Conversion of Forest Residues to a Methane-Rich Gas

Detailed Economic Feasibility Study

March 1986

**Prepared by Battelle Columbus Laboratories
for Pacific Northwest Laboratory
under Contract DE-AC06-76RLO 1830
with the U.S. Department of Energy**

**Pacific Northwest Laboratory
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CONVERSION OF FOREST RESIDUES
TO A METHANE-RICH GAS

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Pacific Northwest Laboratory
Richland, Washington 99352

SUMMARY

An economic evaluation of the application of the multi-solid fluid reactor design to wood gasification was completed in March 1979. The processing options examined in this evaluation were:

- Plant capacity
- Production of a high-Btu (1006 Btu/SCF HHV) gas versus an intermediate-Btu gas (379 Btu/SCF HHV)
- Operating pressure.

Installed equipment costs for a 500-ton-per-day (dry wood) plant operating at 100 psig and designed to produce an intermediate-Btu gas and a 5000 ton-per-day plant operating at 500 psig and designed to produce a high-Btu gas were estimated. The total plant investment costs developed for these two cases were then used to predict the estimates of total plant investments for 4 additional cases.

The total capital requirement for a 500 TPD intermediate-Btu (379 Btu/SCF) plant (Case 1 - described in following tabulation) was estimated to be \$10.2 million and the calculated average gas cost using a utility financing method is \$2.83/MM Btu. An increase in the plant capacity to 1000 TPD (Case 2) results in a reduction in the gas cost to \$2.34/MM Btu. The total capital requirement for this case is \$16.7 million. A significant reduction in the total capital requirement can be made with the elimination of the gas quenching or scrubbing operation to produce a gas for on-site use. Elimination of the gas quenching operation reduces the total capital requirement of a 500 TPD plant (Case 3) to \$8.8 million. The average gas cost is \$2.59/MM Btu. The total capital requirement for a 1000 TPD plant with no gas quench (Case 4) is \$14.2 million and the average gas cost is \$2.09/MM Btu.

The plant designed to produce a high-Btu gas (1060 Btu/SCF) has a capacity of 5000 TPD and an operating pressure of 500 psig (Case 5). The total capital requirement is \$84.3 million and the estimated average gas cost is \$2.45/MM Btu. Some savings can be gained by operating the gasification and combustion sections at 100 psig and compressing the raw gas to

500 psig prior to the shifting and methanation operations (Case 6). The total requirement in this instance is \$80.7 million and the average gas cost is \$2.41/MM Btu. These results are summarized below.

<u>Summary of Plant Designs Included in Study</u>			
<u>Item</u>	<u>Description</u>	<u>Total Capital Requirements, Millions of dollars</u>	<u>Average Gas Cost Utility Financing, \$/MM Btu</u>
Case 1	500 TPD, Intermediate-Btu, 100 psig; quench	10.2	2.83
Case 2	1000 TPD, Intermediate-Btu, 100 psig; quench	16.7	2.34
Case 3	500 TPD, Intermediate-Btu, 100 psig; no quench	8.8	2.69
Case 4	1000 TPD, Intermediate-Btu, 100 psig; no quench	14.2	2.09
Case 5	5000 TPD, High-Btu; 500 psig	84.3	2.45
Case 6	5000 TPD, High-Btu; 100 psig with intermediate compression	80.7	2.41

As expected, the average gas costs are significantly lower at higher capacities. At a capacity of 5000 TPD, a high-Btu gas can be produced for \$0.28/MM Btu less than an intermediate-Btu gas at a capacity of 500 TPD (Case 6 vs Case 3). However, the advantage of the small capacity plant is the relatively low investment required.

The reduction in the average gas cost at lower pressure (Case 6 vs Case 5) is primarily due to the lower equipment cost required for the feed system. At design pressures higher than 150 psig, lockhoppers are required in the feed system and this factor doubles the cost of the feed system.

In addition to the gas quenching and feeding system, the other capital intensive areas in the plant where significant improvements in design may reduce the costs are ash disposal, air compression, electrical generation, acid gas removal, and methanation.

The product gas costs were initially evaluated with a delivered wood chip price of \$10/green ton (40 percent moisture) or \$1.00/MM Btu.

Because this price may fluctuate due to factors external to the gasification process, each case was evaluated to determine the effects of doubling the wood chip price to \$20/green ton. The average gas costs for each case calculated with the increased wood price are listed below.

<u>Item</u>	Average Gas Cost (Utility financing) <u>\$/MM Btu (Wood chip price = \$20/green ton)</u>
Case 1	4.02
Case 2	3.53
Case 3	3.92
Case 4	3.29
Case 5	3.87
Case 6	3.83

Thus, an increase in wood price of \$1.00/MM Btu of wood feed to the process results in an increase in the average gas cost of \$1.42 to \$1.19/MM Btu.

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INTRODUCTION

The primary objective of the current program is to improve the technology and economics of the conversion of forest residue materials to a clean, transportable gas. To meet this objective, a novel reactor system developed by Battelle, the Multi-Solid Fluid-Bed (MSFB) has been proposed as a means of exploiting the high reactivity of biomass materials during gasification. The MSFB reactor design permits high gas velocities, thus allows higher throughputs, and therefore, takes advantage of the high reactivity of wood.

The purpose of this task is to determine the major capital intensive areas in the application of the MSFB reactor to wood gasification. The processing options which are factored into the evaluation are the cost of upgrading the product gas to SNG, the gasification operating pressure, and the plant size. In addition, the sensitivity of the estimated gas cost to the price of wood is evaluated. In order to intensify the effect of these parameters in the economics of the process only the capital cost requirements for the battery limits plant were estimated. Offsite costs such as site development, land, utility construction, etc. were not included. These costs were felt to be site specific and are, generally, estimated in this type of evaluation as factors of the battery limits cost. In addition, holdup time, and therefore equipment, between plant operations has been minimized or eliminated. Treatment requirements for emissions have also been minimized. Therefore, the costs presented in this evaluation do not represent actual construction costs of grass roots facilities but do represent comparison of alternative processing.

It was originally intended to examine the effect of catalytic treatment on process economics. However, the reactivity improvements achieved by aqueous slurry impregnation of calcium oxide and other catalytic agents have not been substantial enough to justify the cost (in both plant energy requirements and capital investment) of catalytic treatment. The most effective catalytic system will be that employing wood ash itself. More experimental work will be required to generate the data required to establish the cost/benefits of wood ash catalysis on wood gasification.

In the meantime, the non-catalyzed gasification process economics reported here will establish a basis for comparison as well as provide for the more rapid estimation of subsequent process configurations involving various wood ash catalysis options.

Intermediate-Btu Gasification

A wood gasification plant designed to produce an intermediate-Btu gas was evaluated to determine the effects of plant size and quenching the product gas on gas costs. Detailed energy and material balances for a nominal 500 ton-per-day plant with an operating pressure of 100 psig and product gas quenching (Case 1) were completed as a basis for specifying equipment and determining installed costs. The investment costs which were derived for Case 1 were then used to predict a gas cost for: (1) a 1000 ton-per-day plant with an operating pressure of 100 psig and product gas quenching (Case 2); (2) a 500 ton-per-day plant with an operating pressure of 100 psig and no product gas quenching (Case 3); and, (3) a 1000 ton-per-day plant with an operating pressure of 100 psig and no product gas quenching (Case 4).

Conceptual Design of Case 1

The conceptual design of the process evaluated in Case 1 is shown in Figure 1 and the energy and material balances are listed in Table 1. The assumptions made in the derivation of these balances are listed in Table A-1 in the Appendix. The results of the gasification model are listed in Table A-2.

Wood chips are delivered to the plant suitable for direct feeding to the gasifier. The nominal size of the chips is 1-1/2" x 1-1/2" x 1/4"-1/8" and, therefore, feed preparation equipment has not been charged to the plant. The feed system includes unloading hoppers, weighing and feeding conveyors, and storage and pressurizing hoppers. A detailed description of the feed system is given in the Appendix.

FIGURE 1. PROCESS FLOW DIAGRAM FOR INTERMEDIATE-BTU GAS FROM 500 TPD (DRY FEED) PLANT - CASE 1

TABLE 1. MATERIAL BALANCE FOR INTERMEDIATE-BTU GAS FROM 500 TPD PLANT - CASE 1
(OPERATING PRESSURE - 100 psig)

	① Wet Feed (40% H ₂ O)	② 10% H ₂ O in Feed	③ Steam to Gasifier	④ Steam to Dryer	⑤ Make Gas	⑥ Gas from Dryer	⑦ Sand Circ.	⑧ Sand Circ.	⑨ Gas to Scrubber	⑩ Product Gas	⑪ Water to Scrubber	⑫ Water Treatment
lb moles/hr												
CH ₄					273.1	273.1			273.1	273.1		
C ₂ H ₄					3.5	3.5			3.5	3.5		
C ₂ H ₆					38.9	38.9			38.9	38.9		
CO					674.1	674.1			674.1	674.1		
CO ₂					312.8	312.8			312.8	312.8		
H ₂					721.6	721.6			721.6	721.6		
N ₂					3.0	3.0			3.0	3.0		
H ₂ S					0.7	0.7			0.7	0.7		
H ₂ O					254.5	1,540.5			1,540.5	83.7		
lbs/hr												
Sand							539,400	539,400				
Wood	41,667	41,667										
Ash												
Carbon												
H ₂ O	27,778	4,630	4,881	10,411							5,300	31,500
Temperature (°F)	80	350	1,000	1,000	1,382	350	1,787	1,382	350	150	150	250
Pressure (psia)	14.7	114.7	300	300	114.7	114.7	114.7	114.7	100	90	14.7	90

TABLE 1. (continued)

