.4 CO Shift

After leaving the particulate control unit, steam is injected into the gas stream and the CO in the syngas is shifted to hydrogen and CO_2 in the shift converter utilizing a sulfur-tolerant shift catalyst. The shift catalyst also promotes the COS hydrolysis reaction. Heat is removed from the gas stream following the shift, the gases are cooled, sour water is condensed, and the gas stream is sent to the sulfur removal unit.

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. The two reactors in parallel are required due to the high gas mass flow rate. Feed to the shift converter is first preheated by hot effluent from the second converter, heated by hot effluent from the first converter, and fed to the top of the two first-stage converters in parallel. Effluent from the first stage is cooled and fed to the top of the second-stage converters. Effluent from the second stage is cooled by exchanging heat with incoming feed, by an air cooler and finally by a water cooler.

2.2.5 Amine Unit/Acid Gas Concentrator

The purpose of the amine unit is to remove H_2S from the fuel gas stream. This step is necessary in order to minimize plant sulfur emissions. The solvent used in this case is a proprietary formulation based on MDEA. A traditional absorber/stripper arrangement will be used.

Cool, dry, and particulate-free synthesis gas enters the absorber unit at 353 psia and 105°F (41°C). In the absorber, H_2S along with some CO_2 is removed from the fuel gas stream. Clean fuel gas exits the top of the absorber and is then routed to the saturator column.

The rich solution leaving the bottom of the absorber is regenerated in a stripper through the indirect application of thermal energy via condensing low-pressure steam in a reboiler. The stripper acid gas stream, consisting of 16 percent H_2S and 78 percent CO_2 (with the balance mostly H_2O), requires further treatment before being sent to the sulfuric acid plant.

Typically, for good performance and operation, the minimum H_2S concentration in the acid gas feed to an acid plant should be above 27 percent. In this case, an acid gas concentrator was used to further concentrate the H_2S stream.

2.2.6 Hydrogen Purification

The product hydrogen stream exits the absorber 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 HRSG to be fired with oxygen.

Treated gas from the amine unit absorber is fed directly to the PSA unit where hydrogen is purified up to approximately 99.9 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 adsorbent beds at 338 psia, and then the impurities are purged from the beds at 17 psia. Purge gas is sent to the gas-fired heat recovery unit for steam generation. Purified hydrogen is produced at 310 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.

2.3 MAJOR EQUIPMENT LIST FOR BITUMINOUS COAL CASE

This section contains the equipment list corresponding to the power plant configuration shown in Figure 2-1. This list, along with the heat and material balance and supporting performance data, was used to generate plant costs and used in the financial analysis. In the following, all feet (ft) conditions specified for process pumps correspond to feet of liquid being pumped.

ACCOUNT 1 COAL RECEIVING AND HANDLING

Equipment No.	Description	Type	Design Condition	Qty.
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	100 ton	2
2	Feeder	Vibratory	150 tph	2
3	Conveyor 1	54" belt	200 tph	1
4	As-Received Coal Sampling System	Two-stage	N/A	1
5	Conveyor 2	54" belt	200 tph	1
6	Reclaim Hopper	N/A	40 ton	2
7	Feeder	Vibratory	150 tph	2
8	Conveyor 3	48" belt	200 tph	1
9	Crusher Tower	N/A	200 tph	1
10	Coal Surge Bin w/Vent Filter	Compartment	200 ton	1
11	Crusher	Granulator reduction	6"x0 - 3"x0	1
12	Crusher	Impactor reduction	3"x0 - 1"x0	1
13	As-Fired Coal Sampling System	Swing hammer	N/A	2
14	Conveyor 4	48" belt	200 tph	1
15	Transfer Tower	N/A	200 tph	1
16	Tripper	N/A	200 tph	1
17	Coal Silo w/Vent Filter and Slide Gates	N/A	400 ton	2

ACCOUNT 2 COAL-WATER SLURRY PREPARATION AND FEED

Equipment No.	Description	Type	Design Condition	Qty.
1	Feeder	Vibrating	80 tph	2
2	Weigh Belt Feeder		48" belt	2
3	Rod Mill	Rotary	80 tph	2
4	Slurry Water Pumps	Centrifugal	180 gpm @ 500 ft	2
5	Slurry Water Storage Tank	Vertical	1,500 gal	1
6	Rod Mill Product Tank	Vertical	35,000 gal	2
7	Slurry Storage Tank with Agitator	Vertical	150,000 gal	2
8	Coal-Slurry Feed Pumps	Positive displacement	700 gpm @ 1,250 ft	2
9	LT Slurry Heater	Shell and tube	20 x 10^6 Btu/h	2
10	HT Slurry Heater	Shell and tube	7 x 10^6 Btu/h	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS BOP SYSTEMS

ACCOUNT 3A CONDENSATE AND FEEDWATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cond. Storage Tank	Vertical, cylindrical, outdoor	200,000 gal	1
2	Condensate Pumps	Vert. canned	1,000 gpm @ 400 ft	2
3	Deaerator	Horiz. spray type	1,100,000 lb/h 205°F to 240°F	1
4	IP Feed Pump	Horiz. centrifugal single stage	400 gpm/1,850 ft	2
5	HP Feed Pump	Barrel type, multi-staged, centr.	500 gpm @ 4,000 ft	2

ACCOUNT 3B MISCELLANEOUS EQUIPMENT

Equipment No.	Description	Type	Design Condition	Qty
1	Auxiliary Boiler	Shop fab., water tube	400 psig, 650°F 70,000 lb/h	1
2	Service Air Compressors	Recip., single stage, double acting, horiz.	100 psig, 750 cfm	2
3	Inst. Air Dryers	Duplex, regenerative	750 cfm	1
4	Service Water Pumps	Horiz. centrifugal, double suction	200 ft, 1,200 gpm	2
5	Closed Cycle Cooling Water Pumps	Horizontal, centrifugal	70 ft, 1,200 gpm	2
6	Fire Service Booster Pump	Two-stage horiz. centrifugal	250 ft, 1,200 gpm	1
7	Engine-Driven Fire Pump	Vert. turbine, diesel engine	350 ft, 1,000 gpm	1
8	Raw Water Pumps	SS, single suction	60 ft, 300 gpm	2
9	Filtered Water Pumps	SS, single suction	160 ft, 120 gpm	2
10	Filtered Water Tank	Vertical, cylindrical	15,000 gal	1
11	Makeup Demineralizer	Anion, cation, and mixed bed	650 gpm	2
12	Sour Water Stripper System	Vendor supplied	300,000 lb/h sour water	1
13	Liquid Waste Treatment System	Vendor supplied	600 gpm	1

ACCOUNT 4 GASIFIER AND ACCESSORIES

ACCOUNT 4A GASIFICATION

Equipment No.	Description	Type	Design Condition	Qty
1	Gasifier	Pressurized entrained bed/syngas cooler	2,500 std (dry-coal basis) @ 425 psia	2
2	Raw Gas Cooler Steam Generator	Fire tube boiler	1,500 psig/600°F 132.7 MMBtu/h	2
3	Raw Gas Cooler Steam Generator	Shell and tube	800 psig/518°F 135.1 MMBtu/h	2
4	Medium-Temperature Candle Filter	Sintered stainless	400 psia, 645ºF	2
5	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	600,000 lb/h, medium-Btu gas	1

ACCOUNT 4B AIR SEPARATION PLANT

Equipment No.	Description	Type	Design Condition	Qty
1	Air Compressor	Centrifugal, multi-stage	75,000 scfm, 67 psia discharge pressure	2
2	Cold Box	Vendor supplied	2,150 ton/day O_2	1
3	Oxygen Compressor	Centrifugal, multi-stage	17,000 scfm, 600 psig discharge pressure	2
4	Liquid Oxygen Storage Tank	Vertical	60' dia x 80' vert	1

ACCOUNT 5 SYNGAS SHIFT AND CLEANUP

ACCOUNT 5A WATER-GAS SHIFT AND RAW GAS COOLING

Equipment No.	Description	Type	Design Condition	Qty
1	High Temperature Shift Reactor	Fixed bed	400 psia, 750ºF	1
2	HP Steam Generator	Shell and tube	70 x 10^6 Btu/h @ 1700 psia and 613ºF	1
3	IP Steam Generator	Shell and tube	45 x 10^6 Btu/h @ 800 psia and 518ºF	1
45	LP Steam Generator	Shell and tube	20 x 10^6 Btu/h @ 200 psia and 382ºF	1
5	Raw Gas Coolers	Shell and tube with condensate drain	150 x 10^6 Btu/h	2
67	Raw Gas Knock Out Drum	Vertical with mist eliminator	400 psia, 130ºF	1

ACCOUNT 5B ACID GAS REMOVAL AND HYDROGEN PURIFICATION

Equipment No.	Description	Type	Design Condition	Qty
1	Amine Absorber 1	Column	100,000 scfm (4,900 acfm), 353 psia, 103ºF	2
2	Amine Regenerator 1	Column	Vendor design	1
3	Sulfuric Acid Plant	Vendor design	230 tpd sulfuric acid	1
4	PSA Unit	Fixed bed	112 MMscfd H_2 @ 310 psia	1

ACCOUNT 6 COMBUSTION TURBINE/ACCESSORIES

Not applicable.

ACCOUNT 7 WASTE HEAT BOILER, DUCTING, AND STACK

Equipment No.	Description	Type	Design Condition Drums	Qty
1	Heat Recovery Steam Generator	Fired drum	1700 psig/1000°F 600,000 lb/h 200 x 10^6 Btu/h	1
2	Stack	Carbon steel plate, type 409 stainless steel liner	213 ft high x 28 ft dia.	1

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition (per each)	Qty
1	78 MW Steam Turbine Generator	TC2F26	1800 psig 1000°F/1000°F	1
2	Bearing Lube Oil Coolers	Plate and frame		2
3	Bearing Lube Oil Conditioner	Pressure filter closed loop		1
4	Control System	Digital electro-hydraulic	1600 psig	1
5	Generator Coolers	Plate and frame		2
6	Hydrogen Seal Oil System	Closed loop		1
7	Surface Condenser	Single pass, divided waterbox	1,030,000 lb/h steam @ 2.4 in. Hga	1
8	Condenser Vacuum Pumps	Rotary, water sealed	2500/25 scfm (hogging/holding)	2

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition (per each)	Qty
1	Circ. Water Pumps	Vert. wet pit	40,000 gpm @ 60 ft	2
2	Cooling Tower	Mechanical draft	100,000 gpm	1

ACCOUNT 10 ASH/SPENT SORBENT RECOVERY AND HANDLING

ACCOUNT 10A SLAG DEWATERING AND REMOVAL

Equipment No.	Description	Type	Design Condition	Qty
1	Slag Dewatering System	Vendor proprietary	272 tpd	1

2.4 CAPITAL COST

The total plant cost for the plant producing 313 tons of hydrogen per day from Pittsburgh No. 8 coal is $376.1 million in 2001 dollars. The capital cost summary is included in Table 2-3.

