PNL-SA-16615
CONF-890236--1

Received by OSTI

APR 05 1989

PNL-SA--16615

DE89 009157

A TECHNICAL AND ECONOMIC ANALYSIS OF DIRECT BIOMASS LIQUEFACTION

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February 1989

Presented at the
Energy from Biomass and Waste XIII
New Orleans, Louisiana
February 13-17, 1989

Work supported by
the U.S. Department of Energy
under Contract DE-AC06-76RLO 1830

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ABSTRACT

This paper is based on the results of a technoeconomic assessment of direct biomass liquefaction processes converting wood and peat to gasoline and diesel fuels. The assessment was carried out by the Working Group of the International Energy Agency, Direct Biomass Liquefaction Activity, in which Canada, Finland, Sweden, and the United States participated. The processes chosen for detailed analysis were Atmospheric Flash Pyrolysis (AFP) and Liquefaction In Pressurized Solvent (LIPS). The assessment covered three steps for each process from feed to final product:

1. primary liquefaction to a crude oil product
2. catalytic hydrotreating to upgrade the crude product to a deoxygenated product oil
3. refining the deoxygenated product to gasoline and diesel fuel.

Present technology cases and potential future technology cases were evaluated. A consistent analytical basis was used throughout to allow comparison of the processes. This assessment shows that AFP is more economical than LIPS both for the production of boiler fuel oil as the primary liquefaction product and for the production of gasoline and diesel fuel products. The potential for future cost reduction through research and development is also clearly demonstrated.

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*Operated for the U.S. Department of Energy by Battelle Memorial Institute under Contract DE-AC06-76RLO 1830

A TECHNICAL AND ECONOMIC ANALYSIS OF DIRECT BIOMASS LIQUEFACTION

INTRODUCTION

Liquid fuels from biomass are expected to play a major role in the energy future of industrialized countries[21]. These fuels offer many advantages over raw biomass as an energy product. For example, the liquid hydrocarbon product is compatible with the hydrocarbon fuels currently marketed that are produced from petroleum. Removal of oxygen and adjustment of the hydrogen to carbon ratio during the liquefaction and upgrading steps results in a product mixture of true hydrocarbons that could be substituted directly into the existing fuels market. The energy density of liquid fuels is higher than raw biomass because of both chemical and physical characteristics. The heat of combustion of dry wood, for example, is half that of gasoline, while the bulk density of solid biomass is much less than conventional liquid fuels since plant structures have high porosity. Other advantages of liquid fuels include stability during long storage periods and ease of transportation.

This paper is based on the results of Stage 2 of a cooperative activity carried out by Canada, Finland, Sweden, and the USA[13]. In an earlier study (Stage 1), compiled in 1983[14], process concepts studied in the four countries were reviewed and evaluated. The results of that study indicated that direct liquefaction of biomass could be used to produce liquid boiler fuels with an overall energy efficiency of 50% to 60%. It was also concluded that upgrading the biomass crude oil to gasoline-like products would be possible, although much less experimental work had been done in that field. Since the earlier study, much progress has been made in the field of biomass liquefaction[3,20].

APPROACH FOR TECHNOECONOMIC ASSESSMENT

The production cost for synthetic gasoline or diesel fuel from the biomass feedstock was determined for two alternative process concepts Atmospheric Flash Pyrolysis (AFP) and Liquefaction In Pressurized Solvent (LIPS). For the processes the conversion to gasoline required three steps: liquefaction, upgrading, and refining. The production cost was determined by estimating the capital and operating costs for the liquefaction, upgrading, and refining plants. Where applicable, credits were included for process byproducts. The capital and operating costs for processing the plant emissions and effluents to conform with environmental restrictions in the participating countries were also included. The plant capacities were based on processing 1000 dry ton/day of biomass feed.

Capital Costs

The capital costs for each plant were determined by summing the costs related to the major pieces of equipment. The equipment was sized from material and energy balances completed for each process. The equipment costs were then estimated using published data from a reference base compiled by the Working Group[13]. In a few cases where published data were not available, equipment costs were determined by obtaining quotes from manufacturers. These cases are noted in the reference base. The costs for plant instrumentation and control were determined by summing the instrumentation costs associated with each piece of equipment. These costs are also listed in the reference base.

The total capital investment was determined using the factor method proposed by Guthrie[12]. The base equipment costs were multiplied by a direct cost factor, a building factor, a site improvement factor, and a utilities factor to obtain the total direct cost for a piece of equipment. This cost was then multiplied by an indirect cost factor, covering engineering costs and construction overheads, and the resulting sum for each piece of equipment gave the Fixed Capital Investment (FCI) for the plant.

The direct cost factor used was dependent on the particular piece of equipment; these factors were obtained from Cran[4]. The indirect cost factors varied from 1.0 to 1.3, depending on whether the base equipment cost data were for a turnkey unit, an installed unit, or the base equipment only. The factors used for the required buildings, site improvements, and utilities are listed below:

-Buildings	1.15
-Site improvement	1.06
-Utilities	1.16

The Total Capital Requirement (TCR) for a plant was determined by adding to the FCI the startup costs (10% of FCI), the working capital (5% of FCI), and the interest during a 3-year construction period. The costs of land and contingencies were not included in the cost estimate.

Operating Costs

The majority of the fixed operating costs (maintenance labor, maintenance materials, overhead, and insurance) was based on percentages of the FCI. These percentages and the basis for the variable operating costs (feedstock, electricity, and catalyst costs) are listed in the section entitled Bases For Assessment (see Table 1).

The capital charges were added to the fixed and variable operating costs to determine the production cost of the product for each plant. The capital charges are determined by the annuity method, which produces levelized costs; a 20-year plant life was assumed. For the upgrading and refining plants, the production cost of the intermediate product became the feedstock cost for the next step.

Refining Costs

Refining costs were estimated differently as average costs in refineries built to handle 5 million tons of various oils per annum. The cost estimate was based on data supplied by Frumkin[11] on refining oils produced by liquefaction of coal and oils from shale. The refining operation was subdivided into separate processes, for instance, catalytic reforming, hydroprocessing, and wastewater treatment. Costs were estimated separately for each of the processes. A power of scale of 0.68 was used to calculate costs for different sizes. The yields in each process were based on petroleum refining, and were changed for the different elemental compositions and molecular weight distributions while assuming similar chemical reactivity.

BASES OF ASSESSMENT

Table 1 shows the assumptions and costs of the various parameters used to determine production costs. The feedstock used in the evaluation was eitherwood chips with 50% moisture or milled peat with 50% moisture. Two sensitivity case analyses are included in this paper. First, the effect of the feedstock price was considered because it varied among the participating countries. Secondly, variations in the capital costs (up to 20% high or low) were analyzed to account for the uncertainty of the capital cost estimates.

TABLE 1
PARAMETERS USED IN COST BASES

Plant Capacity	1000 dry tonnes/day biomass feed
Time	September 1987
Place	US Gulf Coast
Currency	United States Dollars
Annual Operating Time	8000 Hours
Labor Rate	20 US$/h including payroll burden
Maintenance Labor	1% of FCI
Maintenance Materials	3% of FCI
Overhead	2% of FCI
Insurance	2% of FCI
Other Fixed Operating Costs	1% of FCI
Catalyst Price	Specified for process operation
Feedstock Price	
Wood chips, 50% moisture	30 US$/tonne
Milled peat, 50% moisture	20 US$/tonne
Electricity Price	0.065 US$/kWh
Interest Rate	10%
Plant Life	20 years

PROCESSES EVALUATED

The results from the analysis of the AFP and LIPS processes, using a wood or a peat feedstock, are presented here. In the process concepts, catalytic hydrotreating was used in upgrading the products from both processes. The complexity of the refining step, where the finished products were produced, was dependent on the oxygen content of the product from the upgrading step.

Process designs were determined for the evaluation by scaleup of the process concepts under development at the laboratory-scale and at pilot-scale. Two process design cases were prepared for each process studied. The first case, referred to as "Present" technology, was based on actual experimental data available and utilized equipment already commercially proven for identical or similar applications. The second case, referred to as "Potential" technology, was based on process results determined as feasible to obtain in the future and/or using equipment applications not currently demonstrated at the commercial scale. The achievement of the assumptions used in the Potential cases is thought to be reasonable through future experimental and developmental work.