	(13) Steam Generated	(14) Air to Compressor	(15) Compressed Air	(16) Air to Combustor	(17) Flue Gas to Cyclone	(18) Ash to Quench	(19) Water to Quench	(20) Sludge Disposal	(21) Flue Gas to Preheater	(22) Gas to Turbine	(23) Flue Gas to Stack	(24) Char to Combustor
lb moles/hr												
CH ₄												
C ₂ H ₄												
C ₂ H ₆												
CO												
CO ₂					379.7				379.7	379.7	379.7	
H ₂		1,573.9	1,573.9	1,573.9	1,573.9				1,573.9	1,573.9	1,573.9	
N ₂		1,573.9	1,573.9	1,573.9	1,573.9				1,573.9	1,573.9	1,573.9	
H ₂ S												
H ₂ O		24.2	24.2	24.2	24.2				24.2	24.2	24.2	
O ₂		418.4	418.4	418.4	38.0				38.0	38.0	38.0	
SO ₂					0.7				0.7	0.7	0.7	
lbs/hr												
Sand												
Wood												
Ash						2,200		2,220				2,208
Carbon												2,485
H ₂ O	15,490						4,735	4,735				
Sulfur in Char												21
Temperature (°F)	1,000	80	307	1,737	1,787	1,787	1,787	100	1,787	500	86	1,382
Pressure (psia)	300	14.7	121	121	115	115	115	14.7	112	110	15	114.7

Wood chips are fed at 40 percent moisture to a fluid bed dryer which operates at 350 F. Hot make gas from the gasifier is used to partially dry the feed material. Steam tubes are incorporated into the dryer to supply additional evaporation capacity because the sensible heat of the make gas is insufficient to reduce the moisture content to the required level. The feed material enters the gasifier at 10 percent moisture. The gasifier operates at 750 C (1382 F) and 540,000 lb/hr of sand is circulated from the combustor at 1787 F to provide additional heat. Steam is added at the rate of 4881 lb/hr. The energy and material balances for the system were completed using the basis of 78 percent carbon conversion in the gasifier.

Product gas leaves the dryer at 350 F and entrained solids are removed by a series of cyclones. The solids are returned to the gasifier. The product gas is scrubbed in a water quench tank to remove tars and oils and is then stored for transmission as an intermediate-Btu fuel gas. The heating value of the gas is 379 Btu/SCF and the gas production rate is 7288 Btu per pound of dry wood fed to the gasifier.

Char from the gasifier is burned in the combustor to generate process steam and to supply heat to the gasifier through sand circulation. Sand and ash particles entrained in the hot (1787 F) flue gas are removed in cyclones. The solids are cooled in a quench tank and removed to a disposal pond. The hot flue gas is used to preheat the compressed air and then expanded in a gas turbine to generate the power required for the compressor.

Raw water requirements for the plant are 1061 lb/hr. The water distribution requirements for the plant are illustrated in Figure 2.

Total plant electrical needs are 2418 hp, of this 1997 hp are supplied by the gas turbine and steam turbines (Figure 3). The additional power is supplied by a generator which can also be used for start-up. The additional quantity of wood required to produce this electricity is

The plant thermal efficiency is 86.3 percent. It is assumed that there is no excess steam or other by-products produced in the process.

<u>IN</u>		<u>OUT</u>	
	<u>lb/hr</u>		<u>lb/hr</u>
Scrubber	31,600	Ash Disposal	4,735
Raw Water Makeup	1,061	Scrubber	5,300
	<hr/>	Evaporative Losses	5,882 (15%)
	32,561	BFW	4,881
		C.T. Blowdown	11,763 (10%)
			<hr/>
			32,561

FIGURE 2. PLANT WATER SYSTEM CASE 1

FIGURE 3. ELECTRICAL BALANCE - CASE 1

Cost Evaluation of Case 1

The total installed costs for Case 1 are listed in Table 2. The contractor's fee is 11.1 percent of the total installed cost and the contingencies are 15 percent of the subtotal. The estimated total plant investment is \$6.8 million (Table 3).

The annual operating costs include raw materials, purchased water, labor (operating, maintenance, and supervision), administration, supplies, and local taxes and insurance. These items are summarized in Table 4. Total maintenance is equivalent to 6 percent of the total plant investment and maintenance labor is equal to 60 percent of the total maintenance costs. Maintenance supplies account for the remaining 40 percent of the total maintenance costs. The estimated annual operating costs for Case 1 are \$5.5 million. The total capital requirement is \$10.2 million (Table 5).

Based on these figures, the gas costs were estimated assuming a wood cost of \$10/green ton (40 percent moisture). Assuming private investor financing, the average gas cost is \$3.24/MM Btu. Assuming utility financing, the average cost is \$2.93/MM Btu and the first-year gas cost is \$3.18/MM Btu (Table 6). The assumptions used in these calculations are listed in Table A-3 in the Appendix.

Conceptual Design of Case 2

The plant evaluated in Case 2 incorporates the same process flow-sheet as that in Case 1 (see Figure 1) but is designed to process 1000 tons per day of dry wood. Therefore, the energy and material balances represented in Case 1 also apply to Case 2 with increased capacity.

Cost Evaluation of Case 2

The total installed costs which were derived for Case 1 were used to predict the installed costs for Case 2 (1000 tons per day capacity plant) by the exponent method. The following procedure was used to determine the total plant investment costs of Case 2.

TABLE 2. CASE 1 - INSTALLED COST BREAKDOWN

Plant Section	Units	Design Basis	Unit Size or Capacity	Equipment Cost	Installation Factor	Installed Cost	References
1) Feed Storage	2	20 days storage	833,000 ft ³	41,500	1.0	41,500	Battelle
2) Feed System	1	Wood chips - $1\frac{1}{2}" \times 1\frac{1}{2}" \times \frac{1}{4}"$, 100 psig	833 wet ton/day				
Unloading hoppers		100 psig	" "	2,600	1.0	2,600	Battelle
Weighing conveyor		" "	" "	126,000	1.0	126,000	Jeffrey Manuf.
Feed conveyor		" "	" "	130,000	1.0	130,000	Kanyer, Inc.
Storage hoppers		" "	" "	700,000	1.0	700,000	Petrocarb, Inc.
3) Gasification							
Gasifier	1	200 psig, skin temp. 414 F, superficial gas vel = 15 ft/sec	4.2' ID x 40' H	27,700	4.16	151,100	Guthrie
Cyclones	2	4333 ACFM		---	---	75,000	Battelle
Dryer	1	200 psig, 350 F skin temp., superficial gas vel = 5 ft/sec	4.8' ID x 35' H	29,000	4.16	120,600	Guthrie
4) Gas Quench							
Quench tank	1	5 minute retention time	18' ID x 60' H	171,300	4.16	712,500	Guthrie
Recycle water cooler	1	Shell and tube	1,900 ft ²	19,000	3.17	60,100	Guthrie
5) Combustion							
Combustor	1	200 psig, 6720 lb/hr char, 15,490 lb/hr steam generated	5.5' ID x 35' H	55,600	4.16	231,300	Guthrie
Cyclones	2	6775 ACFM		---	---	75,000	Battelle
6) Ash Disposal							
Screens		Vibrating double	53 ft ²	---	---	193,000	Guthrie
Hopper		Refractory lined, 1 hr retention time	2' ID x 6' H	2,800	4.16	11,600	Guthrie
Quench tank		20 minutes retention time	5' ID x 15' H	9,375	4.16	39,000	Guthrie
Cooler		Shell and tube	430 ft ²	8,076	3.17	25,600	Guthrie
Disposal pond	1	2200 lb/hr ash	---	---	---	234,000	Guthrie
7) Air Compressor/Gas Expansion							
Air preheater	1	Shell and tube	41,280 ft ²	158,100	3.17	501,200	Guthrie
Gas turbine	1	17 lbs/sec	1,954 hp	---	---	200,000	Battelle
8) Water Treatment	1	7 day storage	656,000 gal	---	---	290,700	Guthrie
9) Electrical Generation	1	---	319 kW	510,000	1.462	745,600	Guthrie
10) Cooling Tower	1	15 degree temp. rise	2,356 GPM	39,100	1.7	66,500	Guthrie

TABLE 3. TOTAL PLANT INVESTMENT - CASE 1

Plant Section	\$
Feed Storage	41,000
Feed System	958,600
Gasification	346,100
Gas Quench	772,600
Combustion	306,300
Ash Disposal	503,200
Compression	701,200
Water Treatment	290,700
Electrical Generation	745,600
Cooling Tower	66,500
Pumps	64,000
General Utilities (10% of Installed Cost)	532,900
Total Installed Cost	5,329,200
Contractor's Fee (11.1%)	591,500
Subtotal	5,920,700
Contingency (15%)	888,100
TOTAL PLANT INVESTMENT	6,808,800

TABLE 4. CASE 1 - ANNUAL OPERATING COSTS

Raw Materials ($840.8 \text{ T/D} \times \$10/\text{ton} \times 365 \text{ days} \times 0.9$)	2,762,000
Purchased Water ($1 \times 10^6 \text{ gal} \times \$0.70/\text{M gal}$)	700
Labor	
Process Operating Labor ($12.5 \text{ man/shift} \times 8760 \text{ man-hr/yr} \times \$8.00/\text{man-hr}$)	876,000
Maintenance Labor (Total maintenance $\times 60\%$)	245,100
Supervision (20% operating labor & maintenance labor)	224,200
Administration (60% of labor)	807,200
Supplies	
Operating (30% of process operating labor)	262,800
Maintenance (Total maintenance $\times 40\%$)	163,400
Local Taxes and Insurance (2.7% of T.P.I.)	183,800
TOTAL ANNUAL OPERATING COSTS	5,525,200

TABLE 5. CASE 1 - TOTAL CAPITAL REQUIREMENT

Total Plant Investment	\$6,808,800
Allowance for Funds Used During Construction (T.P.I. x 3 yrs x 0.09)	1,838,400
Startup Costs (20% total gross annual operating costs)	1,105,000
Working Capital	
1) Raw materials (14-day inventory wood - \$10/green ton)	117,700
2) Materials and supplies (0.9% T.P.I.)	61,300
3) Net receivables (1/24 annual gas revenue)	311,200
TOTAL CAPITAL REQUIREMENT	\$10,242,400

TABLE 6. CASE 1 - GAS COSTS

Utility Financing	
First year cost	\$3.10/MM Btu
Average cost	\$2.83/MM Btu
Private Investor Financing	
Gas cost	\$3.12/MM Btu

$$\frac{\text{Total Plant Investment of Case 1}}{\left(\frac{\text{Capacity of Case 2}}{\text{Capacity of Case 1}}\right)^{0.7}} = \text{Total Plant Investment of Case 2}$$

The annual operating costs are \$9.2 million and the total capital requirement is \$16.7 million (see Tables 7 and 8).

