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BRITISH COLUMBIA
HYDRO AND POWER AUTHORITY

HAT CREEK COAL UTILIZATION STUDY

November 1981

By

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APPENDIX D.

BRITISH COLUMBIA HYDRO AND POWER AUTHORITY
HAT CREEK COAL UTILIZATION STUDY

EXECUTIVE SUMMARY

1. The studies reported below were performed by North American Mining Consultants, Inc. in association with companies of the Stone & Webster Organization. Canadian Resourcecon Limited were sub-contracted to prepare data on the Canadian liquid fuel and petrochemical markets for use in the study.

2. The purpose of the study was to identify and investigate potential uses for Hat Creek coal which could be considered as alternative or complementary to its use for the production of electric power; to describe those processes which appeared to offer technically and economically viable possibilities; and to describe potential markets for the coal and its conversion products.

3. The ground base for this study was laid by a similar study completed in 1977 [Ref. 1] and this report is essentially an update of that earlier study taking into account the additional data that has become available from mining studies; investigations of coal beneficiation; coal quality and properties related to conversion processing; major changes that have occurred in the Canadian energy resource situation; and significant developments in coal processing technology.

4. The work reported here is based upon information and reports supplied by B.C. Hydro and Power Authority, literature reviews, interviews and enquiries. No field work at Hat Creek and no laboratory, pilot or demonstration work has been undertaken.

5. Attention is drawn to the following general remarks:

(i) Technology

Since the previous report, major research and development work has continued in the United States, West Germany, Great Britain, Australia, and South Africa, with increasing contributions by other countries including Canada. However, the preponderant developments, particularly to commercialization of new coal gasification and liquefaction processes, have been in the United States and West Germany. South Africa has completed the Lurgi-based SASOL II and is building SASOL III; and has developed the Koppers-Totzek based ammonia plant of AECI to reliable operation. Every effort has been made to keep abreast of progress in these countries and to present the status as it exists at March 1981.

(ii) Environmental Engineering

The Government of British Columbia has not yet issued specific regulations governing the design and operation of coal conversion plants. Specific discussions have been held with members of the environmental departments of the provincial government and B.C. Hydro and Power Authority,and Section 6 of this report is partly the outcome of those discussions and partly the outcome of applying what is presently regarded as the best available control technology.

(iii) Capital and Operating Costs

Apart from the plants in South Africa, no full-scale commercial coal conversion plants have been built since completion of the previous study. However, some demonstration scale plants have been constructed in the United States and Germany, and a number of organizations, including the NAMCO parent organizations, have produced cost estimates for the majority of the processes selected for investigation in this report. This data, after suitable amendment for scale, inflation, and location at Hat Creek, has formed the basis of the present economic evaluation. But it must be emphasized that, in the absence of construction and operation of full-scale commercial installations in North America, the costs employed in this report are not equivalent to those derived from detailed and final engineering design and construction for any of the cases considered.

(iv) Product Markets

In considering markets, prices, and revenues for potential coal-conversion products, provincial, North American, and world reference frames have been employed as appropriate to the particular product considered. The impact of the most recent Canadian energy policy statement, as outlined by the Federal Finance Minister (October 1980), has been taken into account, but, and again it must be emphasized, no detailed market studies were carried out for any of the products covered in this report.

6. Conversion Processes - Basis of Selection

The processes selected for economic evaluation in this study are listed in Table 3.1. After preliminary discussions with B.C. Hydro it was determined that the various processes to be selected for evaluation should be to a common base equivalent to 316.5 TJ/day, equivalent to 8000 m^3/day (50,000 b/d) liquid fuels, this size approximating the coal energy requirements of a 2000 MW_2 electric power generating plant. This list is the result of a systematic elimination of a large number of potential processes for converting Hat Creek coal to upgraded solid, liquid or gaseous products. The reasons for elimination have been discussed separately in Sections 2, 3, and 4 of this report, but are summarized as follows:

Table 3.1 Processes Selected for Evaluation

			Appendix D	
Process Type	Process Description	Case	Figure	Table
A. Direct hydrogenation*	H-Coal	A1	D1.1	3.3
	EDS	A2	D1.2	3.3
	SRC-II	A3	D1.3	3.3
	SRC-I	A4	D1.4	3.3
B. Fischer-Tropsch synthesis	Texaco gasification	B1	D1.5	3.4
	Koppers gasification	B2	D1.6	3.4
	Winkler gasification	B3	D1.7	3.4
	Lurgi (Sell Fines) gasification	B4	D1.8	3.4
	Lurgi (Maximum Power) gasification	B5	D1.9	3.4
	Lurgi & Texaco combination	B6	D1.10	3.4
	Lurgi & Koppers combination	B7	D1.11	3.4
C. Methanol synthesis	Texaco gasification	C1	D1.12	3.6
	Koppers gasification	C2	D1.13	3.6
	Winkler gasification	C3	D1.14	3.6
	Lurgi (Sell Fines)	C4	D1.15	3.6
	Lurgi (Maximum Power)	C5	D1.16	3.6
	Lurgi & Texaco combination	C6	D1.17	3.6
	Lurgi & Koppers combination	C7	D1.18	3.6
D. Methanol to Gasoline	Texaco gasification	D1	D1.19	3.7
	Koppers gasification	D2	D1.20	3.7
	Winkler gasification	D3	D1.21	3.7
	Lurgi (Sell Fines)	D4	D1.22	3.7
	Lurgi (Maximum Power)	D5	D1.23	3.7
	Lurgi & Texaco combination	D6	D1.24	3.7
	Lurgi & Koppers combination	D7	D1.25	3.7
E. Production of synthetic natural gas (SNG)	Methanation based on Lurgi gasification	E1	D1.26	3.10

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*Cases A1, A2 and A3 produce liquid fuels and Case A4 (SRC-I) produces mainly clean solid boiler fuel by hydrogenation.

A. Coal-related factors

Any conversion process considered must be compatible with or adaptable to the particular properties of Hat Creek coal.

B. Process-related factors

The time frame of the study requires that any processes considered must be already commercially proven or at such an advanced stage of demonstration that engineering design of a commercial plant could reasonably be undertaken at the present time.

C. Market-related factors

The primary questions addressed in considering market opportunities for Hat Creek coal conversion products have been to determine whether a demand presently exists or will develop in the foreseeable future, the size and location of that demand, and the potential revenues. These considerations have been influenced by the extremely large reserves of natural gas that have been discovered in the Province, the small and diminishing market for heavy fuel oils, the excess capacity situation in the world nitrogenous fertilizer market, and the continuing concern over the future supplies of transportation fuels. Other factors have been the absence of a developed industrial market in the immediate vicinity of Hat Creek and the necessity for transportation of the products to the Lower Mainland. In general, it has appeared clear that the primary thrust for conversion processes, as an alternative to electric power generation, should be directed to production of liquid fuels.

Table 5.0 presents a summarized breakdown of the procedures employed for process elimination and selection.

Table 5.0 Process Selection - Utilization of Hat Creek Coal

Type of Process	Related Factors	Decision
I Solid Application		
a) Direct Combustion	Studied in separate evaluation	Study has objective of evaluation options which are competitive with combustion of Hat Creek coal.
b) Beneficiation	Coal unsuitable for beneficiation	Beneficiation studies not included.
c) Active Carbon	Application of coal to water treatment taking advantage of ion exchange capability.	Study excluded because of lack of local market for application.
d) SRC	Solid solvent refined coal (SRC) is produced by the hydrogenation of coal at low severity.	Rejected in favor of more severe hydrogenation processes which make liquid products.
II Gaseous Fuel Production		
a) Low BTU Gas	Lack of local industrial market for products rules against production from Hat Creek coal.	Rejected LBG as method of utilization.
b) High BTU Gas		Rejected HBG as method of utilization
c) Substitute Natural Gas	Appears to be an unattractive option in view of abundant supplies in B.C. at present.	Evaluate comparative economics of SNG vs. coal liquefaction. Base study on Lurgi gasification process.

Table 5.0 (continued)

Type of Process	Related Factors	Decision
II Gaseous Fuel Production (continued)		
Gasification Processes		
i) Lurgi Process	Hat Creek coal is reported, on the basis of laboratory tests, to be a satisfactory feedstock for Lurgi Fixed Bed gasifiers but this is contingent on the removal of fines smaller than 13 mm from the feed. These constitute a substantial proportion of total coal to plant and economic factors are critically dependent on satisfactory utilization of the fines.	Evaluate indirect coal liquefaction based on: a) combustion of excess fine coal and export of any excess power b) sale of excess fine coal. c) gasification of excess fine coal by the Koppers and Texaco processes. Evaluate SNG production based on Lurgi gasification process.
ii) Koppers Process	Koppers entrained bed process would be satisfactory for the gasification of Hat Creek coal. However, low pressure gasification is expected to be uneconomical as basis for indirect coal liquefaction.	The Koppers Process could be used as a feasible alternative to the Lurgi process. Perform economic evaluation based on best available data for indirect coal liquefactio
iii) Texaco Process	The Texaco, entrained bed, high pressure process based on water slurry feed is in the developmental phase. Application to Hat Creek coal is a possibility but this could only be proved based on a pilot test comparing it with other gasifiers.	Estimate economics of indirect coa liquefaction using the best available data. Perform evaluation in order to recognize if this second generation technology offers potential future advantages to B.C. Hydro.
iv) Winkler Process	Hat Creek coal is potentially gasifiable by a fluidized bed process and closely related data based on lignite gasification is available.	Develop order of magnitude comparative economics based on Winkler gasification applied to indirect coal liquefaction.

Table 5.0 (continued)

Type of Process	Related Factors	Decision
II Gaseous Fuel Production (continued)		
Gasification Processes		
v) Other Processes		
British Gas/Lurgi Slagging Gasifier Cogas Shell Koppers		Excluded from consideration because these processes are not available for application to Hat Creek coal in the near future.
III Liquid Fuel Production		
i) Direct coal Liquefaction	Hat Creek coal has been shown to liquefy readily and with high carbon conversion efficiencies in laboratory tests of direct hydrogenation. It is a potential feedstock for processes derived from Potte Broche (uncatalyzed) and Bergius (catalyzed) processes. However this is contingent on satisfactory removal of the mineral residues (ash).	Include direct hydrogenation by SRC-I, SRC-II, H-Coal and EXXON Donor Solvent processes. Exclude long-term prospects such as the DOW coal liquefaction process and the EPRI short residence time approach. Though it could be applicable, the Consol (Conoco) process was excluded from consideration because of the lack of current activity in its development.
NCB Processes Super Critical Gas Extraction and Liquid Solvent Extraction	Preliminary investigations have been carried out of applicability to Hat Creek coal. Only the supercritical extraction results appeared to be unsatisfactory as a basis for commercial application.	Economic study not carried out because processes are not available for commercial application in the near future.

Table 5.0 (continued)

Type of Process	Related Factors	Decision
ii) Indirect Coal Liquefaction Processes for production of liquid fuels based on synthesis gas feedstocks.		
a) Fischer-Tropsch	Process is well established in commercial service based on low grade coal (South Africa). Process lacks capability to satisfy specific synthetic fuel needs. However, it could assume national importance as a source of synthetic fuels.	Evaluate order of magnitude economics of Fischer-Tropsch technology applied in association with above selected gasifiers viz. Lurgi, Koppers, Texaco and Winkler.
b) Methanol	Methanol production processes are well established. For all practical purposes methanol production based on coal equals Fisher-Tropsch technology in commercial feasibility.	Evaluate order of magnitude economics of methanol technology with above selected gasifiers (as for Fisher-Tropsch).
c) Methanol Conversion to Gasoline	In view of the lack of a general methanol based economy, the conversion of methanol to gasoline may be the best near term policy for utilizing coal gasification. Process has excellent specifications for producing motor fuels.	Evaluate order of magnitude impact of producing gasoline rather than methanol. Use best available published data on the Mobil process.
IV Pyrolysis	Poor yield of liquid products obtained in Fisher Assay test.	Rejected pyrolysis as method of utilization of Hat Creek coal.

7. Coal Quality

The coal quality specification, provided by B.C. Hydro, on which the study has been based, is shown in Table 2.1.

Other coal properties important in coal processing are discussed in detail in Section 2.

8. Comparative Process Performance

Comparative process performance data for these conversion processes listed in Table 3.1 are summarized in Table X. The values shown are based on one tonne of Performance Blend coal.

Where no value is shown for electrical power, the process is self-sufficient. A positive value indicates a small export surplus, values in parantheses indicate a net import requirement. In all cases, emergency power generation necessary to ensure smooth shutdown of plant is provided within the battery limits.