Table 2-3
Capital Cost Summary – Hydrogen Production from Pittsburgh No. 8 Coal

Client: DEPARTMENT OF ENERGY
Project: NETL H_2 Production Facility
Report Date: 03-Jun-2002 01:19 PM

TOTAL PLANT COST SUMMARY

Case: Bituminous H_2 Plant w/o CO_2 Capture
Plant Size: 312.6 H_2 TPD
Estimate Type: Conceptual
Cost Base (Dec) 2001 ($X1000 & $X1000/TPD)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $	TOTAL PLANT COST $/TPD
1	COAL & SORBENT HANDLING	5,354	1,099	3,955	277		**$10,686**	855		1,154	**$12,694**	41
2	COAL & SORBENT PREP & FEED	8,987	1,307	4,905	343		**$15,542**	1,733		1,728	**$19,003**	61
3	FEEDWATER & MISC. BOP SYSTEMS	4,506	1,339	2,679	187		**$8,711**	697		941	**$10,348**	33
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries (E	51,616		21,976	1,538		**$75,130**	9,016		8,415	**$92,560**	296
4.2	Syngas Cooling	w/4.1		w/ 4.1	w/ 4.1			w/ 4.1		w/ 4.1		
4.3	ASU/Oxidant Compression	29,284		w/equip.			**$29,284**	2,343		3,163	**$34,789**	111
4.4-4.9	Other Gasification Equipment	6,881	5,691	4,811	337		**$17,720**	1,447		1,917	**$21,084**	67
	SUBTOTAL 4	*87,781*	*5,691*	*26,787*	*1,875*		*$122,134*	*12,805*		*13,494*	*$148,433*	*475*
5	HYDROGEN SEPARATION/GAS CLEAN	59,135	4,205	21,176	1,482		**$85,999**	10,111		9,611	**$105,721**	338
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Expander Turbine/Generator											
6.2-6.9	Combustion Turbine Accessories											
	SUBTOTAL 6											
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	4,533		551	39		**$5,123**	410		553	**$6,086**	19
7.2-7.9	HRSG Accessories, Ductwork and Stack	561	209	335	23		**$1,129**	90		122	**$1,341**	4
	SUBTOTAL 7	*5,094*	*209*	*886*	*62*		*$6,251*	*500*		*675*	*$7,427*	*24*
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	7,612		1,061	74		**$8,747**	700		945	**$10,392**	33
8.2-8.9	Turbine Plant Auxiliaries and Steam Pip	3,470	106	1,611	113		**$5,300**	424		572	**$6,296**	20
	SUBTOTAL 8	*11,082*	*106*	*2,672*	*187*		*$14,047*	*1,124*		*1,517*	*$16,688*	*53*
9	COOLING WATER SYSTEM	2,375	1,163	1,901	133		**$5,572**	446		602	**$6,619**	21
10	ASH/SPENT SORBENT HANDLING SYS	5,147	653	2,196	154		**$8,150**	888		904	**$9,941**	32
11	ACCESSORY ELECTRIC PLANT	5,029	2,077	4,075	285		**$11,467**	917		1,238	**$13,623**	44
12	INSTRUMENTATION & CONTROL	5,292	1,257	3,903	273		**$10,725**	858		1,158	**$12,741**	41
13	IMPROVEMENTS TO SITE	1,566	900	2,593	182		**$5,241**	419		566	**$6,226**	20
14	BUILDINGS & STRUCTURES		2,663	2,711	190		**$5,564**	445		601	**$6,610**	21
	TOTAL COST	**$201,347**	**$22,671**	**$80,439**	**$5,631**		**$310,088**	**$31,798**		**$34,189**	**$376,074**	**1203**

2.5 DETERMINATION OF COST OF HYDROGEN FROM PITTSBURGH NO. 8 COAL USING PRELIMINARY ASSUMPTIONS

For this economic analysis, the capital and operating costs for the two plants being evaluated have been upgraded to 2001 dollars. Coal cost has been estimated at $1.00 per MMBtu.

2.5.1 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 2-4, 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.

2.5.2 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.

Table 2-4
Operating Cost Data

OPERATING LABOR REQUIREMENTS		
BITUMINOUS H_2 PLANT W/O CO_2 CAPTURE		
Operating Labor Rate (base)	$26.15/hour	
Operating Labor Burden	30.00% of base	
Labor OH Charge Rate	25.00% of labor	
Operating Labor Requirements (OJ) per Shift		
Category	**1 Unit/Mod**	**Total Plant**
Skilled Operator	1.0	1.0
Operator	8.0	8.0
Foreman	1.0	1.0
Lab Techs, etc.	1.0	1.0
TOTAL – OJs	11.0	11.0

CONSUMABLES, BYPRODUCTS & FUELS DATA			
BITUMINOUS H_2 PLANT W/O CO_2 CAPTURE			
	Consumption		**Unit**
Item/Description	**Initial**	**/Day**	**Cost**
Water (/1,000 gal)		846	0.80
Chemicals			
Water Treatment (lb)	61,434	2,048	0.15
Limestone (ton)			16.25
Shift Catalyst (lb)	15,221	507.4	5.25
Amine (lb)	8,640	288.0	0.63
Other			
Supplemental Fuel (MMBtu)			
Purchased Power (MWh)			
LP Steam (/1,000 lb)			
Waste Disposal			
Sludge (ton)			
Slag (ton)		272	10.00
Byproducts & Emissions			
Sulfuric Acid (tons)		229	75.00
Excess Electric Generation (MWh)		901	30.00
Fuel (ton)		2,659	24.90

2.5.3 Consumables

Shift Catalyst:

- Change-out every 3 years
- 0.0045 pound of catalyst per 1,000 standard cubic feet of hydrogen
- 250 tons initial charge
- 85 tons per year annual cost

Proprietary Amine:

- 12 pounds per hour

- 100,000 pounds per year

PSA Sorbent:

- Periodic change-out with scheduled maintenance

SO_2 Conversion Catalyst:

- Periodic change-out with scheduled maintenance

2.5.4 Byproduct Credits

The production of 229 tons of sulfuric acid per day is taken as a byproduct credit at $75 per ton.

2.5.5 Financial Assumptions

The cost of hydrogen was determined based on financial assumptions typically used by Parsons. These are summarized in Table 2-5.

Table 2-5
Financial Parameters

Levelized capacity factor	90%
Design/construction period	4 years
Plant startup date	January 2005
Land area/Unit cost	100 acres @ 41,500/acre
Project book life	20 years
Project tax life	20 years
Tax depreciation method	Accelerated based on ACRS class
Property tax rate	1.0% per year
Insurance tax rate	1.0% per year
Federal income tax rate	34.0%
State income tax rate	4.2%
Capital structure	
Common equity	20% @ 16.50% annum
Debt	80% @ 6.30% annum
Weighted cost of capital (after tax)	6.49%

2.5.6 Cost Results

Applying the financial parameters from Table 2-5, the cost of hydrogen was estimated to be $6.01/MMBtu ($2.06/Mcf) for Pittsburgh No. 8 coal. The results of the cost estimating activity are summarized in Table 2-6. These results were used as the price of hydrogen input to the DOE IGCC financial model. The results are shown in Appendix A.

Table 2-6

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY				
TITLE/DEFINITION				
Case:	Bituminous 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:	3 (years)	BookLife:	20	(years)
TPC(Plant Cost) Year:	2001 (Jan.)	TPI Year:	2005	(Jan.)
Capacity Factor:	90 (%)			

CAPITAL INVESTMENT		**$x1000**	**$x1000/$H_2$TPD**
Process Capital & Facilities		310,088	992.0
Engineering(incl.C.M.,H.O.& Fee)		31,798	101.7
Process Contingency			
Project Contingency		34,189	109.4
TOTAL PLANT COST(TPC)		$376,074	1203.1
TOTAL CASH EXPENDED	$376,074		
AFDC	$24,114		
TOTAL PLANT INVESTMENT(TPI)		$400,188	1280.2
Royalty Allowance			
Preproduction Costs		9,621	30.8
Inventory Capital		3,067	9.8
Initial Catalyst & Chemicals(w/equip.)			
Land Cost		150	0.5
TOTAL CAPITAL REQUIREMENT(TCR)		$413,026	1321.3

OPERATING & MAINTENANCE COSTS (2001 Dollars)	**$x1000**	**$x1000/$H_2$TPD**
Operating Labor	3,276	10.5
Maintenance Labor	2,642	8.5
Maintenance Material	3,963	12.7
Administrative & Support Labor	1,479	4.7
TOTAL OPERATION & MAINTENANCE	$11,361	36.3
FIXED O & M		32.71
VARIABLE O & M		3.63
CONSUMABLE OPERATING COSTS,less Fuel (2001 Dollars)	**$x1000**	**$/T H_2-yr**
Water	222	2.17
Chemicals	1,033	10.06
Other Consumables		
Waste Disposal	892	8.69
TOTAL CONSUMABLE OPERATING COSTS	$2,147	20.91
BY-PRODUCT CREDITS (2001 Dollars)	($14,531)	-141.51
FUEL COST (2001 Dollars)	$21,753	211.85

	1st Year (2005 $)	Levelized (Over Book Life $)
PRODUCTION COST SUMMARY	**$/T H_2-yr**	**$/T H_2-yr**
Fixed O & M	99.57	99.57
Variable O & M	3.63	3.63
Consumables	20.91	20.91
By-product Credit/Penalty	-141.51	-141.51
Fuel	206.81	188.64
TOTAL PRODUCTION COST	189.41	171.25
	19,450	
LEVELIZED CARRYING CHARGES(Capital)		563.12
	57,824	
LEVELIZED(Over Book Life)COST/Ton of Syngas		734.36
Equivalent 1st.Yr.$/MSCF / Lev'd $/MMBtu	2.061	6.01

3. WYODAK PRB COAL

This section is dedicated to the design and cost estimate for a hydrogen plant fed with Wyodak Powder River Basin (PRB) sub-bituminous coal. This coal is characterized having low volatility, high ash and moisture content, and a lower as-received heating value. The low sulfur content results in a lesser value-added from the sulfuric acid byproduct.

3.1 HEAT AND MATERIAL BALANCE

The heat and material balance for the IGCC plant is based on the maximum hydrogen production from 2,500 tons per day of dry coal. Ambient operating conditions are indicated in the plant design basis. The pressurized entrained flow E-Gas™ two-stage gasifier uses a coal/water slurry and oxygen to produce a medium heating value fuel gas. The syngas produced in the gasifier first stage at about 2450°F (1343°C) is quenched to 1900°F (1038°C) by reacting with slurry injected into the second stage. The syngas passes through a fire tube boiler syngas cooler and leaves at 1300°F (704°C). A second gas cooler in series cools the gas further to 645°F (341°C). High-pressure saturated steam is generated in the syngas coolers and is joined with the main steam supply. The process flow diagram resulting from the heat and material balance is shown as Figure 3-1.

The gas goes through a series of additional gas coolers and cleanup processes including a scrubber. Slag captured by the syngas scrubber is recovered in a slag recovery unit.

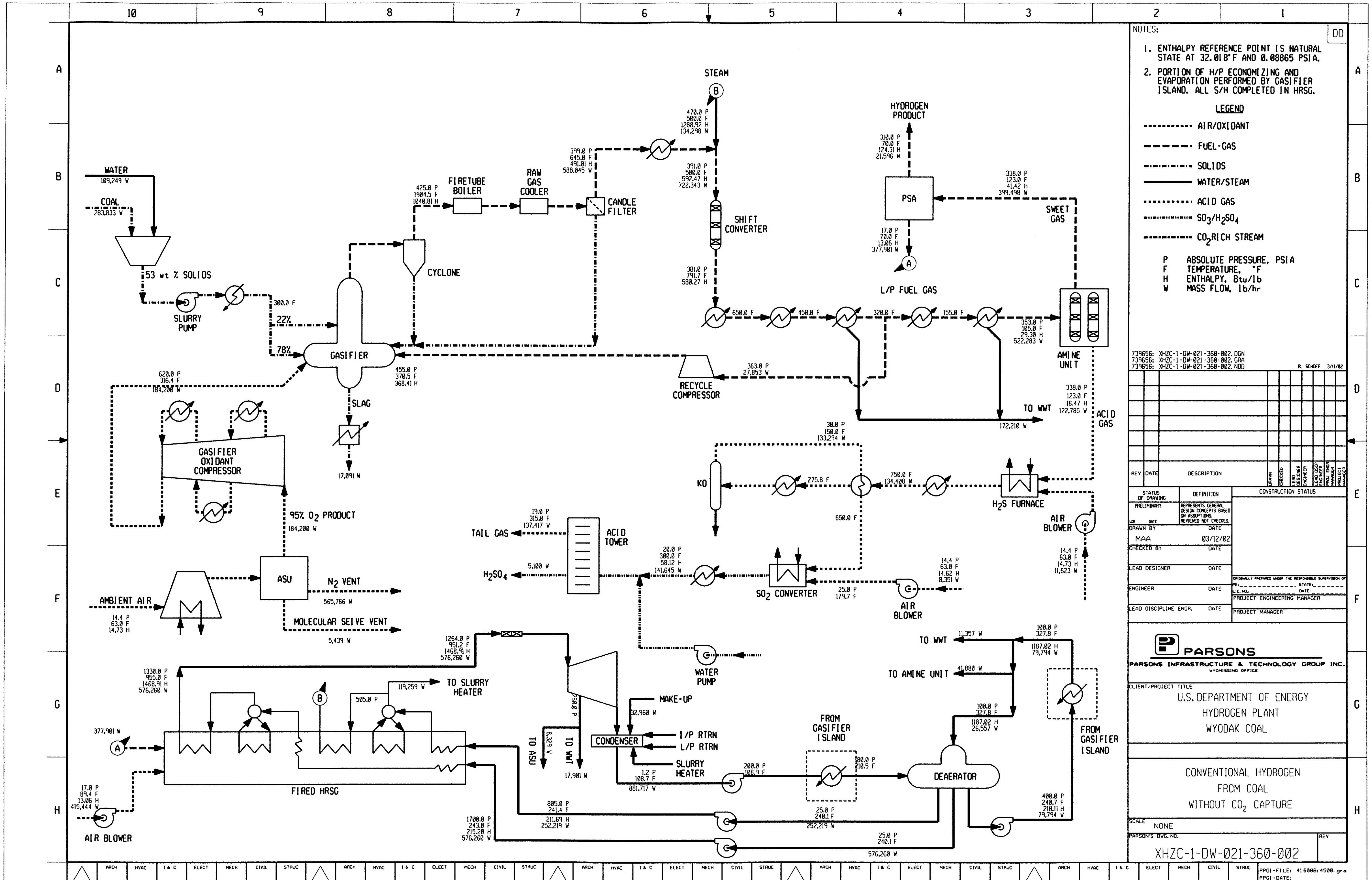
The syngas stream from the syngas scrubber enters the high-temperature shift converter, which contains a bed of sulfided shift catalyst. The shift reaction converts over 80 percent of the CO to hydrogen and CO_2 and hydrolyzes COS to H_2S. Following the shift converter, the cooled gas stream passes through a proprietary amine acid gas removal (AGR) process, which removes H_2S and some of the CO_2. The clean gas stream then passes through the pressure swing adsorption (PSA) for final purification of the hydrogen. Regeneration gas from the PSA contains fuel value, and is fed to the heat recovery steam generator (HRSG). Regeneration gas from the AGR plant is fed to a sulfuric acid plant.