Atmospheric Flash Pyrolysis of Wood

In the Present case of the AFP process, wood fibers (1 mm particle size) are rapidly pyrolyzed at 500°C to vapors and char. The condensed vapors form the primary tar product, which contains approximately 39% oxygen on a dry basis. The flowsheet for the primary liquefaction step is shown in Figure 1. It is based on processes being developed at the

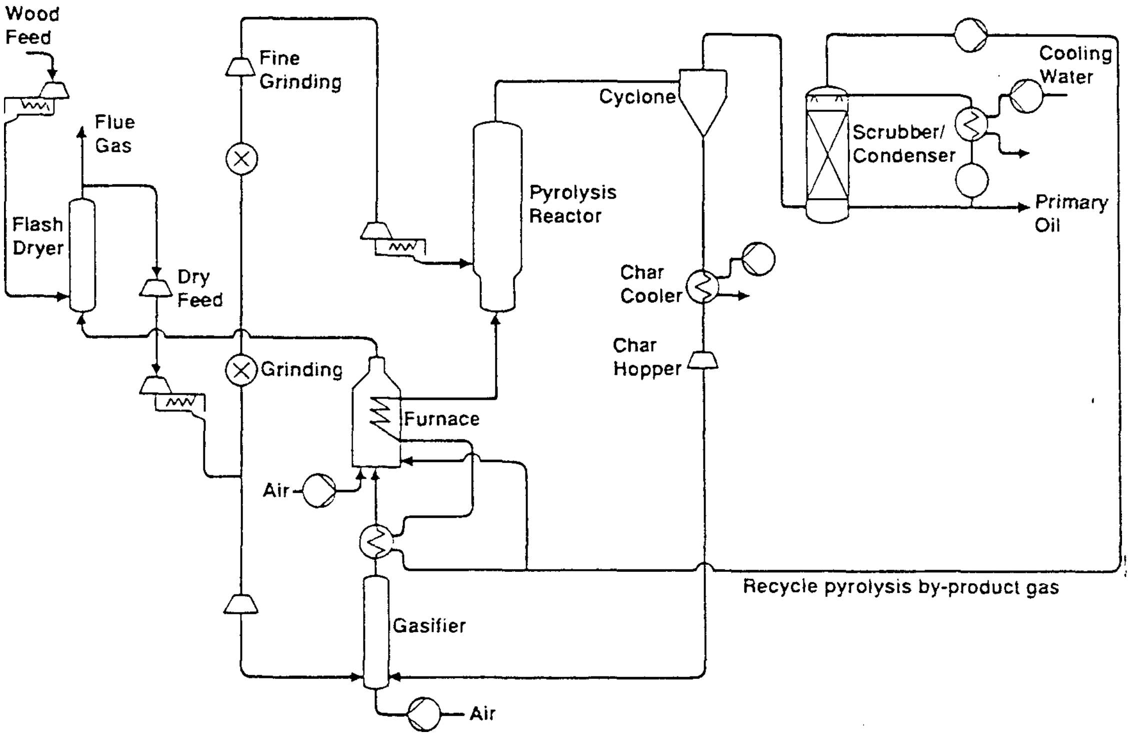

Figure 1. Process Schematic of Wood AFP - Present Technology

Georgia Tech Research Institute, USA[15], and the University of Waterloo, Canada[19]. The material balance is based on data from a continuous 3-kg/h process development unit, operated at the University of Waterloo. The experimental results were normalized to give a material balance closure of 100% for this evaluation.

In the Present case, the primary oil is upgraded using a two-stage catalytic hydrotreating process, shown in Figure 2. The first stage operates at 250°C as a stabilizing step for the second stage, which operates at 350°C. A conventional sulfided cobalt/molybdenum on alumina, petroleum hydrotreating catalyst is used. The material balance for upgrading the primary oil is based on experimental results obtained at Pacific Northwest Laboratory (PNL) using oil from the Georgia Tech Research Institute[8].

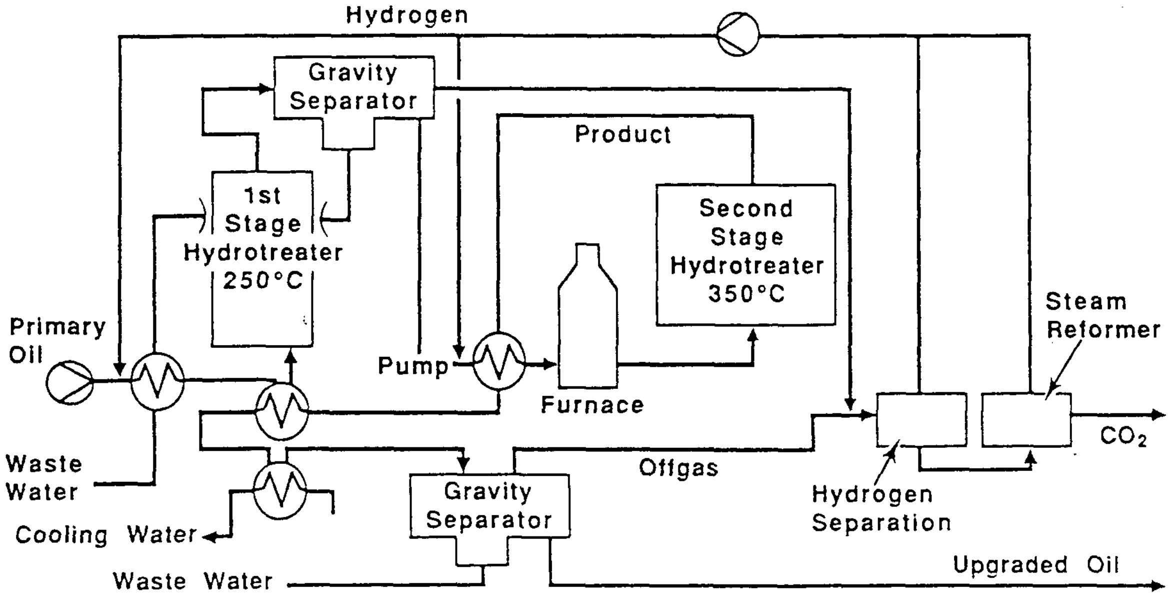

Figure 2. Upgrading AFP Primary Oil, Two-Stage Hydrotreating

Refining the upgraded product consists of multiple steps including: distillation of the upgraded product, hydrodeoxygenation of the light fraction, hydrocracking of the heavy fraction, catalytic reforming of the gasoline product, and steam reforming of the hydrocarbon gas product to produce hydrogen, as shown in Figure 3. A gasoline and diesel product slate is produced.

The flowsheet for the Potential case primary liquefaction step is shown in Figure 4. In this case, the pyrolysis takes place in a circulating, fluidized-bed reactor, which has the advantage of a much greater throughput compared with the fluidized-bed reactor used in the Present case concept. It is assumed that the same physical properties could be obtained in this reactor as those used in the experimental unit, allowing the same material balance and product properties to be used as in the Present case.