9. Coal Conversion Product Markets

(i) The supply, demand, capacity, and prices of non-renewable energy resources with which coal-conversion products could supplement, substitute or compete are discussed in Section 4. The products discussed are -

Crude Oil
Natural Gas
Liquified Petroleum Gases
Ethylene
Motor and Aviation Gasolines
Diesel Fuels
Methanol
Methyl Tertiary-Butyl Ether (MTBE)
Benzene, Toluene, and Xylenes
Petroleum Coke
Ammonia

TABLE 2.1 HAT CREEK COAL QUALITY

	"PERFORMANCE" BLEND		"DESIGN WORST" BLEND		LURGI PLANT SUPPLY					
					GASIFIER (+13MM)		FINES (-13MM) STEAMCOAL		TOTAL	COAL
	AS REC'D BASIS	DRY BASIS	AS REC'D BASIS	DRY BASIS	AS REC'D BASIS	DRY BASIS	AS REC'D BASIS	DRY BASIS	AS REC'D BASIS	DRY BASIS
YIELD	100	76.5	100	77.5	64	48.64	36	27.90	100	76.5
PROXIMATE ANALYSIS:										
Moisture	23.5	-	22.5	-	24.0	-	22.5	-	23.5	-
Volatile Matter	25.2	32.94	24.0	30.97	26.60	35.00	22.86	29.5	25.2	32.94
Ash	25.6	33.46	29.5	38.06	22.23	29.25	31.78	41.0	25.6	33.46
Fixed Carbon	25.7	33.60	24.0	30.97	27.17	35.75	22.86	29.5	25.7	33.60
	100.0	100.00	100.0	100.00	100.00	100.00	100.00	100.0	100.00	100.00
GROSS CALORIFIC VALUE (HHV)										
MJ/Kg	13.85	18.10	12.71	16.40	14.88	19.58	12.07	15.58	13.85	18.10
Btu/lb.	5,955	7,784	5,465	7,052	6,398	8,419	5,190	6,699	5,955	7,784
ULTIMATE ANALYSIS:										
Carbon	35.30	46.14	32.40	41.82	37.29	49.06	31.70	40.91	35.30	46.14
Hydrogen	2.80	3.66	2.70	3.48	2.96	3.89	2.52	3.25	2.80	3.66
Nitrogen	0.70	0.92	0.70	0.90	0.74	0.98	0.63	0.81	0.70	0.92
Sulfur	0.39	0.51	0.55	0.71	0.41	0.54	0.35	0.45	0.39	0.51
Chlorine	0.02	0.03	0.03	0.04	0.03	0.04	0.02	0.03	0.02	0.03
Oxygen (Difference)	11.69	15.28	11.62	14.99	12.34	16.24	10.50	13.55	11.69	15.28
(Moisture)	(23.50)	-	(22.50)	-	(24.00)	-	(22.50)	-	(23.50)	-
(Ash)	(25.60)	33.46	(29.50)	38.06	(22.23)	29.25	(31.78)	41.00	(25.60)	33.46
	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00

[Ref. 3]

Table X Performance Data Summary - Estimates Based on One Tonne of Performance Blend Coal

	Thermal Efficiency Percent	← Product Fuels GJ/tonne →					Water Consumption m^3/tonne
		Gas	Liquid	Solid	Coal Fines	Elec. Power (consumed)	
Direct Coal Liquefaction							
H-Coal	60	-	8.3	-	-	-	0.76
EDS	50	-	6.9	-	-	-	0.69
SRC-II	60	-	8.3	-	-	-	0.76
SRC-I	63	-	1.74	6.98	-	-	0.69
Indirect Coal Liquefaction							
Fischer-Tropsch							
Based on Texaco	35.2	0.14	4.38	-	-	0.12	0.74
Based on Koppers	31.3	0.14	4.38	-	-	(0.12)	0.72
Based on Winkler	30.7	0.14	4.27	-	-	(0.18)	0.57
Based on Lurgi (max. power)	36.4	0.15	4.62	-	-	0.10	0.75
Based on Lurgi (sell fines)	35.1	0.15	4.62	-	1.95	(0.59)	0.69
Based on Lurgi & Texaco	35.6	0.17	5.23	-	-	(0.46)	0.80
Based on Lurgi & Koppers	35.1	0.17	5.23	-	-	(0.53)	0.80
Methanol							
Based on Texaco	50.9	-	6.67	-	-	0.14	0.73
Based on Koppers	42.1	-	6.63	-	-	(0.63)	0.72
Based on Winkler	48.3	-	7.13	-	-	(0.29)	0.58
Based on Lurgi (max. power)	49.0	-	6.79	-	-	0	0.90
Based on Lurgi (sell fines)	49.0	-	6.79	-	1.90	(0.67)	0.76
Based on Lurgi & Texaco	49.4	-	7.65	-	-	(0.57)	0.90
Based on Lurgi & Koppers	47.6	-	7.69	-	-	(0.78)	0.90

Table X (continued)

Indirect Coal Liquefaction (continued)

	Thermal Efficiency Percent	← Product Fuels GJ/tonne →					Water Consumption m^3/tonne
		Gas	Liquid	Solid	Coal Fines	Elec. Power (consumed)	
Methanol to Gasoline Conversion							
Based on Texaco	47.1	1.07	5.21	-	-	0.09	0.60
Based on Koppers	39.8	1.06	5.18	-	-	(0.68)	0.60
Based on Winkler	44.7	1.12	5.47	-	-	(0.31)	0.40
Based on Lurgi (max. power)	46.6	1.04	5.08	-	-	0.12	0.67
Based on Lurgi (sell fines)	45.2	1.04	5.08	-	2.06	(0.62)	0.62
Based on Lurgi & Texaco	45.6	1.20	5.84	-	-	(0.56)	0.64
Based on Lurgi & Koppers	44.4	1.21	5.88	-	-	(0.74)	0.64
Substitute Natural Gas							
Based on Lurgi (max. power)	60.0	8.31	-	-	-	-	1.3

The data cannot be conveniently summarized in a single table, and reference is made to Tables 4.1 - 4.50 of Section 4.

(ii) The marketing studies are in general terms and are not specific and detailed for any single product. However, they support the conclusion that conversion to light liquid (transportation) fuels and chemical intermediates is the preferred conversion option. In accordance with the Canadian Energy Policy (October 1980) it is assumed that product prices will correspond with world posted crude oil prices by the time production would commence (1988/1989).

(iii) There is a limited and diminishing market for heavy fuel oils in the Province, and indeed in Canada as a whole as a result of recently pursued policies to restrict oil consumption in electric power generation. It is assumed that this progression will not now be reversed, and has an important bearing on reducing the attractiveness of the processes for coal conversion to liquids by direct hydrogenation.

(iv) The appreciable and continuing surplus of world capacity over demand for ammonia and nitrogenous fertilizers renders ammonia production at Hat Creek unattractive. Any ammonia produced will meet a highly competitive market situation based mostly on natural gas feedstocks.

10. Comparative Process Economics

(i) The method and requirements of the economic analysis are outlined in the Terms of Reference (Appendix A) and the Financial Economic Criteria (Appendix B).

(ii) The latter required that economic comparison is based on Net Present Values of Gross Profit Margins for project lives of 30 years at discount rates variable in the range 3 - 10 percent. B.C. Hydro indicated that these interest rates, similarly applied to coal mining costs, would increase delivered coal costs in the range $10 - $14 per tonne, and levels of $10, $12, and $14 have been employed.

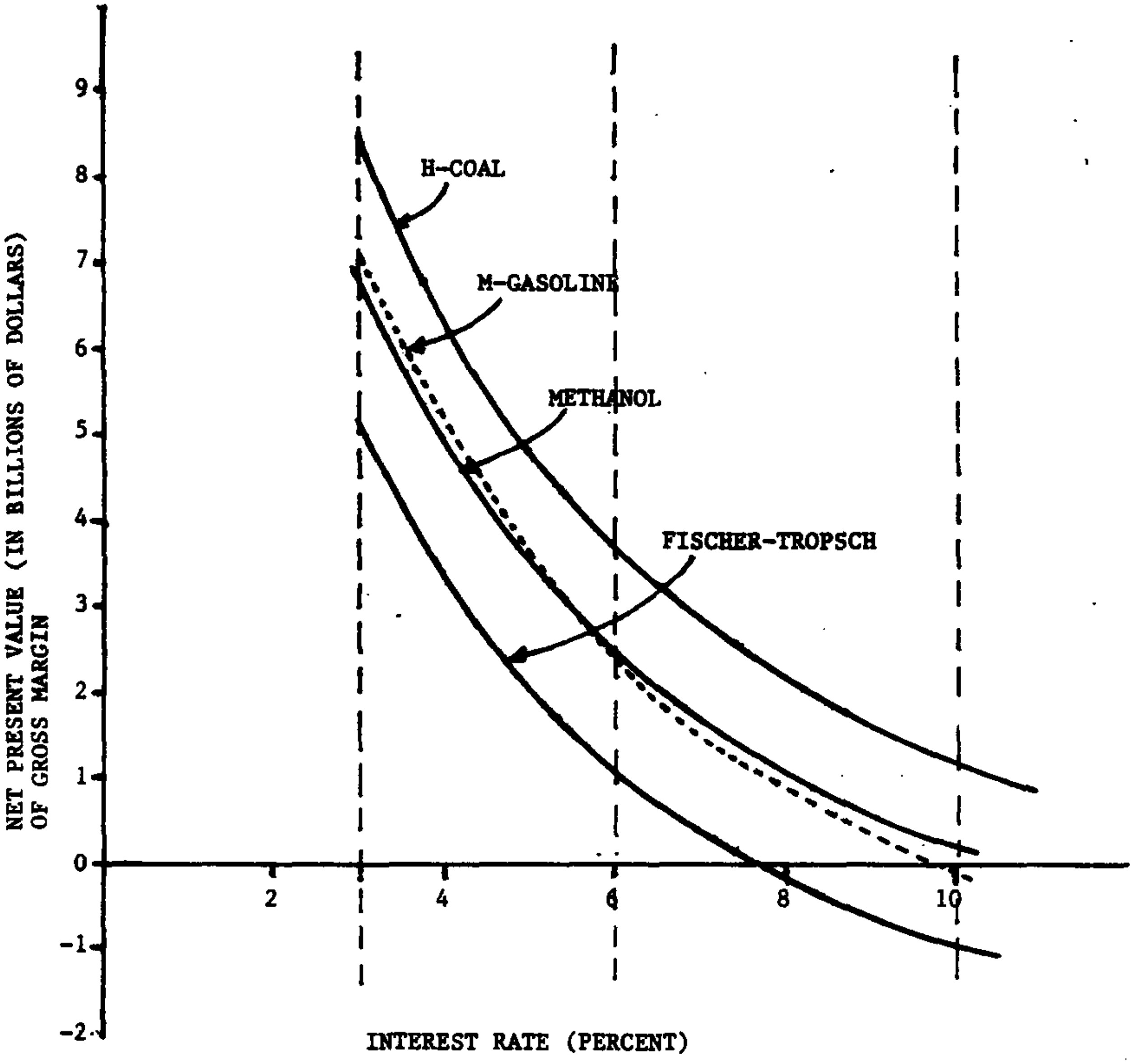
(iii) Additionally, preliminary sensitivity analyses have been performed which allow increases of up to +100 percent in capital costs estimates and reductions of up to -40 percent in revenues estimates.

(iv) A complete listing of financial criteria and assumptions employed is provided in Section 5.2.

(v) The financial results of applying these criteria and assumptions to the twenty-six cases listed in Table 3.1 are shown in Tables 5.6 to 5.9A of the report. The cases providing the largest surpluses of revenues over operating costs (gross margin) are as follows -

Process Type	Process Description	Case No.
Direct Liquefaction	H-Coal	A1
Indirect Liquefaction		
Fischer-Tropsch	Lurgi (Max Power)	B5
Methanol	" " "	C5
Mobil-Gasoline	" " "	D5

These cases were, therefore, selected for a more detailed financial sensitivity analyses, the results of which are summarized in Tables 5.10 - 5.13 and Figures 5.1 - 5.8 of the report, and in Figure 5.9 on the following page.



PROCESS	CAPITAL COST	ANNUAL SALES
H-Coal	$2208 million	$ 991 million
Fischer-Tropsch	$4347 million	$1060 million
Methanol	$3117 million	$ 979 million
M-Gasoline	$3623 million	$1066 million

ALL INDIRECT LIQUEFACTION CASES ARE BASED ON LURGI (MAKE POWER) PROCESS ECONOMICS.

B. C. HYDRO

FIG. 5.9

EFFECT OF INTEREST RATE ON NET PRESENT VALUE BASE CASE CAPITAL COSTS AND SALES REVENUE

NJ 1024

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(vi) A similar previous study, completed in 1977, had found that conversion of Hat Creek coal to synthetic natural gas (SNG) offered advantages over other processes in the then existing state of energy resources and markets. Since that time, the natural gas resources outlook has immensely improved but, nevertheless, a financial analysis was carried out in direct comparison with those cases listed in the previous paragraph. The results are summarized in Table 5.14, assuming that the SNG would be sold at a price 4.27/GJ ($4.50 MMBTU) comparable with the current export price.

(vii) In interpreting the results in these tables and figures it should be borne in mind that the financial analysis does not make provisions for taxation rates applicable to corporations and does not allow for depreciation. With these provisos, the results indicate that financial viability of the selected processes may be ranked

H-Coal
Methanol
M-Gasoline
Fischer-Tropsch.

(viii) For reasons stated earlier, the present study assumes that each process listed is equally technically viable at the present time.

(ix) Also as previously stated, the H-Coal Process is considered less viable in marketing terms because it produces substantial proportions of less-marketable heavy fuel oil.

(x) The ultimate ranking is therefore -

Methanol
M-Gasoline
Fischer-Tropsch

This order compares favorably with current views in the U.S.A. where coal conversion to methanol and M-Gasoline appears to be more favorably regarded, judging by the numbers of proposed projects, to that based on Fischer-Tropsch synthesis.

11. Environmental Impacts

(i) The British Columbian government has not yet issued regulations which deal specifically with the control of emissions from coal conversion plants. This study has, accordingly, considered those regulations governing thermal electrical power generation and petroleum industry processes.

(ii) Processes for conversion of Hat Creek coal studied in this report are considered to be operable within these environmental control regulations, and this study has further considered the best available control technology procedures for estimating the attendant costs.

(iii) It should be noted that one consequence of operating a coal-conversion plant as an alternative to a thermal electric power generating plant at Hat Creek will be a reduction by 30 - 50 percent in the incidence of low grade thermal pollution in the Hat Creek region for similar rates of coal utilization. This difference represents the difference in thermal efficiency and is accounted for by the increased energy content of the conversion products transported from Hat Creek as compared with the electrical energy transmitted.

(iv) In contradistinction, the impact on the environment by importing a large work force for construction will be two to three times greater in the case of coal conversion plant compared with a power station.

(v) The most important items concerned with environmental impact and control are summarized in Table 6.17.

Table 6.17 Comparison of Coal Conversion* and Power Generation** Facilities Operating Parameters

Parameter (t/d)	Koppers-Methanol	Lurgi-Fisher Tropsch (Max. Power)	Direct Liquefaction	Power Generation
Coal Requirement	48,030	66,395	38,095	40,500
Water Requirement	34,600	51,600	28,800	100,200
Particulate Emission	3.4	14.6	3.1	17
Sulphur Emission (as S)	11.4	42.3	10.1	75
Solid Wastes (dry)	12,350	17,290	9,800	10,830

*Plant sizes equivalent to 316.5 TJ/day of coal conversion products

**Equivalent to 2000 MW_e

12 Conclusions

12.1 This study finds that the conversion of Hat Creek coal to light liquid fuels -

(i) is best suited for future energy and petrochemical market needs in British Columbia

(ii) may be technically viable by several alternative process routes but additional investigations and development work will be required in confirmation

(iii) is economically viable within the Terms of Reference and economic/financial criteria imposed on the study.

This utilization offers a technically and economically viable alternative to the use of the coal for thermal electric power generation.

N.B. It should be noted, however, that the provision of a thermal electric power generation plant and/or a coal conversion plant are not mutually exclusive concepts for utilization of the huge coal deposits at Hat Creek. It is considered feasible, given suitably expanded mining plans, to produce sufficient coal to supply both such facilities simultaneously. And while these may be best operated under separate managements there could, by suitable integration of services, be synergistic effects tending to improve the overall utility and economics of the complex. In this context it may be noted that some of the alternative processes considered in this report require local generation of about 700 MWe of electric power.

12.2 The most attractive of these alternate uses and the economic limits within which they are viable, are illustrated in Figures 5.1 - 5.9. These processes, ranked in descending order are -

- (i) H-Coal Process
- (ii) Lurgi/Methanol Synthesis Process
- (iii) Mobil Methanol-to-Gasoline Process
- (iv) Lurgi/Fischer Tropsch Process.