The cryogenic oxygen plant supplies 99 percent pure oxygen to the gasifiers at the rated pressure. A dedicated air compressor provides air supply for the oxygen plants.

The steam cycle is based on maximizing heat recovery from the gasifier cooler and HRSG, as well as utilizing steam generation opportunities in the shift process.

The steam turbine selected to match this cycle is a two-casing, reheat, double-flow (exhaust) machine, exhausting downward to the condenser. The HP and IP turbine sections are contained in one casing, with the LP section in a second casing. The steam turbine drives a 3600 rpm hydrogen-cooled generator. The turbine exhausts to a single-pressure condenser operating at a nominal 1.2 psia at the 100 percent load design point. Two 50 percent capacity, motor-driven pumps are provided for feedwater and condensate.

Figure 3-1


NOTES:
1. ENTHALPY REFERENCE POINT IS NATURAL STATE AT 32.018°F AND 0.08865 PSIA.
2. PORTION OF H/P ECONOMIZING AND EVAPORATION PERFORMED BY GASIFIER ISLAND. ALL S/H COMPLETED IN HRSG.
LEGEND
AIR/OXIDANT
FUEL-GAS
SOLIDS
WATER/STEAM
ACID GAS
SO3/H2SO4
CO2 RICH STREAM
P ABSOLUTE PRESSURE, PSIA
F TEMPERATURE, °F
H ENTHALPY, Btu/lb
W MASS FLOW, lb/hr
WATER
COAL
SLURRY PUMP
GASIFIER
CYCLONE
FIRETUBE BOILER
RAW GAS COOLER
CANDLE FILTER
STEAM
SHIFT CONVERTER
HYDROGEN PRODUCT
PSA
SWEET GAS
L/P FUEL GAS
AMINE UNIT
RECYCLE COMPRESSOR
TO WWT
ACID GAS
SLAG
GASIFIER OXIDANT COMPRESSOR
95% O2 PRODUCT
ASU
AMBIENT AIR
N2 VENT
MOLECULAR SEIVE VENT
TAIL GAS
ACID TOWER
H2SO4
KO
H2S FURNACE
AIR BLOWER
SO2 CONVERTER
WATER PUMP
TO SLURRY HEATER
MAKE-UP
CONDENSER
I/P RTRN
L/P RTRN
SLURRY HEATER
TO ASU
TO WWT
FIRED HRSG
TO AMINE UNIT
FROM GASIFIER ISLAND
DEAERATOR
PARSONS
PARSONS INFRASTRUCTURE & TECHNOLOGY GROUP INC.
U.S. DEPARTMENT OF ENERGY
HYDROGEN PLANT
WYODAK COAL
CONVENTIONAL HYDROGEN FROM COAL WITHOUT CO2 CAPTURE
SCALE NONE
XHZC-1-DW-021-360-002
DRAWN BY MAA 03/12/02

Overall performance for the entire plant is summarized in Table 3-1, which includes auxiliary power requirements. The net plant output power, after plant auxiliary power requirements are deducted, is nominally 42 MW_e. The overall plant thermal effective efficiency (thermal value of hydrogen and power produced) is 59.7 percent, on an HHV basis.

Table 3-1
Plant Performance Summary
Hydrogen Production from Wyodak Coal

Plant Output		
Steam Turbine Power	81,450	kW_e
Total	**81,450**	kW_e
Hydrogen Production		
Hydrogen Product	21,600	lb/h
Auxiliary Load		
Gasifier O_2 Compressor	9,590	kW_e
ASU Air Compressor	21,470	kW_e
Gasifier Slurry Pump	80	kW_e
Coal Handling	270	kW_e
Slag Handling	400	kW_e
Amine Unit	300	kW_e
Recycle Compressor	180	kW_e
H_2SO_4 Plant	100	kW_e
H_2S Furnace Air Blower	230	kW_e
Boiler Feedwater Pumps	1,420	kW_e
Steam Turbine Auxiliaries	250	kW_e
Cooling Tower	1,140	kW_e
Circulating Water Pumps	1,830	kW_e
Miscellaneous Balance of Plant	750	kW_e
Condensate Pumps	200	kW_e
Flue Gas Burner Air Blower	910	kW_e
Wastewater Treatment	500	kW_e
Total	39,620	kW_e
Plant Performance		
Net Auxiliary Load	39,620	kW_e
Net Plant Power	41,830	kW_e
Net Plant Efficiency (HHV)[1]	59.7%	
Net Plant Heat Rate (HHV)	5,820	Btu/kWh
Coal Feed Flowrate	283,833	lb/h
Thermal Input[2]	717,872	kW_e
Condenser Duty	593.7	MMBtu/h

1 – Efficiency calculation includes thermal value of hydrogen and power produced.
2 – HHV of as-fed Wyodak coal is 8,630 Btu/lb.

Figure 3-2 is a block flow diagram for the plant, and is accompanied by Table 3-2, which includes detailed process stream composition and state points.

Figure 3-2
Process Block Flow Diagram
E-Gas™ Gasifier-Based Hydrogen Production Plant – Wyodal Coal

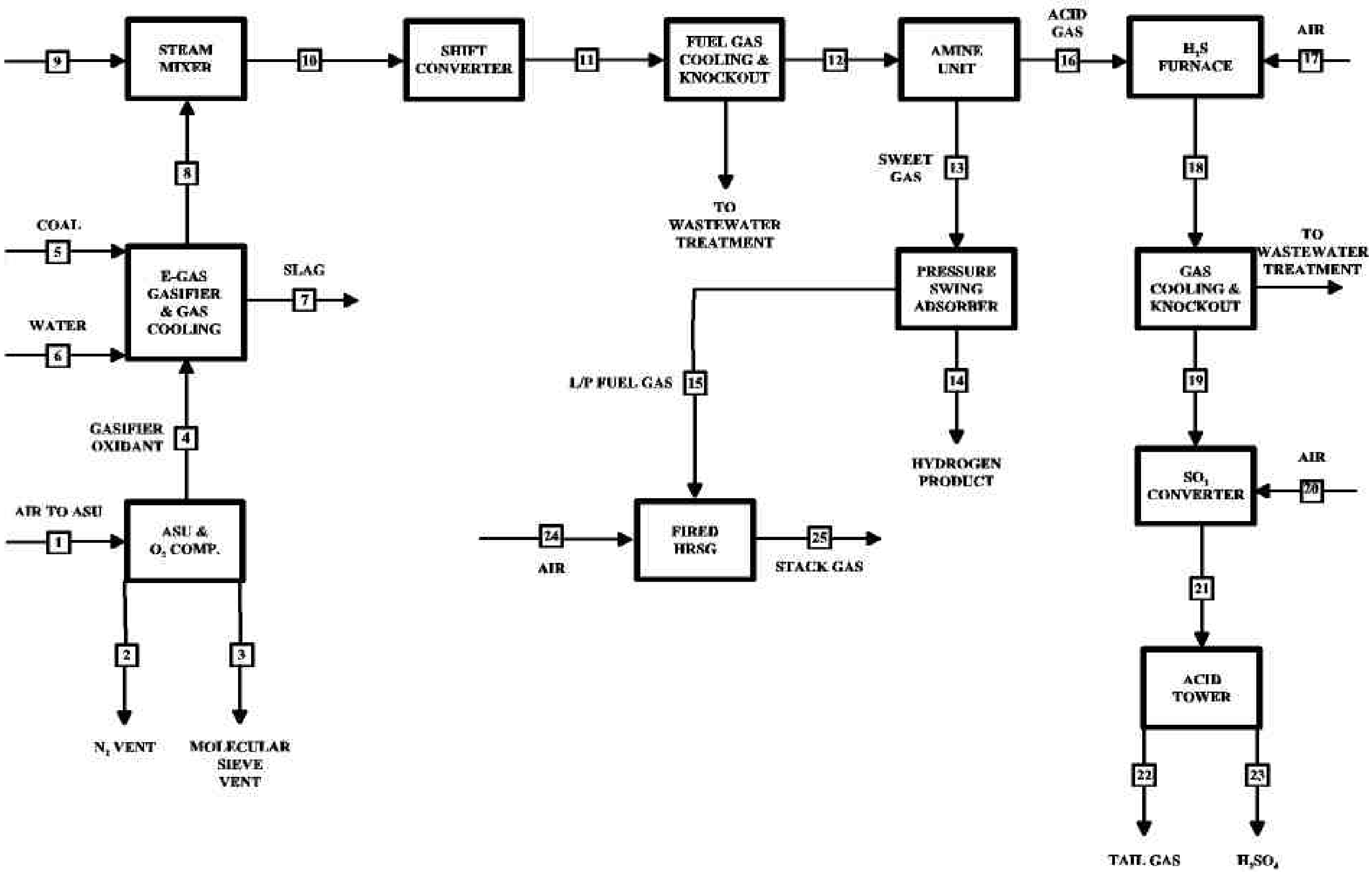


Table 3-2
Process Stream Compositions and State Points – Wyodak Coal

STREAM NUMBER	1	2	3	4	5	6	7	8	9	10	11	12	13
Vapor - Liquid													
Mole Fraction													
Ar	0.0094	0.0027	0.0000	0.0360	0.0000	0.0000	0.0000	0.0075	0.0000	0.0059	0.0059	0.0082	0.0092
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0017	0.0000	0.0014	0.0014	0.0019	0.0021
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2661	0.0000	0.2111	0.0619	0.0852	0.0958
CO_2	0.0003	0.0000	0.0266	0.0000	0.0000	0.0000	0.0000	0.1604	0.0000	0.1272	0.2765	0.3795	0.3042
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2834	0.0000	0.2248	0.3740	0.5148	0.5792
H_2O	0.0108	0.0000	0.9734	0.0000	0.0000	1.0000	0.0000	0.2717	1.0000	0.4222	0.2730	0.0027	0.0031
H_2SO_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.7719	0.9973	0.0000	0.0140	0.0000	0.0000	0.0000	0.0052	0.0000	0.0041	0.0041	0.0057	0.0064
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0021	0.0000	0.0016	0.0016	0.0000	0.0000
O_2	0.2076	0.0000	0.0000	0.9500	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
COS (ppmv)	0	0	0	0	0	0	0	69	0	55	0	0	0
H_2S (ppmv)	0	0	0	0	0	0	0	1,870	0	1,483	1,538	2,082	5
SO_2 (ppmv)	0	0	0	0	0	0	0	0	0	0	0	0	0
SO_3 (ppmv)	0	0	0	0	0	0	0	0	0	0	0	0	0
Total V-L Flow (lb_{mol}/hr)	26,184	20,173	291	5,715	0	6,064	0	28,705	7,455	36,162	36,160	25,243	22,350
Total V-L Flow (lb/hr)	755,405	565,766	5,439	184,200	0	109,249	0	588,045	134,298	722,343	722,346	522,283	399,498
Solids Flow													
Coal	0	0	0	0	283,833	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0	0	0
Slag	0	0	0	0	0	0	17,091	0	0	0	0	0	0
Temperature (°F)	63	60	100	316		60		500	500	500	792	105	123
Pressure (psia)	14.4	14.7	87.5	620.0	14.7	14.7	15.0	391.0	470.0	391.0	381.0	353.0	338.0