Assuming a wood cost of \$10/green ton and private investor financing, the gas cost is \$2.58/MM Btu. Using the utility financing method, the first year cost is \$2.56/MM Btu and the average gas cost is \$2.34/MM Btu (Table 9).

Conceptual Design of Case 3

Case 3 represents a plant which is designed to produce an intermediate-Btu gas for on-site use. The process flowsheet is identical to that of Case 1 without the gas quenching operation (see Figure 4). The plant capacity is 500 ton per day and the operating pressure is 100 psig. The material and energy balances for this process are listed in Table 10.

Cost Evaluation of Case 3

The total installed costs for the plant evaluated in Case 3 were estimated using the costs developed for Case 1; the gas quench section was eliminated from the estimate.

The total plant investment costs for Case 3 are estimated to be \$5.7 million (Table 11). The annual operating costs are \$5.4 million (Table 12) and the total capital requirement is \$8.8 million (Table 13).

With private investor financing the average gas cost is \$2.95/MM Btu. Assuming the utility financing method, the first year gas cost is \$2.93/MM Btu and the average cost is \$2.69/MM Btu (Table 14).

TABLE 7. CASE 2 - ANNUAL OPERATING COSTS

Raw Materials (1681.6 T/D x \$10/ton x 365 days x 0.9)	\$5,524,100
Purchased Water (2 x 10 ⁶ gal x \$0.70/M gal)	1,400
Labor	
Process Operating Labor (15 men/shift x 8760 man-hr/yr x \$8.00/man-hr)	1,051,200
Maintenance Labor (Total maintenance x 60%)	398,200
Supervision (20% operating labor & Maintenance labor)	289,900
Administration (60% of labor)	1,044,000
Supplies	
Operating (30% of process operating labor)	135,400
Maintenance (Total maintenance x 40%)	265,400
Local Taxes and Insurance (2.7% of T.P.I.)	298,600
TOTAL ANNUAL OPERATING COSTS	\$9,188,200

TABLE 8. CASE 2 - TOTAL CAPITAL REQUIREMENT

Total Plant Investment	\$11,060,000
Allowance for Funds Used During Construction (T.P.I. x 3 yrs x 0.09)	2,986,000
Startup Costs (20% total gross annual operating costs)	1,837,600
Working Capital	
1) Raw materials (14-day inventory wood - \$10/green ton)	233,300
2) Materials and supplies (0.9% T.P.I.)	100,000
3) Net receivables (1/24 annual gas revenue)	514,700
TOTAL CAPITAL REQUIREMENT	\$16,731,600

TABLE 9. CASE 2 - GAS COSTS

Utility Financing	
First year cost	\$2.56/MM Btu
Average cost	\$2.34/MM Btu
Private Investor Financing	
Gas cost	\$2.58/MM Btu

FIGURE 4. PROCESS FLOW DIAGRAM FOR INTERMEDIATE-BTU GAS FROM 500 TPD (DRY FEED) PLANT - CASE 3

TABLE 10. MATERIAL BALANCE FOR INTERMEDIATE-BTU GAS FROM 500 TPD PLANT - CASE 3
(OPERATING PRESSURE - 100 psig)

	① Wet Feed (40% H ₂ O)	② 10% H ₂ O in Feed	③ Steam to Gasifier	④ Steam to Dryer	⑤ Make Gas	⑥ Gas from Dryer	⑦ Sand Circ.	⑧ Sand Circ.	⑨ Product Gas
lb moles/hr									
CH ₄					273.1	273.1			273.1
C ₂ H ₄					3.5	3.5			3.5
C ₂ H ₆					38.9	38.9			38.9
CO					674.1	674.1			674.1
CO ₂					312.8	312.8			312.8
H ₂					721.6	721.6			721.6
N ₂					3.0	3.0			3.0
H ₂ S					0.7	0.7			0.7
H ₂ O					254.5	1,540.5			1,540.5
lbs/hr									
Sand							539,400	539,400	
Wood	41,667	41,667							
Ash									
Carbon									
H ₂ O	27,778	4,630	4,881	10,411					
Temperature (F)	80	350	1,000	1,000	1,382	350	1,787	1,382	350
Pressure (psia)	14.7	114.7	300	300	114.7	114.7	114.7	114.7	100

TABLE 10. (continued)

	(10) Steam Generated	(11) Air to Compressor	(12) Compressed Air	(13) Air to Combustor	(14) Flue Gas to Cyclone	(15) Ash to Quench	(16) Water to Quench	(17) Sludge Disposal	(18) Flue Gas to Preheater	(19) Gas to Turbine	(20) Flue Gas to Stack	(21) Char to Combustor
lb moles/hr												
CH ₄												
C ₂ H ₄												
C ₂ H ₆												
CO												
CO ₂					379.7				379.7	379.7	379.7	
H ₂		1,573.9	1,573.9	1,573.9	1,573.9				1,573.9	1,573.9	1,573.9	
N ₂		1,573.9	1,573.9	1,573.9	1,573.9				1,573.9	1,573.9	1,573.9	
H ₂ S												
H ₂ O		24.2	24.2	24.2	24.2				24.2	24.2	24.2	
O ₂		418.4	418.4	418.4	38.0				38.0	38.0	38.0	
SO ₂					0.7				0.7	0.7	0.7	
lb/hr												
Sand												
Wood												
Ash						2,200		2,220				2,208
Carbon												2,485
H ₂ O	15,490						4,735	4,735				
Sulfur in Char												21
Temperature (°F)	1,000	80	307	1,737	1,787	1,787	1,787	100	1,787	500	86	1,382
Pressure (psia)	300	14.7	121	121	115	115	115	14.7	112	110	15	114.7

TABLE 11. TOTAL PLANT INVESTMENT - CASE 3

Plant Section	\$
Feed Storage	41,500
Feed System	958,500
Gasification	346,100
Combustion	306,300
Ash Disposal	503,200
Compression	701,200
Water Treatment	290,700
Electrical Generation	745,600
Cooling Tower	66,500
Pumps	64,000
General Utilities	447,100
Installed Costs	4,470,700
Contractor's Fee	496,200
Subtotal	4,966,900
Contingency	745,000
TOTAL PLANT INVESTMENT	5,711,900

TABLE 12. CASE 3 - ANNUAL OPERATING COSTS

Raw Materials	2,762,000
Purchased Water	700
Labor	
Process Operating Labor	876,000
Maintenance Labor	205,600
Supervision	216,300
Administration	778,700
Supplies	
Operating	262,800
Maintenance	137,100
Local Taxes and Insurance	154,200
TOTAL ANNUAL OPERATING COSTS \$5,393,400	

TABLE 13. CASE 3 - TOTAL CAPITAL REQUIREMENT

Total Plant Investment	\$5,711,900
Allowance for Funds Used During Construction	1,544,200
Startup Costs	1,078,700
Working Capital	
1) Raw materials	117,700
2) Materials and supplies	51,400
3) New receivables	294,300
TOTAL CAPITAL REQUIREMENT	\$8,798,700

TABLE 14. CASE 3 - GAS COSTS

Utility Financing	
First year cost	\$2.93/MM Btu
Average cost	\$2.59/MM Btu
Private Investor Financing	
Gas Cost	\$2.95/MM Btu

Conceptual Design of Case 4

The plant evaluated in this example is similar to Case 3 in that the product gas is not quenched and the sensible heat in the gas is conserved. The capacity of the plant evaluated in Case 4 is 1000 tons per day versus a capacity of 500 tons per day in Case 3.

Cost Evaluation of Case 4

The total plant investment for this example is estimated by using a scale factor with the Total Plant Investment (T.P.I.) estimated for Case 3. Using this method, the T.P.I. for Case 4 is \$9.3 million. The total annual operating costs are \$8.3 million (Table 15) and the total capital investment is \$14.2 million (Table 16). With utility financing, the first year gas cost is \$2.28 and the average gas cost is \$2.09/MM Btu. The average gas cost with private investor financing is \$2.30 (Table 17).

High-Btu Gasification

The gas upgrading operations in a high-Btu gasification process represent a significant portion of the total capital investment. At higher capacities, the economy of scale somewhat reduces the effect on gas costs of high-investment operations. Therefore, a 5000 ton-per-day plant was evaluated in Case 5 to determine the costs of the production of a high-Btu gas relative to that of an intermediate-Btu gas. In addition, the effect of operating pressure was evaluated by comparing a system with an operating pressure of 500 psig (Case 5) to a system with an operating pressure of 100 psig in the gasification and combustion sections and intermediate gas compression to 500 psig (Case 6).

TABLE 15. CASE 4 - TOTAL ANNUAL OPERATING COSTS

Raw Materials	\$5,524,100
Purchased Water	1,400
Labor	
Process Operating Labor	1,051,200
Maintenance	334,100
Supervision	277,100
Administration	332,500
Supplies	
Operating	315,400
Maintenance	222,700
Local Taxes and Insurance	250,500
TOTAL ANNUAL OPERATING COSTS	\$8,309,000

TABLE 16. CASE 4 - TOTAL CAPITAL REQUIREMENT

Total Plant Investment	\$9,279,000
Allowance for Funds Used During Construction	2,505,300
Startup Costs	1,551,800
Working Capital	
1) Raw materials	235,400
2) Materials and supplies	83,500
3) Net receivables	458,900
TOTAL CAPITAL REQUIREMENT	\$14,223,900

TABLE 17. CASE 4 - GAS COSTS

Utility Financing	
First year cost	\$2.28/MM Btu
Average cost	\$2.09/MM Btu
Private Investor Financing	
Gas cost	\$2.30/MM Btu

Conceptual Design of Case 5

The plant evaluated in Case 5 was designed to process 5000 TPD of dry wood at 500 psig and to produce 6094 Btu of SNG per pound of dry wood (Figure 5). The gasification, combustion, and air compression sections of the plant are similar to those described for Case 1. Pressurizing lock hoppers have been added to the feed system used in Case 1 to achieve a feed pressure of 500 psig.