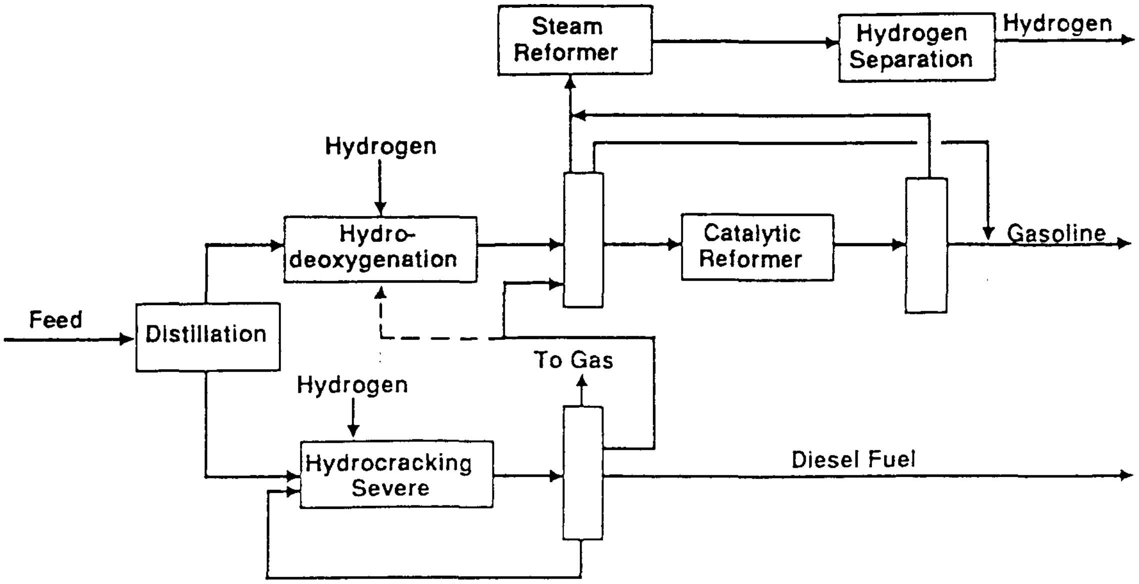

Figure 3. Refining of Upgraded Oil from Wood AFP

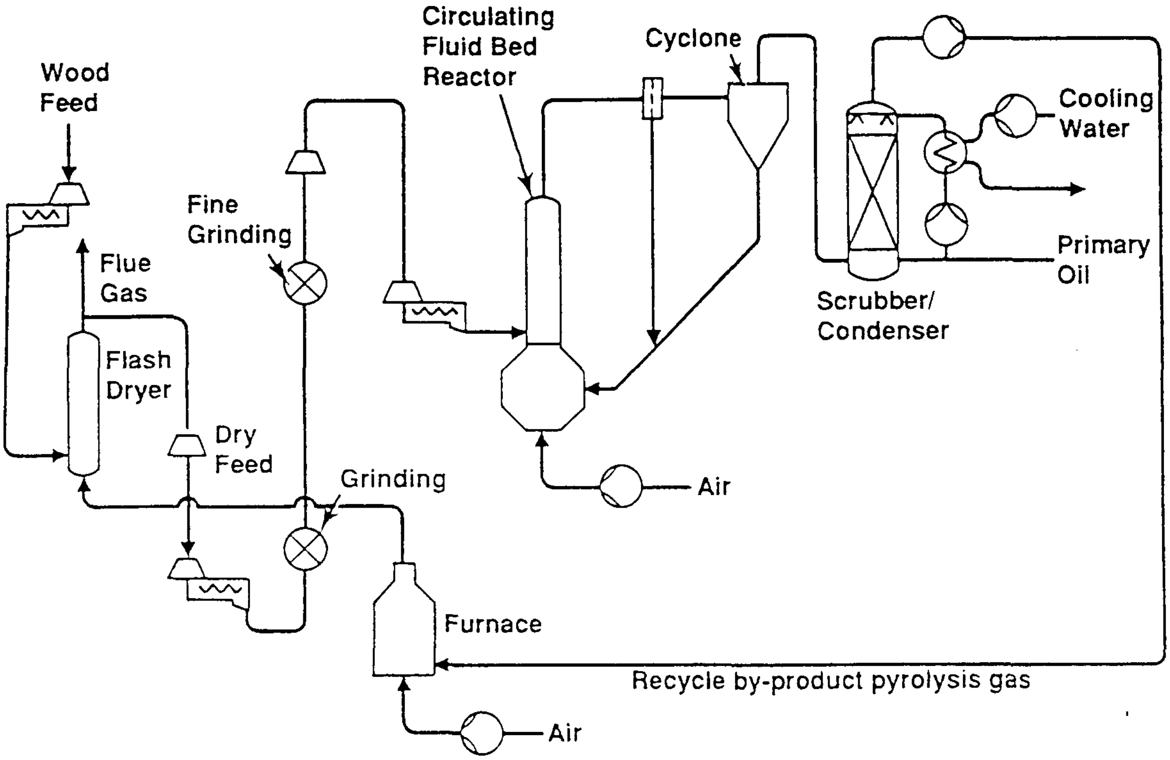

Figure 4. Process Schematic Wood AFP - Potential Technology

In the upgrading step for the Potential case, catalytic hydrotreating takes place under nonisothermal conditions in three reactors operated in series, at 300°C, 350°C, and 400°C, respectively, see Figure 5. Process improvements are assumed, through catalyst development, including a stoichiometrically-balanced mass balance where sufficient byproduct gas is produced to provide for needed

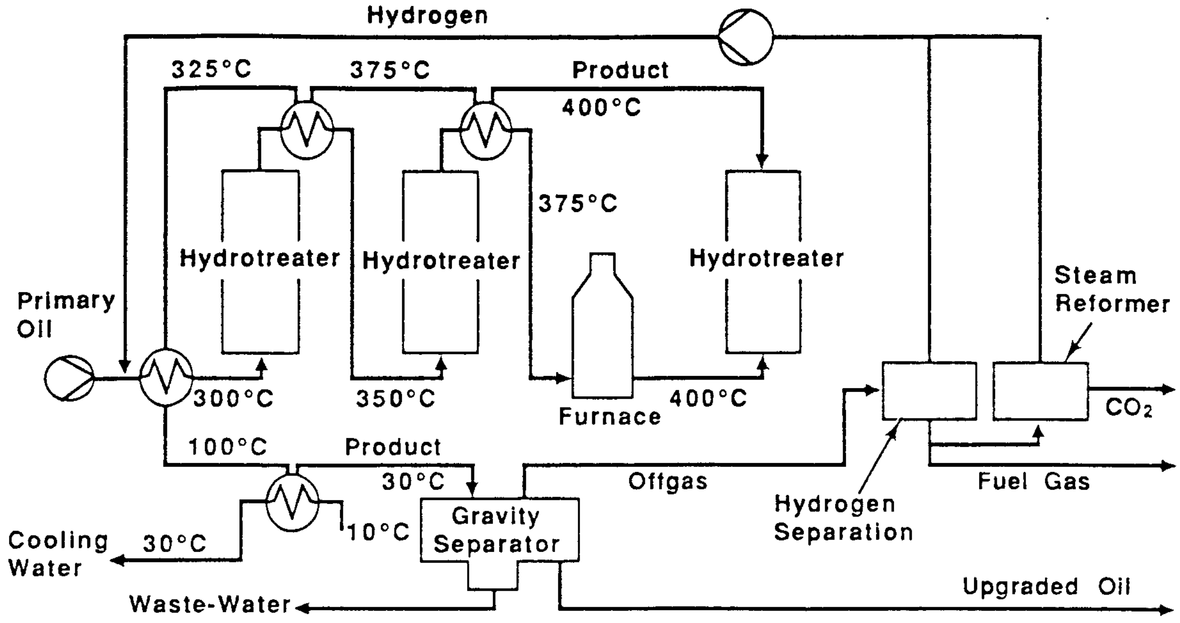

Figure 5. Nonisothermal Catalytic Hydrotreatment for Upgrading AFP Primary Oils

hydrogen production. In addition, a higher reactor space velocity is used. The upgraded product is assumed to be primarily a high-octane gasoline requiring no further processing other than distillate fractionation.

Liquefaction of Wood In Pressurized Solvent

The Present case of the LIPS process is based on tests at the Biomass Liquefaction Experimental Facility, Albany, Oregon, USA, in the PERC mode[22]. A flowsheet of the process is shown in Figure 6. Wood chips are ground to less than 0.5 mm and mixed with recycled wood-derived oil. A sodium carbonate catalyst solution and CO/H_2 syngas are added to the slurry prior to preheating. Liquefaction takes place in an upflow tubular reactor at 350°C, 20.5 MPa, with a 20-minute residence time. Gas is flashed from the reactor effluent stream. A portion of the liquid is recycled. Water is separated from the primary oil product and is treated before discharge. The syngas required for the liquefaction and for preheating the slurry is produced by gasifying a portion of the wood feed in an oxygen-blown gasifier. The material balance for the primary liquefaction is based on data from Run TR-12 at the Biomass Liquefaction Experimental Facility. The primary product oil contains 12.7% oxygen on a dry basis.

The primary oil is upgraded by catalytic hydrotreatment as shown in Figure 7. Processing conditions include 400°C, 2000 psig, and liquid hourly space velocity of 0.3 with a sulfided cobalt/molybdenum on alumina catalyst. The material balance is based on experimental results obtained at PNL using product oil from the Albany plant, Run TR-12[1].