Table 3-2 (Cont'd)
Process Stream Compositions and State Points – Wyodak Coal

STREAM NUMBER	14	15	16	17	18	19	20	21	22	23	24	25
Vapor - Liquid												
Mole Fraction												
Ar	0.0000	0.0171	0.0000	0.0094	0.0012	0.0012	0.0094	0.0019	0.0020	0.0000	0.0094	0.0142
CH_4	0.0000	0.0039	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.1780	0.0003	0.0000	0.0003	0.0003	0.0000	0.0003	0.0003	0.0000	0.0000	0.0000
CO_2	0.0010	0.5641	0.9791	0.0003	0.8640	0.8811	0.0003	0.8125	0.8260	0.0000	0.0003	0.3735
H_2	0.9976	0.2205	0.0018	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0000	0.0057	0.0000	0.0108	0.0194	0.0000	0.0108	0.0009	0.0000	0.0103	0.0108	0.1237
H_2SO_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.9897	0.0000	0.0000
N_2	0.0014	0.0107	0.0000	0.7719	0.0978	0.0997	0.7719	0.1580	0.1606	0.0000	0.7719	0.4679
NH_3	0.0000	0.0000	0.0001	0.0000	0.0001	0.0001	0.0000	0.0001	0.0001	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.2076	0.0008	0.0008	0.2076	0.0108	0.0110	0.0000	0.2076	0.0207
COS (ppmv)	0	0	0	0	0	0	0	0	0	0	0	0
H_2S (ppmv)	0	9	18,618	0	0	0	0	0	0	0	0	0
SO_2 (ppmv)	0	0	0	0	16,429	16,754	0	46	47	0	0	0
SO_3 (ppmv)	0	0	0	0	0	0	0	15,403	0	0	0	0
Total V-L Flow (lb_{mol}/hr)	10,316	12,034	2,849	403	3,263	3,200	289	3,486	3,328	52	14,400	24,034
Total V-L Flow (lb/hr)	21,596	377,901	122,785	11,623	134,408	133,294	8,351	141,645	137,417	5,100	415,444	793,224
Solids Flow												
Coal	0	0	0	0	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	0	0	0
Slag	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	70	70	123	331	750	650	180	300	315	315	89	280
Pressure (psia)	310.0	17.0	338.0	45.0	42.0	29.5	25.0	20.0	19.0	19.0	17.0	14.7

3.2 PROCESS DESCRIPTION

Following are more detailed descriptions of the key process elements:

3.2.1 Gasifier

For this application to produce hydrogen, a dual-train E-Gas™ gasifier of the Wabash River configuration, consisting of two 50 percent gasifiers, 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 1900°F goes through an internal cyclone that separates entrained particles from the gas for recycle to the gasifier, followed by a fire-tube boiler and gas cooler to reduce gas temperature to 645°F (341°C). Following the cooler, remaining particulates are removed from the gas with a metallic 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 ASU. The air separation plant is designed to produce a nominal output of 2,200 tons/day of 99 percent pure O_2. The high-pressure plant is designed with one production train, 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 sintered metal candle configuration, operating at the relatively low temperature of 645°F (341°C). The vessel and candle array is similar to the configuration used at the Wabash River demonstration plant. A particulate removal vessel is provided for each gasifier train.

3.2.4 CO Shift

After leaving the particulate control unit, steam is injected into the gas stream and the CO in the syngas is shifted to hydrogen and CO_2 in the shift converter utilizing a sulfur-tolerant shift catalyst. The shift catalyst also promotes the COS hydrolysis reaction. Heat is removed from the gas stream following the shift, the gases are cooled, sour water is condensed, and the gas stream is sent to the sulfur removal unit.

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. The two reactors in parallel are required due to the high gas mass flow rate. Feed to the shift converter is first preheated by hot effluent from the second converter, heated by hot effluent from the first converter, and fed to the top of the two first-stage converters in parallel. Effluent from the first stage is cooled and fed to the top of the second-stage converters. Effluent from the second stage is cooled by exchanging heat with incoming feed, by an air cooler and finally by a water cooler.

3.2.5 Amine Unit/Acid Gas Concentrator

The purpose of the amine unit is to remove H_2S from the fuel gas stream. This step is necessary in order to minimize plant sulfur emissions. The solvent used in this case is a proprietary formulation based on MDEA. A traditional absorber/stripper arrangement will be used.

Cool, dry, and particulate-free synthesis gas enters the absorber unit at 353 psia and 105°F (41°C). In the absorber, H_2S along with some CO_2 is removed from the fuel gas stream. Clean fuel gas exits the top of the absorber and is then routed to the saturator column.

The rich solution leaving the bottom of the absorber is regenerated in a stripper through the indirect application of thermal energy via condensing low-pressure steam in a reboiler. The stripper acid gas stream, consisting of 16 percent H_2S and 78 percent CO_2 (with the balance mostly H_2O), requires further treatment before being sent to the sulfuric acid plant.

Typically, for good performance and operation, the minimum H_2S concentration in the acid gas feed to an acid plant should be above 27 percent. In this case, an acid gas concentrator was used to further concentrate the H_2S stream.

3.2.6 Hydrogen Purification

The product hydrogen stream exits the absorber 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 HRSG to be fired with oxygen.

Treated gas from the amine unit absorber is fed directly to the PSA unit where hydrogen is purified up to approximately 99.9 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 adsorbent beds at 338 psia, and then the impurities are purged from the beds at 17 psia. Purge gas is sent to the gas-fired heat recovery unit for steam generation. Purified hydrogen is produced at 310 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.

3.3 MAJOR EQUIPMENT LIST FOR WYODAK COAL CASE

This section contains the equipment list corresponding to the power plant configuration shown in Figure 3-1. This list, along with the heat and material balance and supporting performance data, was used to generate plant costs and used in the financial analysis. In the following, all feet (ft) conditions specified for process pumps correspond to feet of liquid being pumped.

ACCOUNT 1 COAL RECEIVING AND HANDLING

Equipment No.	Description	Type	Design Condition	Qty.
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	100 ton	2
2	Feeder	Vibratory	300 tph	2
3	Conveyor 1	54" belt	450 tph	1
4	As-Received Coal Sampling System	Two-stage	N/A	1
5	Conveyor 2	54" belt	450 tph	1
6	Reclaim Hopper	N/A	40 ton	2
7	Feeder	Vibratory	150 tph	2
8	Conveyor 3	48" belt	300 tph	1
9	Crusher Tower	N/A	300 tph	1
10	Coal Surge Bin w/Vent Filter	Compartment	300 ton	1
11	Crusher	Granulator reduction	6"x0 - 3"x0	1
12	Crusher	Impactor reduction	3"x0 - 1"x0	1
13	As-Fired Coal Sampling System	Swing hammer	N/A	2
14	Conveyor 4	48" belt	300 tph	1
15	Transfer Tower	N/A	300 tph	1
16	Tripper	N/A	300 tph	1
17	Coal Silo w/Vent Filter and Slide Gates	N/A	600 ton	2

ACCOUNT 2 COAL-WATER SLURRY PREPARATION AND FEED

Equipment No.	Description	Type	Design Condition	Qty.
1	Feeder	Vibrating	100 tph	2
2	Weigh Belt Feeder		48" belt	2
3	Rod Mill	Rotary	100 tph	2
4	Slurry Water Pumps	Centrifugal	180 gpm @ 500 ft	2
5	Slurry Water Storage Tank	Vertical	1,500 gal	1
6	Rod Mill Product Tank	Vertical	35,000 gal	2
7	Slurry Storage Tank with Agitator	Vertical	150,000 gal	2
8	Coal-Slurry Feed Pumps	Positive displacement	700 gpm @ 1,250 ft	2
9	LT Slurry Heater	Shell and tube	20 x 10^6 Btu/h	2
10	HT Slurry Heater	Shell and tube	7 x 10^6 Btu/h	2

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS BOP SYSTEMS

ACCOUNT 3A CONDENSATE AND FEEDWATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cond. Storage Tank	Vertical, cylindrical, outdoor	200,000 gal	1
2	Condensate Pumps	Vert. canned	1,000 gpm @ 400 ft	2
3	Deaerator	Horiz. spray type	1,100,000 lb/h 205°F to 240°F	1
4	IP Feed Pump	Horiz. centrifugal single stage	200 gpm/1,850 ft	2
5	HP Feed Pump	Barrel type, multi-staged, centr.	500 gpm @ 4,000 ft	2

ACCOUNT 3B MISCELLANEOUS EQUIPMENT

Equipment No.	Description	Type	Design Condition	Qty
1	Auxiliary Boiler	Shop fabricated, water tube	400 psig, 650°F 70,000 lb/h	1
2	Service Air Compressors	Recip., single stage, double acting, horiz.	100 psig, 750 cfm	2
3	Inst. Air Dryers	Duplex, regenerative	750 cfm	1
4	Service Water Pumps	Horiz. centrifugal, double suction	200 ft, 1,200 gpm	2
5	Closed Cycle Cooling Water Pumps	Horizontal, centrifugal	70 ft, 1,200 gpm	2
6	Fire Service Booster Pump	Two-stage horiz. centrifugal	250 ft, 1,200 gpm	1
7	Engine-Driven Fire Pump	Vertical turbine, diesel engine	350 ft, 1,000 gpm	1
8	Raw Water Pumps	SS, single suction	60 ft, 300 gpm	2
9	Filtered Water Pumps	SS, single suction	160 ft, 120 gpm	2
10	Filtered Water Tank	Vertical, cylindrical	15,000 gal	1
11	Makeup Demineralizer	Anion, cation, and mixed bed	70 gpm	2
12	Sour Water Stripper System	Vendor supplied	200,000 lb/h sour water	1
13	Liquid Waste Treatment System	Vendor supplied	400 gpm	1

ACCOUNT 4 GASIFIER AND ACCESSORIES

ACCOUNT 4A GASIFICATION

Equipment No.	Description	Type	Design Condition	Qty
1	Gasifier	Pressurized entrained bed/syngas cooler	2,500 std (dry-coal basis) @ 425 psia	2
2	Raw Gas Cooler Steam Generator	Fire tube boiler	1,500 psig/600°F 160.0 MMBtu/h	2
3	Raw Gas Cooler Steam Generator	Shell and tube	800 psig/518°F 163.3 MMBtu/h	2
4	Medium-Temperature Candle Filter	Sintered stainless	400 psia, 645°F	2
5	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	770,000 lb/h, medium-Btu gas	1

ACCOUNT 4B AIR SEPARATION PLANT

Equipment No.	Description	Type	Design Condition	Qty
1	Air Compressor	Centrifugal, multi-stage	75,000 scfm, 67 psia discharge pressure	2
2	Cold Box	Vendor supplied	2,200 ton/day O_2	1
3	Oxygen Compressor	Centrifugal, multi-stage	17,000 scfm, 600 psig discharge pressure	2
4	Liquid Oxygen Storage Tank	Vertical	60' dia x 80' vert	1

ACCOUNT 5 SYNGAS SHIFT AND CLEANUP

ACCOUNT 5A WATER-GAS SHIFT AND RAW GAS COOLING

Equipment No.	Description	Type	Design Condition	Qty
1	High Temperature Shift Reactor 1	Fixed bed	400 psia, 500°F	1
2	High Temperature Shift Reactor 2	Fixed bed	400 psia, 750°F	1
3	HP Steam Generator	Shell and tube	70 x 10^6 Btu/h @ 1700 psia and 613°F	1
4	IP Steam Generator	Shell and tube	45 x 10^6 Btu/h @ 800 psia and 518°F	1
5	LP Steam Generator	Shell and tube	30 x 10^6 Btu/h @ 200 psia and 382°F	1
6	Raw Gas Coolers	Shell and tube with condensate drain	100 x 10^6 Btu/h	2
7	Raw Gas Knock Out Drum	Vertical with mist eliminator	400 psia, 130°F	1

ACCOUNT 5B ACID GAS REMOVAL AND HYDROGEN PURIFICATION

Equipment No.	Description	Type	Design Condition	Qty
1	Amine Absorber 1	Column	90,000 scfm (4,400 acfm), 353 psia, 103°F	2
2	Amine Regenerator 1	Column	Vendor design	1
3	Sulfuric Acid Plant	Vendor design	61 tpd sulfuric acid	1
4	PSA Unit	Fixed bed	93 MMscfd H_2 @ 310 psia	1

ACCOUNT 6 COMBUSTION TURBINE/ACCESSORIES

Not applicable.