Cyclones are used to remove entrained solids from make gas leaving the fluid bed dryer. Tars and oils which may be present in the gas are removed in a sand filter.

Prior to methanation, the hydrogen to carbon monoxide ratio is increased to 3.06/1 in the shift reactor. Seventy-four percent of the total gas stream is passed through the reactor and 50 percent of the carbon monoxide is reacted to achieve the required ratio. (The operating temperature is very close to the dew point of the gas--490 F vs. 486 F--and, to prevent condensation, it may be necessary to preheat the gas stream prior to the shift conversion unit.) After shift conversion, carbon dioxide and hydrogen sulfide are removed from the gas. A Benfield acid gas removal system was included in this evaluation because operating and cost data were available; however, another system may be more suitable to this application.

Following acid gas removal, the gas stream is preheated to 550 F before entering the zinc oxide beds which remove any traces of sulfur compounds. The gas stream is then split into three streams which are introduced into each of the three methanation reactors. The reaction temperature is controlled by a combination of hot gas recycle and interstage cooling for heat recovery. The methanation product is cooled to 125 F and compressed to 1000 psig for transmission in existing pipeline distribution systems.

The heating value of the product gas is 1060 Btu per SCF of gas.

Water which remains in the gas stream after the methanation step is removed with a glycol drying system. The material balance for this system design is listed in Table 18. The results of the gasification model used to derive this balance are shown in Table A-4 in the Appendix.

FIGURE 5. PROCESS FLOWSHEET FOR GASIFICATION OF FOREST RESIDUES AND CONVERSION TO SNG - CASE 5

TABLE 18. MATERIAL BALANCE FOR 5000 TPD HIGH BTU GASIFICATION PLANT - CASE 5

	1	2	3	4	5	6	7	8	9	10	11	12
	Wet Feed	Dried Wood	Wood to Gasifier	Make Gas	Gas to Cyclone	Circ. Sand	Circ. Sand	Air to Compressor	Steam to Dryer	Compressed Air	Compressed Air	Flue Gas to Cyclone
lb mole/hr												
CH ₄				3,901.1	3,901.1							
C ₂ H ₄				50.3	50.3							
C ₂ H ₆				556.3	556.3							
CO				2,661.7	2,661.7							
CO ₂				5,069.7	5,069.7							3,960.5
H ₂				4,152.2	4,152.2							
N ₂				29.7	29.7			16,413.2		16,413.2	16,413.2	16,413.2
H ₂ S				6.5	6.5							
H ₂ O				6,011.6	18,871.7			252.3		252.3	252.3	252.3
O ₂								4,363.0		4,363.0	4,363.0	396.6
SO ₂												5.9
lbs/hr												
Wood	416,667	416,667										
Sand						2,410,000	2,410,000					
Ash												
Carbon												
H ₂ O	277,778	46,296	108,292						170,946			
Temperature (°F)	80	500	1,000	1,382	500	1,787	1,382	80	1,000			
Pressure (psia)	14.7	515	1,025	515	515	515	515	14.7	1,025			

TABLE 18. (continued)

	13	14	15	16	17	18	19	20	21	22	23	24
	Flue Gas to Preheater	Ash	Water to Ash Disposal	Ash Disposal	Steam Produced	Flue Gas to Turbine		Gas to Sand Filter				
lb mole/hr												
CH ₄								3,901.1	2,882.9	2,882.9	2,882.9	1,018.2
C ₂ H ₄								50.3	37.2	37.2	37.2	13.1
C ₂ H ₆								556.3	411.1	411.1	411.1	145.2
CO								2,661.7	1,966.9	1,966.9	983.5	694.8
CO ₂	3,960.5					3,960.5	3,960.5	5,069.7	3,746.5	3,746.5	4,729.9	1,323.2
H ₂								4,152.2	3,068.5	3,068.5	4,051.9	1,083.7
N ₂	16,413.2					16,413.2	16,413.2	29.2	21.9	21.9	21.9	7.8
H ₂ S								6.5	4.8	4.8	4.8	1.7
H ₂ O	252.3					252.3	252.3	18,871.7	13,946.2	13,946.2	13,962.8	4,925.5
O ₂	396.6					396.6	396.6					
SO ₂	5.9					5.9	5.9					
lbs/hr												
Wood												
Sand												
Ash		19,875		19,875								
Carbon												
H ₂ O			41,918	41,918	806,258							
Temperature (F)	1,787	1,787	80	100	1,000	1,150	255	500	490	550	700	490
Pressure (psia)	500	500	14.7	14.7	1,025	500	20	510	500	500	500	500

TABLE 18. (continued)

	25	26	27	28	29	30	31	32	33	34	35	36
<hr/>												
lb mole/hr												
CH ₄	2,882.9	3,901.1	3,901.1	3,901.1	3,901.1	1,300.4	5,578.2	11,156.4	5,578.2	5,578.2	5,578.2	5,578.2
C ₂ H ₄	37.2	50.3	50.3	50.3	50.3	16.8	50.4	100.8	50.4	50.4	50.4	50.4
C ₂ H ₆	411.1	556.3	556.3	556.3	556.3	185.4	556.3	1,112.4	556.3	556.3	556.3	556.3
CO	983.5	1,678.3	1,678.3	1,678.3	1,578.3	559.4	--	--	--	--	--	--
CO ₂	4,729.9	6,053.1	6,053.1	2.3	2.3	0.8	2.3	4.8	2.3	2.3	2.3	2.3
H ₂	4,051.9	5,135.6	5,135.6	5,129.6	5,129.6	1,709.9	94.8	189.6	94.8	94.8	94.8	94.8
N ₂	21.9	29.7	29.7	29.7	29.7	9.9	29.7	59.4	29.7	29.7	29.7	29.7
H ₂ S	4.8	6.5	6.5	--	--	--	--	--	--	--	--	--
H ₂ O	12,962.8	17,888.3	1,038.5	51.2	51.2	17.1	1,729.5	3,459.0	31.9	31.9	7.4	--
O ₂												
SO ₂												
<hr/>												
lbs/hr												
Wood												
Sand												
Ash												
Carbon												
H ₂ O												
Temperature (F)	650	600	250	122	550	550	640	650	125	264	100	100
Pressure (psia)	500	500	500	495	490	490	480	490	480	1,015	1,015	1,010
<hr/>												

With the assumptions used in this evaluation, the quantity of steam which is produced in the plant is more than that required for operation. Therefore, it was assumed that excess steam would be used to generate electricity which then can be sold as a by-product.

The water distribution system for the plant is shown in Figure 6. Because of the large quantity of water entering with the wood feed, additional water intake to the plant is not required.

The electrical distribution system for the plant is shown in Figure 7. The gas expansion turbine produces 47,000 hp for air compression. Excess steam from the process is used to produce 228,708 hp of which 199,740 hp is a by-product to the plant.

The cold gas efficiency is 72.8 percent and the plant thermal efficiency is 87.4 percent.

Cost Evaluation for Case 5

The detailed installed equipment costs for Case 5 are listed in Table 19. The total plant investment is \$56.7 million and is itemized in Table 20. The total gross annual operating costs are \$38.8 million (Table 21). A net value of \$17.6 million is received from the surplus electrical power which is generated by the steam turbines. This value is calculated by assuming the power can be sold to a utility or nearby user for \$0.015/kWh. The credit for by-product electrical power is not included in the gas cost calculations because capital investments costs have not been included in the overall estimate for power distribution equipment. In addition, in an actual plant the by-product power would be minimized and would not have an effect on the average gas cost.

The total capital requirement is \$84.3 million (Table 22). Assuming a raw wood cost of \$10/ton and with private investor financing, the gas cost is \$2.75/MM Btu; with utility financing the first year gas cost is \$2.71/MM Btu and the average gas cost is \$2.45/MM Btu (Table 23).

FIGURE 6. CASE 5 - PLANT WATER SYSTEM

FIGURE 7. CASE 5 - ELECTRICAL BALANCE

TABLE 19. CASE 5 - INSTALLED COST BREAKDOWN

Plant Section	Units	Design Basis	Unit Size or Capacity	Equipment Cost, \$	Installation Factor	Installed Cost, \$	References
1) Feed Storage	1	20 days storage	$8.3 \times 10^6 \text{ ft}^3$	208,325	1.0	208,300	Battelle
2) Feed System		Wood chips - $1\frac{1}{2}'' \times 1\frac{1}{2}'' \times \frac{1}{4}''$	8333 wet tons/day				
Unloading hoppers				15,600	1.0	16,000	Battelle
Weighing conveyor				252,000	1.0	252,000	Jeffrey Manuf.
Feed conveyor				780,000	1.0	780,000	Kanya, Inc.
Storage hoppers				4,200,000	1.0	4,200,000	Petrocarb
Lock hoppers				4,200,000	1.0	4,200,000	Petrocarb
3) Gasification		Superficial gas velocity 12.2 ft/sec, 513 F skin temp.					
Gasifier	1	600 psi	6' ID x 45' H	99,000	4.16	412,000	Guthrie
Cyclones	2	15,919 ACFM		115,250	1.2	138,300	
Dryer	1	Superficial gas velocity 5 ft/sec, 600 psi	8.5' ID x 50' H	142,800	4.16	594,000	Guthrie
Sand Filter	1	Space velocity = 10,000 ft ³	8' ID x 20' H	61,200	4.16	254,600	Guthrie
4) Shift Conversion							
Heat exchanger #1	1	Shell and tube	14,976 ft ²	70,000	3.17	221,800	Guthrie
Shift reactor	1	700 P, 50% conversion	14' ID x 18' H	122,400	4.16	509,200	Guthrie
5) Acid Gas Removal							
Heat exchanger #2	1	Shell and tube	17,570 ft ²	104,100	3.17	330,100	Guthrie
Acid gas removal unit	1	Benfield unit	17,410 moles/hr moisture-free gas	---	--	5,428,000	Benfield
6) Methanation							
Zinc oxide bed	1	Space velocity = 10,000	6' ID x 15' H	33,150	4.16	137,900	Guthrie
Methanation reactors	3	Space velocity = 10,000	Spherical 12' ID	53,600	4.16	222,800	Guthrie
Recycle Compressor	1	76 lbs/sec	10,990 hp	---	--	1,142,100	Battelle
Condensor	1	Shell and tube	141,200 ft ²	520,200	3.17	1,648,900	Guthrie
Heat exchanger #3	1	Shell and tube	10,400 ft ²	79,100	3.17	250,800	Guthrie
Heat exchanger #4	1	Shell and tube	149 ft ²	4,600	3.17	14,700	Guthrie
Heat exchanger #5	1	Shell and tube	770 ft ²	11,900	3.17	37,800	Guthrie
Heat exchanger #6	1	Shell and tube	1,212 ft ²	16,940	3.17	53,700	Guthrie
Heat exchanger #7	1	Shell and tube	141,756 ft ²	334,100	3.17	1,059,000	Guthrie

