### 4.2.4 SULFURIC ACID PLANT

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

### 4.2.5 HYDROGEN SEPARATION/CONVENTIONAL TURBINE EXPANDER

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

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

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

**Table 4-5**
**Plant Water Balance**

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

### 4.2.6 EFFECTIVE THERMAL EFFICIENCY

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

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

$$\text{ETE} = \frac{33{,}336 \text{ lb } H_2\text{/h x } 61{,}095 \text{ Btu/lb} + 13{,}900 \text{ kW x } 3{,}414 \text{ Btu/kWh}}{283{,}833 \text{ lb coal/h x } 8{,}630 \text{ Btu/lb} + 31{,}537 \text{ lb sawdust/h x } 5{,}165 \text{ Btu/lb}}$$

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

## 4.3 COST ESTIMATE

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

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

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

**CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY**

**TITLE/DEFINITION**

| | | | | |
|---|---|---|---|---|
| Case: | Hydrogen Fuel Facility (Wy/Bio Blend) w/600C Hot Gas Desulfurization | | | |
| Plant Size: | 400.0 $H_2$ TPD | HeatRate: | | (Btu/kWh) |
| Primary/Secondary Fuel(type): | Wyodak/Biom | Cost: | 0.65 | ($/MMBtu) |
| Design/Construction: | 2.5 (years) | BookLife: | 20 | (years) |
| TPC(Plant Cost) Year: | 2000 (Jan.) | TPI Year: | 2005 | (Jan.) |
| Capacity Factor: | 80 (%) | | | |

| CAPITAL INVESTMENT | | $x1000 | $x1000/$H_2$TPD |
|---|---|---|---|
| Process Capital & Facilities | | 297,459 | 743.6 |
| Engineering(incl.C.M.,H.O.& Fee) | | 28,793 | 72.0 |
| Process Contingency | | 6,169 | 15.4 |
| Project Contingency | | 33,242 | 83.1 |
| TOTAL PLANT COST(TPC) | | $365,662 | 914.1 |
| TOTAL CASH EXPENDED | $365,662 | | |
| AFDC | $23,446 | | |
| TOTAL PLANT INVESTMENT(TPI) | | $389,109 | 972.7 |
| Royalty Allowance | | | |
| Preproduction Costs | | 9,229 | 23.1 |
| Inventory Capital | | 2,304 | 5.8 |
| Initial Catalyst & Chemicals(w/equip.) | | | |
| Land Cost | | 150 | 0.4 |
| TOTAL CAPITAL REQUIREMENT(TCR) | | $400,792 | 1001.9 |

| OPERATING & MAINTENANCE COSTS (2000 Dollars) | $x1000 | $x1000/$H_2$TPD |
|---|---|---|
| Operating Labor | 3,871 | 9.7 |
| Maintenance Labor | 2,604 | 6.5 |
| Maintenance Material | 3,906 | 9.8 |
| Administrative & Support Labor | 1,619 | 4.0 |
| TOTAL OPERATION & MAINTENANCE | $12,000 | 30.0 |
| FIXED O & M | | 24.00 |
| VARIABLE O & M | | 6.00 |
| **CONSUMABLE OPERATING COSTS,less Fuel (2000 Dollars)** | **$x1000** | **$/T $H_2$-yr** |
| Water | 88 | 0.75 |
| Chemicals | 1,066 | 9.13 |
| Other Consumables | | |
| Waste Disposal | 603 | 5.17 |
| TOTAL CONSUMABLE OPERATING COSTS | $1,757 | 15.04 |
| **BY-PRODUCT CREDITS (2000 Dollars)** | ($4,251) | -36.39 |
| **FUEL COST (2000 Dollars)** | $11,900 | 101.87 |

| | 1st Year (2005 $) | Levelized (Over Book Life $) | |
|---|---|---|---|
| **PRODUCTION COST SUMMARY** | $/T $H_2$-yr | $/T $H_2$-yr | |
| Fixed O & M | 82.18 | 82.18 | 0.6725696 |
| Variable O & M | 6.00 | 6.00 | 0.0490976 |
| Consumables | 15.04 | 15.04 | 0.1230982 |
| By-product Credit/Penalty | -36.39 | -36.39 | -0.29785 |
| Fuel | 98.85 | 90.17 | 0.7379375 |
| TOTAL PRODUCTION COST | 165.68 | 157.00 | 1.2848532 |
| **LEVELIZED CARRYING CHARGES(Capital)** | | 480.35 | 3.931157 |
| **LEVELIZED(Over Book Life)COST/Ton of $H_2$** | | 637.34 | |
| **Equivalent $/MMBtu** | | 5.22 | |