However, the H-Coal Process has not yet been commercially demonstrated to the same degree as the Methanol and Fischer-Tropsch Processes, and its heavier oil products are not as readily marketable.

12.3 The limited market in Western Canada for heavy fuel oils renders the Solvent Refined Coal processes unattractive.

12.4 The conversion of Hat Creek coal to Synthetic Natural Gas (SNG) is not economically viable at current gas export prices.

12.5 There appears to be no opportunities for usefully converting Hat Creek coal to upgraded solid products.

12.6 The low yields of liquid products on pyrolysis of Hat Creek coal preclude its use for combined pyrolysis/thermal electric power generation applications.

12.7 The processes for conversion of Hat Creek coal to light liquid fuels listed in 7.2 can be operated within current environmental control regulations,and this study has included the best available control technology procedures, with attendant costs.

12.8 There have been no developments in coal-cleaning techniques during the past five years which are encouraging for the cleaning and beneficiation of Hat Creek coal in the future. All uses of the coal must, therefore, continue to be predicated on a low-rank, low-grade quality feedstock.

12.9 Although Hat Creek coal exhibits adsorptive and ion exchange properties, its use as a medium for effluent treatment and wastewater purification is not recommended for the following reasons:

(i) These properties are not extraordinary as compared with other commercially available water treatment materials

(ii) the presence of swelling clays in the coal will, themselves, create a serious water treatment problem through dispersion in the water being treated

(iii) the coal cannot be economically regenerated and it is a low-grade fuel

(iv) the coal cannot be prepared and transported to industrial regions for use in water treatment at costs competitive with existing materials and methods.

12.10 The chemically and structurally bound water in the minerals present in Hat Creek coal, by interfering with the Standard method of analyses, may cause misleading proximate and ultimate analyses.

BRITISH COLUMBIA HYDRO AND POWER AUTHORITY
HAT CREEK COAL UTILIZATION STUDY

1. INTRODUCTION

1.1 The studies reported below were performed by North American Mining Consultants, Inc. in association with companies of the Stone & Webster Organization. Canadian Resourcecon Limited were sub-contracted to prepare data on the Canadian liquid fuel and petrochemical markets for use in the study

1.2 The purpose of the study was to identify and investigate potential uses for Hat Creek coal which could be considered as alternative or complementary to its use for the production of electric power; to describe those processes which appeared to offer technically and economically viable possibilities; and to describe potential markets for the coal and its conversion products.

1.3 The ground base for this study was laid by a similar study completed in 1977 [Ref. 1] and this report is essentially an update of that earlier study taking into account the additional data that has become available from mining studies; investigations of coal beneficiation; coal quality and properties related to conversion processing; major changes that have occurred in the Canadian energy resource situation; and significant developments in coal processing technology.

1.4 The Terms of Reference for the work are shown in Appendix A.

1.5 The Economic/Financial Criteria governing the economic analysis are set out in Appendix B.

1.6 A computer program was produced as part of the work to provide an extended sensitivity analysis and to allow rapid estimations to be made of the effect of variation in major elements of the economics of coal conversion. This program is set out in Appendix C.

1.7 The work reported here is based upon information and reports supplied by B.C. Hydro and Power Authority, literature reviews, interviews and enquiries. No fieldwork at Hat Creek and no laboratory, pilot or demonstration work has been undertaken. As recorded in the previous Study [Ref. 1] the available information remains insufficient in some important areas of the work and a number of additional laboratory and pilot investigations will be required before progress from the present conceptual-only stage can be made. Subjects where additional information is required are indicated in the technical reviews. Some of the necessary investigations are in progress for B.C. Hydro by certain other organizations at the present time but final results are not yet available.

1.8 Attention is drawn to the following general remarks:

(i) Technology

Since the previous report, major research and development work has continued in the United States, West Germany, Great Britain, Australia and South Africa with increasing contributions by other countries including Canada. However the preponderant developments, particularly to commercialization of new coal gasification and liquefaction processes, have been in the United States and West Germany. South Africa has completed the Lurgi-based SASOL II and is building SASOL III; and has developed the Koppers-Totzek-based ammonia plant of AECI to reliable operation. Every effort has been made to keep abreast of progress in these countries and to present the status as it exists at March 1981.

(ii) Environmental Engineering

The government of British Columbia has not yet issued specific regulations governing the design and operation of coal conversion plants. Specific discussions have been held with members of the environmental departments of the provincial government and B.C. Hydro and Power Authority and Section 6 of this report is partly the outcome of those discussions and partly the outcome of applying what is presently regarded as the best available control technology.

(iii) Capital and Operating Costs

Apart from the plants in South Africa no full-scale commercial coal conversion plants have been built since completion of the previous Study. However some demonstration scale plants have been constructed in the United States and Germany, and a number of organizations, including the NAMCO parent organizations, have produced cost estimates for the majority of the processes selected for investigation in this report. This data, after suitable amendment for scale, inflation and location at Hat Creek, has formed the basis of the present economic evaluation. But it must be emphasized that, in the absence of construction and operation of full scale commercial installations in North America, the costs employed in this Report do not correspond to those derived from detailed and final engineering design and construction for any of the cases considered.

(iv) Product Markets

In considering markets, prices and revenues for potential coal conversion products, provincial, North American and World reference frames have been employed as appropriate to the particular product considered. The impact of the most recent Canadian energy policy statement, as outlined by the Federal Finance Minister (October 1980) has been taken into account but, and again it must be emphasized, no detailed market studies were carried out for any of the products covered in this Report.

2. GENERAL TECHNICAL CONSIDERATIONS

2.1 Coal Properties

2.1.1 Specification of Coal Quality

Since the previous study [Ref. 1] in which the available analysis and test data on Hat Creek coal were reviewed and interpreted in detail, additional data have become available from the continuing series of laboratory investigations, pilot and full scale trials by a number or organizations on behalf of B. C. Hydro. This additional test work has included hydrogenation liquefaction; pilot scale pulverized coal combustion; commercial scale thermal generating plant trials; pilot washing trials; laboratory scale beneficiation trials employing gravity separation in a dry, fluidized cascade system; and laboratory scale beneficiation trials employing dry, electrostatic separation methods.

Much of the work has confirmed the unusual nature of the Hat Creek coal deposits. In some cases peripheral tests carried out for one purpose have been instrumental in providing explanations for unusual behaviour encountered in another.

A comprehensive review of all the data pertaining to the specification and use of the coal as a boiler fuel was recently reported. [Ref. 2]. This review covered all available analytical and test results for the different quality zones within the mineable deposit and took into account the proposed method of mining [Ref. 3]. On this basis B. C. Hydro provided specifications for Hat Creek Coal Quality shown in Table 2.1.

Referring to Table 2.1, the Design Worst Blend quality, which corresponds to coal from Zones A, B and C, formed the basis for calculating the capacities of the various units of the coal conversion processes considered. After preliminary discussions with B. C. Hydro it was determined that the various processes to be

TABLE 2.1 HAT CREEK COAL QUALITY

	"PERFORMANCE" BLEND		"DESIGN WORST" BLEND		LURGI PLANT SUPPLY					
					GASIFIER (+13MM)		FINES (-13MM) STEAMCOAL		TOTAL	COAL
	AS REC'D BASIS	DRY BASIS	AS REC'D BASIS	DRY BASIS	AS REC'D BASIS	DRY BASIS	AS REC'D BASIS	DRY BASIS	AS REC'D BASIS	DRY BASIS
YIELD	100	76.5	100	77.5	64	48.64	36	27.90	100	76.5
PROXIMATE ANALYSIS:										
Moisture	23.5	-	22.5	-	24.0	-	22.5	-	23.5	-
Volatile Matter	25.2	32.94	24.0	30.97	26.60	35.00	22.86	29.5	25.2	32.94
Ash	25.6	33.46	29.5	38.06	22.23	29.25	31.78	41.0	25.6	33.46
Fixed Carbon	25.7	33.60	24.0	30.97	27.17	35.75	22.86	29.5	25.7	33.60
	100.0	100.00	100.0	100.00	100.00	100.00	100.00	100.0	100.00	100.00
GROSS CALORIFIC VALUE (HHV)										
MJ/Kg	13.85	18.10	12.71	16.40	14.88	19.58	12.07	15.58	13.85	18.10
Btu/lb.	5,955	7,784	5,465	7,052	6,398	8,419	5,190	6,699	5,955	7,784
ULTIMATE ANALYSIS:										
Carbon	35.30	46.14	32.40	41.82	37.29	49.06	31.70	40.91	35.30	46.14
Hydrogen	2.80	3.66	2.70	3.48	2.96	3.89	2.52	3.25	2.80	3.66
Nitrogen	0.70	0.92	0.70	0.90	0.74	0.98	0.63	0.81	0.70	0.92
Sulfur	0.39	0.51	0.55	0.71	0.41	0.54	0.35	0.45	0.39	0.51
Chlorine	0.02	0.03	0.03	0.04	0.03	0.04	0.02	0.03	0.02	0.03
Oxygen (Difference)	11.69	15.28	11.62	14.99	12.34	16.24	10.50	13.55	11.69	15.28
(Moisture)	(23.50)	-	(22.50)	-	(24.00)	-	(22.50)	-	(23.50)	-
(Ash)	(25.60)	33.46	(29.50)	38.06	(22.23)	29.25	(31.78)	41.00	(25.60)	33.46
	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00

[Ref. 3]

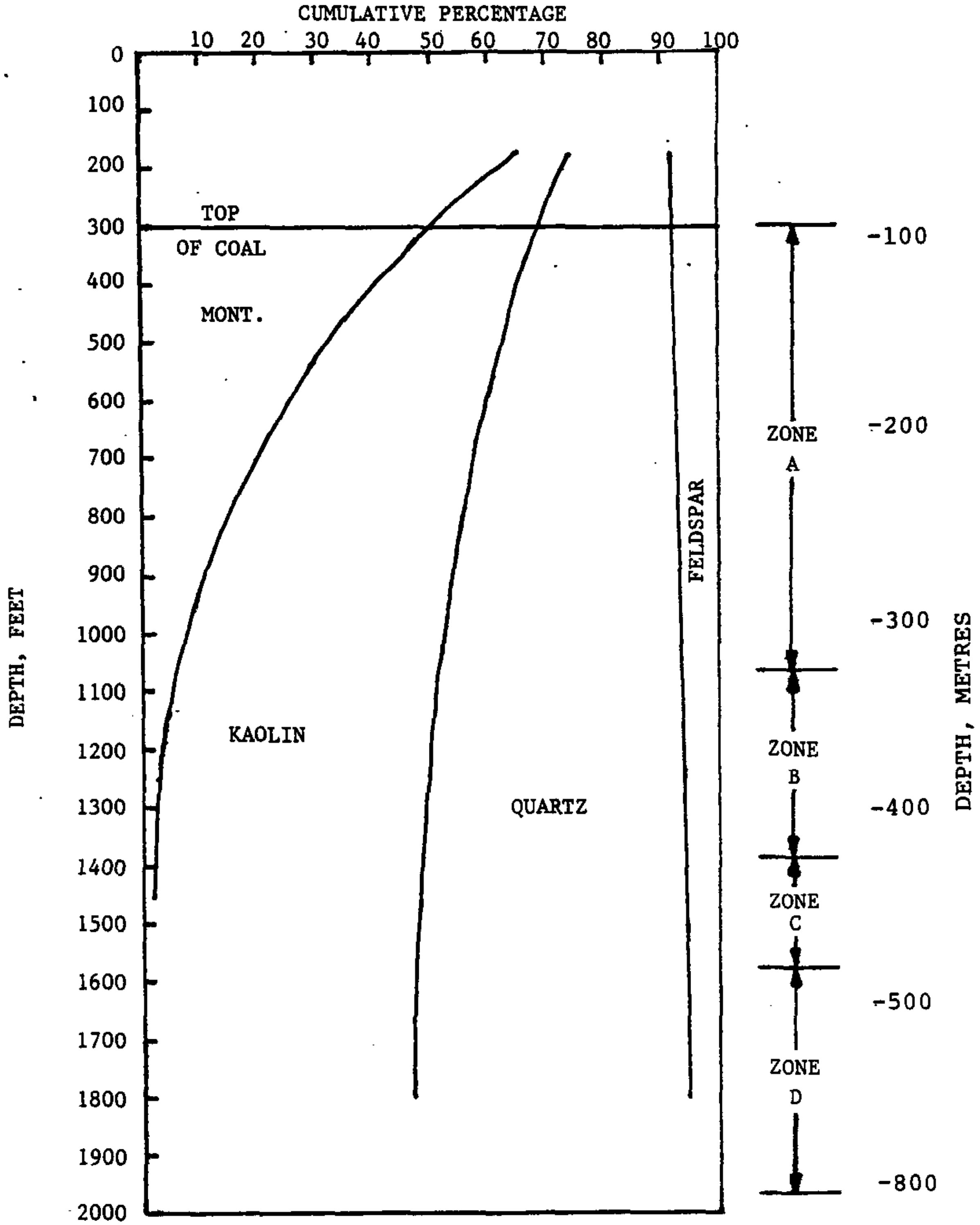
selected for evaluation should be to a common base equivalent to 316.5 TJ/day, equivalent to 8,000 m^3/d (50,000 b/d) liquid fuels. The Performance Blend quality formed the basis of the calculation of the mass and energy balances except for the cases involving Lurgi Gasification. In the latter case the fixed-bed type of gasification process is unable to handle fine coal. Normally it is necessary to remove fines smaller than 6 mm (1/4 inch) from the gasifier feed. However the presence of considerable quantities of clays in the raw coal feed to the screening plant is expected to cause rapid blinding of 6 mm screen decks so that B.C. Hydro advised that 13 mm (1/2 inch) was the smallest screen size considered practical. Screening at this size is expected to provide a yield of 64 percent of the raw coal as gasifier feedstock. The -13 mm fines fraction is rejected as a lower grade. Various methods for utilizing the fines have been considered for the selected process options and are discussed later. Table 2.1 also indicates the anticipated qualities of gasifier feedstock (+13 mm) and fines rejects obtained on screening. Utilization of the coal in a variety of potential coal conversion processes requires consideration of properties additional to those shown in Table 2.1, and in some cases requires interpretation of those shown according to a different set of criteria compared with combustion as dried, pulverized coal in boilers.

2.1.2 Proximate Analysis

Analyses available from a number of differing sources reveal discrepancies in reported values for volatile matter, fixed carbon and calorific values. Difference between laboratories in values for Fuel Ratio, calculated as a ratio of the fixed carbon and the total volatile matter content were commented upon in the previous Study [Ref. 1]. More recently, differences in the first degree linear equations derived by various laboratories for the relationship between calorific value (HHV) and ash contents have been noted by other investigators [Ref. 2].