TABLE 19. (continued)

Plant Section	Units	Design Basis	Unit Size or Capacity	Equipment Cost, \$	Installation Factor	Installed Cost, \$	References
7) Final Compression and Gas Drying							
Compressor	1	30 lb/sec	2,420 hp	---	--	454,000	Battelle
Heat exchanger #8	1	Shell and tube	8,187 ft ²	67,300	3.17	213,400	Guthrie
Glycol drying unit	1	133 lb/hr water removed	--	---	--	584,200	C.F. Braun
8) Combustion							
Combustor	1	600 psi, skin temp. 486 F, superficial gas vel = 5 ft/sec	10'2" ID x 50' H	173,200	4.16	720,700	Guthrie
Cyclones	2	15,777 ACFM		115,250	1.2	138,300	
9) Ash Disposal							
Screens	1	Vibrating double	120 ft ³ /hr ash	1,311,200	1.34	1,757,000	Guthrie
Hopper	1	600 psi, 60 min retention time	5' 10 x 9' H	17,900	4.16	74,300	Guthrie
Quench tank	1	20 minutes retention time	5' ID x 7' H	10,200	4.16	42,400	Guthrie
Cooler	1	Shell and tube	3,806 ft ²	33,200	3.17	105,200	Guthrie
Disposal pond	1	19,875 lb/hr	--	---	--	2,114,200	Battelle
10) Air Compression/Gas Expansion							
Air preheater	1	Shell and tube	134,835 ft ²	499,800	3.17	1,584,400	Guthrie
Gas turbine		168 lb/sec	47,000 hp	---	--	2,515,000	Battelle
11) Water Treatment							
	1	1 day storage, 14,000,000 gallons	--	---	--	1,045,200	Guthrie
12) Electrical Generation							
	1	Steam turbines 95% efficiency	13,400 kW	3,457,000	1.462	5,054,000	Guthrie
13) Cooling Tower							
	1	15 degree temp. rise	8,748 CPM	357,000	1.97	703,300	Guthrie

TABLE 20. TOTAL PLANT INVESTMENT - SNG

Feed Storage	\$208,300
Feed System	9,448,000
Gasifier	412,000
Gasifier Cyclones	138,300
Dryer	594,000
Sand Filter	254,600
Shift Reactor	509,200
H.E. #1	221,800
H.E. #2	330,100
Acid Gas Removal System	5,428,000
H.E. #3	250,700
Zinc Oxide Bed	137,900
Methanators	700,200
H.E. #4	14,700
H.E. #5	37,800
H.E. #6	53,700
H.E. #7	1,059,000
Recycle Compressor	1,142,100
Condensor	1,648,900
Final Compressor	454,500
H.E. #8	213,400
Glycol Drying System	584,200
Combustor	720,700
Combustor Cyclones	138,300
Ash Disposal System	1,981,400
Ash Disposal Pond	2,114,200
Air Preheater	1,584,400
Compressor/Turbine	2,515,000
Cooling Tower	703,300
Water Treatment	1,045,200
Electrical Generation	5,054,000
Pumps	187,000
General Utilities (10% of Installed Cost)	4,431,700
Installed Cost	\$44,316,600
Contractor's Fee (15%)	4,874,800
Subtotal	49,191,400
Contingency (15)	7,378,700
TOTAL PLANT INVESTMENT	\$56,570,100

TABLE 21. ANNUAL OPERATING COSTS -SNG

Raw Materials (8333.3 T/D x \$10/ton x 365 days x 0.9)	\$27,375,000
Labor	
Process Operating Labor (30 men x 8760 man-hr/yr x \$8/man hr)	2,102,400
Maintenance Labor (Total maintenance x 0.6)	2,036,500
Supervision (20% operating and maintenance labor)	827,800
Administration (60% of labor)	2,980,000
Supplies	
Operating (30% of operating labor)	630,700
Maintenance (Total maintenance x 0.4)	1,357,700
Local Taxes and Insurance	1,527,400
TOTAL ANNUAL OPERATING COSTS	\$38,837,500

TABLE 22. TOTAL CAPITAL REQUIREMENT - SNG

Total Plant Investment	\$56,570,100
Allowance for Funds Used During Construction (T.P.I. x 3 yrs x 0.09)	15,273,900
Startup Costs (20% total annual gross operating costs)	7,767,500
Working Capital	
1) Raw materials (14-day inventory)	1,166,700
2) Materials and supplies (0.9% T.P.I.)	509,100
3) Net receivables (1/24 annual gas revenue)	2,294,000
(by-product revenue from surplus electricity)	733,800
TOTAL CAPITAL REQUIREMENT	\$84,315,100

TABLE 23. CASE 5 - GAS COSTS

Utility Financing	
First year gas cost	\$2.71/MM Btu
Average gas cost	\$2.45/MM Btu
Private Investor Financing	
Gas cost	\$2.75/MM Btu

Conceptual Design of Case 6

The operating pressure of the gasification and combustion sections in the plant evaluation in Case 6 is 100 psig. The gas is compressed to 500 psig prior to the shift conversion operation (see Figure 8). The gas cleanup sections of the plant are identical to those described in Case 5. The plant capacity is 5000 tons per day. The material balance for this system is listed in Table 24 and the electrical requirements are illustrated in Figure 9. The power requirements are increased due to the added compression requirements.

Cost Evaluation for Case 6

The total plant investment for Case 6 was estimated by scaling the costs of the gasification and combustion section estimated in Case 1 which has a capacity of 500 TPD to 5000 TPD. A seven-tenths scale factor was used. The gas cleanup section of the plant is identical to that in Case 5. Intermediate compression of the gas from 100 psig to 500 psig is required between the gasification and gas cleanup operations.

The total plant investment in Case 6 is estimated to be \$53.8 million which is approximately 5 percent less than the total plant investment for Case 5 (Table 25). The lower operating pressure (100 psig vs. 500 psig) for the feeding, gasification and combustion sections is reflected in the lower estimated plant investment.

The estimated annual operating costs for Case 6 are \$38.5 million (Table 26). The total capital requirement is \$80.7 million (Table 27). Assuming utility financing, the first year gas cost is \$2.66 and the average gas cost is \$2.41/MM Btu. Based on private investor financing the average gas cost is \$2.69 (Table 28).

FIGURE 8. PROCESS FLOWSHEET FOR GASIFICATION OF FOREST RESIDUES AND CONVERSION TO SNG - CASE 6

TABLE 24. MATERIAL BALANCE FOR 5000 HIGH BTU GASIFICATION PLANT - CASE 6

	1	2	3	4	5	6	7	8	9	10	11	12
	Wet Feed	Dried Wood	Wood to Gasifier	Make Gas	Gas to Cyclone	Circ. Sand	Circ. Sand	Air to Compressor	Steam to Dryer	Compressed Air	Compressed Air	Flue Gas to Cyclone
lb mole/hr												
CH ₄				3,901.1	3,901.1							
C ₂ H ₄				50.3	50.3							
C ₂ H ₆				556.3	556.3							
CO				2,661.7	2,661.7							
CO ₂				3,069.7	3,069.7							3,960.3
H ₂				4,152.2	4,152.2							
N ₂				29.7	29.7			16,413.2		16,413.2	16,413.2	16,413.2
H ₂ S				6.3	6.3							
H ₂ O				6,011.6	18,871.7			252.3		252.3	252.3	252.3
O ₂								4,363.0		4,363.0	4,363.0	396.6
SO ₂												5.9
lbm/hr												
Wood	416,667	416,667										
Sand						2,410,000	2,410,000					
Ash												
Carbon												
H ₂ O	277,778	46,296	108,292						170,946			
Temperature (°F)	80	500	1,000	1,382	500	1,787	1,382	80	1,000	420	1,737	1,787
Pressure (psia)	14.7	115	1,025	115	115	115	115	14.7	1,025	120	120	115

TABLE 24. (continued)

	13	14	15	16	17	18	19	20a	20	21	22	23	24
	Flue Gas to Preheater	Ash	Water to Ash Disposal	Ash Disposal	Steam Produced	Flue Gas to Turbine		Gas to Compressor	Gas to Sand Filter				
lb mole/hr													
CH ₄								3,901.1		2,882.9	2,882.9	2,882.9	1,018.2
C ₂ H ₄								50.3		37.2	37.2	37.2	13.1
C ₂ H ₆								556.3		411.1	411.1	411.1	145.2
CO								2,661.7		1,966.9	1,966.9	983.5	694.8
CO ₂	3,960.5					3,960.5	3,960.5	5,069.7		3,746.5	3,746.5	4,729.9	1,323.2
H ₂								4,152.2		3,068.5	3,068.5	4,051.9	1,083.7
N ₂	16,413.2					16,413.2	16,413.2	29.2		21.9	21.9	21.9	7.8
H ₂ S								6.5		4.8	4.8	4.8	1.7
H ₂ O	252.3					252.3	252.3	18,871.7		13,946.2	13,946.2	13,962.8	4,925.5
O ₂	396.6					396.6	396.6						
SO ₂	5.9					5.9	5.9						
lbs/hr													
Wood													
Sand													
Ash		19,875		19,875									
Carbon													
H ₂ O			41,918	41,918	806,258								
Temperature (°F)	1,787	1,787	80	100	1,000	1,750	255	500	500	490	550	700	490
Pressure (psia)	115	115	14.7	14.7	1,025	115	20	115	510	500	500	500	500

TABLE 24. (continued)