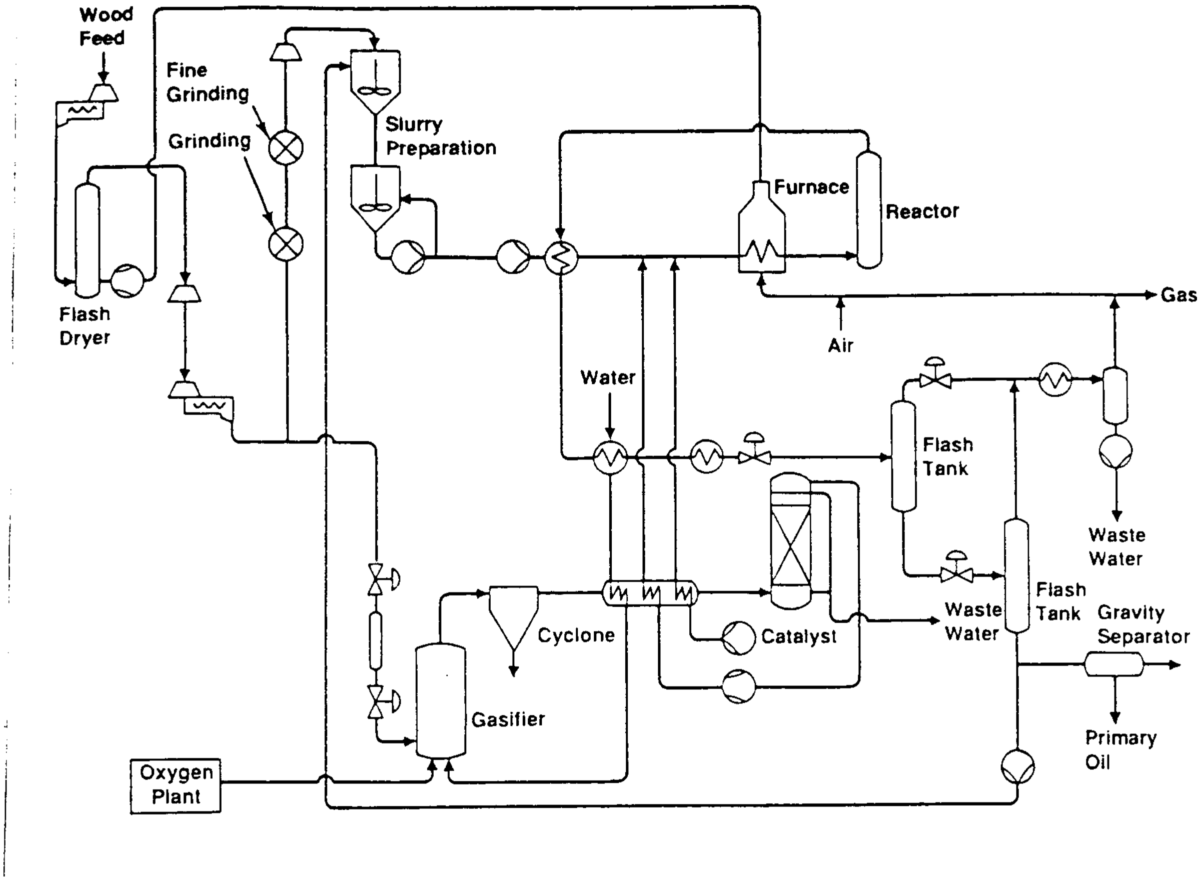

Figure 6. Process Schematic Wood LIPS - Present Technology

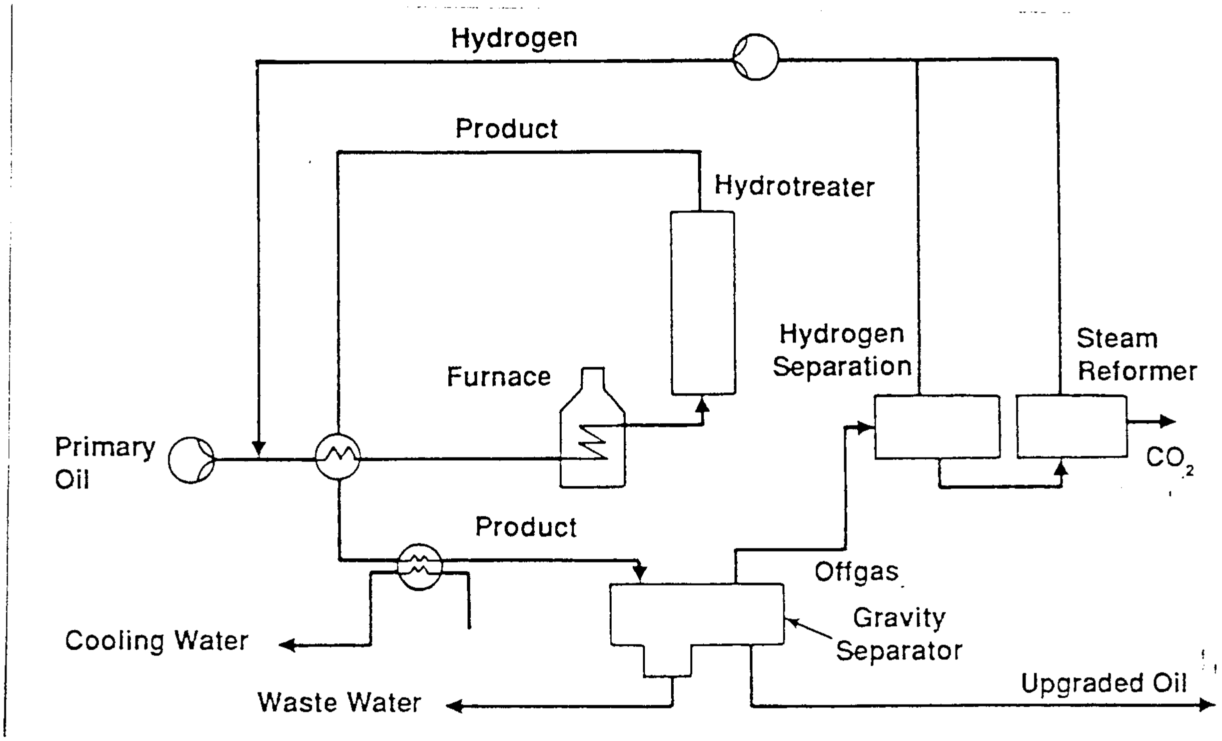

Figure 7. Upgrading Wood LIPS Primary Oil

Refining the upgraded product to a finished product slate involves multiple steps including: distillation of the upgraded product, hydrocracking of the heavy fraction, catalytic reforming of the gasoline fraction, and steam reforming of the hydrocarbon product gases to produce hydrogen, as shown in Figure 8. A gasoline and diesel product slate is produced.

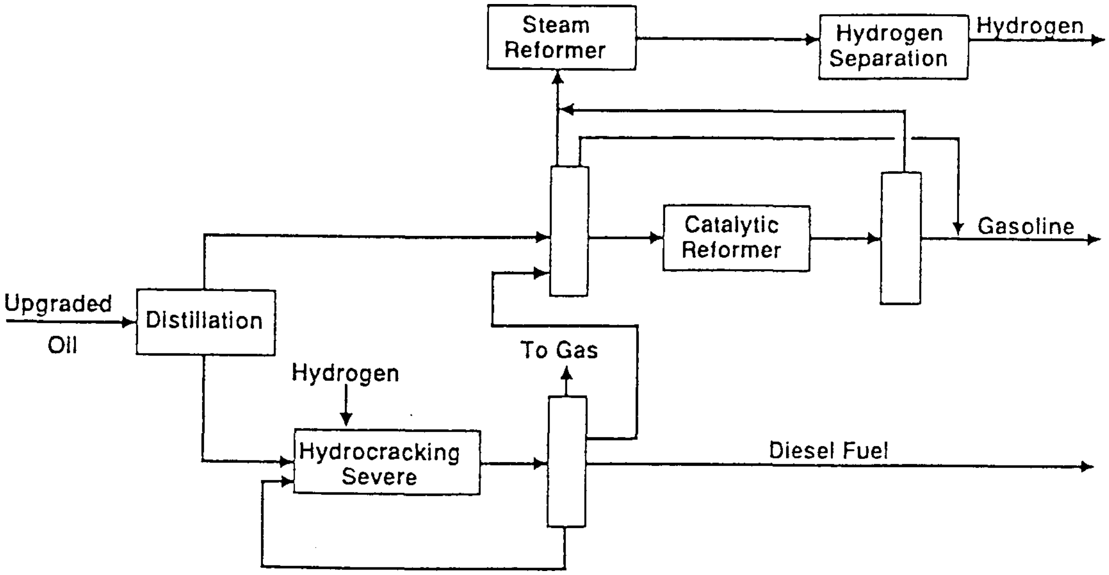

Figure 8. Refining of Upgraded Oil from Wood LIPS

The Potential case for this process is based on the use of an extruder to feed high concentration wood slurries, as tested at the University of Arizona[23]. The flowsheet for this process is shown in Figure 9. The oil phase of the slurry consists of recycled vacuum distillate bottoms. Superheated steam is added to the reactor to provide the reactor heat requirement. Sodium carbonate is used in this process and is assumed to be recycled entirely in the distillate bottoms. No reducing gas is added to the reactor. The liquid product stream is separated into a distillate product and recycled bottoms in a vacuum distillation tower.