All these problems are attributable to the presence of substantial proportions of hydrated clays and carbonates in the raw coal, their presence seriously interfering with the standard methods employed for the analysis. The overall effect is serious and important because it is not possible to determine and state the actual quantity of coal substance (dry, mineral matter free) in the coal as delivered to a process plant, or to state the composition of the coal substance. The analyses shown in Table 2.1 are subject to these undetermined errors.

A better understanding of the interference by intercalated and inherent clays has been provided indirectly following an investigation of electrostatic beneficiation of Hat Creek coal at the University of Western Ontario, [Ref. 4]. Figure 2.1, taken from that report shows the averaged mineralogical composition of the coal mineral matter for various depths in the deposit. Figure 2.1 shows that the proportion of hydrated swelling clays (montmorillonite, kaolinite) varies in the range 65-40 percent for depths to 500 meters and between 70-60 percent for depths to 200 meters. The relative proportions of montmorillonite to kaolinite vary from 90:10 to 0:100, being about 45:55 at the 200 meter level. However, the proportion of hydrated clays does not vary consistently with depth, being more closely related to the coal-bearing sequence, which also varies with depth for different areas of the deposit. Therefore, while being a useful but general indicator, Figure 2.1 is unreliable for predicting proportions of clay minerals. Figure 2.2, taken from the same source, shows the dehydration of these minerals on heating. It is noteworthy that very little of the absorbed and structural water is lost at a temperature of 107 ± 3°C, the standard for determination of total moisture content in coal. Montmorillonite loses its water more or less regularly between 100 - 800°C, by which time 9 percent of its



SOURCE: [REF. 4]

B. C. HYDRO

FIG. 2.1

AVERAGED COMPOSITION OF MINERAL MATTER (MINERALOGICAL DATA, COURTESY B.C. HYDRO)

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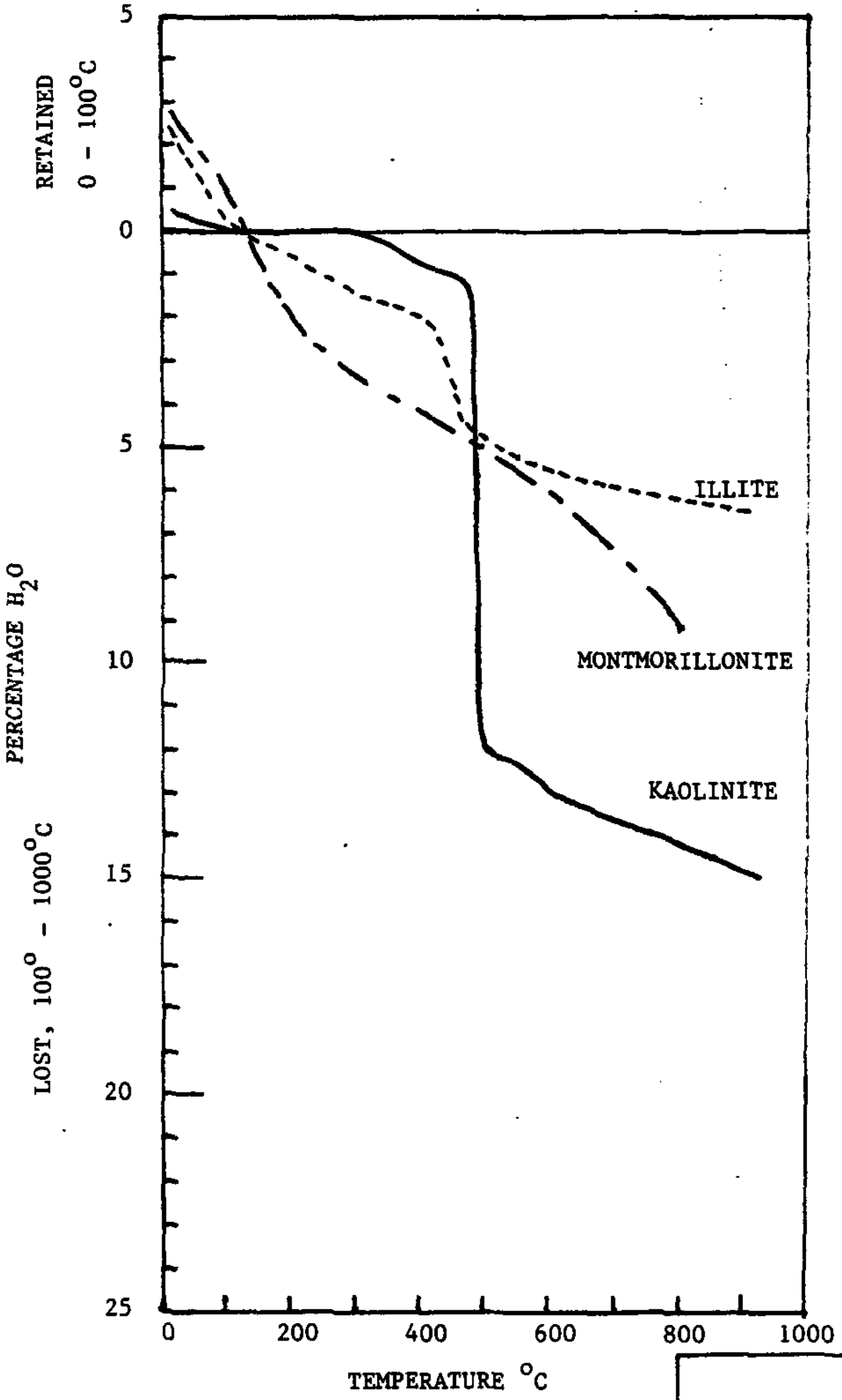


SOURCE: [REF. 4]

B. C. HYDRO	
FIG. 2.2 TYPICAL DEHYDRATION CURVES OF SELECTED CLAY MINERALS [ADAPTED FROM JACKSON, 1956]	
NJ 1024	NORTH AMERICAN MINING CONSULTANTS, INC.

mass is lost. Kaolinite loses very little water until a temperature of 500°C is attained, at which temperature a rapid release of about 14 percent by weight of water, probably accompanying a phase change, is released.

Virtually all coals start to decompose on heating to 300°C, decomposition becoming rapid at temperatures above 350°C.

The combined effect of these observations is that it is impossible to drive off all the water in raw Hat Creek coal at temperatures below the coal decomposition point. For Performance Blend coal containing 25.6 percent ash (as received), the proportion of swelling clays falls in the range 12-18 percent. The actual water not determined as moisture content associated with these hydrated clays lies in the range 1.8 - 2.0 percent on the as received basis and represents water accounted as volatile matter in the reported analyses figures, and is hence accounted as coal substance.

It should be noted that the presence of this undetermined moisture does not affect the apparent repeatabilities or "precision" of the standard determinations of moisture content. Thus it is reported that 121 samples showed an average moisture content of 21.86 percent with a standard deviation of 4.14 percent and a standard error of 0.38 percent. These values are a measure of the repeatability but, in this instance, not the accuracy of the actual water content. [Ref. 5].

The overall impact of this undetermined water on the proximate and ultimate analysis is as follows:

Volatile Matter

The undetermined water reports as volatile matter leading to high values.

A further complication arises from the simultaneous presence of carbonates which decompose, completely or substantially under the conditions of test, liberating carbon dioxide. This carbon dioxide, from inorganic materials present in the raw coal, is therefore accounted as volatile matter and hence as coal substance. Separate determination of the carbon dioxide in the coal permits correction for the effect of carbonates on the volatile content but assumes their complete decomposition under the conditions of the standard volatile matter determination. If complete decomposition does not take place the correction is itself in error. This could be determined by appropriate determination of carbon dioxide in the residual "coke button." (In the case of Hat Creek coal no coke button results and the determination would be made on the residual char.) The reported carbon dioxide content of the Performance Blend is 1.8 weight percent, which corresponds to a carbonate content of 4.0 weight percent (as $CaCO_3$), dry coal basis.

Ash and Fixed Carbon

Ash is the ignited residue of the minerals present in the coal. In general, the ash content is equated with the original mineral matter, weight changes resulting from decomposition of small quantities of carbonates, or oxidation of iron pyrites to iron oxides, being small enough to be ignored. In the case of Hat Creek coal the presence of substantial quantities of dehydrated montmorillonite and kaolinite in the ash results in appreciable hygroscopicity. Errors of 4 percent in the measured ash contents have been observed due to reabsorption of atmospheric moisture. Accurate estimations of ash therefore require that the standard procedures require dry cooling over dessicant and rapid weighing be strictly observed.

This assumption, that ash may be equated with mineral matter, thus allows the calculation of Fixed Carbon according to the formulae,

F.C.% = 100 - (% Moisture + %V.M. + % Ash) (As Received Basis)

F.C.% = 100 - (% Volatile Matter + % Ash) (Dry Basis)

The fixed carbon, being a difference value, therefore accumulates errors that occur in the determinations of moisture, ash and volatile matter contents. As has been shown, these can be substantial in the case of Hat Creek coal, so that the results of the Proximate Analyses must be regarded with reserve.

2.1.3 Calorific Value (HHV)

The differences in the first degree linear relationships between HHV and ash reported by various laboratories has been mentioned earlier and discussed in other reports [Refs. 1, 2, 5]. These differences can be significant, for example.

Source	Reported Relationships
Dolmage-Campbell [Ref. 6]	kJ/kg = 30225 - 353.27 x (ash percent) BTU/lb = 13003 - 151.98 x (ash percent)
EMR [Ref. 5]	kJ/kg = 27998 - 321 x (ash percent) BTU/lb = 12045 - 138 x (ash percent)

These equations differ by about 6 percent of HHV over the range of ash contents expected to be encountered. A recent report [Ref. 2] has suggested that the carbon content contributing to the heat values was different in the high ash material, defined as being greater than 60 percent, from that in the low ash material, and suggested at least two different correlations between HHV and ash.

However the fact is that the mineral matter in the coal undergoes substantial decomposition on ignition at all levels of measured "ash", partly from loss of residual water of hydration and partly from decomposition of carbonates. These effects clearly become greater at all levels of increasing ash and affect the reliability of the HHV/ash relationships.

2.1.4 Ultimate Analysis

The problem of correcting observed values of carbon, hydrogen, nitrogen, sulfur (and forms of sulfur) and chlorine for the minerals present is similar to that discussed above for proximate analysis and calorific value.

Other factors also intrude. For example the carbon in carbonates is measured together with that in the coal substance unless specifically corrected for the actual carbon dioxide present in the analysis sample.

The analyses for forms of sulfur indicated that the bulk of the sulfur present is organic and hence can only be associated with the coal substance. Removal of mineral matter by beneficiation will result in a concentration of sulfur in the cleaned coal in real terms and in greater proportion than the concentration of calorific value. There appears to be no grounds for expecting that beneficiation would lead to significant reduction in sulfur emissions in fuel value terms.

The large measure of agreement that existed between laboratories reporting ultimate analyses was noted in the previous report. The probability of errors in interpretation described above therefore appear to be generally uniformly operative. If the undetermined water associated with the clays is assumed to be 2 percent in the Performance Blend, the "corrected" proximate and ultimate analysis would be as shown in Table 2.2.

Table 2.2 Hat Creek Coal Quality Corrected for Undetermined Moisture Content

	PERFORMANCE		"CORRECTED"	
	As rec'd Basis	Dry Basis	As rec'd Basis	Dry Basis
PROXIMATE ANALYSIS				
Moisture	23.5	-	25.5	-
Volatile matter	25.2	32.94	23.2	31.14
Ash	25.6	33.46	25.6	34.36
Fixed carbon	25.7	33.60	25.7	34.50
	100.0	100.00	100.0	100.00
GROSS CALORIFIC VALUE				
MJ/Kg	13.85	18.10	13.85	18.59
BTU/lb	5,955	7,784	5,955	7,993
ULTIMATE ANALYSIS				
Carbon	35.30	46.14	35.30	47.38
Hydrogen	2.80	3.66	2.80	3.76
Nitrogen	0.70	0.92	0.70	0.94
Sulfur	0.39	0.51	0.39	0.52
Chlorine	0.02	0.03	0.02	0.03
Oxygen (difference)	11.69	15.28	9.69	13.01
(Moisture)	(23.50)	-	(25.50)	-
(Ash)	(25.60)	33.46	(25.60)	34.36

The primary effects are on Volatile Matter in the proximate analysis, since the undetermined moisture reports are a volatile loss in this determination; and on oxygen in the case of the ultimate analysis, since this value is determined by the difference from 100 percent of the other measured components.