ACCOUNT 7 WASTE HEAT BOILER, DUCTING, AND STACK

Equipment No.	Description	Type	Design Condition Drums	Qty
1	Heat Recovery Steam Generator	Fired drum	1700 psig/1000°F 600,000 lb/h 200 x 10^6 Btu/h	1
2	Stack	Carbon steel plate, type 409 stainless steel liner	213 ft high x 28 ft dia.	1

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Design Condition (per each)	Qty
1	81 MW Steam Turbine Generator	TC2F26	1800 psig 1000°F/1000°F	1
2	Bearing Lube Oil Coolers	Plate and frame		2
3	Bearing Lube Oil Conditioner	Pressure filter closed loop		1
4	Control System	Digital electro-hydraulic	1600 psig	1
5	Generator Coolers	Plate and frame		2
6	Hydrogen Seal Oil System	Closed loop		1
7	Surface Condenser	Single pass, divided waterbox	880,000 lb/h steam @ 2.4 in. Hga	1
8	Condenser Vacuum Pumps	Rotary, water sealed	2500/25 scfm (hogging/holding)	2

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition (per each)	Qty
1	Circ. Water Pumps	Vertical wet pit	40,000 gpm @ 60 ft	2
2	Cooling Tower	Mechanical draft	100,000 gpm	1

ACCOUNT 10 ASH/SPENT SORBENT RECOVERY AND HANDLING

ACCOUNT 10A SLAG DEWATERING AND REMOVAL

Equipment No.	Description	Type	Design Condition	Qty
1	Slag Dewatering System	Vendor proprietary	205 tpd	1

3.4 CAPITAL AND OPERATING COSTS FOR HYDROGEN FROM WYODAK PRB COAL

The total plant cost for the plant producing 313 tons of hydrogen per day from Wyodak PRB coal is $364.6 million in 2001 dollars. The capital cost summary is included in Table 3-3.

Table 3-3
Capital Cost Summary – Hydrogen Production from Wyodak Coal

Client: DEPARTMENT OF ENERGY — Report Date: 03-Jun-2002 03:16 PM
Project: NETL H_2 Production Facility

TOTAL PLANT COST SUMMARY

Case: Sub-Bituminous H_2 Plant w/o CO_2 Capture
Plant Size: 259.2 H_2 TPD — Estimate Type: Conceptual — Cost Base (Dec) 2001 ($X1000 & $X1000/TPD)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $	TOTAL PLANT COST $/TPD
1	COAL & SORBENT HANDLING	6,242	1,282	4,611	323		$12,457	997		1,345	$14,799	57
2	COAL & SORBENT PREP & FEED	10,581	1,539	5,776	404		$18,299	2,040		2,034	$22,373	86
3	FEEDWATER & MISC. BOP SYSTEMS	4,259	1,253	2,561	179		$8,253	660		891	$9,804	38
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries (E	54,709		23,270	1,629		$79,607	9,553		8,916	$98,076	378
4.2	Syngas Cooling	w/4.1		w/ 4.1	w/ 4.1			w/ 4.1		w/ 4.1		
4.3	ASU/Oxidant Compression	29,814		w/equip.			$29,814	2,385		3,220	$35,419	137
4.4-4.9	Other Gasification Equipment	7,665	5,742	5,067	355		$18,829	1,538		2,037	$22,404	86
	SUBTOTAL 4	*92,188*	*5,742*	*28,336*	*1,984*		*$128,250*	*13,476*		*14,173*	*$155,899*	*602*
5	HYDROGEN SEPARATION/GAS CLEAN	46,572	3,701	16,976	1,188		$68,437	8,005		7,644	$84,086	324
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Expander Turbine/Generator											
6.2-6.9	Combustion Turbine Accessories											
	SUBTOTAL 6											
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	4,533		551	39		$5,123	410		553	$6,086	23
7.2-7.9	HRSG Accessories, Ductwork and Stack	496	185	296	21		$997	80		108	$1,185	5
	SUBTOTAL 7	*5,029*	*185*	*847*	*59*		*$6,120*	*490*		*661*	*$7,271*	*28*
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	7,843		1,094	77		$9,013	721		973	$10,707	41
8.2-8.9	Turbine Plant Auxiliaries and Steam Pip	3,567	109	1,656	116		$5,447	436		588	$6,471	25
	SUBTOTAL 8	*11,409*	*109*	*2,749*	*192*		*$14,460*	*1,157*		*1,562*	*$17,178*	*66*
9	COOLING WATER SYSTEM	2,438	1,193	1,951	137		$5,719	458		618	$6,795	26
10	ASH/SPENT SORBENT HANDLING SYS	4,322	559	1,845	129		$6,855	745		760	$8,360	32
11	ACCESSORY ELECTRIC PLANT	4,974	2,052	4,028	282		$11,335	907		1,224	$13,466	52
12	INSTRUMENTATION & CONTROL	5,045	1,199	3,720	260		$10,224	818		1,104	$12,146	47
13	IMPROVEMENTS TO SITE	1,465	842	2,425	170		$4,902	392		529	$5,824	22
14	BUILDINGS & STRUCTURES		2,630	2,702	189		$5,521	442		596	$6,559	25
	TOTAL COST	**$194,524**	**$22,285**	**$78,527**	**$5,497**		**$300,832**	**$30,586**		**$33,142**	**$364,560**	**1407**

3.5 DETERMINATION OF COST OF HYDROGEN FROM WYODAK PRB COAL USING PRELIMINARY ASSUMPTIONS

For this economic analysis, the capital and operating costs for the two plants being evaluated have been upgraded to 2001 dollars. Coal cost has been estimated at $0.60 per MMBtu.

3.5.1 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 3-4, 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 3-4
Operating Cost Data

OPERATING LABOR REQUIREMENTS		
BITUMINOUS H_2 PLANT W/O CO_2 CAPTURE		
Operating Labor Rate (base)	$26.15/hour	
Operating Labor Burden	30.00% of base	
Labor OH Charge Rate	25.00% of labor	
Operating Labor Requirements (OJ) per Shift		
Category	**1 Unit/Mod**	**Total Plant**
Skilled Operator	1.0	1.0
Operator	8.0	8.0
Foreman	1.0	1.0
Lab Techs, etc.	1.0	1.0
TOTAL – OJs	11.0	11.0

CONSUMABLES, BYPRODUCTS & FUELS DATA			
BITUMINOUS H_2 PLANT W/O CO_2 CAPTURE			
	Consumption		**Unit**
Item/Description	**Initial**	**/Day**	**Cost**
Water (/1,000 gal)		884	0.80
Chemicals			
Water Treatment (lb)	64,217	2,141	0.15
Limestone (ton)			16.25
Shift Catalyst (lb)	12,785	426.2	5.25
Amine (lb)	8,640	288.0	0.63
Other			
Supplemental Fuel (MMBtu)			
Purchased Power (MWh)			
LP Steam (/1,000 lb)			
Waste Disposal			
Sludge (ton)			
Slag (ton)		205	10.00
Byproducts & Emissions			
Sulfuric Acid (tons)		61	75.00
Excess Electric Generation (MWh)		1,004	30.00
Fuel (ton)		3,406	10.36

3.5.2 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.

3.5.3 Consumables

Shift Catalyst:

- Change-out every 3 years
- 0.0045 pounds of catalyst per 1,000 standard cubic feet of hydrogen
- 210 tons initial charge
- 70 tons per year annual cost

Proprietary Amine:

- 12 pounds per hour
- 100,000 pounds per year

PSA Sorbent:

- Periodic change-out with scheduled maintenance

SO_2 Conversion Catalyst:

- Periodic change-out with scheduled maintenance

3.5.4 Byproduct Credits

The production of 61 tons of sulfuric acid per day is taken as a byproduct credit at $75 per ton.

3.5.5 Financial Assumptions

The cost of hydrogen was determined based on financial assumptions typically used by Parsons. These are summarized in Table 3-5.

Table 3-5
Financial Parameters

Levelized capacity factor	90%
Design/construction period	4 years
Plant startup date	January 2005
Land area/Unit cost	100 acres @ 41,500/acre
Project book life	20 years
Project tax life	20 years
Tax depreciation method	Accelerated based on ACRS class
Property tax rate	1.0% per year
Insurance tax rate	1.0% per year
Federal income tax rate	34.0%
State income tax rate	4.2%
Capital structure	
Common equity	20% @ 16.50% annum
Debt	80% @ 6.30% annum
Weighted cost of capital (after tax)	6.49%

3.5.6 Cost Results

Applying the financial parameters from Table 3-5, the cost of hydrogen was estimated to be \$6.44/MMBtu (\$2.20/Mcf) for Wyodak PRB coal. The results of the cost estimating activity are summarized in Table 3-6. These results were used as the price of hydrogen input to the DOE IGCC financial model. The results are shown in Appendix B.

Table 3-6

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY

TITLE/DEFINITION

Case:	Sub-Bituminous H_2 Plant w/o CO_2 Capture		
Plant Size:	259.2 TPD-Synga	HeatRate:	(Btu/kWh)
Primary/Secondary Fuel(type):	Wyodak	Cost:	0.60 ($/MMBtu)
Design/Construction:	3 (years)	BookLife:	20 (years)
TPC(Plant Cost) Year:	2001 (Jan.)	TPI Year:	2005 (Jan.)
Capacity Factor:	90 (%)		

CAPITAL INVESTMENT		**$x1000**	**$x1000/$H_2$TPD**
Process Capital & Facilities		300,832	1160.8
Engineering(incl.C.M.,H.O.& Fee)		30,586	118.0
Process Contingency			
Project Contingency		33,142	127.9
TOTAL PLANT COST(TPC)		$364,560	1406.7
TOTAL CASH EXPENDED	$364,560		
AFDC	$23,376		
TOTAL PLANT INVESTMENT(TPI)		$387,936	1496.9
Royalty Allowance			
Preproduction Costs		9,096	35.1
Inventory Capital		2,102	8.1
Initial Catalyst & Chemicals(w/equip.)			
Land Cost		150	0.6
TOTAL CAPITAL REQUIREMENT(TCR)		$399,284	1540.7

OPERATING & MAINTENANCE COSTS (2001 Dollars)	**$x1000**	**$x1000/$H_2$TPD**
Operating Labor	3,276	12.6
Maintenance Labor	2,549	9.8
Maintenance Material	3,824	14.8
Administrative & Support Labor	1,456	5.6
TOTAL OPERATION & MAINTENANCE	$11,105	42.8
FIXED O & M		38.56
VARIABLE O & M		4.28

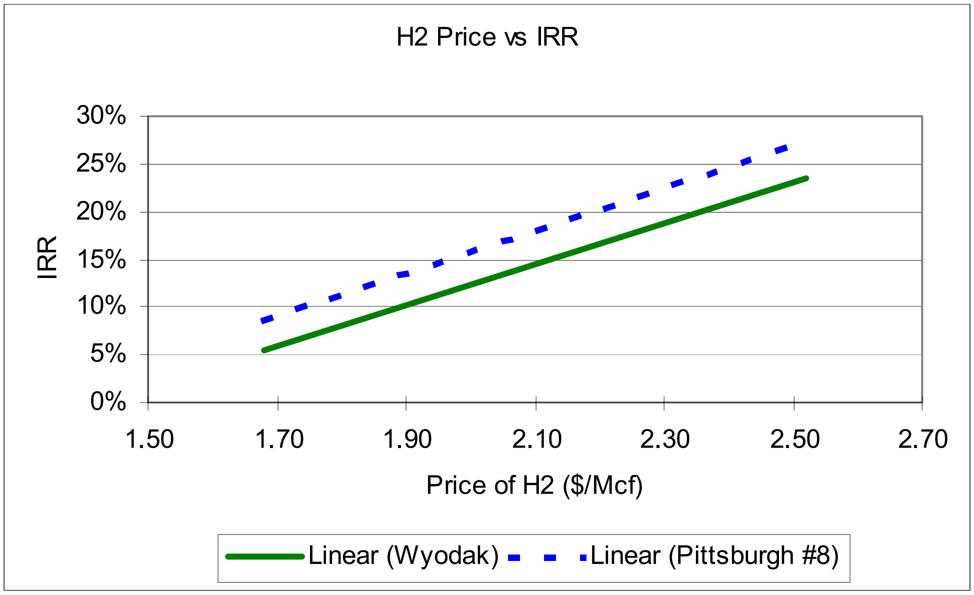
CONSUMABLE OPERATING COSTS,less Fuel (2001 Dollars)	**$x1000**	**$/T H_2-yr**
Water	232	2.73
Chemicals	897	10.54
Other Consumables		
Waste Disposal	674	7.91
TOTAL CONSUMABLE OPERATING COSTS	$1,803	21.18
BY-PRODUCT CREDITS (2001 Dollars)	($11,401)	-133.93
FUEL COST (2001 Dollars)	$11,587	136.11

PRODUCTION COST SUMMARY	1st Year (2005 $) **$/T H_2-yr**	Levelized (Over Book Life $) **$/T H_2-yr**
Fixed O & M	117.40	117.40
Variable O & M	4.28	4.28
Consumables	21.18	21.18
By-product Credit/Penalty	-133.93	-133.93
Fuel	132.87	121.20
TOTAL PRODUCTION COST	141.81	130.14
LEVELIZED CARRYING CHARGES(Capital)	12,072	656.63
LEVELIZED(Over Book Life)COST/Ton of Syngas	55,900	786.76
Equivalent 1st.Yr.$/MSCF / Lev'd $/MMBtu	2.202	6.44

4. COST OF HYDROGEN USING THE DOE IGCC FINANCIAL MODEL

The DOE IGCC financial model was utilized to determine the sensitivity of internal rate of return (IRR), which would be obtained as the selling price of hydrogen varies. The tariff values for hydrogen were those estimated in Sections 2 and 3. The results are shown on Figure 4-1. The summaries of the financial model analyses are shown in Appendices A and B, respectively.