Although the extruder feeding method and reactor system are based on research work by the University of Arizona, sufficient material balance data were not available for this evaluation. Therefore, the material balance for the reactor was based on batch results from the University of Toronto Steam Pyrolysis System[2]. The primary product oil is assumed to have an oxygen content of 23.5% on a dry basis.

Catalytic hydrotreatment is used to upgrade the primary oil under similar processing conditions as the Present case, depicted in Figure 7, based on results obtained

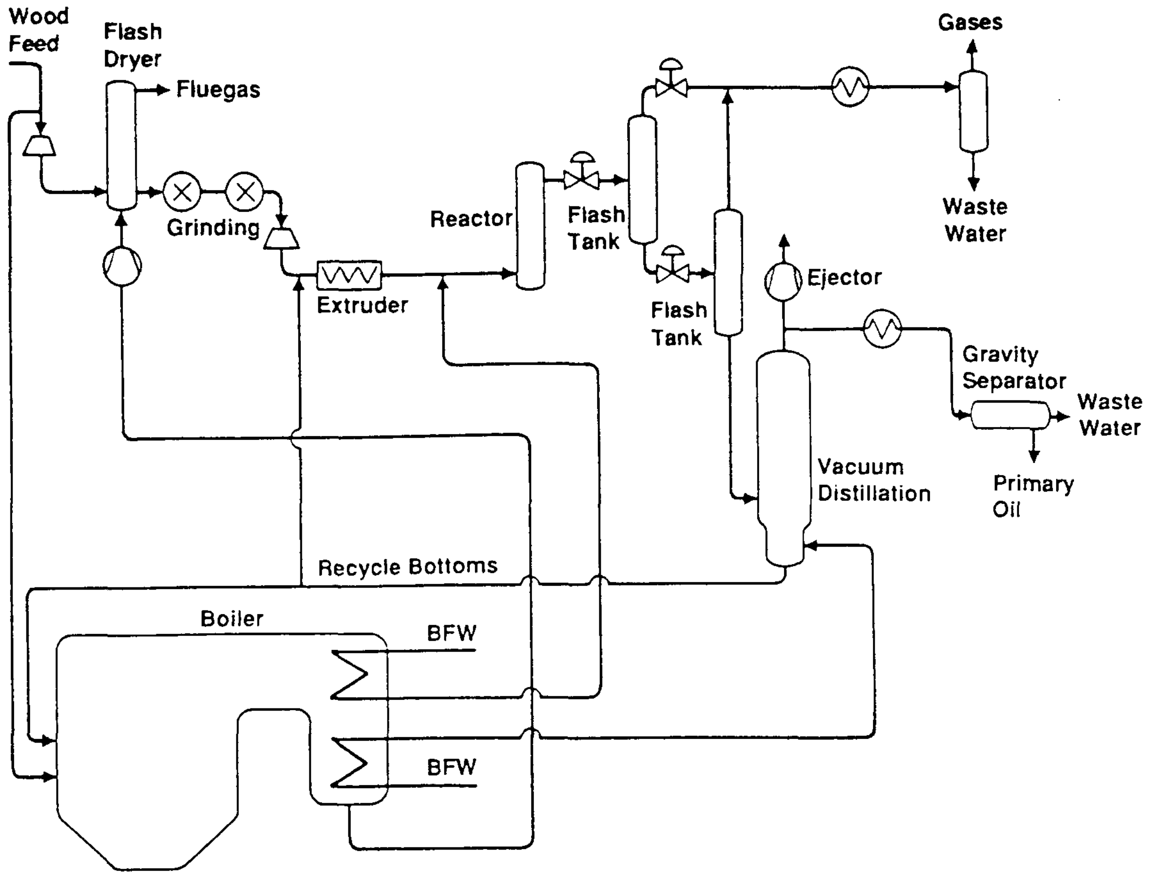

Figure 9. Process Schematic Wood LIPS - Potential Technology

at PNL[7]. Process improvements again are assumed through catalyst development, including a stoichiometrically balanced mass balance. A higher reactor space velocity also is used. In this Potential case, the catalyst lifetime is extended to a more conventional turnover basis because of the separation of the heavy ends from the primary product oil through distillation. As a result of the assumed catalyst improvements and the differences in the primary product oil, the upgraded product is primarily a high-octane gasoline requiring no further processing other than distillate fractionation.

Atmospheric Flash Pyrolysis of Peat

The Present technology process for AFP of peat is virtually the same as that described for wood. Peat fibers (dried and ground to less than 1 mm) are rapidly pyrolyzed at 500°C to vapors and char. The condensed vapors form the primary product, which consists of two phases. The tar phase is upgraded, while the aqueous phase is sent to wastewater treatment. The flowsheet for the primary liquefaction step is shown in Figure 10. It is based on process development work at the University of Waterloo, Canada[18]. The material balance is based on data from a continuous 3-kg/h process development unit.

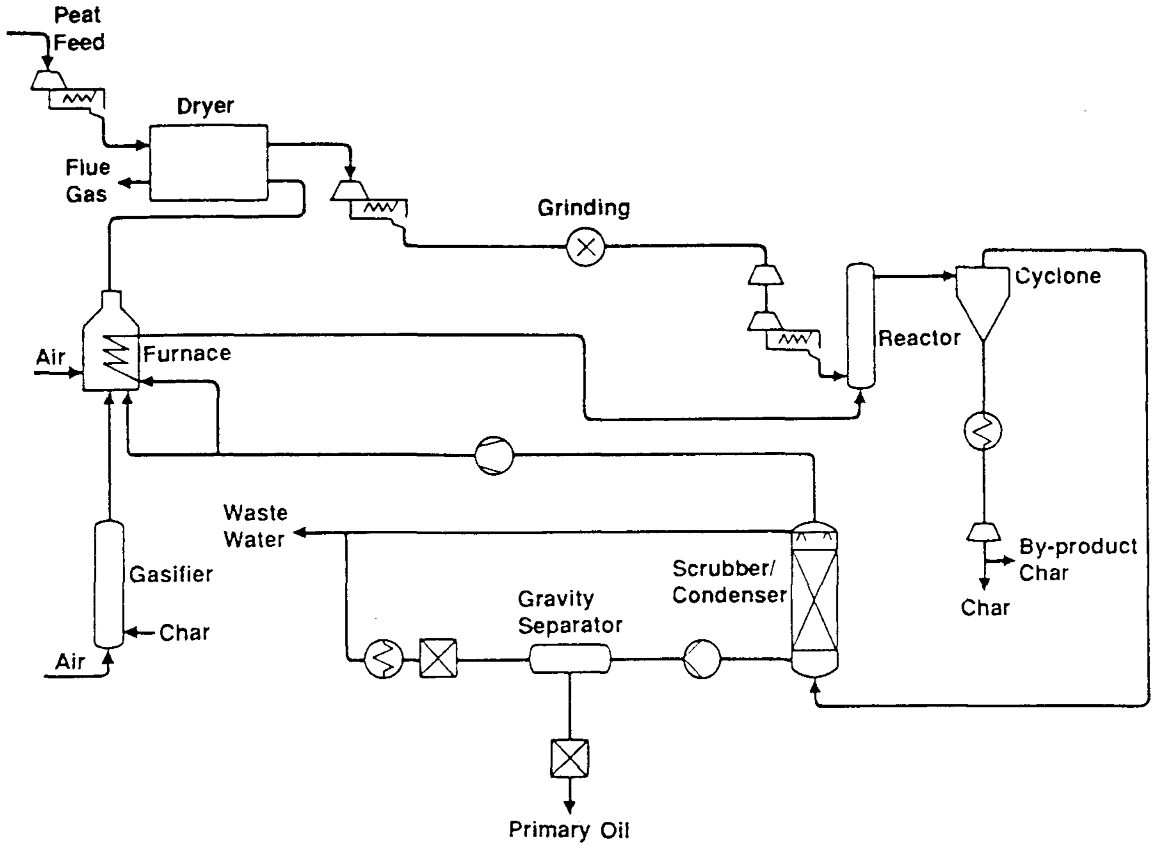

Figure 10. Process Schematic Peat AFP - Present Technology

The upgrading step for the Present case takes place in the nonisothermal catalytic hydrotreater configuration shown in Figure 5. The mass balance is based on experimental results achieved with peat-derived oil in tests at PNL[9].