2.1.5 Petrographic Analysis

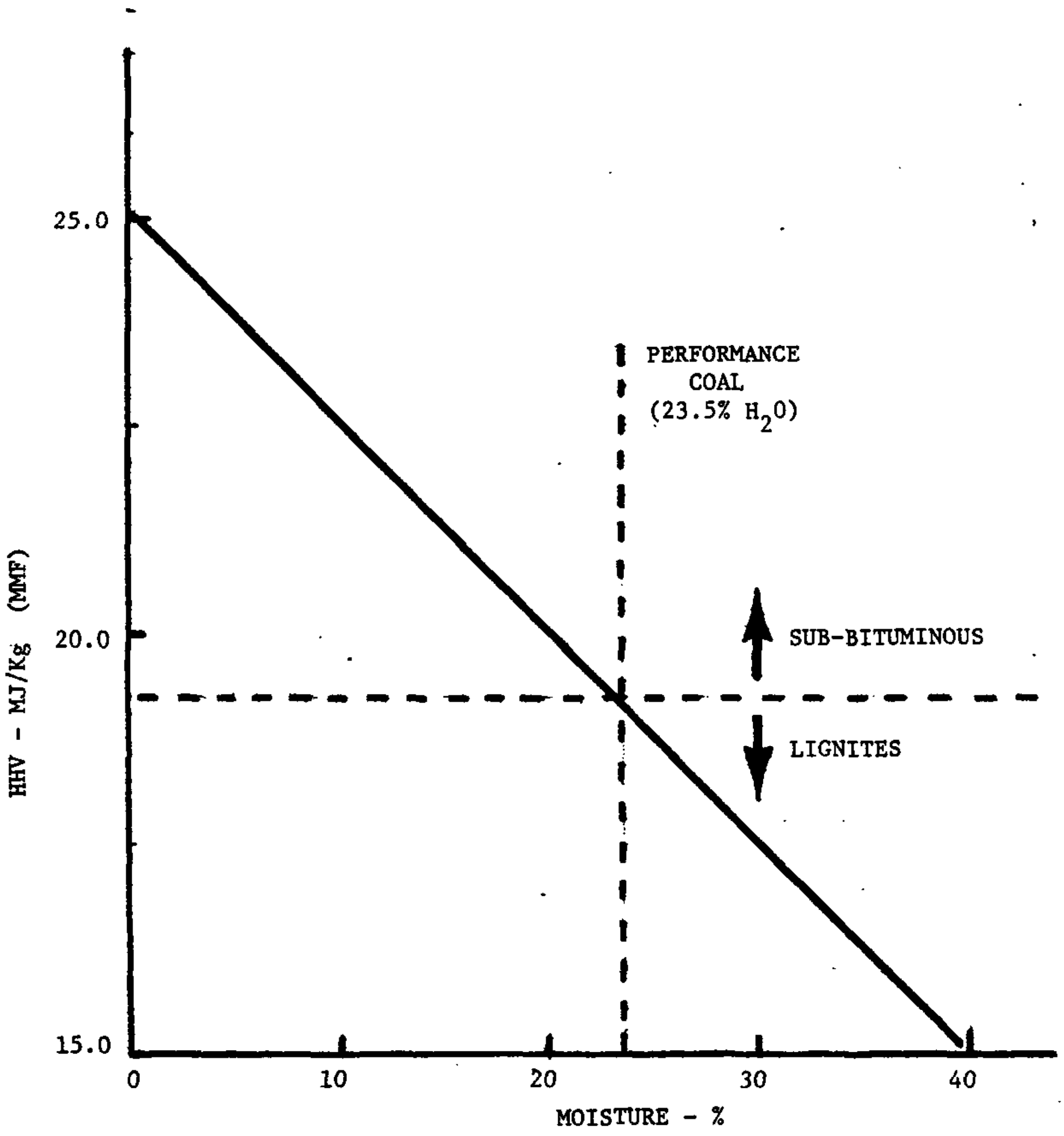
Since the earlier study additional petrographic analysis have been reported by CANMET [Ref. 7]. This reference also contains the results of petrographic analysis carried out at Bergbau-Forschung, Essen, West Germany. The results serve to confirm further the non-homogeneity and variablity of the coal substance within the deposit. The relatively large proportions of low rank vitrinite (mean maximum reflectances 0.34 -0.46) and the low concentrations of other "reactive" materials (exinites) confirms and explains the mutual absence of caking and agglutination properties. The relatively high reactivity of the coal in combustion tests and gasification tests (Pressure Reick) noted in the earlier study is expected from the low rank of the vitrinites present.

2.1.6 Coal Rank

Application of the ASTM Classification of Coals by Rank "ASTM D 388-77" to the properties of the Performance Blend indicates the coal to be Sub-bituminous C. The relationship between gross calorific value, calculated on the mineral-matter-free basis, and various levels of moisture content is shown in Figure 2.3. The calculations are made according to the Parr Formulas (ASTM D 388-77) with conversion to SI units.

2.1.7 Size Consist

Normal raw coal preparation before delivery to the battery limits of the conversion plant will crush run-of-mine coal to below 50 mm (2") top size by two-stage crushing. The expected size consist of freshly wrought coal, and coal which has been stockpiled and recovered for use, are shown in Table 2.3. This table illustrates the size consist to be expected for delivery to all the coal conversion processes to be considered except those which involve Lurgi gasification. Lurgi gasifiers, being fixed bed types, are unable to receive fines, generally specified as -3 mm (1/8"), without serious increase



B. C. HYDRO	
FIG. 2.3 HAT CREEK COAL - PERFORMANCE BLEND RELATIONSHIP BETWEEN CALORIFIC VALUE (MINERAL-MATTER-FREE BASIS) AND MOISTURE CONTENT	
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in pressure drop across bed. Feed coal is therefore normally screened at 6 mm, oversize passing to the gasifiers and the fines passing to some alternative use, usually as boiler fuel. However, because of the presence of substantial quantities of free clays in Hat Creek coal it is expected that 6 mm screens will blind rapidly. B.C. Hydro has therefore indicated that if Lurgi Gasification is used as a process step the run-of-mine coal will receive alternative special crushing and screening treatment designed to minimize excess crushing and production of fines. Screening will be carried out on 13 mm screen decks to avoid blinding. The combination of selective crushing and screening is expected to produce a yield of 64 percent oversize gasifier feedstock and 36 percent fines. The normal preparation corresponding to Table 2.3 would yield 43 percent oversize and 57 percent fines under the same screening conditions, or 38 percent oversize and 62 percent fines in the case of crushed coal recovered from stockpiles. The possibility of achieving this reduction in fines by special treatment is of great importance in assessing the potential application of Lurgi Gasification as a process step. Failure to achieve it will have serious economic consequences.

Since, on screening, there is an accumulation of higher ash materials in the fines, the actual feed to Lurgi Gasifiers will be better than the Performance Blend. The expected improvement, at the expense of the fines quality is illustrated in Table 2.1.

Table 2.3 Size Consist of Hat Creek Coal Supplied to Battery Limits of Conversion Plant

Size mm	Normal Coal Weight Percent	Stored Coal Weight Percent
50 - 25	10	7[1]
25 - 13	16	15
13 - 6	17	16
6 - 3	15	15
3 - 1.5	13	10
1.5 - 0.6	14	12
0.6 - 0	15	25
Total	100	100

2.1.8 Ash Composition and Properties

The ash properties are summarized in Table 2.4. The previous report commented upon the high ash fusion temperatures and the need for taking the slag viscosity data into account in the specification of slagging bottom furnaces and gasifiers. A recent report [Ref. 2] further draws attention to these properties and points out the alkali oxides content, as measured by laboratory methods, may be lower than observed values,with some consequent effect on raising ash fusion temperatures.

In work carried out before 1976, Lurgi Mineraloeltechnik GmbH had commented that the particular ash properties were probably helpful to the operation of dry-bottom rotating grate gasifiers and the ash levels in the feed coal were not a serious hindrance to their process. At that time no data concerning the effects of ash

---

[1] Effective top size 40 mm or less

Table 2.4 Ash Composition and Properties of Hat Creek Coal

Composition		Performance Blend percent	Design Worst Blend percent
SiO_2		53.6	51.8
Al_2O_3		28.1	28.9
TiO_2		1.0	1.0
Fe_2O_3		8.3	8.6
CaO		3.5	3.1
MgO		1.5	1.7
Na_2O		2.21	1.78
K_2O		0.62	0.80
P_2O_5		0.17	0.25
SO_3		1.75	1.80
Mn_3O_4		0.16	1.13
V_2O_5		0.06	0.56
Ash Fusibilities		°C	°C
Reducing	IT	1170-1500+	1140-1500+
	ST	1210-1500+	1150-1500+
	HT	1250-1500+	1230-1500+
	FT	1290-1500+	1270-1500+
Oxidizing	IT	1310-1500+	1300-1500+
	ST	1330-1500+	1330-1500+
	HT	1340-1500+	1360-1500+
	FT	1360-1500+	1390-1500+
250 Poise Viscosity		1500	1500

Note: IT - initial softening temperature
ST - softening temperature
HT - hemispherical temperature
FT - final temperature

properties on other candidate gasification processes were available. The Winkler process is a dry bottom, fluidized bed process for which it was thought that high ash fusion temperatures were beneficial. However, the discharge of dry ash from the fluidized bed always involves carbon losses, reported to be as high as 12 percent, and these losses are clearly higher in the case of high ash coals. In the case of the Koppers gasification process, which is an entrained flow slagging gasifier, doubts were expressed about the impact of the high ash fusion temperatures and slag viscosity characteristics on the operation of the gasifiers. Reports on operations at Modderfontein, S.A. were indicating that problems relating to ash were causing low gasifier availabilities. Accordingly, during the course of this Study, direct discussions have been held with the Koppers Company at which the coal properties and the ash characteristics of Hat Creek coal were specifically discussed. The Koppers company advised that the coal is regarded as possessing adequate characteristics for its use in Koppers gasifiers.

B.C. Hydro has discussed the coal quality and ash properties with the Texaco Company and its suitability as a feedstock for the Texaco Coal gasification process, which is an entrained flow gasifier currently undergoing commercial demonstration trials in the U.S.A. and West Germany [Refs. 11, 12, 13]. Texaco advised that the coal was considered a doubtful feedstock at the present stage of process development because of its low rank and grade. However recent papers claim that the process has operated successfully on coals ranging from lignites to anthracites and since estimates of capital and operating costs from the Texaco gasification process are becoming available it has been included in the general process comparisons in this study.

2.1.9 Grindability

Typically for low rank coals, the Hardgrove Grindability indices reported fell in the range 35-55. The coal is therefore comparatively hard. Additionally, the minerals present contain quartz, feldspar and other abrasive components. These properties must be included in consideration of the specification of mill capacities and constructional materials.

2.1.10 Pyrolysis

Apart from the Fischer Assay carried out by Lurgi before 1976 and included in the previous study report, no further investigation of the behavior of the coal undergoing pyrolysis has become available. The results of the previous tests are repeated in Table 2.5. They are most notable for the very low yields of liquid hydrocarbons (tar) obtained. This fact is consistent with the relatively high oxygen content of the coal and the presence of substantial proportions of oxidized coal, which has a well known effect of inhibiting tar yields during pyrolysis and of increasing the gas yields. It should also be observed that the conditions under which the Fischer Assay is performed in the laboratory can be expected to give higher liquid yields than is obtainable in commercial scale operation. Hat Creek coal must, therefore, be considered a poor candidate for commercial pyrolysis processes. However, given the growing interest in these processes for application to Western Canadian low rank coals, the anticipated results from the Lurgi Ruhrgas Process are included. The results are based on the work reported in the previous Study corrected for Performance Blend quality. (See Appendix D. Section 1.6)

Somewhat higher yields might result from Flash Hydropyrolysis, in which the coal is rapidly heated in hydrogen atmospheres at high pressures. [See, for example, Ref. 10.] However,

Table 2.5 **Carbonization Assay and Coking Properties of Hat Creek Coal**

	As Received Basis	Dry Basis	Dry Ash Free Basis	Moist, Mineral Matter Free Basis
CARBONIZATION ASSAY (FISCHER)				
Gas liquor %	25.0	3.2	5.6	26.8
Tar %	3.1	3.9	6.8	5.3
Gas %	4.5	5.8	9.9	7.7
Char %	67.4	87.1	77.7	60.2
	100.00	100.0	100.0	100.0

COKING AND CAKING INDICES:

Free Swelling Index	0
Gray-King Coke Type	A
Gieseler Plastometer	Non-fluid
Ruhr Dilatometer -	
Max Expansion	Nil
Contraction @ 500°C	10%

such developments are a long way from commercial demonstration and are not considered further in the present Study.

In addition to noting the very low yields of liquid hydrocarbons, it is necessary to consider the principal product produced which is a very high ash char. Viable operation will require that this material be used as primary fuel for thermal power generation. It will have a HHV of about 14.50 MJ/Kg dry basis and will present a difficult grinding and pulverizing problem. However, given the low rank of the parent coal, the char can be expected to be reactive and exhibit good ignition and flame characteristics.

2.2 Coal Beneficiation

Table 2.1 indicates that the ash content of the Performance Blend is greater than 33 percent, dry basis. Concentrations at this level impose significant mass and volume transfer loads on coal conversion processes, with attendant increases in capital and operating cost requirements, so that substantial improvement in economic performance would be the general result for any major removal of minerals from the coal ahead of the coal conversion operations. However, the overall requirement is that the ultimate disposal of the ash would be, environmentally, less troublesome and less costly. Accordingly, the prospects for reducing this ash burden by prior coal beneficiation have been reexamined in the light of various investigations reported since the earlier study [Ref. 1].

2.2.1 Wet Cleaning

The range of options commercially available for beneficiation at the present time are summarized in Figure 2.4.

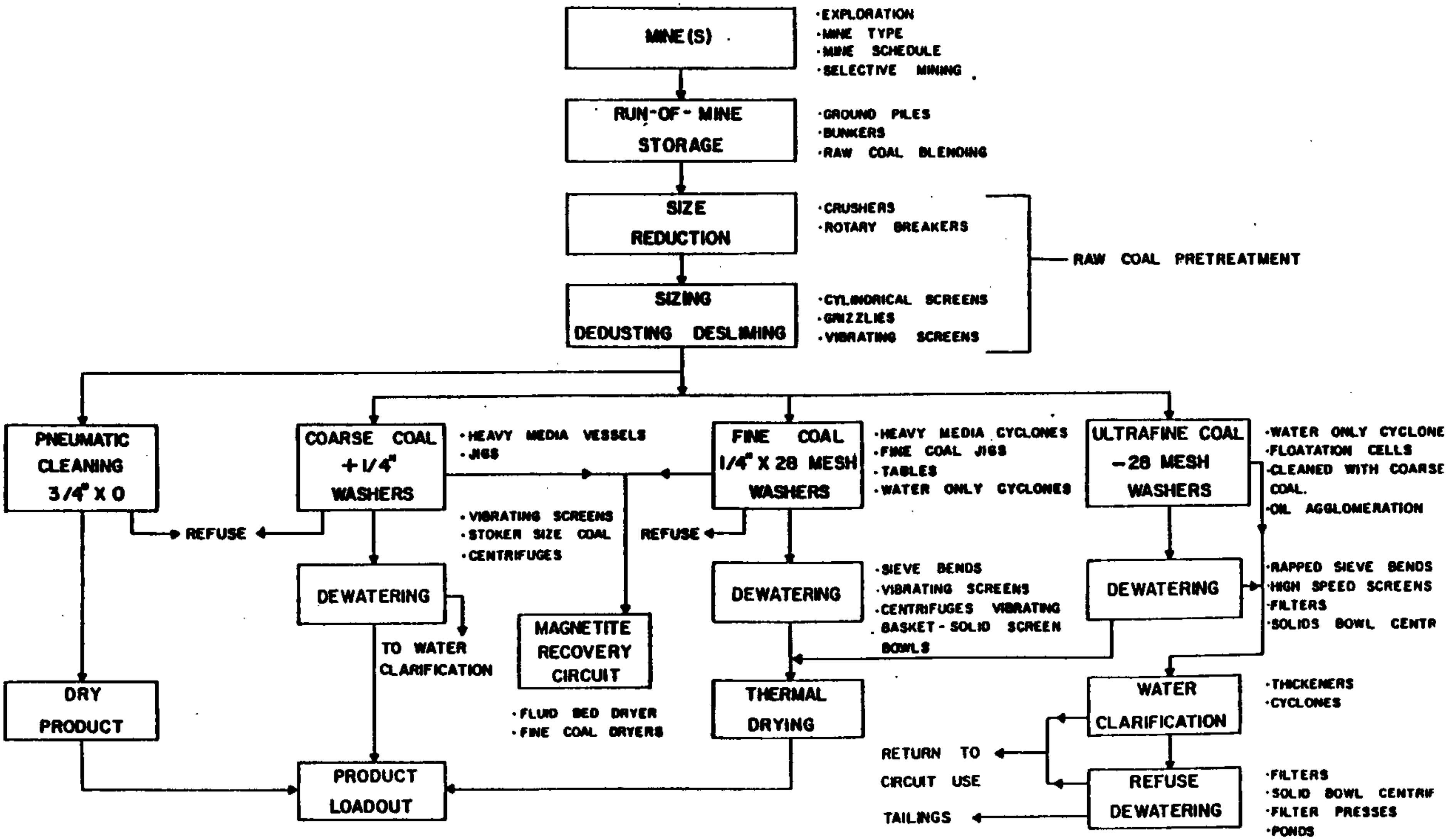

FIG. 2.4 GENERAL COAL PREPARATION PLANT FLOW SHEET

RANGE OF OPTIONS

Inspection of this figure shows that the principal processes available,with the sole exception of pneumatic cleaning, involve wet cleaning methods.