	25	26	27	28	29	30	31	32	33	34	35	36
lb mole/hr												
CH ₄	2,882.9	3,901.1	3,901.1	3,901.1	3,901.1	1,300.4	5,578.2	11,156.4	5,578.2	5,578.2	5,578.2	5,578.2
C ₂ H ₄	37.2	50.3	50.3	50.3	50.3	16.8	50.4	100.8	50.4	50.4	50.4	50.4
C ₂ H ₆	411.1	556.3	556.3	556.3	556.3	185.4	556.3	1,112.4	556.3	556.3	556.3	556.3
CO	983.5	1,678.3	1,678.3	1,678.3	1,578.3	559.4	—	—	—	—	—	—
CO ₂	4,729.9	6,053.1	6,053.1	2.3	2.3	0.8	2.3	4.8	2.3	2.3	2.3	2.3
H ₂	4,051.9	5,135.6	5,135.6	3,129.6	5,129.6	1,709.9	94.8	189.6	94.8	94.8	94.8	94.8
N ₂	21.9	29.7	29.7	29.7	29.7	9.9	29.7	59.4	29.7	29.7	29.7	29.7
H ₂ S	4.8	6.5	6.5	—	—	—	—	—	—	—	—	—
H ₂ O	12,962.8	17,888.3	1,038.5	51.2	51.2	17.1	1,729.5	3,459.0	31.9	31.9	7.4	—
O ₂												
SO ₂												
lbs/hr												
Wood												
Sand												
Ash												
Carbon												
H ₂ O												
Temperature (°F)	650	600	250	122	550	550	640	650	125	264	100	100
Pressure (psia)	500	500	500	495	490	490	480	490	480	1,015	1,015	1,010

FIGURE 9. CASE 6 - ELECTRICAL BALANCE

TABLE 25. CASE 6 - TOTAL PLANT INVESTMENT

Plant Section	\$
Feed Storage Through Combustion	20,165,800
Sand Filter Through Final Compression	13,040,800
Intermediate Compression	4,709,100
General Utilities	4,212,900
Total Installed Cost	42,128,600
Contractor's fee (11.1%)	4,676,300
Subtotal	46,804,900
Contingency (15%)	7,020,700
TOTAL PLANT INVESTMENT	53,825,600

TABLE 26. ANNUAL OPERATING COSTS - CASE 6

Raw Materials	\$27,375,000
Labor	
Process Operating Labor	2,102,400
Maintenance Labor	1,937,700
Supervision	808,000
Administration	2,908,900
Supplies	
Operating	630,700
Maintenance	1,291,800
Local Taxes and Insurance	1,453,300
TOTAL ANNUAL OPERATING COSTS	\$38,507,800

TABLE 27. CASE 6 - TOTAL CAPITAL REQUIREMENT

Total Plant Investment	\$53,825,600
Allowance for Funds Used During Construction	14,532,900
Startup Costs	7,701,600
Working Capital	
1) Raw materials	1,166,700
2) Materials and supplies	484,400
3) Net receivables	2,273,900
(By-product revenue)	733,800
TOTAL CAPITAL REQUIREMENT	\$80,688,900

TABLE 28. CASE 6 - GAS COSTS

Utility Financing	
First year cost	\$2.66/MM Btu
Average cost	\$2.41/MM Btu
Private Investor Financing	
Gas cost	\$2.69/MM Btu

Sensitivity of Gas Costs to Raw Wood Price

The price of wood chips has an effect on the final gas cost through both the operating costs and the total capital requirement. The actual price of wood chips is dependent on site-specific and economic factors external to the plant cost. Therefore, the effect on gas costs of doubling the price of wood chips was determined by calculating the average gas cost (using the utility financing method) for each case. In the initial cost evaluations, the wood price was assumed to be \$10/ton of green wood which is equivalent to \$1/MM Btu of wood. This price was increased to \$20/ton of green wood (\$2.00/MM Btu) and the average gas costs were recalculated. The average gas costs for Cases 1-6 are compared in Table 29. An increase in the wood price results in an increase in the average gas cost of \$1.19 to \$1.42 per million Btu of gas produced.

The annual operating costs and total capital requirements for Cases 1-6 from which the average gas costs were derived at a wood price of \$20/ton are listed in Tables A-5 through A-6.

CONCLUSIONS

- (1) As expected, due to the economy of scale, increasing the plant size reduces the average gas cost. A 5000 TPD capacity plant can produce a high-Btu gas for approximately the same gas cost as a small 500 TPD plant which produces an intermediate-Btu gas. Based on utility financing and a wood chip price of \$10/green ton, the average cost of a high-Btu gas produced by a 5000 TPD plant with an operating pressure of 500 psig is \$2.45 per million Btu of product gas. The 500 TPD capacity plant with a quenching operation can produce an intermediate-Btu gas for \$2.83 per million Btu of product gas. However, the advantage of a small plant is the relatively low capital investment required.

TABLE 29. COMPARISON OF AVERAGE GAS COSTS OF CASES 1-6
WITH VARIABLE RAW WOOD PRICES (UTILITY
FINANCING METHOD)

Case	Description of Plant	Raw Wood, price	
		\$20/ton	\$10/ton
		Gas cost \$/MM Btu	Gas Cost, \$/MM Btu
1	(Intermediate-Btu, 500 TPD, Quench)	4.02	2.83
2	(Intermediate-Btu, 1000 TPD, Quench)	3.53	2.34
3	(Intermediate-Btu, 500 TPD, No Quench)	3.92	2.69
4	(Intermediate-Btu, 1000 TPD, No Quench)	3.29	2.09
5	(High-Btu, 5000 TPD, 500 psig)	3.87	2.45
6	(High-Btu, 5000 TPD, 100 psig)	3.83	2.41

- (2) The quenching operation in an intermediate-Btu gasification plant represents 15 percent of the total plant investment and is an operation that can be completely eliminated if the product gas can be used on-site. Elimination of the quenching operation reduces the average gas cost to \$2.69 per million Btu compared to \$2.83 per million Btu.
- (3) At design pressures higher than 150 psig, lockhoppers are required in the dry wood feeding system. This factor almost doubles the cost of the feed system which represents 20 percent of the total installed costs for a plant that operates at 500 psig and produces a high-Btu gas. By reducing the operating pressure to 100 psig, the lockhoppers can be eliminated from the feed system. Compression of the gas prior to the gas upgrading operations does add to the investment costs; however, the overall investment costs are 5 percent less at lower pressure and the average gas costs are \$2.41 per million Btu vs \$2.45 per million Btu at an operating pressure of 500 psig.
- (4) In addition to the quenching and feeding system, other capital intensive areas in the process where significant improvements may be made are:
- o Ash disposal - equivalent to 9 percent of the total installed costs (T.I.C.)
 - o Air compression/gas expansion - equivalent to 13 percent of T.I.C. in an intermediate-Btu process; over 5 percent in a high-Btu process.
 - o Electrical generation - 11-13 percent of T.I.C.
 - o Acid gas removal (high-Btu production) - 12 percent of T.I.C.
 - o Methanation (high-Btu production) - 7 percent of T.I.C.

- (5) An increase in the price of wood chips fed to the plant affects the final gas cost through both increased operating costs and total capital requirement. Wood chip prices vary according to location and generally fall between \$10 to \$20 per ton of green wood. At a wood price of \$10/ton, the average gas cost (using the utility financing method) of a 500 TPD plant producing an intermediate-Btu gas with a quench system is \$2.83 per million Btu. At a wood price of \$20/ton, the gas cost is increased to \$4.02 per million Btu. In the specific cases evaluated, an increase of \$10/ton in the wood price resulted in increases in the final gas costs between \$1.42 per million Btu and \$1.19 per million Btu. Therefore in order to reduce the final gas cost, the feed preparation costs must be minimized.

APPENDIX A

Detailed Description of the
Wood Chip Feed System to the Gasifiers

The feed system designed for the wood gasification process includes unloading, weighing, and feed sections, and storage pressurizing, and feed hoppers. The unloading section consists of receiving hoppers where the wood chips are dumped from trucks or train cars. The weighing section includes a belt conveyor (36 to 48 inches in length) and weighing scale. A rotary plug valve in the feed conveyor system could be used as a metering valve to eliminate the cost of the weighing conveyor.

The feed section contains a belt conveyor which can feed up to 2800 TPD per conveyor train. The 5000 TPD plant will require 3 trains.

Three trains are required for the storage, pressurizing, and feed hoppers with each train handling 90 ton/hr (150 tph wet). The vessels are 12 ft diameter maximum. A single system allows pressurizing to 150 psig. Above this pressure, a second system is required which essentially represents a second stage and doubles the cost.

Installed Cost by Process Feed Rate

<u>Feed Section</u>	<u>500 tpd</u>	<u>1000 tpd</u>	<u>5000 tpd</u>	<u>Reference</u>
Unloading Hoppers	\$ 2,600	\$ 3,200	\$ 15,600	Battelle
Weighing Conveyor with Scale	126,000	252,000	252,000	Jeffrey Manuf.
Feed Conveyor	130,000	260,000	780,000	Kanya, Inc.
Storage, Pressurizing Feed Hoppers	100,000	1,400,000	4,200,000	Petrocarb, Inc.
(0-150 psig)				
(150-500)	1,400,000	2,800,000	8,400,000

FIGURE A.1. FEED SYSTEM FOR WOOD GASIFICATION

TABLE A-1. BASIS OF CALCULATIONS FOR MATERIAL
AND ENERGY BALANCES

-
1. Single train plant
 2. Location - Gulf coast area
 3. Composition of wood to plant gate:

	<u>Wt. %</u>
C	29.82
H	3.24
S	0.06
O	23.58
N	0.12
Ash	3.18
Moisture	40.00
 4. Moisture-free wood heating value: 8364.5 Btu/lb (4.65 kcal/g)
 5. 90 percent operating factor
 6. Carbon conversion in gasifier:
 - 78 percent intermediate-Btu gasification
 - 74.5 percent high-Btu gasification
 7. Gasifier temperature - 750 C
 8. Combustor temperature - 975 C
 9. 10 percent excess air in combustor
 10. Heat transfer coefficients for heat exchangers:
 - Gas-gas - 5 Btu/hr ft² F
 - Gas-liquid - 100 Btu/hr ft² F
 - Liquid-liquid - 200 Btu/hr ft² F
 11. Compression efficiency - 88 percent
 12. Feed storage - 20 days
 13. Water storage pond - 7 days
 14. Blowdown from scrubber - 10 percent of circulating water
 15. Steam turbines - 95 percent efficiency
 16. All excess steam used to generate electricity
-