Refining the upgraded product consists of multiple steps including: distillation of the upgraded product, hydrodeoxygenation of the light fraction, hydrocracking of the heavy fraction, catalytic reforming of the gasoline product, and steam reforming of the hydrocarbon gas product to produce hydrogen, as shown in Figure 11. A gasoline and diesel product slate is produced.

The flowsheet for the Potential case primary liquefaction step is shown in Figure 12. In this case, the pyrolysis takes place in a circulating, fluidized-bed reactor, which has the advantage of a much greater throughput compared with the fluidized-bed reactor used in the Present case concept. It is assumed that the same physical properties could be obtained in this reactor as those in the experimental unit, allowing the same material balance and product properties to be used as in the Present case.

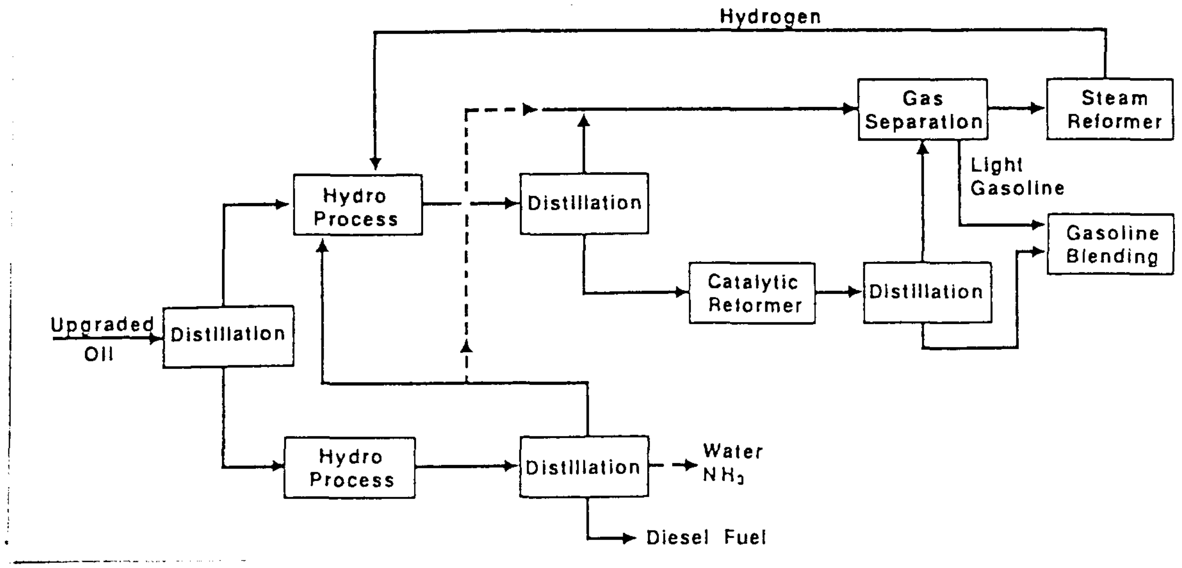

Figure 11. Refining of Upgraded Oil from Peat AFP

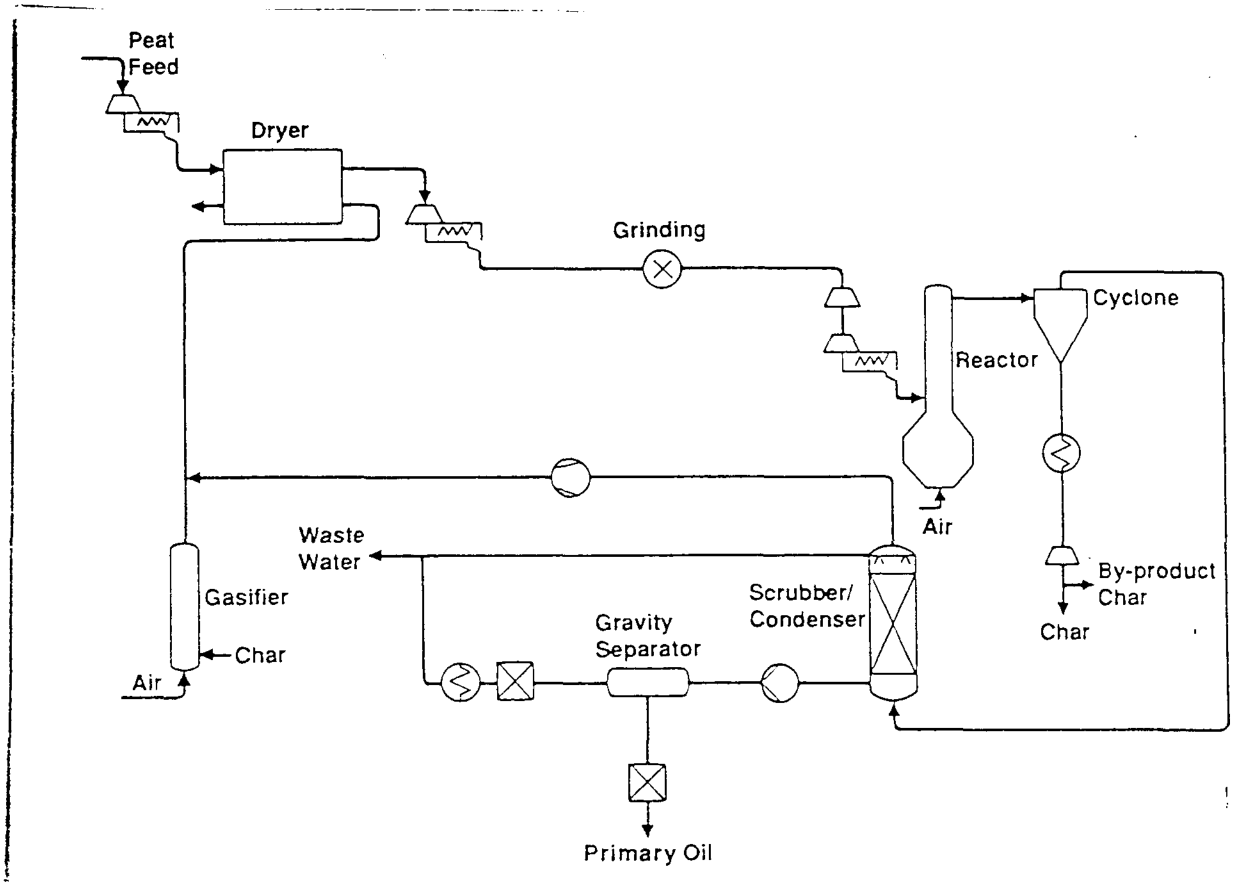

Figure 12. Process Schematic Peat AFP - Potential Technology

The upgrading step for the Potential case also takes place under nonisothermal catalytic hydrotreating conditions as seen in Figure 5. Process improvements are assumed, through catalyst development, including a stoichiometrically-balanced mass balance where sufficient byproduct gas is produced to provide for needed hydrogen production. In addition, a higher reactor space velocity is used. The upgraded product is assumed to be primarily a high-octane gasoline requiring no further processing other than distillate fractionation.

Other Processes

In addition to the processes evaluated in this paper, other direct biomass liquefaction processes are under development in a number of laboratories around the world. These processes could not be included in the assessment because of the limited resources of the cooperative activity. Four of these processes are summarized below.

Ablative Fast Pyrolysis. Under development at the Solar Energy Research Institute, Golden, Colorado, this process rapidly pyrolyzes wood sawdust, less than 3 mm particle size, on the heated wall of a vortex reactor tube[6].

Rapid Thermal Processing. In this process, wood particles are rapidly pyrolyzed at a residence time of 400 to 500 milliseconds. The biomass feedstock is mixed with a solid heat transfer medium in a cyclonic reactor to achieve high heatup rates. This process in under development by Ensyn Engineering, Ottawa, Canada[10].