The earlier work revealed a wide variation in the results of standard washability tests carried out on various coal samples from different locations in the deposit. It was noted, early on, that the results were influenced by the pretreatment of the samples and that wetting the coal led to substantial structural degradation and release of clays. This phenomena was convincingly demonstrated by a trial at EMR [Ref. 5] in which a sample of the coal was dispersed in water and pumped repeatedly around a closed circuit containing a hydrocyclone. Extensive, progressive attrition occurred. The separated fine coal approached a "limiting" ash content of about 16-17 percent. However the minerals present contain large proportions of montmorillonite, kaolinite and lesser quantities of quartz, feldspar, cristobalite and siderite. The former are strongly swelling clays which produce a bulky, gelatinous, thixotropic mass that is very difficult to consolidate or dewater and produce huge volumes of semi-fluid waste. Weighed against this water problem the improvement in ash content to be obtained was not considered by B.C. Hydro to be a significant benefit in use of the coal for thermal power generation, and has played the major role in their decision to reject wet cleaning methods.

Similar considerations apply for the majority of the primary coal conversion processes. Thus, except for conventional Lurgi gasification or its slagging modification, the attrition of the coal itself to finer sizes is not necessarily a serious drawback, because these processes require finely divided coal feedstocks. However, the problem of the dewatering and disposal of wet clays is considered to be so great that it outweighs the other potential benefits.

Cleaning by wet methods is therefore rejected as a viable alternative given the present status of water clarification and sludge disposal methods; which are, nevertheless, highly developed so that the problem is clearly an intractable one.

2.2.2 Dry Cleaning

It is clear that, to be successful in cleaning Hat Creek coal, it is necessary to achieve unambiguous separation of the coal and clay particles. It is also clear that, given the structural composition of the coal such separation can only occur after sub-division to fine sizes generally below 1-3 mm. This requirement rules out commercially available air tables, which normally operate on sizes up to 20 mm and which tend to eliminate all particles less than 1 mm as dust regardless of composition. Successful cleaning of Hat Creek coal, or other western low rank coals similarly contaminated by substantial quantities of clays, therefore requires the development and commercialization of new processes. This problem has been recognized for some time. Several programs are in progress in Canada, receiving federal, provincial and industry support, aimed at developing dry methods for cleaning low rank coals. Recently, two methods have been actively investigated, one being electrostatic beneficiation [Ref. 4], the other being a gravity controlled pneumatic separation in fluidized beds (Fluidized Cascade) [Ref. 8]. The status of development of these processes were assessed in direct discussions with the investigators. They are both at laboratory pilot stage scale.

(i) Electrostatic Beneficiation:

Electrostatic beneficiation is a dry process which depends on the movement of small particles of opposite charge in opposite directions in an electric field; positively charged particles will move in the direction of the field and negatively charged particles will move in a direction opposite to the field. Thus, separation will depend primarily on the charging process used and the ability of different constituents of a mixture to charge with opposite sign.

The technique has a long history, the first practical separator for minerals having been patented in 1899 (L.I. Blake and L.N. Morscher. U.S. Patent Nos. 668.791: 668.792). An article in Zeitschrift fuer Elektrochemie for 1907 contains references to fourteen patented, electrostatic separators. Probably the best known commercial machine was the Huff Separator which was applied to minerals and coals until froth flotation became the dominantly preferred process in the 1920's. However development work was continued in Germany until 1945 and a full scale plant was constructed in the Ruhr,but was destroyed by bomb damage before commissioning. Since that time no known commercial plant for electrostatic beneficiation of coal is recorded.

The early separators usually consisted of a rotating metal drum over which the material was fed and subjected to a high voltage electric field or a corona discharge providing gaseous ions. The behavior of various coals depended on their electrical properties, some were repelled far from the drum, some fell near and some adhered and had to be scraped off. The fine ash particles present generally behaved in the opposite sense. Efficient separation depended upon finely grinding the coal, generally to sizes less than 200 mesh (150 microns). Attempts to develop and further improve this process, especially to effect removal of pyritic sulfur from high-sulfur coals, are presently being conducted by Advanced Energy Dynamics at Natick, Massachusetts.

The work at the University of Western Ontario employs a laboratory scale electrostatic tower which acts on a falling stream of the finely divided coal to separate the particles according to their mass-to-charge ratios. The feed coal is electrically charged by triboelectrificiation in a fluidized bed device before passing to the tower. The coal separation experiments indicated that with carefully controlled humidity, high electric field strengths and multistage processing a product containing somewhat less than 30 percent ash at 90 percent BTU recovery could be achieved. This degree of beneficiation is of little importance for reducing the ash burden on coal conversion plants to a stage where substantial reductions in capital costs would result.

(ii) Counter-Current Fluidized Cascade Beneficiation:

The counter-current gas fluidized cascade (CFC) is claimed to be a novel technique for pneumatic separation of mixed granular solids according to specific gravity in a fluidized bed under conditions where the fluidizing velocity is only moderately in excess of the minimum fluidizing velocity. In the CFC,vertical partial segregation is magnified in a horizontal direction by use of counter-current enrichment obtained by creating opposite horizontal motions in the upper and lower strata of the fluidized bed by means of baffles or paddles attached to an endless chain.

In work carried out in a laboratory scale unit at the University of Western Ontario for B.C. Hydro in 1978 the following results were obtained [Ref. 9].

Table 2.6 Average Cleaning Results Obtained with Laboratory CFC Unit

	Ash
Feed Coal	49.81% (d.b.)
Cleaned Coal	43.6 % (d.b.)
Rejects	53.3 % (d.b.)
BTU Recovery	77.0 %

Source: [Ref. 9]

For this process also the degree of beneficiation is minimal in relation to the requirements of the coal conversion processes.

Further development of the CFC has now been transferred to the Alberta Mining Research Center. From discussions with staff at the Center it is not expected that the CFC beneficiation process will be developed to commercial scale in time for application to Hat Creek coal in the time frame of the present Study, i.e., where design and construction could be carried out before 1985 to 1990.

2.2.3 Other Cleaning Methods

One other possible method for beneficiating coal that avoids wetting it with water involves the use of certain organic liquids as the dense medium of separation. By appropriate selection of specific gravity generally in the range of 1.4 - 1.8, coal floats and mineral matter sinks under conditions where the clays present cannot gel and disperse. Certain halogenated hydrocarbons exhibit specific gravities in this range and are available as bulk chemicals. One such process, OTISCA*, is currently under development in the U.S.A. with the main aim of achieving desulfurization of fine coal by precise control of the density of separation [Ref. 14]. A major problem to be solved is to achieve a high recovery of the expensive liquids from the coal and refuse products. However, the presence of clays in the minerals can lead to unacceptable losses of medium by absorption and for this reason its use for cleaning Hat Creek coal is not regarded as a candidate process at the present stage of development. Successful development of any dry cleaning process will, generally, produce very large volumes of dry dusts for disposal. In the scale of the present study the quantities of ash ultimately requiring disposal will be 10,000 - 15,000 t/d and if dry cleaning is introduced about half this quantity will require disposal as dry dusts, the remainder as ignited, quenched fly ash or clinker residues. In environmental terms, disposal of ignited ash residues may prove to be the easier alternative. It is necessary to bear in mind that in regard to the overall mine/utility complex, coal beneficiation cannot reduce the total quantities of solid wastes requiring disposal, only their form. Additionally, the most economically feasible mining method

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*Otisca Industries, Lafayette, New York

precludes, for many years, the disposal of solid wastes back to the mine [Ref. 3] so that provision of a separate solid wastes disposal area is necessary.

2.3 Coal Preparation for Conversion Processes

2.3.1 Thermal Drying

The coal, as received at a conversion plant, will contain 20-25 percent of water which constitutes a large, inert mass load for most processes to be considered. In some cases totally dried coal is the preferred feedstock, as in the direct hydrogenation liquefaction processes. In others, where steam is a process reactant, the presence of some moisture in the coal is not necessarily a disadvantage. The Winkler, Koppers and Lurgi Gasification processes all employ oxygen and steam as reactants and hence, a certain level of moisture in the coal feedstock can be tolerated. However, for low rank, low grade coals such as Hat Creek, this level does not exceed about 9-10 percent and also assumes that the coal is not sticky and difficult to handle at this level. Some measure of removal of moisture therefore appears desirable and thermal drying units are included in the costs of coal preparation for the processes to be considered later.

Even when all surface moisture has been evaporated from low rank coals it is customary to observe that high levels of residual water, bound structurally in the coal substance, remains. Removal of this "bound" water requires greater heat imput to the drier per unit weight of water evaporated and drying costs increase markedly. In addition, low rank coals are more susceptible to spontaneous ignition during drying than are bituminous coals and hence must receive less vigorous treatment with hot gases. The combined effects of these factors is that it is not generally practicable to dry low rank coals below some minimum moisture content that has been termed Measured Critical Moisture (MCM) [Ref. 15].

Drying trials have not been conducted on Hat Creek coals but from consideration of the long term dehydration behavior in air, and in discussions with B.C. Hydro, this MCM value is assumed to be 11-13 percent.

A further feature of low rank coals is that, after drying, they usually exhibit strong tendencies to reabsorb water with little change in external volume. In the majority of cases this is of small consequence except to ensure that the coal is passed to the processing units directly after thermal drying, crushing and pulverizing. The case of Texaco gasification is,however, an exception. This process injects a coal/water slurry into the gasifier under pressure. Process developments require that the water content of the slurry be as low as possible and slurries containing as much as 70 percent of pulverized coal (75 percent through 200 mesh) in 30 percent water have been achieved. Pulverizing the coal requires first that it be dried and, in the case of low rank coals, this means thermal drying to the MCM values. However, subsequent slurrying with water leads to substantial reabsorption without volume change of the coal, so that it is not possible to attain pumpable slurries containing coal water ratios of 70:30. In the case of Hat Creek coal, if it is assumed that reabsorption will increase the moisture content from MCM (~13%) to the equilibrium moisture (23.5%),the resulting coal/water slurry cannot contain more than about 50 percent coal. This fact, coupled with the high ash content, is a reason why Hat Creek coal is not considered a good feedstock for the Texaco gasification process.

2.3.2 Crushing and Pulverizing

The general requirements for crushing and pulverizing related to the coal conversion processes considered in the Study are summarized in Table 2.7. The need for screening out fines below 13 mm in the case of Lurgi gasification has been discussed above. (Par. 2.1.1). The costs of providing coal preparation systems to produce coal feedstocks of appropriate moisture and size characteristics have been included in the overall coal conversion plant costs.

2.4 Coal Conversion

2.4.1 Process Selection Criteria

Because the present study requires comparative marketing and economic analysis of possible processes applicable at Hat Creek, attention must be focused on those processes which are already commercially proven or are at advanced stages of demonstration on commercial scale. A further requirement is that the product slates are compatible with existing or foreseen market requirements. Progress towards the development and construction of commercial coal conversion projects during the past five years has been slow, no major commercial coal conversion plant has been completed in North America and some projects that had been commenced are presently stalled. The following paragraphs outline areas where progress has been made.

2.4.2 Upgraded Solid Products

(i) Cleaned Coal

There has been an extension of wet cleaning practices, especially for cleaning fine and superfine sizes (below 0.5 mm). There has

Table 2.7 **Coal Preparation for Conversion Processes**

	Direct Liquefaction Processes	Gasification Processes				
		Winkler	Koppers	Lurgi		Texaco
As received (mm)	100 x 0	100 x 0	100 x 0	100 x 0		100 x 0
Screening	-	-	-	+13mm	-13mm	-
Crushing (mm)	8 x 0	8 x 0	8 x 0	-	8 x 0	8 x 0
Drying	MCM*	-	MCM*	-	MCM*	MCM*
Pulverizing (200 mesh, 150 microns)	>75%	-	>75%	-	>75%	>75%
Coal/water slurry	-	-	-	-	-	70/30+

*Measured Critical Moisture Content

+Probably unobtainable with Hat Creek Coal been renewed interest in oil agglomeration of fine coal both as a recovery and as a cleaning mechanism. Some progress towards the use of coal in hybrid liquid fuels have been made, including coal oil mixtures, coal oil water mixtures (e.g. COALIQUID*) and coal water liquids (e.g. CARBOGEL**). The combustion of these products is presently receiving attention, work on the development of suitable burners going ahead at a number of locations. It is expected that these hybrid fuels will find a limited market for industrial steam generation but their use in large quantities by the electric utilities is not expected. Part of the advantages claimed for these fuels is that the ash and sulfur contents of the coal component can be reduced to low levels by first grinding the coals and cleaning the resulting pulps by froth flotation or oil agglomeration. Such methods are known to work for some coals but are not generally applicable, and while, for bituminous coals, substantial reduction in ash is usually possible the simultaneous reduction in pyritic sulfur is much less certain. In any case, these processes are not effective in reducing the organic sulfur component of the coal. The methods are not generally applicable to low rank coals because the fundamental surface properties of the coal particles resist recovery by froth flotation; and the methods are not applicable to Hat Creek Coal.

(ii) Hat Creek Coal as Solid Fuel

Reference to Table 2.1 shows that the "inerts" contents of the coal is about 50 percent. This

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* COALIQUID INC., Louisville, Kentucky, U.S.A

** AB CARBOGEL, Helsingborg, Sweden

presents a serious obstacle to transportation of the coal for any distances requiring loading and discharge from railcars because of the costs involved, which are usually made on a ton-mile basis irrespective of coal quality. Additionally, there are ample reserves of high grade bituminous coals in British Columbia awaiting development which will present powerful competition for any market for solid fuels that may develop.

The only conceivable use remaining as a solid fuel is for steam electric power generation at the Hat Creek site. These uses are considered only to the extent that they are necessary adjuncts to coal conversion processes. The principal utilization of the coal for thermal electric power generation is specifically excluded from the Scope of Work of this Study.