Figure 4-1
Sensitivity of IRR to Hydrogen Costs



Note: Due to protected cells in the financial model, it was necessary to classify the sulfuric acid byproduct as elemental sulfur. The elemental sulfur cell valued at $75 per ton is equal to the commercial value of sulfuric acid.

APPENDIX A

PITTSBURGH NO. 8 COAL WITH DOE IGCC FINANCIAL MODEL

IGCC Project Analysis For

Pitts.#8 H2 w/o CO2

Refresh

United States
Department of Energy
Science, Security and Energy: Powering the 21st Century

Project Description

The Pitts.#8 H2 w/o CO2 involves the construction of a Coal-fired 37.55 MW integrated gasification combined cycle (IGCC) plant that has a total initial capital cost of $440,772 (in thousand dollars). The Pitts.#8 H2 w/o CO2 is located in the U.S. and, on an annual basis, will generate the following outputs (see shaded box to the right):

> 296,867 MWh of Electricity
> 37,597 MMcf of Hydrogen
> 75,541 Tons of ~~Elemental~~ Sulfur

Project Timing

The scheduled start date for plant constructior is January 1, 2002. The Pitts.#8 H2 w/o CO2 has an estimated construction period of 36 months. Plant start-up will commence on January 1, 2005.

Project Financial Structure and Key Results

The capital structure for Pitts.#8 H2 w/o CO2 is 80% debt and 20% equity. Key project financial results include:

> 18% Internal Rate of Return
> $ 122,020 Net Present Value (in Thousand Dollars) at a 6% Discount Rate
> 1.4 Benefit to Cost Ratio (BCR)
> The Payback Year (on equity) is 2011

Additional Comments Default Data

Gasification Financial Model: Summary Financial Results

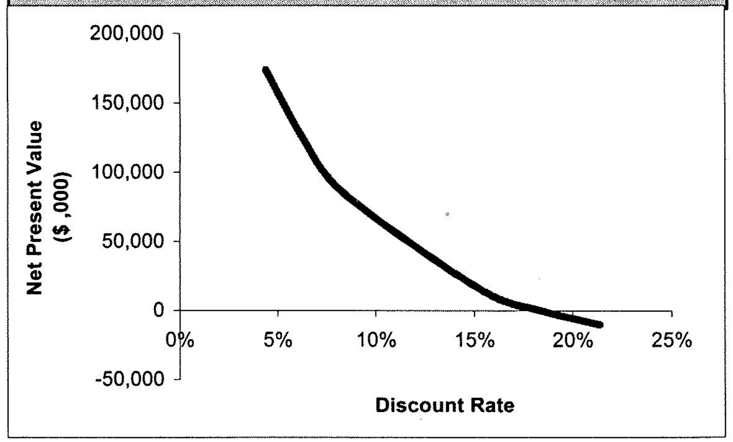
Pitts.#8 H2 w/o CO2
the U.S.

Default Data

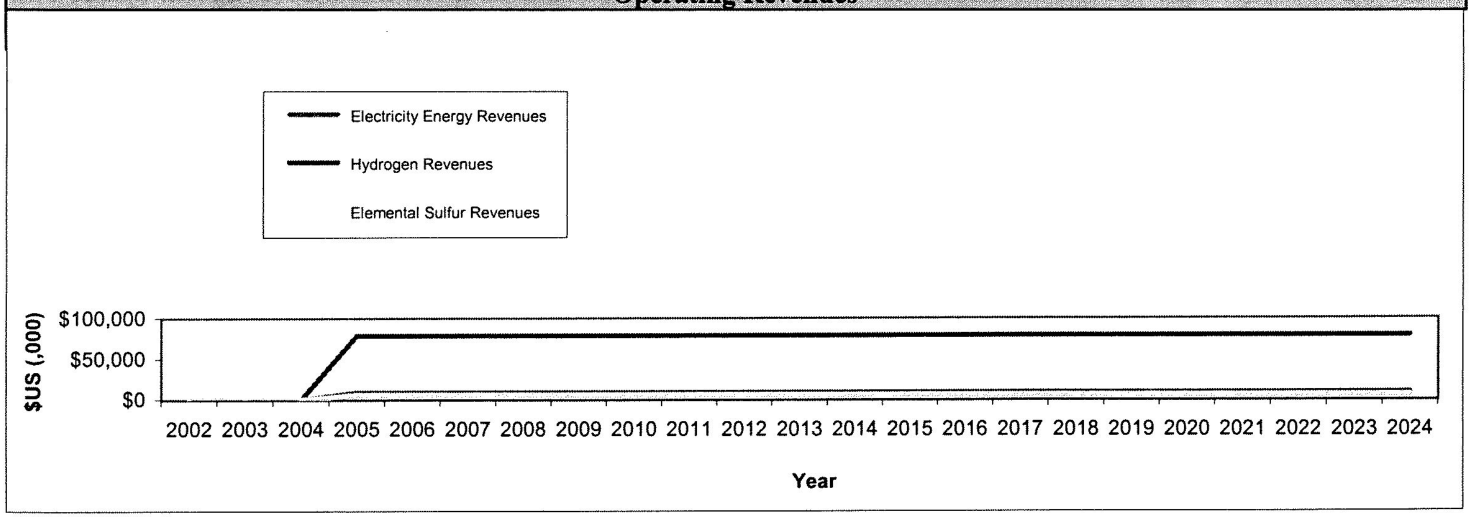
Key Financial Results	
Internal Rate of Return (IRR)	18%
Payback Year (on equity investment)	2011
Benefit to Cost Ratio	1.43

NPV (in Thousand Dollars)	
NPV at 4% discount rate	$173,716
NPV at 6% discount rate	$122,020
NPV at 8% discount rate	$84,556

IRR: 18%



Operating Revenues



Gasification Financial Model:
Plant Performance Summary

Default Data

Plant Capacity	
Syngas Capacity (MMcf/Day)	0
Net Electric Power Capacity (MW)	38
Steam Capacity (Tons/Hr)	0
Hydrogen Capacity (MMcf/Day)	114
Carbon Monoxide Capacity (MMcf/Day)	0
Elemental Sulfur Capacity (Tons/Day)	229
Slag Ash Capacity (Tons/Day)	0
Fuel (Tons/Day)	0
Chemicals (Tons/Day)	0
Environmental Credit (Tons/Day)	0
Other (Tons/Day)	0

Annual Plant Output (adjusted for both planned and forced outages)	
Syngas (MMcf)	0
Power Generation (MWh)	296,867
Steam Generation (Tons)	0
Hydrogen Generation (MMcf)	37,597
Carbon Monoxide (MMcf)	0
Elemental Sulfur (Tons)	75,541
Slag Ash (Tons)	0
Fuel (Tons)	0
Chemicals (Tons)	0
Environmental Credit (Tons)	0
Other (Tons)	0

Gasification Financial Model:
Plant Performance Summary

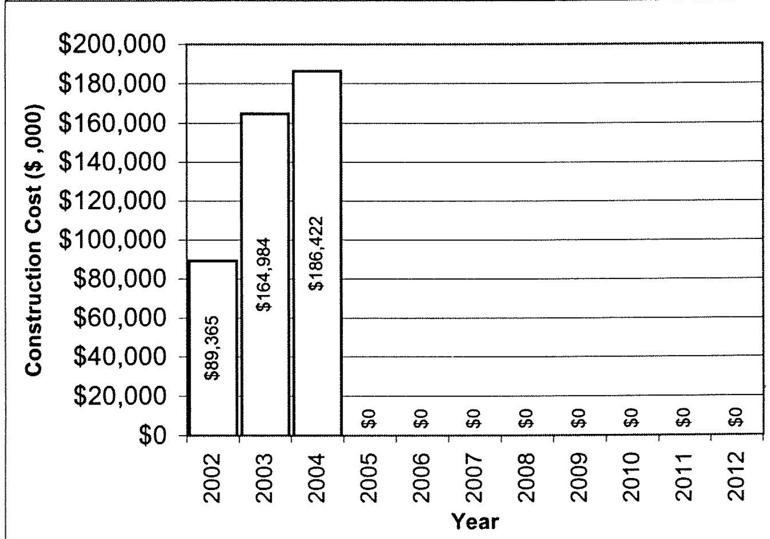
Plant Performance Results	
Primary Project Output/Application	Multiple Outputs
First Year of Operation (plant start-up date)	2005
Primary Fuel Heat Rate (Btu/kWh) based on HHV	69,809
Secondary Fuel Heat Rate (Btu/kWh) based on HHV	0
Plant Operating Efficiency	Not Applicable
Primary Fuel Heat Input (MMBtu)	21,814,797
Secondary Fuel Heat Input (MMBtu)	0
Primary Fuel Type	Coal
Secondary Fuel Type	None
Annual Primary Fuel Consumption (Thousand tons)	876
Not Applicable	0
Operating Hours (8760 hours - Planned Maintenance)	8,322
Guaranteed Availability (adjusts for forced outages)	95%
Overall Plant Availability (adjusts for planned and forced outages)	90%

Gasification Financial Model:
Cost Summary

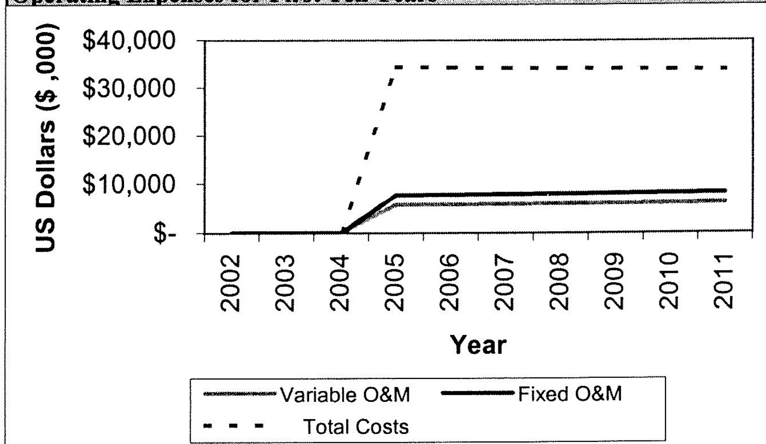
Pitts.#8 H2 w/o CO2
the U.S.
Default Data

Construction/Project Cost (in Thousand Dollars)		
Capital Costs	**Category**	**Percentage**
EPC Costs	$348,173	79%
Initial Working Capital	$6,547	1%
Owner's Contingency (% of EPC Costs)	$34,189	8%
Development Fee (% of EPC Costs)	$0	0%
Start-up (% of EPC Costs)	$9,621	2%
Initial Debt Reserve Fund	$0	0%
Owner's Cost (in thousand dollars)	$0	0%
Additional Capital Cost	$0	0%
Total Capital Costs	$398,530	90%
Financing Costs		
Interest During Construction	$31,911	7%
Financing Fee	$10,331	2%
Additional Financing Cost #1	$0	0%
Additional Financing Cost #2	$0	0%
Total Financing Costs	$42,242	10%
Total Project Cost/Uses of Funds	$440,772	100%
Sources of Funds		
Equity	$88,154	20%
Debt	$352,617	80%
Total Sources of Funds	$440,772	100%