Vacuum Pyrolysis. In this process wood chips are pyrolyzed under vacuum in a multiple hearth reactor. Condensation of individual streams from the hearths allows primary separation of the liquid products. This process is under development at Université Laval, Québec, Canada[16].

Catalytic Cracking of Pyrolyzates. Also under development at the Solar Energy Research Institute is an alternative upgrading method utilizing ZSM-5 zeolite catalyst to convert pyrolyzates to gasoline-range aromatic hydrocarbon at low pressure and without added hydrogen[5]. Previous tests with this type of processing with LIPS-type oils did not give good results[17].

RESULTS OF THE ASSESSMENT

Detailed cost estimates for the direct biomass liquefaction processes were made based on the procedures described above. Estimates for both capital and operating costs were prepared. With process designs balanced for both mass and energy, total product costs could be calculated using the predetermined assessment bases.

Capital Costs

A summary of the capital costs for the eight process cases is given in Tables 2 and 3. From these numbers, several important conclusions can be drawn. Considering first Table 2 and the wood processes, the largest difference is found in the potential for process improvement as demonstrated by the difference between the two Present cases and the two Potential cases. This difference results from improvements

TABLE 2
SUMMARY OF CAPITAL COSTS FOR WOOD LIQUEFACTION PROCESS CASES

	AFP PRESENT	AFP POTENTIAL	LIPS PRESENT	LIPS POTENTIAL
	Millions, US$			
Fixed Capital Investment (FCI)				
Primary Liquefaction	38.9	20.6	65.8	37.8
Crude Upgrading	36.4	26.8	20.9	20.3
Product Finishing	11.4	0.5	11.9	0.6
Total	86.7	47.9	98.6	58.7
Total Capital Requirement (TCR)				
Primary Liquefaction	49.8	26.4	84.2	48.4
Crude Upgrading	46.6	34.3	26.8	26.0
Product Finishing	14.5	0.7	15.3	0.7
Total	111.0	61.3	126.2	75.1

TABLE 3
SUMMARY OF CAPITAL COSTS FOR PEAT LIQUEFACTION PROCESS CASES

	AFP PRESENT	AFP POTENTIAL
	Millions, US$	
Fixed Capital Investment (FCI)		
Primary Liquefaction	40.4	30.5
Crude Upgrading	27.2	26.0
Product Finishing	7.5	2.4
Total	75.1	58.9
Total Capital Requirement (TCR)		
Primary Liquefaction	51.7	39.0
Crude Upgrading	34.8	33.3
Product Finishing	9.6	3.0
Total	96.1	75.3

in the primary liquefaction step and, through improvement in the upgrading step, near elimination of the product finishing requirements. The difference in capital cost between AFP and LIPS is not large when considering the complete process, from wood to gasoline. The capital cost of the primary liquefaction step is much higher for the high-pressure (LIPS) process. However, due to the higher crude product quality (less oxygen) and lower volume flow, the upgrading step requires less equipment and has a lower capital cost compared with the AFP process. Product finishing is nearly the same for both processes.

The peat processing data is based on a more limited experimental base. In any case, the AFP process shows less potential for improvement than the wood AFP due to two reasons. First, the Present case upgrading is designed around nonisothermal reactor data for peat as is the Potential case, consequently, the improvement from the two-stage hydrotreating process is not included. Secondly, the peat is not as easily hydroprocessed because of sulfur and nitrogen that must be removed.

TABLE 4
PRODUCTION COST SUMMARY FOR WOOD LIQUEFACTION PROCESS CASES

	AFP PRESENT	AFP POTENTIAL	LIPS PRESENT	LIPS POTENTIAL
	Millions US$ per year			
Fixed Operating Costs				
Operating Labor	6.68	6.43	5.58	4.73
Maintenance Labor	0.86	0.49	0.99	0.59
Overheads	1.74	0.96	1.98	1.18
Maintenance Materials	2.60	1.44	2.96	1.76
Taxes & Insurance	1.74	0.96	1.98	1.18
Others	0.86	0.49	0.99	0.59
Total	14.48	10.77	14.48	10.03
Variable Operating Costs				
Feedstock	20.00	20.00	20.00	23.30
Electricity	4.89	3.57	9.72	10.14
Catalyst	0.30	0.02	2.36	0.01
Sludge Disposal	0.16	0.02	0.04	0.09
Fuel Cost	0.39	0.06	1.32	0.06
Total	25.74	23.67	33.44	33.60
Capital Charges	12.96	7.17	14.75	8.78
Total Production Cost, Finished Product	53.18	41.61	62.67	52.39

Wood Price, US$/tonne, (50% moisture)	30
Interest Rate, %	10
Electricity, US$/kWh	0.065

Product Costs

Annual operating costs for the six cases are summarized in Tables 4 and 5. The basis for the calculation (feedstock price, interest rate, and electricity price) is given in the table and is defined as the base case. Feedstock cost is the single largest cost item. The cost of capital is the next largest operating cost but is less than 1/4 of the total in the Present cases and decreases to about 1/6 of total operating cost in the Potential cases.

TABLE 5
PRODUCTION COST SUMMARY FOR PEAT LIQUEFACTION PROCESS CASES

	AFP PRESENT	AFP POTENTIAL
	Millions US$ per year	
Fixed Operating Costs		
Operating Labor	6.63	6.43
Maintenance Labor	0.74	0.59
Overheads	1.50	1.18
Maintenance Materials	2.25	1.77
Taxes & Insurance	1.50	1.18
Others	0.74	0.59
Total	13.36	11.74
Variable Operating Costs		
Feedstock	13.33	13.33
Electricity	4.54	2.93
Catalyst	0.25	0.26
Sludge Disposal	0.05	0.05
Fuel Cost	0.08	0.05
Total	18.25	16.62
Capital Charges	11.23	8.81
Total Production Cost, Finished Product	42.79	37.17

Peat Price, US$/tonne, (50% moisture)	20
Interest Rate, %	10
Electricity, US$/kWh	0.065

Product costs are presented in Table 6. These costs clearly show the advantage of the AFP process for producing a low-cost primary oil product, and this conclusion carries through to comparisons of product cost on an energy basis and on a product-value basis. The costs of the finished products are much closer in value, but the AFP process still has a

TABLE 6
COMPARISON OF BIOMASS LIQUEFACTION COSTS - BASE CASE ASSUMPTIONS

	WOOD				PEAT	
	AFP PRESENT	AFP POTENTIAL	LIPS PRESENT	LIPS POTENTIAL	AFP PRESENT	AFP POTENTIAL
Product Costs, US$/tonne						
Primary Product	150.0	131.9	423.3	298.1	132.9	101.8
Final Product	673.8	514.5	815.0	584.7	665.2	418.4
Lower Heating Values, GJ/tonne						
Primary Product	16.1	19.1	31.5	24.3	17.2	17.2
Final Product	41.5	39.6	41.7	39.6	41.6	41.7
(GASOLINE FRACTION)	(0.8)	(0.9)	(0.7)	(0.9)	(0.7)	(0.9)
Energy Costs, US$/GJ						
Primary Product	9.32	6.91	13.44	12.27	7.73	5.92
Final Product	16.24	12.99	19.54	14.77	15.99	10.03
Product Energy Value(1), US$/GJ						
Primary Product	6.70	6.70	5.20	5.20	5.20	5.20
Final Product	8.30	8.45	8.15	8.45	8.15	8.45
Product Cost/Value Ratio						
Primary Product	1.39	1.03	2.58	2.36	1.49	1.14
Final Product	1.96	1.54	2.40	1.75	1.96	1.19

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(1)Average spot market price from 1977 to 1987 for comparable petroleum-based liquid fuel.

lower cost. Since there is no significant difference in the composition of the finished products, comparisons of finished product cost on the basis of tonnage, energy, or value result in the same conclusion. For both processes, the Potential technology case demonstrates a significant advantage in lower product cost.

Process Energy Efficiencies

Calculated energy efficiencies for the four process cases are given in Table 7. Efficiencies of nearly 70% are achievable in the primary liquefaction. Finished product efficiencies of 50% to 60% are expected. The AFP process appears to be more efficient in all cases. The data also show no significant energy value in the byproducts from the wood liquefaction processes, as the efficiencies are the same whether considering liquid products alone or all products. The peat processes, however, produce a larger amount of byproduct char.