TABLE A-1. (Continued)

-
17. All carbon steel vessels
 18. Raw wood chips delivered to plant - $1\frac{1}{2}" \times 1\frac{1}{2}" \times \frac{1}{4}"$ (no feed preparation required.)
 19. Physical properties of wood char:
 - $D_p = 1.05$ inches (diameter of equivalent sphere)
 - $\rho_p = 30$ lb/ft³
 20. Physical properties of sand:
 - $D_p = 0.0177$ in. (48 mesh)
 - $\rho_p = 165$ lb/ft³
 - $c_p = 0.191$ Btu/lb F
 21. Physical properties of dense material:
 - $\rho_p = 250$ lb/ft³
 22. 500 TPD capacity plant:
 - (1) Sand circulation rate - 267.7 tons/hr
 - (2) Gasifier product gas volumetric flow rate - 108 ACF/sec
 - (3) Gasifier superficial gas velocity - 15.29 ft/sec
 - (4) Combustor superficial gas velocity - 5.15 ft/sec
 - (5) Combustor flue gas volumetric flow rate - 116.5 ACF/sec
 - (6) Dryer superficial gas velocity - 5 ft/sec
 - (7) Dryer effluent gas flow - 75.1 ACF/sec
 23. 5000 TPD capacity plant:
 - (1) Sand circulation rate - 1197.4 tons/hr
 - (2) Gasifier superficial gas velocity - 12.2 ft/sec
 - (3) Combustor superficial gas velocity - 5 ft/sec
 - (4) Dryer superficial gas velocity - 5 ft/sec
 24. Hydrogen/carbon monoxide to methanation = 3.06/1
 25. 100 percent conversion of carbon monoxide in methanation reactors
 26. No product storage capabilities
-

TABLE A-2. GASIFICATION MODEL RESULTS FOR CASE 1

STEAM-O ₂ GASIFICATION MODEL							
WOOD COMPOSITION (WT. PCT.)				CONDITIONS			
C	44.73			C CONVERSION (PCT.)		78.00	
H	4.86			STEAM INPUT TEMP. (C)		537.9	
S	.09			O ₂ INPUT TEMP. (C)		27.0	
O	35.37			REACTOR TEMP. (C)		750.0	
N	.18			OUTLET GAS TEMP. (C)		750.0	
ASH	4.77			PRESSURE (PSIA)		115.	
MOISTURE	10.00			S RETENT. IN ASH (PCT.)		50.00	
				O ₂ STREAM PURITY (PCT.)		98.0	
VOL. MTR.	76.00						
HF WOOD HTG. VALUE = 4.650 KCAL/G							
WOOD INLET TEMP. = 176.0 C							
STEAM DECOMPOSITION = 7.0 PCT.							
ITERATION 6 HEAT BALANCE = 423,696 PCT.							
ITER.	ROOTS OF RS EQUATION		H ₂ PARTIAL PRESS. (PSI)			CH ₄	RO
	ONE USED	OTHER ONE	ASSUMED	CALC.	DIFF.	SELECT.	
1	.10625E+00	-.57596E+00	37.59	35.94	=1.65	20.61	-.0015
2	.10655E+00	-.58418E+00	35.94	36.35	.41	20.23	.0005
3	.10649E+00	-.58252E+00	36.27	36.27	=.00	20.30	.0001
** ITERATION CONVERGED.							
** ITERATION FOR HEAT BALANCE CONVERGED.							

TABLE A-2. (Continued)

	ENTHALPY		G-MOLES /100 G WOOD	GAS COMP. (MOLE PCT.)	
	IN	OUT		WET	DRY
CH ₄		9,581	,58981	11,96	13,47
C ₂ H ₄		12,693	,00761	,15	,17
C ₂ H ₆		16,056	,08411	1,71	1,92
CO		5,368	1,45612	29,54	33,25
CO ₂		8,270	,67567	13,71	15,43
H ₂		5,095	1,55859	31,62	35,59
N ₂	,001	5,311	,00643	,13	,15
H ₂ S		6,925	,00140	,03	,03
H ₂ O	,825	,938	,54972	11,15	
C		,298			
ASH		,197			
WOOD	,065				
O ₂	,001				
TOTAL			4,92945		

WEIGHT RATIOS =	/WOOD	/MF WOOD	/MAF WOOD
STEAM	,1065	,1183	,1249
OXYGEN	,0001	,0001	,0001

EXTERNAL HEAT REQUIRED = ,495163 KCAL/G WOOD

PRODUCT GAS HEATING VALUES =	LOW (NET)	HIGH (GROSS)
BTU/LB WOOD (KCAL/G WOOD)	5989, (3,330)	6558, (3,646)
BTU/SCF	320,0	350,4

PRODUCT GAS MOLECULAR WEIGHT = 19,475 , DENSITY = ,1131 LB/FT³

GAS UPGRADING CALCULATIONS =

ULTIMATE HIGH HEATING VALUE = 1051,0 BTU/SCF
G-MOLES CO SHIFTING = ,70244
G-MOLES ACID GAS REMOVAL = 1,37811 CO ₂ + ,00140 H ₂ S = 1,37951
PERCENT PRE-FORMED CH ₄ = 43,9

TABLE A-2. (Continued)

	ENTHALPY		LB-MOLES	GAS COMP. (MOLE PCT.)	
	IN	OUT	/100 LB WOOD	WET	DRY
CH4		9,581	,58981	11,96	13,47
C2H4		12,693	,00761	,15	,17
C2H6		16,056	,08411	1,71	1,92
CO		5,368	1,45612	29,54	33,25
CO2		8,270	,67567	13,71	15,43
H2		5,095	1,55859	31,62	35,59
N2	,001	5,311	,00643	,13	,15
H2S		6,925	,00140	,03	,03
H2O	,825	,938	,54972	11,15	
C		,298			
ASH		,197			
WOOD	,065				
O2	,001				
TOTAL			4,92945		

WEIGHT RATIOS -		/WOOD	/HF WOOD	/MAF WOOD
STEAM		,1065	,1183	,1249
OXYGEN		,0001	,0001	,0001

EXTERNAL HEAT REQUIRED = 891,293 BTU/LB WOOD

PRODUCT GAS HEATING VALUES -	LOW (NET)	HIGH (GROSS)
BTU/LB WOOD (KCAL/G WOOD)	5989, (3,330)	6558, (3,646)
BTU/SCF	320,0	350,4

PRODUCT GAS MOLECULAR WEIGHT = 19,475 , DENSITY = ,1131 LB/FT3

GAS UPGRADING CALCULATIONS -

ULTIMATE HIGH HEATING VALUE = 1051,0 BTU/SCF

LB-MOLES CO SHIFTING = ,70244

LB-MOLES ACID GAS REMOVAL = 1,37811 CO2 + ,00140 H2S = 1,37951

PERCENT PRE-FORMED CH4 = 43,9

TABLE A-2. (Continued)

HEAT QUANTITIES (10 ³ BTU/TON WOOD)-	
FROM WOOD CHAR	2635,5
FOR GASIFICATION	-1782,6
FROM METHANATION	0,0
FROM CO SHIFTING	0,0
FOR DRYING WOOD	-1437,0
FOR GASIFIER STEAM	+305,9
FOR CO ₂ /H ₂ S REMOVAL	0,0
FROM GAS SENSIBLE HEAT	943,6

BALANCE	53,7

TABLE A-3. BASIS OF GAS COST CALCULATIONS

-
-
1. Total Plant Investment = installed equipment costs + contractor's fees
+ contingencies
 2. Contractor's Fee's = 11.1 percent of installed equipment costs
 3. Contingencies = 15 percent of installed equipment costs + contractor's
fees
 4. Annual operating costs:
 - Raw materials
 - Purchased water
 - Labor
 - Administration
 - Supplies
 - Local taxes and insurance
 5. Total Maintenance Costs = 6 percent of the total plant investment
 6. Labor = operating labor + maintenance labor + supervision
 7. Maintenance Labor = 60 percent of total maintenance
 8. Operating Labor = \$8/man-hour (8304 hours/year)
 9. Supervision = 20 percent of operating and maintenance labor
 10. Administration = 60 percent of labor
 11. Supplies = operating supplies + maintenance supplies
 12. Operating Supplies = 30 percent of operating labor
 13. Maintenance Supplies = 40 percent of total maintenance
 14. Local Taxes and Insurance = 2.7 percent of total plant investment
 15. Total Capital Requirement = total plant investment + allowance for
funds used during construction +
startup costs + working capital
 16. Interest on Funds Used During Construction = 9 percent; 3 year
construction time
 17. Start-up Costs = 20 percent total annual gross operating costs
 18. Working capital = raw materials + materials and supplies + net
receivables
-

TABLE A-3. (Continued)

19. 14-Day Inventory for Raw Materials

20. Materials and Supplies = 0.9 percent of total plant investment

21. Net Receivables = $1/24$ annual gas revenue at calculated sales price +
 $1/24$ annual by-product revenue

22. Utility Financing Basis:

20 year project life

5 percent per year straight line depreciation on total capital
 requirement excluding working capital