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

**CAPITAL INVESTMENT & REVENUE REQUIREMENT SUMMARY**

**TITLE/DEFINITION**

| | | | |
|---|---|---|---|
| Case: | Hydrogen Fuel Facility (Wy/Bio Blend) w/600C Hot Gas Desulfurization | | |
| Plant Size: | 400.0 $H_2$ TPD | HeatRate: | (Btu/kWh) |
| Primary/Secondary Fuel(type): | Wyodak/Biom | Cost: | 0.50 ($/MMBtu) |
| Design/Construction: | 2.5 (years) | BookLife: | 20 (years) |
| TPC(Plant Cost) Year: | 2000 (Jan.) | TPI Year: | 2005 (Jan.) |
| Capacity Factor: | 80 (%) | | |

| CAPITAL INVESTMENT | | $x1000 | $x1000/H₂TPD |
|---|---|---|---|
| Process Capital & Facilities | | 297,459 | 743.6 |
| Engineering(incl.C.M.,H.O.& Fee) | | 28,793 | 72.0 |
| Process Contingency | | 6,169 | 15.4 |
| Project Contingency | | 33,242 | 83.1 |
| TOTAL PLANT COST(TPC) | | $365,662 | 914.1 |
| TOTAL CASH EXPENDED | $365,662 | | |
| AFDC | $23,446 | | |
| TOTAL PLANT INVESTMENT(TPI) | | $389,109 | 972.7 |
| Royalty Allowance | | | |
| Preproduction Costs | | 9,160 | 22.9 |
| Inventory Capital | | 2,026 | 5.1 |
| Initial Catalyst & Chemicals(w/equip.) | | | |
| Land Cost | | 150 | 0.4 |
| TOTAL CAPITAL REQUIREMENT(TCR) | | $400,445 | 1001.0 |

| OPERATING & MAINTENANCE COSTS (2000 Dollars) | $x1000 | $x1000/H₂TPD |
|---|---|---|
| Operating Labor | 3,871 | 9.7 |
| Maintenance Labor | 2,604 | 6.5 |
| Maintenance Material | 3,906 | 9.8 |
| Administrative & Support Labor | 1,619 | 4.0 |
| TOTAL OPERATION & MAINTENANCE | $12,000 | 30.0 |
| FIXED O & M | | 24.00 |
| VARIABLE O & M | | 6.00 |

| CONSUMABLE OPERATING COSTS,less Fuel (2000 Dollars) | $x1000 | $/T H₂-yr |
|---|---|---|
| Water | 88 | 0.75 |
| Chemicals | 1,066 | 9.13 |
| Other Consumables | | |
| Waste Disposal | 603 | 5.17 |
| TOTAL CONSUMABLE OPERATING COSTS | $1,757 | 15.04 |
| **BY-PRODUCT CREDITS (2000 Dollars)** | ($4,251) | -36.39 |
| **FUEL COST (2000 Dollars)** | $9,154 | 78.36 |

| PRODUCTION COST SUMMARY | 1st Year (2005 $) $/T H₂-yr | Levelized (Over Book Life $) $/T H₂-yr | |
|---|---|---|---|
| Fixed O & M | 82.18 | 82.18 | 0.6725696 |
| Variable O & M | 6.00 | 6.00 | 0.0490976 |
| Consumables | 15.04 | 15.04 | 0.1230982 |
| By-product Credit/Penalty | -36.39 | -36.39 | -0.29785 |
| Fuel | 76.04 | 69.36 | 0.5676442 |
| TOTAL PRODUCTION COST | 142.87 | 136.19 | 1.1145599 |
| LEVELIZED CARRYING CHARGES(Capital) | | 479.93 | 3.9277537 |
| LEVELIZED(Over Book Life)COST/Ton of H₂ | | 616.12 | |
| Equivalent $/MMBtu | | 5.04 | |

## 4.4 SUMMARY AND CONCLUSIONS

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

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

| | 90% Wyodak<br>10% Biomass | Baseline Case<br>Pittsburgh No. 8<br>600°C Membrane |
|---|---|---|
| Coal Feed | 283,833 lb/h | 221,631 lb/h |
| Biomass Feed | 31,537 lb/h | N/A |
| Oxygen Feed (95%) to Gasifier | 186,650 lb/h | 165,818 lb/h |
| Oxygen Feed to Retentate Combustor | 25,300 lb/h | 58,701 lb/h |
| Water to Prepare Feed Slurry | 114,009 lb/h | 94,025 lb/h |
| Hydrogen Product Stream | 33,337 lb/h | 35,903 lb/h |
| $CO_2$ Product Stream | 575,923 lb/h | 582,566 lb/h |
| Sulfuric Acid Product | 5,057 lb/h | 19,482 lb/h |
| Gross Power Production | | |
| Turbine Expander | 55 MW | 84 MW |
| Steam Turbine | 28 MW | N/A |
| Auxiliary Power Requirement | (69 MW) | (77 MW) |
| Net Power Production | 14 MW | 7 MW |
| Net Plant Water Makeup | 100,979 lb/h | 198,150 lb/h |
| Effective Thermal Efficiency (ETE), HHV | 79.8% | 80.4% |
| Capital Cost, $1,000 | $365,662 | $359,791 |
| Hydrogen Product Cost, $/MMBtu | $5.22<br>($0.65 Feedstock)<br>$5.04<br>($0.50 Feedstock) | $5.06 |