Product Cost Sensitivities

Sensitivities of the product costs have been examined in certain cases. The effect of feedstock price is a major consideration for these processes. As noted above, the feedstock is the single largest operating cost and has the largest impact on product cost. In Figure 13, Final Product Costs are plotted for a range of feedstock prices. The base case price of 30 US$/tonne of wood chips (50% moisture) is indicated. The base case feedstock price is typical for Scandinavia and is about double the market rate in North America. As seen in Figure 13, the Final Product Cost is strongly dependent on feedstock price. All four process cases react to changes in the feedstock price in a similar manner.

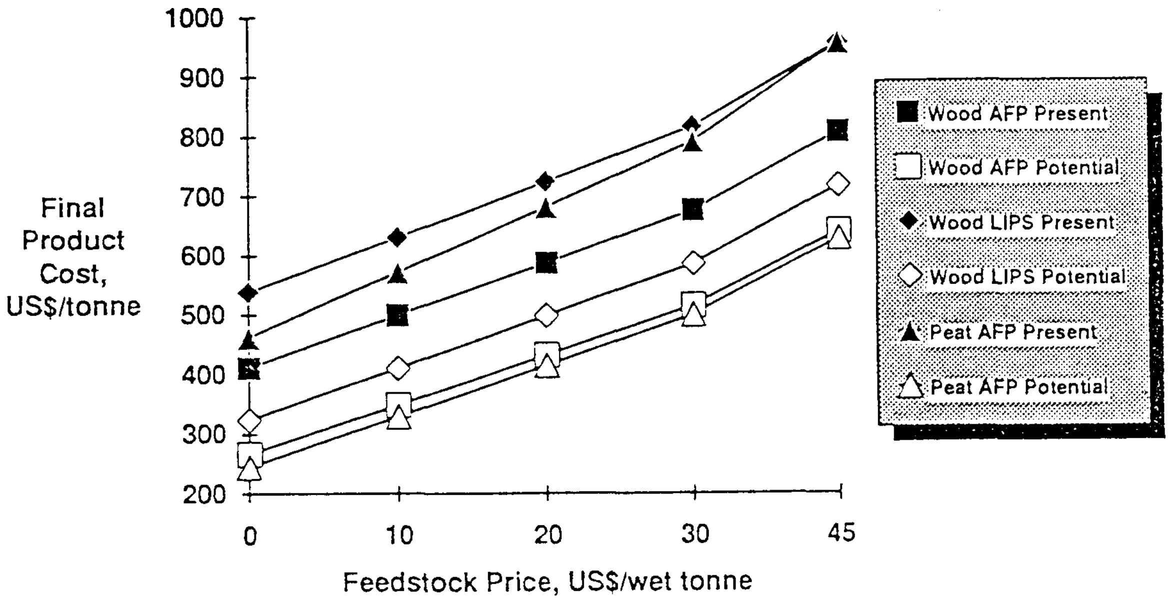

Figure 13. Feedstock Price Sensitivities

TABLE 7
COMPARISON OF PROCESS ENERGY EFFICIENCIES - LOWER HEATING VALUES BASIS

	WOOD				PEAT	
	AFP PRESENT	AFP POTENTIAL	LIPS PRESENT	LIPS POTENTIAL	AFP PRESENT	AFP POTENTIAL
Primary Liquefaction						
Feed to Liquid Product	0.62	0.69	0.59	0.51	0.72	0.72
Total Energy to Liquid Product	0.61	0.68	0.55	0.48	0.69	0.70
Total Energy to Products	0.61	0.68	0.55	0.48	0.83	1.01
Finished Product						
Feed to Liquid Product	0.57	0.55	0.55	0.53	0.56	0.73
Total Energy to Liquid Product	0.52	0.53	0.48	0.49	0.54	0.70
Total Energy to Products	0.52	0.53	0.48	0.49	0.67	1.00

The effect of capital cost (FCI) on the Final Product Cost is plotted in Figure 14. The range of ±20% was selected considering the accuracy of the equipment cost estimates used in this study. Within this range of capital cost variation, the Final Product Costs vary by less than 8% for the Present process cases and 5% for the Potential process cases. Therefore, the Present process cases are somewhat more sensitive to changes in the capital cost than the Potential process cases.

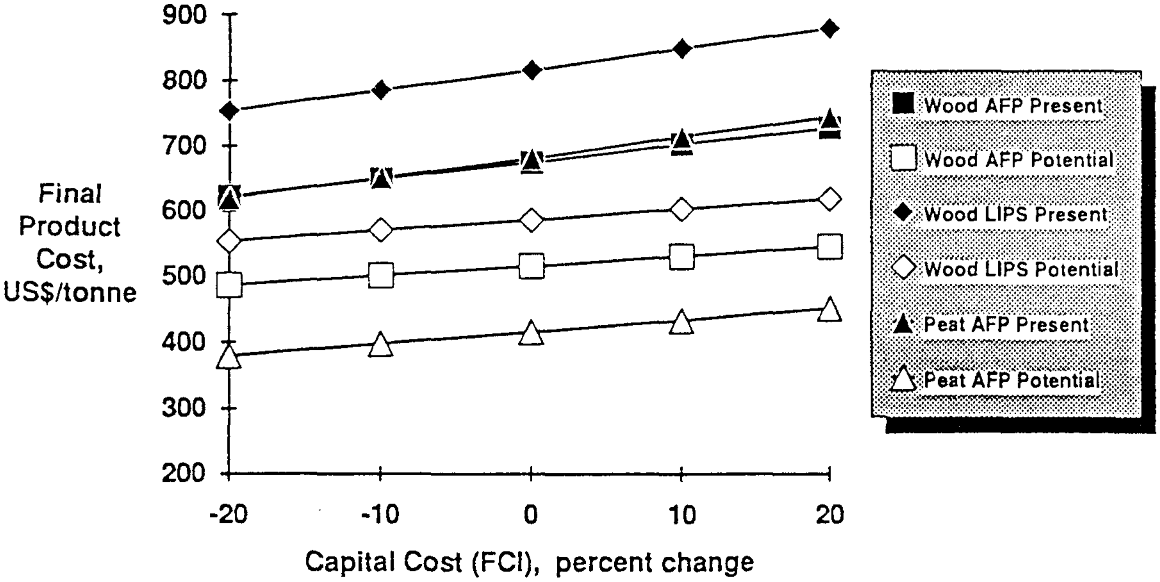

Figure 14. Capital Cost Sensitivities

CONCLUSIONS

With direct liquefaction of biomass, synthetic gasoline or diesel oil can be produced with high efficiency (up to 50 to 60%). This efficiency applies to both the AFP process and the LIPS process.

The most promising Present process case for the production of primary liquid (boiler fuel oil) is the AFP process. Using the average spot market price for fuel oil in the period 1977-1987 as a reference, the ratio of product cost to product value is about 1.4 at a wood chip price of 30 US$/tonne (50% moisture), and decreases to below 0.9 for a lower cost of 10 US$/tonne. That is, in this latter case, the process yields a product that is competitive with an equivalent amount (on an energy basis) of petroleum-derived fuel with similar combustion properties.

For production of gasoline by direct liquefaction, the AFP process also appears to be the most economical. With present technology, the estimated ratio between the product cost and the product value (defined as average spot market price for 1977-1987) is about 2.0 for a feedstock price of 30 US$/tonne (50% moisture).

On the basis of sensitivity studies, all the biomass liquefaction processes appear to be more sensitive to feedstock price than capital cost. Although thermochemical conversion processes are generally viewed to be capital-intensive, the range of study of capital cost variability had little effect on the final product cost. However, the cost of feedstock was found to be the largest operating cost, and the final product cost was found to be highly sensitive to feedstock price.

There appears to be a development potential for both the AFP and LIPS processes, allowing a possible cost reduction of 20% to 30% for the products. Of the Potential liquefaction technologies, the AFP appears as the most proven. For the Potential process based on LIPS, the main uncertainty applies to scaleup from laboratory tests, in particular the possibility to achieve the performance measured in laboratory experiments in full-sized reactors. The main uncertainty about use of the Potential AFP process for gasoline production applies to upgrading the primary product.

As far as the most promising strategy for future research and development is concerned, the Working Group concludes that the most promising process for gasoline production by direct liquefaction of biomass is atmospheric flash pyrolysis. The high-pressure processes may have the same future potential, but the research needs and the uncertainties are much greater. It can therefore be recommended that further cooperative research and development should be focused on the AFP process.

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