(iii) Solvent Refining

Solvent refining to produce solid products (e.g. Gulf SRC-I Process) has been demonstrated. Recent discussions with Gulf established that low rank coals can, in some cases, be suitable candidates but this is generally determined by the oxygen content of the coal, the hydrogen requirement and the yield of refined product. Qualitative investigation of the behavior of Hat Creek coal was carried out by NAMCO for B.C. Hydro in 1978 and showed that the coal was reactive to solution hydrogenation and exhibited high carbon conversion efficiency. However, the conditions of the test did not permit reliable estimates of the hydrogen demand to be determined [Ref. 16]. There is evidence suggesting that the reactivity observed is, to some extent, catalyzed by the ash minerals present. Conversion to SRC-I is one of the processes selected for evaluation in this study but, for convenience, is included in the direct liquefaction processes.

(iv) Pyrolysis

Progress in commercialization of new pyrolysis processes has been very slow. The Lurgi Ruhrgas process, considered in the earlier Study and repeated in this one, continues to be the only major commercial application. Design and construction of a commercial demonstration plant for the COED Process was announced by the U.S. Department of Energy and subsequently cancelled in 1978. The behavior of Hat Creek coal on pyrolysis has been discussed above (2.1.10) and it is not regarded as a suitable candidate for pyrolysis processes.

2.4.3 Gaseous Products

For conversion to upgraded gaseous products, the so-called "first generation" processes - Lurgi, Koppers-Totzek and Winkler continue to be the only fully commercial primary gasification processes available. All three processes produce a nitrogen free synthesis gas by gasification of the coal with oxygen and steam. After suitable shifting, the synthesis gas (carbon monoxide and hydrogen) forms an intermediate for the production of hydrogen (ammonia), synthetic natural gas (methane) or a variety of synthetic liquid products discussed under Indirect Liquefaction below.

The last Winkler installation was made in the 1960's. However Lurgi and Koppers gasifiers have continued to be favored and have been the basis of the two largest installations for coal conversion in recent years, both in South Africa. The AECI 100,000 t/a ammonia plant at Modderfontein, S.A. has six Koppers-Totzek gasifiers. SASOL-II and III are each based on 36 Lurgi gasifiers.

Progress in so-called "third generation" processes for total coal gasification such as Hygas, Bigas, Synthane, etc. is still a long way from commercialization and they provide no competition to the established processes at the present time. In fact, substantial improvements in the Lurgi, Koppers-Totzek and Winkler processes, often referred to as "second generation" processes, have been made which are considered likely to ensure their pre-eminence for the foreseeable future. These improvements may be summarized -

(i) Lurgi	- Extension to gasification of highly caking coals.
	- Development of high temperatures slagging operation by the British Gas Council. Has been offered with full engineering guarantees since October 1980 (BGC/Lurgi Process).
	- Development of a high pressure gasifier to operate at 100 bar, compared with 25-30 bar in standard Lurgi Gasifiers (The Lurgi Ruhr 100 Process).
(ii) Koppers-Totzek	- Demonstration of high pressure operation by Shell (SK Process) in a 150 t/d unit. Design and engineering of 1000 t/d commercial gasifiers is in progress

but the Shell Company has not yet made the technology generally available [Ref. 17].

(iii) Winkler - Rheinische Braunkohlenwerke A.G. has continued the development of the Winkler process to operate at high pressures (up to 10 bar) and temperatures (up to 1100°C). The process (Rheinbraun HTW) has been demonstrated on the semi-commercial scale at 5 bar and 950°C, and a first commercial demonstration unit is scheduled for operation in 1983/1984. [Ref. 20].

Three other developments have reached a stage where commercialization may be imminent and which may be regarded as second generation developments. These are the Texaco, U-Gas and Exxon Catalytic Gasification processes. Of these, the Texaco process is considered to be at the most advanced stage of development and has been selected for inclusion in this Study.

The Texaco Corporation has demonstrated on a pilot scale that its partial oxidation gasification process, initially developed for gasifying heavy oil feedstocks, can be successfully applied to coal/water slurry feedstocks [Ref. 11].

Following the initial demonstration by Texaco, a 6 t/d demonstration plant was constructed by a consortium at Oberhausen-Holten in West Germany (Ruhrchemie AG/Ruhrkohle AG) [Ref. 13] which has been operating since January 1978. Almost immediately, the syngas produced was fed into Ruhrchemie syngas network and from July 1980, part of the syngas was passed to a methanation pilot plant for production of SNG. As expected, both synthetic processes have operated without complication. During 1980, projects in the U.S.A. employing Texaco Coal Gasification for the manufacture of SNG, combined-cycle electric power and methanol have been announced. A Texaco gasifier was included in a pilot ammonia manufacturing plant started up by the Tennessee Valley Authority at Muscle Shoals, Alabama, in October 1980.

One proposal of particular interest is to produce SNG and methanol from syngas produced by a combination of BGC/Lurgi Slagging Gasification and Texaco Coal Gasification [Ref. 19]. Such a combination is synergistic in several ways -

a) Coal fines produced in preparing coal by the BGC/Lurgi gasifiers are unusable by that process but can easily be included in feedstock to a Texaco gasifier.

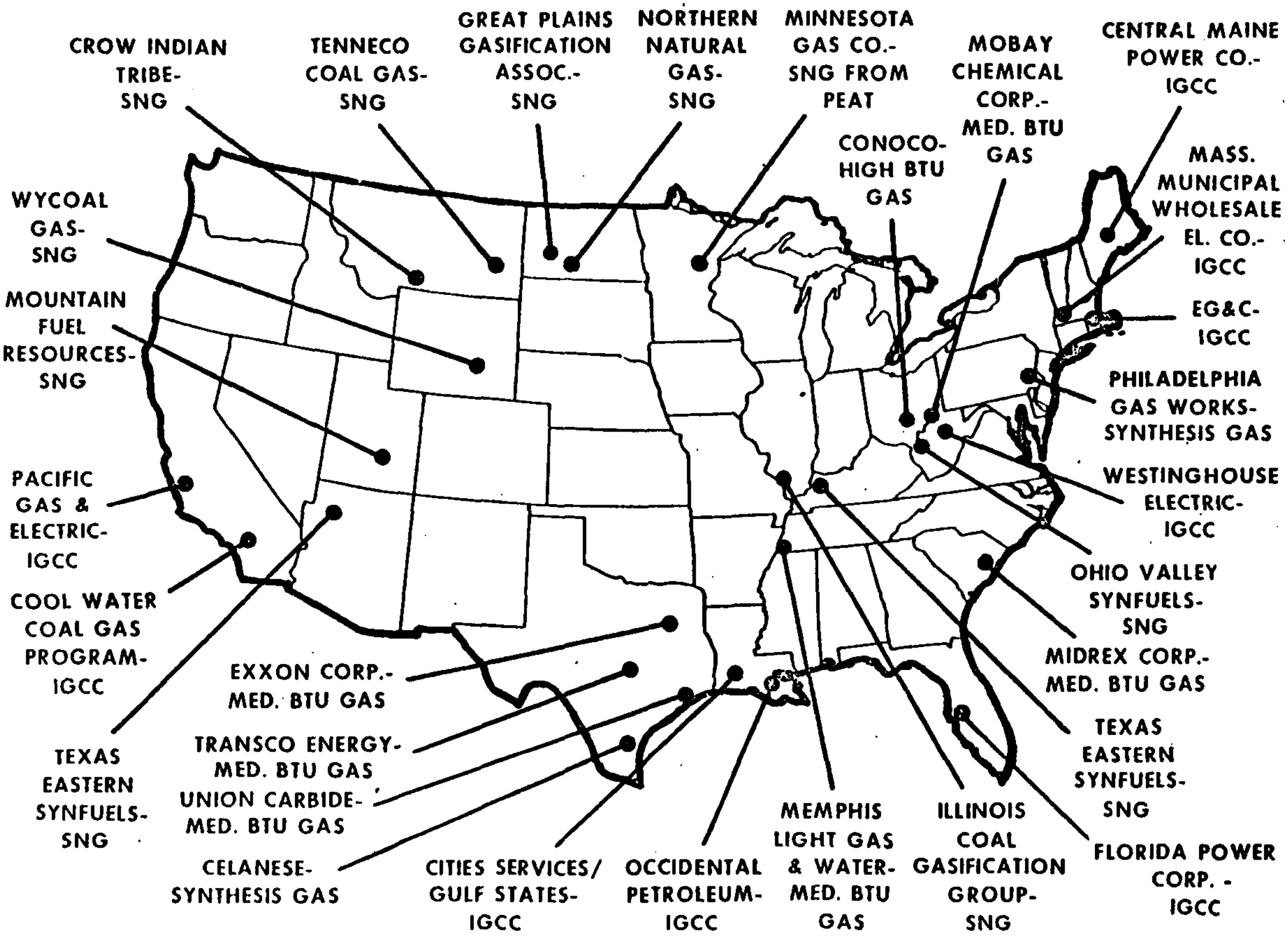
b) Phenolic liquors, a byproduct by BGC/Lurgi gasification, can be used to prepare the Texaco slurry feed - hence avoiding substantial effluent treatment costs.

c) The primary BGC/Lurgi gas contains substantial proportions of primary methane and hence reduces the methanation steps required for upgrading to SNG.

Fines rejected by Lurgi gasifiers could also be fed to Koppers or Winkler gasifiers and such combinations have been included in the present Study.

Figure 2.5 shows some projects currently being considered in the United States for the Production of Synthetic fuel gases. Processes for production of low or medium BTU fuel gases have not been considered in this study because of the

Figure 2.5 U.S. SYNTHETIC FUEL PROJECTS - MEDIUM AND HIGH BTU GAS



absence of any substantial market for such fuels in the vicinity of Hat Creek. Pipeline economics generally rule out the feasibility of transporting such gases to the Lower Mainland industrial regions.

2.4.4 Liquid Products

The production of liquids from coal can be achieved by acting directly on the coal with solvents; by combining solvent extraction with hydrogenation; or by indirect synthesis of liquids from syngas produced by one of the processes described in the previous sub-section.

(i) Solvent Extraction

Work on solvent extraction processes is mainly being pursued in Great Britain. Coal liquefaction research at the Coal Research Establishment (National Coal Board) has led to the development of two separate liquefaction processes for the production of distillate fuels and chemical feedstocks; the Liquid Solvent Extraction Process (L.S.E.) and the Supercritical Gas Extraction Process (S.G.E.). Both processes are two-stage processes - an extraction stage in which a coal extract is separated from mineral matter and undissolved coal and a subsequent extract hydrogenation stage.

Liquid Solvent Extraction Process:

In this process most of the coal is dissolved in an aromatic solvent (digestion) and the residual solids consisting of mineral matter and undissolved coal are removed by filtration: The resultant coal extract solution is fed to a hydrocracker for conversion to premium liquid products and a fraction suitable for re-cycling as solvent. Originally developed for the processing of British bituminous coals the process is thought to be adaptable to the processing of brown coals and lignites.

Under an agreement between the British Columbia Hydro and Power Authority and the National Coal Board (UK), a sample of coal from the Hat Creek deposit has been processed in a 20 Kg per day integrated liquid solvent extraction/hydrocracking continuous pilot plant. The plant, after adjustment of process conditions to accommodate the Hat Creek coal characteristics, was reported to have operated without problems. With production of recycle solvent in balance with requirements a total light oil yield equivalent to 45 percent of coal input (dry, mineral matter-free) was observed. This is equivalent to approximately 21.5 percent on the 'as received' performance blend.

Supercritical Gas Extraction Process:

This process depends on the ability of a compressed supercritical fluid to dissolve relatively high molecular weight substrates. It is operated under conditions where a hydrogen rich portion of the coal is selectively extracted in a short residence time extraction stage, leaving a hydrogen deficient char. The extract produced is separated from the solvent and passes to a hydroprocessing stage whilst the char product is used as a solid fuel for production of heat and power or is gasified to produce process hydrogen requirements. It is reported that an extract yield of the order of 50 percent represents a balanced case in which there is just sufficient char for the provision of these commodities.

Under the existing agreement between the British Columbia Hydro and Power Authority and the National

Coal Board (UK), laboratory tests of the SGE Process on Hat Creek coal have been carried out. Extract yields in the range of 21 to 27 percent on dry, ash free coal basis were obtained. It is reported that yields of 35 to 40 percent (d.a.f.) are obtained from high volatile bituminous coals; and yields of up to 50 percent (d.a.f.) have been obtained with certain lignites. It was therefore concluded that the yields from Hat Creek coal are low, so that the coal does not appear to be a good candidate for this process application.

It should be noted that the LSE Process is at the small, continuous pilot plant stage (20 kg per day coal throughput) and the SGE Process is at laboratory stage. The British Government and the National Coal Board have announced that development to commercialization of both these processes will be pursued but the timetable announced is not expected to result in commercialization before the end of this century. For this reason these processes have not been considered further in this Study.

(ii) Direct Hydrogenation

The current status of development of direct liquefaction processes is summarized in Table 2.8. Commercial demonstration plants have been built and are being commissioned at Catlettsburg, Kentucky (H-Coal Process, Dynalectron Corporation) and Baytown, Texas (Exxon Donor Solvent Process). A projected 6,000 t/d coal demonstration plant for the Gulf SRC-II Process, to be built at Morgantown, West Virginia is at an advanced stage, being supported by German and Japanese funding in addition to major funding by the U.S. Government. In West

Table 2.8 Process Data on Direct Coal Conversion Pilot Plants

		Products			
Process and Contractor	Size of plant (coal feed) and date of completion	Primary	Secondary	Temperature/Pressure	Residence time in reactor
SRC-I Southern Company Services, International Coal Refinery Co.	6,000 t/d 1984	Solid sulphur free fuel	Fuel oil, naphtha	450°C/120 bar	0.5 hours
SRC-II Gulf Oil	6,000 t/d	Fuel oil	Gas, LPG, naphtha	460°C/130 bar	1.0 hour
Exxon Donor Solvent Exxon Oil Company	200 t/d 1980	Fuel oil	LPG, naphtha, gas	450°C/135 bar	36 minutes
H-Coal Hydrocarbon Research	200 to 600 t/d	Fuel oil	Naphtha, gas	355/455°C/205 bar	not available
H-Coal alternative mode	1980	Synthetic crude oil	-	355/455°C/205 bar	not available
New I G Process Ruhrkohle and Veba Oel	200 t/d 1983	Middle distillate 200° - 325°C	Gas, naphtha	450°C/300 bar	not available

Source: [Ref. 18]

Germany a consortium of Ruhrkohle and Veba Oel are constructing a 200 t/d plant to produce gas and light oils by a new I.G. Process. Major plans for direct liquefaction demonstation plants have also been announced for Australia, while it has recently been suggested that a fourth SASOL plant in South Africa will employ hydrogenation instead of the indirect Fischer-Tropsch process [Ref. 21].

