Table 2-18
Plant Water Balance

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

2.5.2 EFFECTIVE THERMAL EFFICIENCY

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

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

$$\text{ETE} = \frac{28{,}563 \text{ lb } H_2\text{/h x } 61{,}095 \text{ Btu/lb} + 47{,}950 \text{ kW x } 3{,}414 \text{ Btu/kWh}}{221{,}631 \text{ lb coal/h x } 12{,}450 \text{ Btu/lb}}$$

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

2.6 COST ESTIMATING

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

2.6.1 APPROACH TO COST ESTIMATING

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

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

Table 2-19

ESTIMATE BASIS/FINANCIAL CRITERIA for REVENUE REQUIREMENT CALCULATIONS			
GENERAL DATA/CHARACTERISTICS			
Case Title:		Hydrogen Fuel Facility	
Unit Size:/Plant Size:		422.5 H_2 TPD	422.5 H_2 TPD
Location:		Middletown, USA	
Fuel:		Pittsburgh	
Levelized Capacity Factor / Preproduction (equivalent months)		80 %	1 month
Capital Cost Year Dollars (Reference Year)		2000 (January)	
Delivered Cost of Coal		1.00$/MBtu	$/MBtu
Design/Construction Period:		3 years	
Plant Startup Date (1st Year Dollars):		2000 (January)	
Land Area/Unit Cost:		150 acres	$1,500/acre
FINANCIAL CRITERIA			
Project Book Life:		20 years	
Book Salvage Value:		%	
Project Tax Rate:		20 year	
Tax Depreciation Method:		Accel. based on ACRS Class	
Property Tax Rate:		% per	
Insurance Tax Rate:		% per	
Federal Income Tax Rate:		34.0%	
State Income Tax Rate:		6.0 %	
Investment Tax Credit/% Eligible:		%	%
Economic Basis:		Over Book Constant Dollars	
Capital Structure:		% of Total	Cost (%)
Common Equity		20	16.5
Preferred Stock			
Debt		80	6.3
Weighted Cost of Capital: (after tax)		6.4 %	
		Over Book Life	2000 to
Escalation Rates:	General	% per year	% per year
	Primary Fuel	-1.1% per year	-0.6% per year
	Secondary Fuel	0.6 % per year	1.27% per year

The IGCC cost model used in the June 1999 Letter Report[5] was the basis for developing the bulk of the balance-of-plant cost portion of the estimate. Use of this model assured consistency in the evaluation of balance-of-plant costs. As before, the capital cost for the gasifiers, gas cleanup including CO_2 removal, and the gas turbine was based on recent studies conducted by Parsons. Destec gasifier pricing was adjusted to reflect the impact of using a total quench in the second stage rather than a firetube boiler, followed by a ceramic candle filter. Balance-of-plant process system costs were estimated from cost curves developed by Parsons based in large part on the results of completed construction projects.

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

- The H_2 ceramic molecular sieve membrane requirement was calculated utilizing the membrane coefficient R&D goal confirmed by ETTP:

  0.1 std cc/minute/cm^2/cm Hg P_{H_2} differential

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

  1.0 lb H_2/h/ft^2

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

2.6.2 PRODUCTION COSTS (OPERATION AND MAINTENANCE)

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

2.6.3 COST RESULTS

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

Table 2-20

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY				
TITLE/DEFINITION				
Case:	Hydrogen Fuel Facility w/Hot Gas Desulfurization			
Plant Size:	422.5 H_2 TPD	HeatRate:	(Btu/kWh)	
Primary/Secondary Fuel(type):	Pitts. #8	Cost:	1.00 ($/MMBtu)	
Design/Construction:	2.5 (years)	BookLife:	20 (years)	
TPC(Plant Cost) Year:	2000 (Jan.)	TPI Year:	2005 (Jan.)	
Capacity Factor:	80 (%)			
CAPITAL INVESTMENT		$x1000	$x1000/$H_2$ TPD	
Process Capital & Facilities		296,792	702.5	
Engineering(incl.C.M.,H.O.& Fee)		28,890	68.4	
Process Contingency		9,270	21.9	
Project Contingency		33,495	79.3	
TOTAL PLANT COST(TPC)		$368,448	872.1	
TOTAL CASH EXPENDED	$368,448			
AFDC	$23,625			
TOTAL PLANT INVESTMENT(TPI)		$392,073	928.1	
Royalty Allowance				
Preproduction Costs		9,466	22.4	
Inventory Capital		3,067	7.3	
Initial Catalyst & Chemicals(w/equip.)				
Land Cost		150	0.4	
TOTAL CAPITAL REQUIREMENT(TCR)		$404,755	958.1	
OPERATING & MAINTENANCE COSTS (2000 Dollars)		$x1000	$x1000/$H_2$ TPD	
Operating Labor		3,574	8.5	
Maintenance Labor		2,733	6.5	
Maintenance Material		4,099	9.7	
Administrative & Support Labor		1,577	3.7	
TOTAL OPERATION & MAINTENANCE		$11,982	28.4	
FIXED O & M			22.69	
VARIABLE O & M			5.67	
CONSUMABLE OPERATING COSTS,less Fuel (2000 Dollars)		$x1000	$/T H_2-yr	
Water		89	0.72	
Chemicals		761	6.17	
Other Consumables				
Waste Disposal		793	6.43	
TOTAL CONSUMABLE OPERATING COSTS		$1,643	13.32	
BY-PRODUCT CREDITS (2000 Dollars)		($8,739)	-70.84	
FUEL COST (2000 Dollars)		$19,337	156.76	

PRODUCTION COST SUMMARY	1st Year (2005 $) $/T H_2-yr	Levelized (Over Book Life $) $/T H_2-yr
Fixed O & M	77.71	77.71
Variable O & M	5.67	5.67
Consumables	13.32	13.32
By-product Credit/Penalty	-70.84	-70.84
Fuel	152.11	138.75
TOTAL PRODUCTION COST	177.97	164.61
LEVELIZED CARRYING CHARGES(Capital)		459.36
LEVELIZED(Over Book Life)COST/Ton of H_2		623.97
Equivalent $/MMBtu		5.11

Table 2-21

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY

TITLE/DEFINITION

Case:	Hydrogen Fuel Facility w/600C Hot Gas Desulfurization		
Plant Size:	430.8 H_2 TPD	HeatRate:	(Btu/kWh)
Primary/Secondary Fuel(type):	Pitts. #8	Cost:	1.00 ($/MMBtu)
Design/Construction:	2.5 (years)	BookLife:	20 (years)
TPC(Plant Cost) Year:	2000 (Jan.)	TPI Year:	2005 (Jan.)
Capacity Factor:	80 (%)		

CAPITAL INVESTMENT		$x1000	$x1000/$H_2$TPD
Process Capital & Facilities		290,035	673.2
Engineering(incl.C.M.,H.O.& Fee)		28,255	65.6
Process Contingency		8,792	20.4
Project Contingency		32,708	75.9
TOTAL PLANT COST(TPC)		$359,791	835.1
TOTAL CASH EXPENDED	$359,791		
AFDC	$23,070		
TOTAL PLANT INVESTMENT(TPI)		$382,861	888.6
Royalty Allowance			
Preproduction Costs		9,266	21.5
Inventory Capital		3,038	7.1
Initial Catalyst & Chemicals(w/equip.)			
Land Cost		150	0.3
TOTAL CAPITAL REQUIREMENT(TCR)		$395,315	917.6

OPERATING & MAINTENANCE COSTS (2000 Dollars)	$x1000	$x1000/$H_2$TPD
Operating Labor	3,574	8.3
Maintenance Labor	2,663	6.2
Maintenance Material	3,994	9.3
Administrative & Support Labor	1,559	3.6
TOTAL OPERATION & MAINTENANCE	$11,790	27.4
FIXED O & M		21.89
VARIABLE O & M		5.47