Construction Costs by Year



Operating Expenses for First Ten Years



Gasification Financial Model:
Plant Input Sheet

Project Inputs	Case A	Case B	Case C	Case D
Project Summary Data				
Project Name	Pitts.#8 H2 w/o CO2			
Project Location	the U.S.			
Project Type/Structure	BO			
Primary Output/Plant Application **(Options: Power, Multiple Outputs)**	Multiple Outputs			
Primary Fuel Type **(Options: Gas, Coal, Petroleum Coke, Other/Waste)**	Coal			
Secondary Fuel Type **(Options: None, Gas, Coal, Petroleum Coke, Other/Waste)**	None			
Plant Output and Operating Data : Note - All ton units are US Short Tons (2000 lbs)				
Syngas Capacity (MMcf/Day)	0			
Gross Electric Power Capacity (MW)	78			
Net Electric Power Capacity (MW)	38			
Steam Capacity (Tons/Hr)	0			
Hydrogen Capacity (MMcf/Day)	114			
Carbon Monoxide Capacity (MMcf/Day)	0			
Elemental Sulfur Capacity (Tons/Day)	229			
Slag Ash Capacity (Tons/Day)	0			
Fuel (Tons/Day)	0			
Chemicals (Tons/Day)	0			
Environmental Credit (Tons/Day)	0			
Other (Tons/Day)	0			
Operating Hours per Year	8,322			
Guaranteed Availability (percentage)	95%			
Enter One of the Following Items(For Each Primary/Secondary Fuel) Depending on Project Type:				
Primary Fuel Heat Rate (Btu/kWh) based on HHV **FOR POWER PROJECTS**				
Secondary Fuel Heat Rate (Btu/kWh) based on HHV **FOR POWER PROJECTS**				
Primary Fuel Annual Fuel Consumption (in MMcf *OR* Thousand Tons) **FOR NON POWER PROJECTS**	876			
Secondary Fuel Annual Fuel Consumption (in MMcf *OR* Thousand Tons) **FOR NON POWER PROJECTS**				
Initial Capital and Financing Costs (enter 'Additional Costs' in thousand dollars)				
EPC (in thousand dollars)	348,173			
Owner's Contingency (% of EPC Costs)	10%			
Development Fee (% of EPC Costs)	0%			
Start-up (% of EPC Costs)	3%			
Owner's Cost (in thousand dollars)	1%			
Additional Capital Cost	0			
Additional Financing Cost #1	0			
Additional Financing Cost #2	0			
Operating Costs and Expenses				
Variable O&M (% of EPC Cost)	1.8%			
Fixed O&M Cost (% of EPC Cost)	2.1%			
Additional Comments	Default Data			

Gasification Financial Model:
Scenario Inputs

Default Data

FINANCIAL ASSUMPTIONS

Capital Structure	
Percentage Debt	80%
Percentage Equity	20%
Total Debt Amount	$352,617

Project Debt Terms	
Loan 1: Senior Debt	
% of Total Project Debt **(total for Loans 1,2, and 3 must = 100%)**	70%
Loan Amount (Thousand $)	$246,832
Interest Rate	6%
Financing Fee	3%
Repayment Term (in Years)	12
Grace Period on Principal Repayment	1
First Year of Principal Repayment	2006
Loan 2: Subordinated Debt	
% of Total Project Debt	30%
Loan Amount (Thousand $)	$105,785
Interest Rate	6%
Financing Fee	3%
Repayment Term (in Years)	15
Grace Period on Principal Repayment	1
First Year of Principal Repayment	2006
Loan 3: Subordinated Debt	
% of Total Project Debt	
Loan Amount (Thousand $)	$0
Interest Rate	
Financing Fee	
Repayment Term (in Years)	
Grace Period on Principal Repayment	
First Year of Principal Repayment	2005

Loan Covenant Assumptions	
Interest Rate for Debt Reserve Fund (DRF)	5%
Debt Reserve Fund Used on Senior Debt **(Options: Yes or No)**	No
Percentage of Total Debt Service used as DRF	50%

Depreciation : "SL" for Straight-Line OR "DB" for 150% Declining Balance		**Method**
Construction (Years) : *Note - DB Method Must be 15 or 20 years*	15	SL

Gasification Financial Model:
Scenario Inputs

Subsidized Tax Rate (used as investment incentive)	0%
Length of Subsidized Tax Period (in Years)	0

FUEL/FEEDSTOCK ASSUMPTIONS

Fuel Prices : For the Base Year, then escalated by fuel factors in B71-B74 above	
Gas ($/Mcf)	4.00
Coal ($/US Short Ton)	24.90
Petroleum Coke ($/US Short Ton)	19.30
Other/Waste ($/US Short Ton)	0.00
Alternatively, use Forecasted Prices (From Fuel Forecasts Sheet)? (Yes/No)	No
Heating Value Assumptions	
HHV of Natural Gas (Btu/cf)	1,022
HHV of Coal (Btu/lb)	12,450
HHV of Petroleum Coke (Btu/lb)	14,819
HHV of Other/Waste (Btu/lb)	10,714

TARIFF ASSUMPTIONS

INITIAL TARIFF LEVEL (In Dollars in the first year of construction)	
Syngas ($/Mcf)	$0
Capacity Payment (Thousand $/MW/Year)	$0
Electricity Payment ($/MWh)	$30
Steam ($/US Short Ton)	$5.00
Hydrogen ($/Mcf)	$2.10
Carbon Monoxide ($/Mcf)	$0
Elemental Sulfur ($/US Short Ton)	$75
Slag Ash ($/US Short Ton)	$0
Fuel ($/US Short Ton)	$0
Chemicals ($/US Short Ton)	$0
Environmental Credit ($/US Short Ton)	$0
Other ($/US Short Ton)	$0

CONSTRUCTION ASSUMPTIONS

	Base Year :	2002					
Construction Schedule	**A**	**B**	**C**	**D**	**E**	**F**	**G**
Construction Start Date	1/1/02	1/1/02	1/1/03	1/1/02	1/1/02	1/1/02	1/1/02
Construction Period (in months)	36	48	36	36	36	36	36
Plant Start-up Date *(must start on January 1)*	1/1/05	1/1/06	1/1/06	1/1/05	1/1/05	1/1/05	1/1/05
EPC Cost Escalation in Effect? (Yes/No)	Yes						

Percentage of Cost for Construction Periods	Five Year Construction Period					Four Year Construction Period				Three Year Period		
Enter for Five, Four or Three Year Periods (To the Right --->)	**Year 1**	**Year 2**	**Year 3**	**Year 4**	**Year 5**	**Year 1**	**Year 2**	**Year 3**	**Year 4**	**Year 1**	**Year 2**	**Year 3**
EPC Costs : Unescalated Allocations	15.0%	25.0%	25.0%	10.0%	25.0%	25.0%	25.0%	25.0%	25.0%	25.0%	42.5%	32.5%
EPC Costs: Escalated Allocations (Use EPC Escalation Sheet to Calculate)	14.2%	24.5%	25.0%	10.2%	26.0%	23.9%	24.9%	25.4%	25.9%	24.5%	42.6%	32.9%
Initial Working Capital	0.0%	0.0%	0.0%	0.0%	100.0%	0.0%	0.0%	0.0%	100.0%	0.0%	0.0%	100.0%

APPENDIX B

WYODAK PRB COAL WITH DOE IGCC FINANCIAL MODEL

IGCC Project Analysis For

Wyodak H2 w/o CO2 | Refresh

United States
Department of Energy
Science, Security and Energy: Powering the 21st Century

Project Description

MW integrated gasification combined cycle (IGCC) plant that has a total initial capital cost of $425,879 (in thousand dollars). The Wyodak H2 w/o CO2 is located in the U.S. and, on an annual basis, will generate the following outputs (see shaded box to the right):

> 313,232 MWh of Electricity
> 30,951 MMcf of Hydrogen
> 20,160 Tons of Elemental Sulfur

Project Timing

The scheduled start date for plant constructior is January 1, 2002. The Wyodak H2 w/o CO2 has an estimated construction period of 36 months. Plant start-up will commence on January 1, 2005.

Project Financial Structure and Key Results

The capital structure for Wyodak H2 w/o CO2 is 80% debt and 20% equity. Key project financial results include:

> 15% Internal Rate of Return
> $ 79,752 Net Present Value (in Thousand Dollars) at a 6% Discount Rate
> 1.3 Benefit to Cost Ratio (BCR)
> The Payback Year (on equity) is 2014

Additional Comments

Default Data

Gasification Financial Model: Summary Financial Results

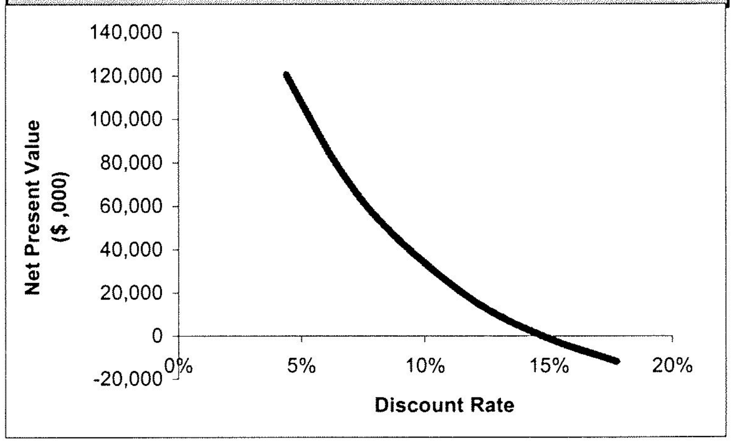
Wyodak H2 w/o CO2 the U.S.

Default Data

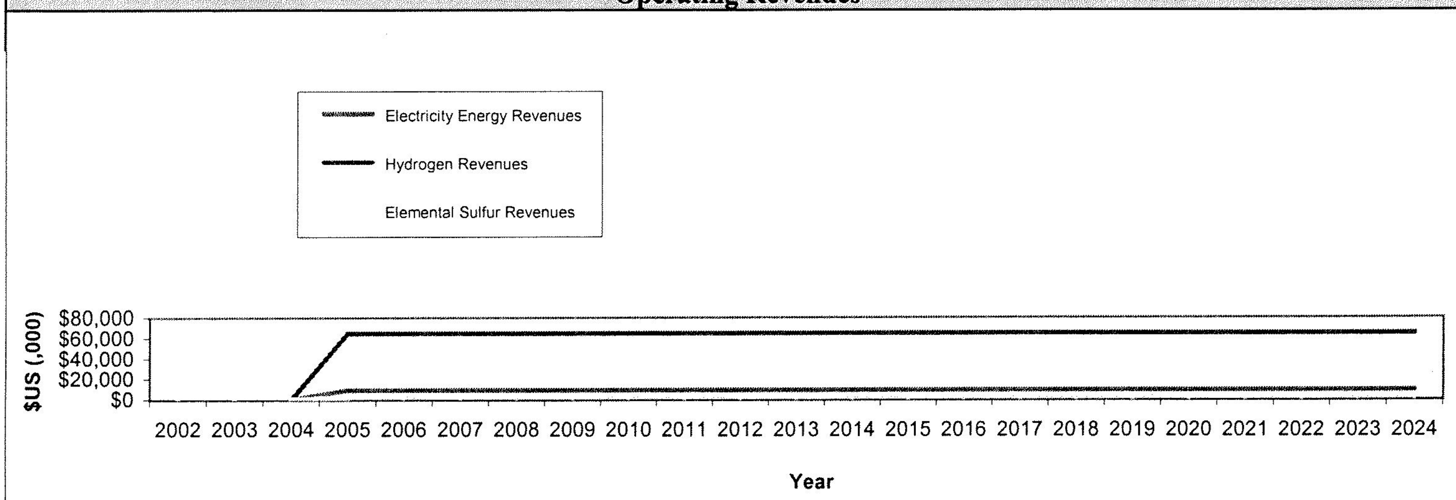
Key Financial Results	
Internal Rate of Return (IRR)	15%
Payback Year (on equity investment)	2014
Benefit to Cost Ratio	1.29

NPV (in Thousand Dollars)	
NPV at 4% discount rate	$120,493
NPV at 6% discount rate	$79,752
NPV at 8% discount rate	$50,495