48 percent federal income tax rate

First year gas cost, \$/MM Btu =

$$N + 0.05(C-W) + 0.01\left[p + \frac{48}{52}(1-d)r\right][C - 0.025(C-W)]$$

Average gas cost, \$/MM Btu =

$$\frac{N + 0.05(C-W) + 0.005\left[p + \frac{48}{52}(1-d)r\right][C+W]}{G}$$

C = Total capital requirement, millions of dollars

W = Working capital, millions of dollars

N = Total net operating cost, millions of dollars per year

G = Annual gas production, trillion Btu/year

d = 75 percent debt fraction

i = Interest on debt, 9 percent per year

r = return on equity, 15 percent per year

p = Return on rate base, 10.5 percent per year

23. Private Investor Financing Basis:

20 year project life

16 year sum-of-the-year's digits depreciation on total plant
 investment

100 percent equity capital

12 percent discounted cash flow return rate

48 percent federal income tax rate

Average gas cost, \$/MM Btu =

$$\frac{N + 0.247 I + 0.1337 S + 0.2305 W}{G}$$

I = Total plant investment, millions of dollars

S = Startup costs, millions of dollars

W = Working capital, millions of dollars

N = Total net operating cost in first year, millions of dollars/
 year

G = Annual gas production, trillion Btu/year

TABLE A-4. GASIFICATION MODEL RESULTS FOR CASE 5

STEAM-O ₂ GASIFICATION MODEL			
WOOD COMPOSITION (WT. PCT.,)		CONDITIONS	
C	44.73	C CONVERSION (PCT.,)	74.50
H	4.86	STEAM INPUT TEMP. (C)	537.9
S	.09	O ₂ INPUT TEMP. (C)	27.0
O	35.37	REACTOR TEMP. (C)	750.0
N	.18	OUTLET GAS TEMP. (C)	750.0
ASH	4.77	PRESSURE (PSIA)	515.
MOISTURE	10.00	S RETENT. IN ASH (PCT.,)	50.00
		O ₂ STREAM PURITY (PCT.,)	98.0
VOL. MTR.	76.00		
HF WOOD HTG. VALUE = 4,650 KCAL/G			
WOOD INLET TEMP. = 260.0 C			
STEAM DECOMPOSITION = 0.0 PCT.			
ITERATION 1	HEAT BALANCE = 235,000 PCT.		
	** ITERATION CONVERGED,		
ITERATION 2	HEAT BALANCE = 240,000 PCT.		
	** ITERATION CONVERGED,		
ITERATION 3	HEAT BALANCE = 161,794 PCT.		
	** ITERATION CONVERGED,		
ITERATION 4	HEAT BALANCE = 177,337 PCT.		
	** ITERATION CONVERGED,		

TABLE A-4. (Continued)

ITERATION 5		HEAT BALANCE = 174,388 PCT.		
** ITERATION CONVERGED,				
** ITERATION FOR HEAT BALANCE CONVERGED,				
	ENTHALPY		LB-MOLES	GAS COMP, (MOLE PCT,)
	IN	OUT	/100 LB WOOD	WET DRY
CH4		9,581	,64263	17,39 23,75
C2H4		12,693	,01087	,22 ,31
C2H6		16,056	,12017	2,48 3,39
CO		5,368	,57492	11,86 16,20
CO2		8,270	1,09506	22,59 30,86
H2		5,095	,89688	18,50 25,28
N2	,001	5,311	,00642	,13 ,18
H2S		6,925	,00140	,03 ,04
H2O	,816	,937	1,29851	26,79
C		,298		
ASH		,197		
WOOD	,108			
O2	,001			
TOTAL			4,84685	
WEIGHT RATIOS =				
		/WOOD	/MF WOOD	/MAF WOOD
STEAM		,2339	,2599	,2745
OXYGEN		-,0001	-,0001	-,0001
EXTERNAL HEAT REQUIRED = 400,172 BTU/LB WOOD				
PRODUCT GAS HEATING VALUES =				
		LOW (NET)	HIGH (GROSS)	
	BTU/LB WOOD (KCAL/G WOOD)	5341, (2,969)	5902, (3,281)	
	BTU/SCF	290,2	320,7	

TABLE A-4. (Continued)

PRODUCT GAS MOLECULAR WEIGHT = 22,109 , DENSITY = ,5762 LB/FT3	
GAS UPGRADING CALCULATIONS *	
ULTIMATE HIGH HEATING VALUE =	1075,6 BTU/SCF
LB-MOLES CO SHIFTING =	,20697
LB-MOLES ACID GAS REMOVAL =	1,30203 CO2 + ,00140 H2S = 1,30343
PERCENT PRE-FORMED CH4 =	69,6
HEAT QUANTITIES (10 ⁶ BTU/TON WOOD) *	
FROM WOOD CHAR	2990,5
FOR GASIFICATION	-800,3
FROM METHANATION	624,8
FROM CO SHIFTING	73,3
FOR DRYING WOOD	-1636,5
FOR GASIFIER STEAM	-668,8
FOR CO2/H2S REMOVAL	-1497,4
FROM GAS SENSIBLE HEAT	935,5

BALANCE	1,1

TABLE A-5. ANNUAL OPERATING COSTS - CASE 1
(RAW WOOD - \$20/TON)

Raw Materials (840.8 T/D x \$20/ton x 365 days x 0.9)	\$5,524,100
Purchased Water (1 x 10 ⁶ gal x \$0.70/M gal)	700
Labor	
Process Operating Labor	876,000
Maintenance	245,100
Supervision	224,200
Administration	807,200
Supplies	
Operating	262,800
Maintenance	163,400
Local Taxes and Insurance	183,800
TOTAL ANNUAL OPERATING COSTS	8,287,300

TABLE A-6. TOTAL CAPITAL REQUIREMENT - CASE 1
(\$20/TON)

Total Plant Investment	\$6,808,000
Allowance for Funds Used During Construction (T.P.I. x 3 yrs x 0.9)	1,838,400
Startup Costs (20% total gross annual operating costs)	1,657,500
Working Capital	
1) Raw materials	235,400
2) Materials and supplies	61,300
3) Net receivables	401,000
TOTAL CAPITAL REQUIREMENT	\$11,001,600

TABLE A-7. ANNUAL OPERATING COSTS - CASE 2
(RAW WOOD - \$20/TON)

Raw Materials (1681.6 T/D x \$20/ton x 365 days x 0.9)	\$11,048,100
Purchased Water (2 x 10 ⁶ gal x \$0.70/M gal)	1,400
Labor	
Process Operating Labor	1,051,200
Maintenance Labor	398,200
Supervision	289,900
Administration	1,044,000
Supplies	
Operating	315,400
Maintenance	265,400
Local Taxes and Insurance	298,600
TOTAL ANNUAL OPERATING COSTS	\$14,712,200

TABLE A-8. TOTAL CAPITAL REQUIREMENT - CASE 2
(RAW WOOD - \$20/TON)

Total Plant Investment	\$11,060,000
Allowance for Funds Used During Construction	2,986,000
Startup Costs	2,942,400
Working Capital	
1) Raw materials	470,800
2) Materials and supplies	99,500
3) Net receivables	704,200
TOTAL CAPITAL REQUIREMENT	\$18,262,900

TABLE A-9. ANNUAL OPERATING COSTS - CASE 3
(RAW WOOD - \$20/TON)

Raw Materials	\$5,524,000
Purchased Water	700
Labor	
Process Operating Labor	876,000
Maintenance Labor	245,100
Supervision	216,300
Administration	807,200
Supplies	
Operating	262,800
Maintenance	137,100
Local Taxes and Insurance	154,200
TOTAL ANNUAL OPERATING COSTS	\$8,223,400

TABLE A-10. TOTAL CAPITAL REQUIREMENT - CASE 3
(RAW WOOD - \$20/TON)

Total Plant Investment	\$5,711,900
Allowance for Funds Used During Construction	1,544,200
Startup Costs	1,644,700
Working Capital	
1) Raw materials	235,400
2) Materials and supplies	51,400
3) Net receivables	391,000
TOTAL CAPITAL REQUIREMENT	\$

TABLE A-11. ANNUAL OPERATING COSTS - CASE 4
(RAW WOOD - \$20/TON)

Raw Materials	\$11,048,100
Purchased Water	1,400
Labor	
Process Operating Labor	1,051,200
Maintenance	334,100
Supervision	277,100
Administration	332,500
Supplies	
Operating	315,400
Maintenance	222,700
Local Taxes and Insurance	250,500
TOTAL ANNUAL OPERATING COSTS	\$13,833,000

TABLE A-12. TOTAL CAPITAL REQUIREMENT - CASE 4
(RAW WOOD - \$20/TON)

Total Plant Investment	\$9,279,000
Allowance for Funds Used During Construction	2,505,000
Startup Costs	2,766,600
Working Capital	
1) Raw materials	470,800
2) Materials and supplies	83,500
3) Net receivables	656,400
TOTAL CAPITAL REQUIREMENT	\$15,761,300

TABLE A-13. ANNUAL OPERATING COSTS - CASE 5
(RAW WOOD - \$20/TON)

Raw Materials (8333.3 x \$20/ton x 365 days x 0.9)	\$54,749,800
Labor	
Process Operating labor	2,102,400
Maintenance Labor	2,036,500
Supervision	827,800
Administration	2,980,000
Supplies	
Operating	530,700
Maintenance	1,357,700
Local Taxes and Insurance	1,527,400
TOTAL ANNUAL OPERATING COSTS	\$66,212,300

TABLE A-14. TOTAL CAPITAL REQUIREMENT - CASE 5
(RAW WOOD - \$20/TON)

Total Plant Investment	\$56,670,100
Allowance for Funds Used During Construction	15,172,900
Startup Costs	13,242,500
Working Capital	
1) Raw materials	2,333,300
2) Materials and supplies	509,100
3) Net receivables	3,225,000
(By-product revenue)	733,800
TOTAL CAPITAL REQUIREMENT	\$91,786,700

TABLE A-15. ANNUAL OPERATING COSTS - CASE 6
(RAW WOOD - \$20/TON)

Raw Materials	\$54,750,000
Labor	
Process Operating Labor	2,102,400
Maintenance Labor	1,937,700
Supervision	808,000
Administration	2,908,900
Supplies	
Operating	630,700
Maintenance	1,291,800
Local Taxes and Insurance	1,453,300
TOTAL ANNUAL OPERATING COSTS	\$65,882,800

TABLE A-16. TOTAL CAPITAL REQUIREMENT - CASE 6
(RAW WOOD - \$20/TON)

Total Plant Investment	\$53,825,600
Allowance for Funds Used During Construction	14,532,900
Startup Costs	13,176,600
Working Capital	
1) Raw materials	2,333,300
2) Materials and supplies	484,400
3) Net receivables	3,191,700
(By-product revenue)	733,800
TOTAL CAPITAL REQUIREMENT	\$88,278,300

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