CONSUMABLE OPERATING COSTS,less Fuel (2000 Dollars)	$x1000	$/T H_2-yr
Water	96	0.77
Chemicals	765	6.08
Other Consumables		
Waste Disposal	793	6.30
TOTAL CONSUMABLE OPERATING COSTS	$1,654	13.15
BY-PRODUCT CREDITS (2000 Dollars)	($6,504)	-51.70
FUEL COST (2000 Dollars)	$19,337	153.71

PRODUCTION COST SUMMARY	1st Year (2005 $) $/T H_2-yr	Levelized (Over Book Life $) $/T H_2-yr	
Fixed O & M	74.97	74.97	0.6135793
Variable O & M	5.47	5.47	0.0447913
Consumables	13.15	13.15	0.107596
By-product Credit/Penalty	-51.70	-51.70	-0.423092
Fuel	149.15	136.05	1.113442
TOTAL PRODUCTION COST	191.05	177.95	1.4563165
LEVELIZED CARRYING CHARGES(Capital)		439.92	3.6003157
LEVELIZED(Over Book Life)COST/Ton of H_2		617.87	
Equivalent $/MMBtu		5.06	

Table 2-22

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY

TITLE/DEFINITION

Case:	Hydrogen Fuel Facility w/300C Hot Gas Desulfurization		
Plant Size:	438.8 H_2 TPD	HeatRate:	(Btu/kWh)
Primary/Secondary Fuel(type):	Pitts. #8	Cost:	1.00 ($/MMBtu)
Design/Construction:	2.5 (years)	BookLife:	20 (years)
TPC(Plant Cost) Year:	2000 (Jan.)	TPI Year:	2005 (Jan.)
Capacity Factor:	80 (%)		

CAPITAL INVESTMENT		$x1000	$x1000/$H_2$ TPD
Process Capital & Facilities		288,192	656.8
Engineering(incl.C.M.,H.O.& Fee)		27,974	63.8
Process Contingency		8,195	18.7
Project Contingency		32,436	73.9
TOTAL PLANT COST(TPC)		$356,797	813.2
TOTAL CASH EXPENDED	$356,797		
AFDC	$22,878		
TOTAL PLANT INVESTMENT(TPI)		$379,675	865.3
Royalty Allowance			
Preproduction Costs		9,053	20.6
Inventory Capital		3,024	6.9
Initial Catalyst & Chemicals(w/equip.)			
Land Cost		150	0.3
TOTAL CAPITAL REQUIREMENT(TCR)		$391,903	893.2

OPERATING & MAINTENANCE COSTS (2000 Dollars)	$x1000	$x1000/$H_2$ TPD
Operating Labor	3,574	8.1
Maintenance Labor	2,608	5.9
Maintenance Material	3,912	8.9
Administrative & Support Labor	1,545	3.5
TOTAL OPERATION & MAINTENANCE	$11,639	26.5
FIXED O & M		21.22
VARIABLE O & M		5.31

CONSUMABLE OPERATING COSTS, less Fuel (2000 Dollars)	$x1000	$/T H_2-yr
Water	195	1.52
Chemicals	808	6.31
Other Consumables		
Waste Disposal	793	6.19
TOTAL CONSUMABLE OPERATING COSTS	$1,797	14.02
BY-PRODUCT CREDITS (2000 Dollars)	($3,861)	-30.14
FUEL COST (2000 Dollars)	$19,337	150.93