IRR: 15%



Operating Revenues



Gasification Financial Model:
Plant Performance Summary

Default Data

Plant Capacity	
Syngas Capacity (MMcf/Day)	0
Net Electric Power Capacity (MW)	40
Steam Capacity (Tons/Hr)	0
Hydrogen Capacity (MMcf/Day)	94
Carbon Monoxide Capacity (MMcf/Day)	0
Elemental Sulfur Capacity (Tons/Day)	61
Slag Ash Capacity (Tons/Day)	0
Fuel (Tons/Day)	0
Chemicals (Tons/Day)	0
Environmental Credit (Tons/Day)	0
Other (Tons/Day)	0

Annual Plant Output (adjusted for both planned and forced outages)	
Syngas (MMcf)	0
Power Generation (MWh)	313,232
Steam Generation (Tons)	0
Hydrogen Generation (MMcf)	30,951
Carbon Monoxide (MMcf)	0
Elemental Sulfur (Tons)	20,160
Slag Ash (Tons)	0
Fuel (Tons)	0
Chemicals (Tons)	0
Environmental Credit (Tons)	0
Other (Tons)	0

Gasification Financial Model:
Plant Performance Summary

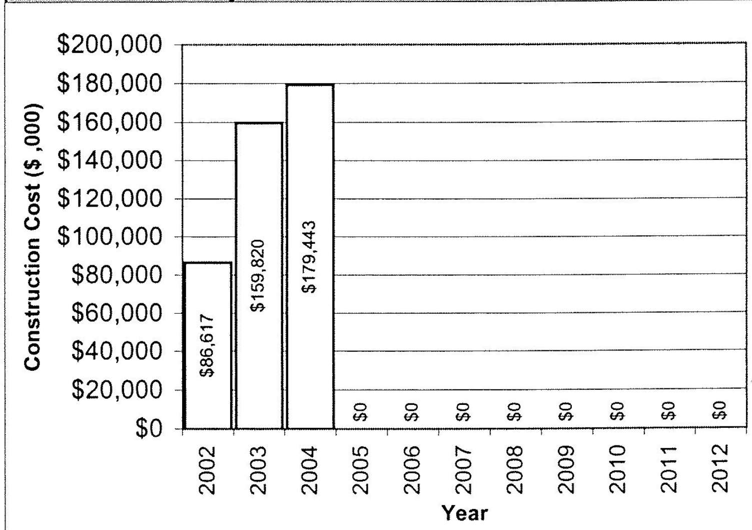
Plant Performance Results	
Primary Project Output/Application	Multiple Outputs
First Year of Operation (plant start-up date)	2005
Primary Fuel Heat Rate (Btu/kWh) based on HHV	58,733
Secondary Fuel Heat Rate (Btu/kWh) based on HHV	0
Plant Operating Efficiency	Not Applicable
Primary Fuel Heat Input (MMBtu)	19,365,334
Secondary Fuel Heat Input (MMBtu)	0
Primary Fuel Type	Coal
Secondary Fuel Type	None
Annual Primary Fuel Consumption (Thousand tons)	1,122
Not Applicable	0
Operating Hours (8760 hours - Planned Maintenance)	8,322
Guaranteed Availability (adjusts for forced outages)	95%
Overall Plant Availability (adjusts for planned and forced outages)	90%

Gasification Financial Model:
Cost Summary

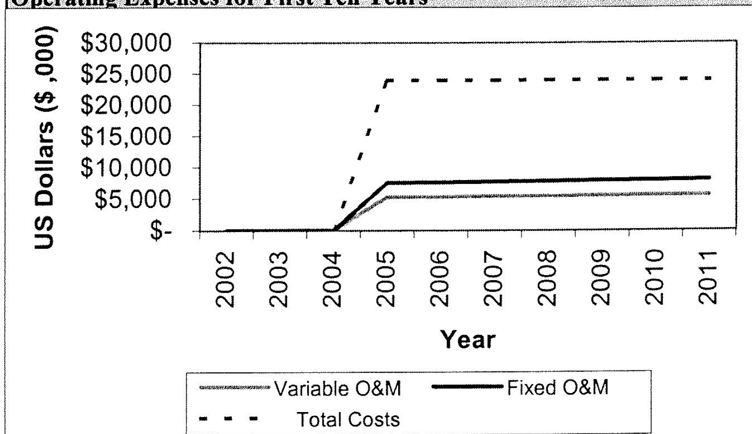
Wyodak H2 w/o CO2
the U.S.
Default Data

Construction/Project Cost (in Thousand Dollars)	Category	Percentage
Capital Costs		
EPC Costs	$337,513	79%
Initial Working Capital	$5,313	1%
Owner's Contingency (% of EPC Costs)	$33,142	8%
Development Fee (% of EPC Costs)	$0	0%
Start-up (% of EPC Costs)	$9,096	2%
Initial Debt Reserve Fund	$0	0%
Owner's Cost (in thousand dollars)	$0	0%
Additional Capital Cost	$0	0%
Total Capital Costs	$385,064	90%
Financing Costs		
Interest During Construction	$30,833	7%
Financing Fee	$9,982	2%
Additional Financing Cost #1	$0	0%
Additional Financing Cost #2	$0	0%
Total Financing Costs	$40,815	10%
Total Project Cost/Uses of Funds	$425,879	100%
Sources of Funds		
Equity	$85,176	20%
Debt	$340,703	80%
Total Sources of Funds	$425,879	100%

Construction Costs by Year



Operating Expenses for First Ten Years



Gasification Financial Model:
Plant Input Sheet

Project Inputs	Case A	Case B	Case C	Case D
Project Summary Data				
Project Name	Wyodak H2 w/o CO2			
Project Location	the U.S.			
Project Type/Structure	BO			
Primary Output/Plant Application **(Options: Power, Multiple Outputs)**	Multiple Outputs			
Primary Fuel Type **(Options: Gas, Coal, Petroleum Coke, Other/Waste)**	Coal			
Secondary Fuel Type **(Options: None, Gas, Coal, Petroleum Coke, Other/Waste)**	None			
Plant Output and Operating Data : Note - All ton units are US Short Tons (2000 lbs)				
Syngas Capacity (MMcf/Day)	0			
Gross Electric Power Capacity (MW)	81			
Net Electric Power Capacity (MW)	40			
Steam Capacity (Tons/Hr)	0			
Hydrogen Capacity (MMcf/Day)	94			
Carbon Monoxide Capacity (MMcf/Day)	0			
Elemental Sulfur Capacity (Tons/Day)	61			
Slag Ash Capacity (Tons/Day)	0			
Fuel (Tons/Day)	0			
Chemicals (Tons/Day)	0			
Environmental Credit (Tons/Day)	0			
Other (Tons/Day)	0			
Operating Hours per Year	8,322			
Guaranteed Availability (percentage)	95%			
Enter One of the Following Items(For Each Primary/Secondary Fuel) Depending on Project Type:				
Primary Fuel Heat Rate (Btu/kWh) based on HHV **FOR POWER PROJECTS**				
Secondary Fuel Heat Rate (Btu/kWh) based on HHV **FOR POWER PROJECTS**				
Primary Fuel Annual Fuel Consumption (in MMcf *OR* Thousand Tons) **FOR NON POWER PROJECTS**	1,122			
Secondary Fuel Annual Fuel Consumption (in MMcf *OR* Thousand Tons) **FOR NON POWER PROJECTS**				
Initial Capital and Financing Costs (enter 'Additional Costs' in thousand dollars)				
EPC (in thousand dollars)	337,513			
Owner's Contingency (% of EPC Costs)	10%			
Development Fee (% of EPC Costs)	0%			
Start-up (% of EPC Costs)	3%			
Owner's Cost (in thousand dollars)	1%			
Additional Capital Cost	0			
Additional Financing Cost #1	0			
Additional Financing Cost #2	0			
Operating Costs and Expenses				
Variable O&M (% of EPC Cost)	1.7%			
Fixed O&M Cost (% of EPC Cost)	2.2%			
Additional Comments	Default Data			

Gasification Financial Model:
Scenario Inputs

Default Data

FINANCIAL ASSUMPTIONS

Capital Structure	
Percentage Debt	80%
Percentage Equity	20%
Total Debt Amount	$340,703

Project Debt Terms	
Loan 1: Senior Debt	
% of Total Project Debt **(total for Loans 1,2, and 3 must = 100%)**	70%
Loan Amount (Thousand $)	$238,492
Interest Rate	6%
Financing Fee	3%
Repayment Term (in Years)	12
Grace Period on Principal Repayment	1
First Year of Principal Repayment	2006
Loan 2: Subordinated Debt	
% of Total Project Debt	30%
Loan Amount (Thousand $)	$102,211
Interest Rate	6%
Financing Fee	3%
Repayment Term (in Years)	15
Grace Period on Principal Repayment	1
First Year of Principal Repayment	2006
Loan 3: Subordinated Debt	
% of Total Project Debt	
Loan Amount (Thousand $)	$0
Interest Rate	
Financing Fee	
Repayment Term (in Years)	
Grace Period on Principal Repayment	
First Year of Principal Repayment	2005

Loan Covenant Assumptions	
Interest Rate for Debt Reserve Fund (DRF)	5%
Debt Reserve Fund Used on Senior Debt **(Options: Yes or No)**	No
Percentage of Total Debt Service used as DRF	50%

Depreciation : "SL" for Straight-Line OR "DB" for 150% Declining Balance		**Method**
Construction (Years) : *Note - DB Method Must be 15 or 20 years*	15	SL

Gasification Financial Model:
Scenario Inputs

Subsidized Tax Rate (used as investment incentive)	0%
Length of Subsidized Tax Period (in Years)	0

FUEL/FEEDSTOCK ASSUMPTIONS

Fuel Prices : For the Base Year, then escalated by fuel factors in B71-B74 above	
Gas ($/Mcf)	4.00
Coal ($/US Short Ton)	10.36
Petroleum Coke ($/US Short Ton)	19.30
Other/Waste ($/US Short Ton)	0.00
Alternatively, use Forecasted Prices (From Fuel Forecasts Sheet)? (Yes/No)	No
Heating Value Assumptions	
HHV of Natural Gas (Btu/cf)	1,022
HHV of Coal (Btu/lb)	8,630
HHV of Petroleum Coke (Btu/lb)	14,819
HHV of Other/Waste (Btu/lb)	10,714

TARIFF ASSUMPTIONS

INITIAL TARIFF LEVEL (In Dollars in the first year of construction)	
Syngas ($/Mcf)	$0
Capacity Payment (Thousand $/MW/Year)	$0
Electricity Payment ($/MWh)	$30
Steam ($/US Short Ton)	$5.00
Hydrogen ($/Mcf)	$2.10
Carbon Monoxide ($/Mcf)	$0
Elemental Sulfur ($/US Short Ton)	$75
Slag Ash ($/US Short Ton)	$0
Fuel ($/US Short Ton)	$0
Chemicals ($/US Short Ton)	$0
Environmental Credit ($/US Short Ton)	$0
Other ($/US Short Ton)	$0

CONSTRUCTION ASSUMPTIONS

	Base Year :	2002					
Construction Schedule	**A**	**B**	**C**	**D**	**E**	**F**	**G**
Construction Start Date	1/1/02	1/1/02	1/1/03	1/1/02	1/1/02	1/1/02	1/1/02
Construction Period (in months)	36	48	36	36	36	36	36
Plant Start-up Date *(must start on January 1)*	1/1/05	1/1/06	1/1/06	1/1/05	1/1/05	1/1/05	1/1/05
EPC Cost Escalation in Effect? (Yes/No)	Yes						

Percentage of Cost for Construction Periods	**Five Year Construction Period**					**Four Year Construction Period**				**Three Year Period**		
Enter for Five, Four or Three Year Periods (To the Right --->)	**Year 1**	**Year 2**	**Year 3**	**Year 4**	**Year 5**	**Year 1**	**Year 2**	**Year 3**	**Year 4**	**Year 1**	**Year 2**	**Year 3**
EPC Costs : Unescalated Allocations	15.0%	25.0%	25.0%	10.0%	25.0%	25.0%	25.0%	25.0%	25.0%	25.0%	42.5%	32.5%
EPC Costs: Escalated Allocations (Use EPC Escalation Sheet to Calculate)	14.2%	24.5%	25.0%	10.2%	26.0%	23.9%	24.9%	25.4%	25.9%	24.5%	42.6%	32.9%
Initial Working Capital	0.0%	0.0%	0.0%	0.0%	100.0%	0.0%	0.0%	0.0%	100.0%	0.0%	0.0%	100.0%