Total plant costs are roughly equal, resulting from a combination of increased and decreased equipment requirements. The cost adjustments to the hydrogen plant due to the change over to the Wyodak/biomass blend are reflected in increased feedstock handling, increased oxygen plant size due to the higher water content (and associated increase in $CO_2$ content), and the need for a steam turbine which produces 28 MW from excess low-pressure steam. The capital costs were lower in sulfur control areas because of the low-sulfur feedstock, resulting in only 61 tpd sulfuric acid production from the blend versus 234 tpd from bituminous coal. This resulted in a lowering of byproduct credits.

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

The amount of hydrogen produced from the Wyodak/biomass blend is lowered by about 7 percent, primarily due to the higher level of $CO_2$ produced in the gasifier. This resulted in a lowered amount of reactive syngas ($H_2$ and CO) available for hydrogen production.

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

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

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

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

| | 100% Wyodak | 100% Pittsburgh No. 8 |
|---|---|---|
| Coal Feed | 283,833 lb/h | 221,631 lb/h |
| Oxygen Feed (95%) | 220.986 lb/h | 252,369 lb/h |
| Water to Prepare Feed Slurry | 109,249 lb/h | 94,025 lb/h |
| Hydrogen Product Stream | 29,221 lb/h | 34,004 lb/h |
| Limestone Sorbent to FGD | 10,583 lb/h | 25,188 lb/h |
| $CO_2$ Product Stream | 538,410 lb/h | 603,324 lb/h |
| Net Plant Water Makeup | 99,960 lb/h | 188,878 lb/h |
| | | |
| Gross Power Production | | |
| ATS Turbine Expander | 102 MW | 120 MW |
| Steam Turbine | 20 MW | N/A |
| Auxiliary Power Requirement | (67 MW) | (78 MW) |
| Net Power Production | 55 MW | 42 MW |
| Effective Thermal Efficiency (ETE), HHV | 80.4% | 79.8% |
| | | |
| Capital Cost, \$1,000 | \$313,597 | \$306,605 |
| Hydrogen Product Cost, \$/MMBtu | \$3.91 (\$0.67/MMBtu Feedstock) | \$4.05 (\$1.00/MMBtu Feedstock) |

# 5. REFERENCES


[1] "Orimulsion: $CO_2$ Capture in Power Generation: Hydrogen Production; and Methanol Production," IEA Greenhouse Gas R&D Programme, Report No. PH2/11, May 1997.

[2] Final Report, Bellefonte Completion Project Fossil Repowering Phase I Screening Study, TVA Contract TV-96889V, 1997.

[3] Nitrogen Fixation by Ammonia Synthesis for $CO_2$ Sequestration, U.S. Department of Energy, Federal Energy Technology Center, Pittsburgh, Pennsylvania, 1997.

[4] Conference call, Parsons Power and Oak Ridge National Laboratory (Rod Judkins, Doug Fain), September 2, 1997.

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

[6] Fain, Douglas E., Research and Development on Hydrogen Separation Technology with Inorganic Membranes, Pittsburgh Coal Conference, 1999.

[7] Rutkowski, M.D. and DeLallo, M.R., Decarbonized Fuel Plants Utilizing Inorganic Membranes for Hydrogen Separation, 12th Annual Conference of Fossil Energy Materials, May 1998.

[8] An Attractive Option for $CO_2$ Control in IGCC Systems: Water Gas Shift with Integrated $H_2/CO_2$ Separation, Netherlands Energy Research Foundation.

[9] Badin, J.S., DeLallo, M.R., Klett, M.G., Rutkowski, M.D., Temchin, J.R., *Decarbonized Fuel from Coal, Plant Review*, Pittsburgh Coal Conference 1999.

[10] Badin, J.S., DeLallo, M.R., Rutkowski, M.D., et al., *Decarbonized Fuel Production Facility – A Technical Strategy for Coal in the Next Century*, 1999 Gasification Technologies Conference, October 1999.

[11] Badin, J.S., DeLallo, M.R., Rutkowski, M.D., Temchin, J.R., *Decarbonized Fuel Production Facility, A Technical Strategy for Coal in the Next Century,* Power-Gen International Conference, December 1998.

[12] Fain, Douglas E., *Research and Development on Hydrogen Separation Technology with Inorganic Membranes*, Pittsburgh Coal Conference, 1999.

[13] Badin, J.S., DeLallo, M.R., Rutkowski, M.D., Temchin, J.R., *Coal Processing Plant for Hydrogen Production with $CO_2$ Capture*, 25th International Technical Conference on Coal Utilization & Fuel Systems, March 2000, Clearwater, Florida.