PRODUCTION COST SUMMARY	1st Year (2005 $) $/T H_2-yr	Levelized (Over Book Life $) $/T H_2-yr
Fixed O & M	72.68	72.68
Variable O & M	5.31	5.31
Consumables	14.02	14.02
By-product Credit/Penalty	-30.14	-30.14
Fuel	146.46	133.59
TOTAL PRODUCTION COST	208.33	195.46
LEVELIZED CARRYING CHARGES(Capital)		428.24
LEVELIZED(Over Book Life)COST/Ton of H_2		623.70
Equivalent $/MMBtu		5.10

Table 2-23

CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY				
TITLE/DEFINITION				
Case:	Hydrogen Fuel Facility w/600C (80% H_2) Hot Gas Desulfurization			
Plant Size:	342.8 H_2 TPD		HeatRate:	(Btu/kWh)
Primary/Secondary Fuel(type):	Pitts. #8		Cost:	1.00 ($/MMBtu)
Design/Construction:	2.5 (years)		BookLife:	20 (years)
TPC(Plant Cost) Year:	2000 (Jan.)		TPI Year:	2005 (Jan.)
Capacity Factor:	80 (%)			
CAPITAL INVESTMENT			$x1000	$x1000/H2 TPD
Process Capital & Facilities			311,678	909.3
Engineering(incl.C.M.,H.O.& Fee)			30,006	87.5
Process Contingency			8,907	26.0
Project Contingency			35,059	102.3
TOTAL PLANT COST(TPC)			$385,650	1125.1
TOTAL CASH EXPENDED		$385,650		
AFDC		$24,728		
TOTAL PLANT INVESTMENT(TPI)			$410,378	1197.3
Royalty Allowance				
Preproduction Costs			9,724	28.4
Inventory Capital			3,132	9.1
Initial Catalyst & Chemicals(w/equip.)				
Land Cost			150	0.4
TOTAL CAPITAL REQUIREMENT(TCR)			$423,384	1235.2
OPERATING & MAINTENANCE COSTS (2000 Dollars)			$x1000	$x1000/H2 TPD
Operating Labor			3,574	10.4
Maintenance Labor			2,855	8.3
Maintenance Material			4,283	12.5
Administrative & Support Labor			1,607	4.7
TOTAL OPERATION & MAINTENANCE			$12,319	35.9
FIXED O & M				28.75
VARIABLE O & M				7.19
CONSUMABLE OPERATING COSTS,less Fuel (2000 Dollars)			$x1000	$/T H2-yr
Water			172	1.72
Chemicals			798	7.97
Other Consumables				
Waste Disposal			793	7.92
TOTAL CONSUMABLE OPERATING COSTS			$1,763	17.61
BY-PRODUCT CREDITS (2000 Dollars)			($15,100)	-150.88
FUEL COST (2000 Dollars)			$19,337	193.21

PRODUCTION COST SUMMARY	1st Year (2005 $) $/T H2-yr	Levelized (Over Book Life $) $/T H2-yr
Fixed O & M	98.47	98.47
Variable O & M	7.19	7.19
Consumables	17.61	17.61
By-product Credit/Penalty	-150.88	-150.88
Fuel	187.48	171.01
TOTAL PRODUCTION COST	159.87	143.40
LEVELIZED CARRYING CHARGES(Capital)		592.24
LEVELIZED(Over Book Life)COST/Ton of H2		735.64
Equivalent $/MMBtu		6.02

2.7 SUMMARY AND CONCLUSIONS

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

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

	1402°F Membrane (761°C)	1112°F Membrane (600°C) Baseline Case	572ºF Membrane (300°C)	1112°F Membrane with 80% Hydrogen Transport
HSD Exit Temperature	1402ºF (761ºC)	1112ºF (600ºC)	572ºF (300ºC)	1112ºF (600ºC)
Coal Feed	221,631 lb/h	221,631 lb/h	221,631 lb/h	221,631 lb/h
Oxygen Feed (95%)	231,218 lb/h	224,519 lb/h	218,657 lb/h	287,917 lb/h
Hydrogen Product Stream	35,205 lb/h	35,903 lb/h	36,564 lb/h	28,562 lb/h
CO_2 Product Stream	581,657 lb/h	582,566 lb/h	585,598 lb/h	583,220 lb/h
Sulfuric Acid Product	19,482 lb/h	19,482 lb/h	19,482 lb/h	19,482 lb/h
Gross Power Production	94 MW	84 MW	71 MW	131 MW
Auxiliary Power Requirement	76 MW	77 MW	76 MW	83 MW
Net Power Production	18 MW	7 MW	(6 MW)	48 MW
Effective Thermal Efficiency, HHV	80.2%	80.4%	80.3%	69.2%
Capital Cost, $1,000 (Year 2000)	$368,448	$359,791	$356,797	$385,650
Hydrogen Product Cost, $/MMBtu	$5.11	$5.06	$5.10	$6.02

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

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

- Hydrogen Flux – The hydrogen flux was based on the R&D goal of 0.1 std cc/minute/ cm^2/cm Hg P_{H_2} differential.
- Separation Factor – The separation determines the hydrogen purity and is high for hydrogen, increasing with higher temperatures. Even at 300°C the separation factor would be above 200.
- Operating Pressure and Temperature – It was assumed that a 950 psi pressure differential can be contained by the inorganic membrane. The operational goal for the membranes is

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

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

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

3. HYDROGEN FROM COAL AND NATURAL GAS-BASED PLANTS

Throughout 1999 and 2000, conceptual systems and cost analyses were developed by Parsons for a coal processing plant to produce hydrogen while recovering carbon dioxide (CO_2) for offsite processing or sequestration. This has been referred to as a hydrogen fuel plant.

This work has been reported in several venues including the June 1999 letter report[5], and U.S. Department of Energy (DOE) sponsored conferences.[9,10,11,12,13] This work has resulted in a baseline plant for production of hydrogen from coal utilizing the ORNL-developed inorganic membrane for separation of hydrogen from syngas.

The purpose of this section is to compare hydrogen cost from conventional methods, with and without CO_2 recovery, against the baseline hydrogen fuel plant.

3.1 CASES 1, 2, AND 3 – HYDROGEN FROM NATURAL GAS WITHOUT AND WITH CO_2 RECOVERY

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

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

3.1.1 NATURAL GAS CONDITIONING

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

3.1.2 NATURAL GAS REFORMER/BOILER

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

$$CH_4 + H_2O = CO + 3H_2$$

$$CO + H_2O = CO_2 + H_2$$

$$CO + 3H_2 = CH_4 + H_2O$$

The reforming reaction is strongly endothermic, with energy supplied by firing the reformer on the outside of the catalyst tubes with recycled syngas from the hydrogen purification process. The metallurgy of the tubes usually limits the reaction temperature to 1400-1700°F. The flue gas path of the fired reformer is integrated with additional boiler surfaces to produce about 700,000 lb/hour steam. Of this, about 450,000 lb/hour is superheated to 450 psia and 750°F, to be added to the incoming natural gas. Additional steam from the boiler is either shipped offsite or used within the plant for regeneration of CO_2 from the acid gas removal process.

The CO-shift and methanation reactions quickly reach equilibrium at all points in the catalyst bed. High steam-to-carbon ratio, low pressure, and high temperature favor the equilibrium composition of the reformed gas. The process generally employs a steam-to-carbon ratio of 3 to 5 at a process temperature of around 1500°F and pressures up to 500 psig to convert more than 70 percent of hydrocarbon to oxides of carbon at the outlet of the reformer so as to ensure a minimum concentration of CH_4 in the product gas. After the reformer, the process gas mixture of CO and H_2 passes through a heat recovery step and is fed into a water-gas shift reactor to produce additional H_2.

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

Table 3-1
Composition of Synthetic Gas

Component	Volume %
CH_4	8
CO	7
CO_2	6
H_2	44
H_2O	35
Total	100

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