The production of direct coal liquids from Hat Creek coals has therefore been included in this study.

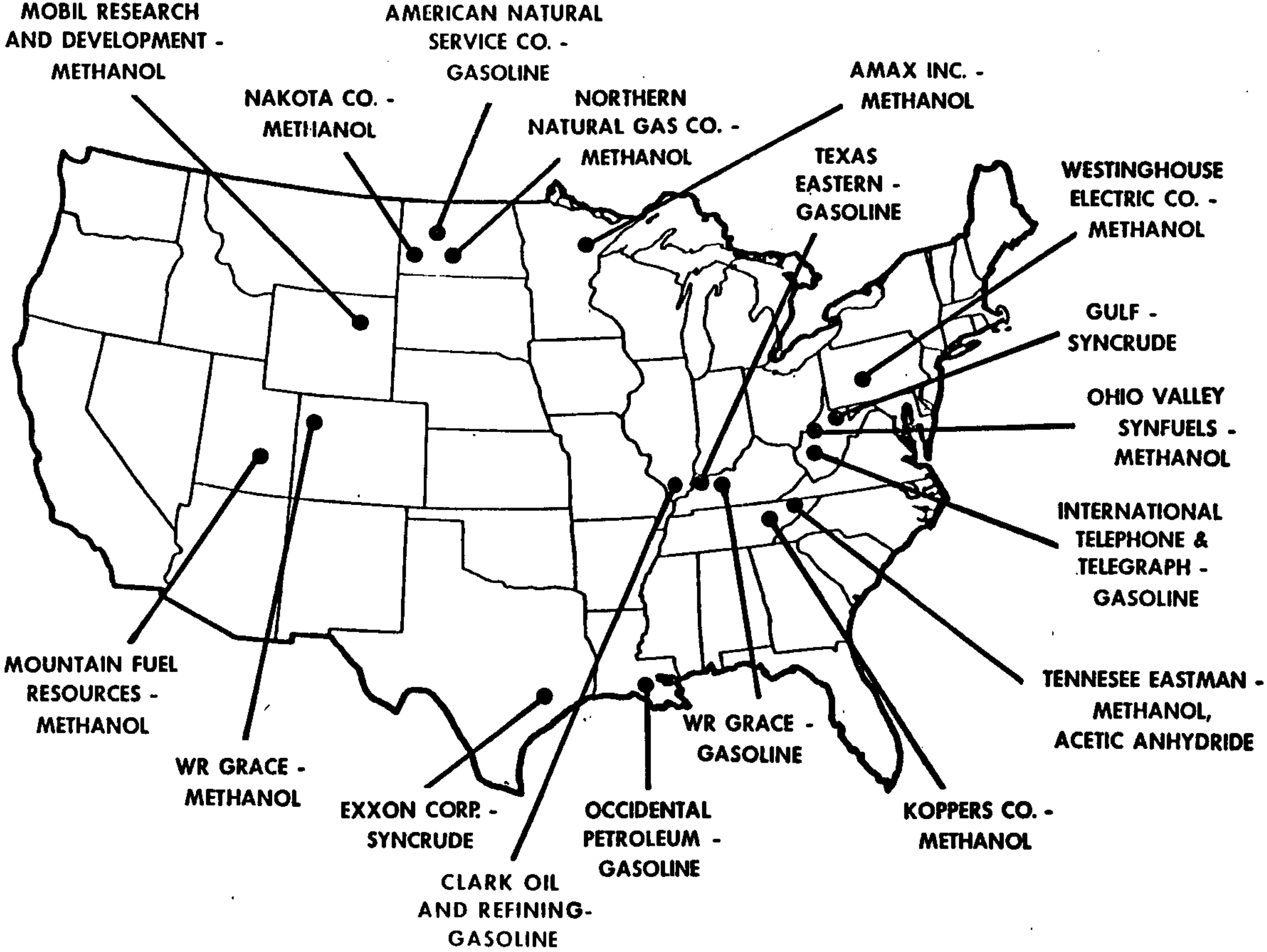
(iii) Indirect Liquefaction

This class of processes, requiring the prior production of syngas, is linked with the developments of the gasification processes already described. Some proposed projects currently receiving consideration in the United States are shown in Figure 2.6. The majority of these projects involve the synthesis of methanol from syngas, in some cases followed by production of gasoline by the Mobil Process e.g. the W.R. Grace project in Tennessee. A few projects are based on Fischer-Tropsch Synthesis, e.g. the Texas Eastern project in western Kentucky. The Fischer-Tropsch synthesis has been amply demonstrated by operations at SASOL where a combination of synthol and ARGE processes is capable of producing an extensive product slate.

More recently, the catalyzed synthesis of methanol in high yield by hydrogenation of carbon monoxide has been demonstrated in processes developed by

FIGURE 2.6

**U.S. SYNTHETIC FUEL PROJECTS
DIRECT AND INDIRECT LIQUEFACTION**



Imperial Chemical Industries, Lurgi and Wentworth Brothers, Inc. These processes are considered fully commercial and are attracting increasing attention as the possibilities of employing methanol as gasoline extenders, gasoline replacement or as a chemical feedstock are realized. More than fifteen projects for production of methanol from coal-based syngas are presently being considered in the U.S.A. and this number is likely to increase (See Figure 2.6).

The Mobil Oil Corporation has demonstrated a Zeolite-based catalytic process for direct conversion of a methanol feedstock to high octane gasoline in high yield. The process is flexible and can be interrupted at an early stage to give various other products, including olefins [Ref. 22, 23].

The utilization of Hat Creek coal in Fischer-Tropsch/ARGE synthesis, methanol synthesis and methanol/gasoline conversion have been included in this Study.

(iv) Methanol

A study by Exxon found coal conversion to methanol is presently the most economically attractive liquefaction process. [Ref. 24]. Costs of these liquids are now estimated at 40 to 60 percent above medium BTU gas, shale oil or imported crude from OPEC. Improvements in technology are likely to bring these costs down.

Methanol from coal is one of the leading candidates for supplementing and eventually replacing gasoline when oil runs short. Methanol blends, however, would present a number of problems if used in a fuel distribution system and vehicles designed for gasoline. Chief among them are:

1. Phase separation on contact with small amounts of water.
2. Vapor lock
3. Cold startability
4. Incompatibility with some fuel system materials.
5. Increased evaporative emissions.

Although methanol has desirable properties as a gas turbine fuel, gas turbines do not yet power highway vehicles. Methanol also is not suitable for diesel engines without excessive amounts of cetane improver or else provision of a separate fuel to initiate combustion.

Methanol and ethanol differ markedly from gasoline in several characteristics that are important to their potential use as transportation fuels, see Table 2.9.

Compared to typical gasoline, methanol requires 44 percent as much air for combustion, produces 49 percent as much energy, and requires 3.7 times as much heat for vaporization.

The automotive and chemical markets are the most attractive for the sale of methanol. However, the transportation market is just starting to develop. Both markets will expand rapidly once an initial supply of coal-based methanol is available.

For most of the industrialized world, methanol from coal could be the alcohol fuel of the future for both

Table 2.9 Fuel Properties of Alcohols and Gasoline

	Gasoline	Ethanol	Methanol
Oxygen Content	0	34.7	49.9
Net Heat of Combustion			
MJ/dm^3	32.2	21.2	15.8
(1,000 BTU/gal.)	(115.4)	(76.0)	(56.6)
Heat of Vaporization			
MJ/dm^3	0.25	0.66	0.93
(1,000 BTU/gal.)	(0.90)	(2.4)	(03.3)
Distillation Temperature			
°C	32-210	78	65
°F	90-410	173	149
Water Solubility	0	∞	∞

economic and supply reasons. As mentioned earlier alcohol fuels pose serious problems for the conventional internal combustion, reciprocating piston engine. Burning significant quantities of methanol would necessitate some complex and expensive engine modifications. Furthermore, there may be safety and emission problems associated with gasohol that have not been identified to date. Materials and corrosion problems with gasohols (methanol or ethanol) involve not only the plastics and rubber compounds commonly found in fuel systems, but also light metal castings, such as aluminum,once the alcohol portion of the gasohol increases.

In a recent survey by the United States Department of Energy, [Ref. 25] the three major U.S. automobile producers listed methanol-derived gasoline using the Mobil M-gasoline process as the preferred fuel of the future. Pure methanol ranked second and methanol blended with gasoline was third. However, the auto makers expect initially to introduce blends because of the present consumers acceptance of gasohol. The survey also included many large electric utilities which indicated their support of methanol conversion. They estimate that if methanol becomes readily available at favorable prices, 75 percent to 100 percent of their peak power turbines may eventually be converted to use methanol. Methanol is an acceptable gas turbine fuel and the technical problems are readily solvable, but the utilities appear to be more concerned about its future security of supply and price.

The survey included ranking of synfuel technologies by leading financial institutions based on their economic outlook. These were ranked as follows:

Table 2.10 Ranking of Synthetic Fuels

Rank	Synfuel	Source
1	Gas	Oil Shale
2	Liquids	Coal
3	Gasoline	Coal
4	Gasoline	Oil Shale
5	Liquids	Oil Shale
6	Gas	Biomass

2.5 Hat Creek Coal in Water Treatment

Coals,especially low rank lignites and sub-bituminous varieties,have long been known to exhibit substantial ion exchange and physico-chemical adsorptive properties and historically have played a limited role in water treatment, usually as an alternative to activated carbon. Such coal can potentially be used as a filter medium, absorbant, ion exchange resin or support medium in biological filters. Investigations of the utilization of coal for such purposes have shown that, in general, coal does not offer any unique advantage over conventional materials and, depending on the function, usually is not as efficient. Problems encountered during investigations which would pertain to Hat Creek coal utilization include blinding of treatment beds and the contamination of initial charges of water by fine particles which are colloidal in nature [Ref. 26, 27, 28].

The only water treatment areas where coal could potentially be competitive with conventional systems appears to be in the limited instances where enhanced suspended solids removal is required or for treatment of industrial waste streams which are heavily contaminated with trace metals. However the presence of substantial quantities of intercalated bands of swelling clays in Hat

Creek coal precludes its use for removal of suspended solids. Its use would likely lead to enhancement of the problem unless substantial pretreatment of the coal was first carried out. However, this would create its own suspended solids problem.

It is reported that the utilization of suitable pretreated coal as a precoat filter could be competitive with other filtration techniques. However bed blinding requires the periodic removal and replacement of surface layers where clogging occurs. It is anticipated that in the absence of suitable methods of regeneration the heating value of the removed coal must be recovered for such a use to be economical.

The adsorptive capacity of coal, in particular Hat Creek coal, is reported to be larger than activated carbon relative to selected trace metals at high influent concentrations and equilibrium levels of 300-400 mg/l. Adsorptive capacity drops off significantly when equilibrium levels of less than 5 mg/l are desired. Large quantities of coal would be necessary relative to required effluent concentrations, bed replacement, and the nonregenerative nature of coal. Unless the coal's heating value is recovered, the economics of coal use as an adsorptive agent would probably be prohibitive due to the base cost of the coal, transportation costs and subsequent disposal as a solid waste.

Utilization of Hat Creek coal for water treatment purposes appears to be quite limited in scope. Its use would be additionally restricted to those applications where facilities are available to recover the heating value of the coal. It is believed that such restricted use could not justify, in itself, mine development or the installation of coal preparation and supporting transportation facilities to produce and prepare the coal in marketable form in competition with existing, conventional supplies of water treatment materials.

3. COAL PROCESSING

3.1 Basis of Process Selection:

The discussion of general technical considerations in the preceding section indicated that the available methods of utilizing Hat Creek coal included manufacturing the following:

a) Principally solid products.

- Combustion fuel for thermal power generation.
- Conversion to solvent refined coal by hydrogenation under high severity conditions.

b) Principally liquid products including:

- Conversion to liquid forms by hydrogenation under suitable conditions of high severity.
- Conversion of synthesis gas produced from coal to liquid forms by Fischer-Tropsch and methanol process technologies.
- Pyrolysis of coal to produce tars, oils and residual char for subsequent use.

c) Principally gaseous products including:

- Conversion to "low BTU gas" based on gasification with air.
- Conversion to "high BTU gas" based on gasification with oxygen.
- Conversion to Substitute Natural Gas (SNG) by methanation of high BTU gas.

The use of the coal as combustion fuel for thermal power generation is outside ths scope of this study. For the other selected processes in the above categories material and energy balances have been estimated based on producing approximately 316.5 TJ/d (equivalent to 3663 MW_t or 50,000 BPD petroleum derived fuel oil) of energy products, exclusive of energy value credit for byproducts sulfur and ammonia.

The major focus of attention in this study is coal conversion to liquids and solids by direct hydrogenation and indirect methods in which the coal is first gasified. Sources of data related to the direct liquefaction of Hat Creek Coal were as follows:

i) H-Coal process data is based on a recent study of that process published by EPRI* and based on data provided by the developer, Hydrocarbon Research Incorporated [Ref. 29]. That work has recently been amplified in additional studies funded by EPRI [30-33].

ii) Hat Creek data estimates for the Exxon Donor Solvent Process are based on application of the process to Wyodak coal, a western USA sub-bituminous coal, as published in reports on work funded by the Department of Energy (USA). [Ref. 37-38]

iii) Data on the SRC-II (liquid fuel product) and SRC-I (solid fuel product) are based on studies performed by the contractor in connection with the Northeast Coal Utilization Program NECUP. [Ref. 34-36]

The SRC-I and SRC-II processes are not generally considered to be suitable for processing coals of the Hat Creek type. However studies by NAMCO indicate that Hat Creek coal is a reactive variety which is possibly suitable for conversion by the SRC type processes [Ref. 16]. Recent discussions with the process developers indicate that SRC type technology is applicable to subbituminous coals of Wyodak type.

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*EPRI - The Electrical Power Research Insititute, Palo Alto, California.

The various processes referred to above related to direct coal hydrogenation are described in the following report material.

The sources of data related to indirect coal liquefaction were as follows: [Ref. 39-56]

i) Private files and estimates of cost and performance data for methanol manufacture and Texaco gasification.

ii) Published sources and previous studies for B. C. Hydro on Fischer-Tropsch technology and Lurgi gasification.

iii) Recently published data on use of Winkler gasification applied to lignites in the manufacture of methanol.

iv) Data on the gasification of sub-bituminous coals by the Koppers process as published by Koppers.

v) Concepts from the literature such as combination methods where the Lurgi process is used as the primary method and coal fines are disposed of in Texaco or Koppers type gasification units.

vi) Data on the Methanol to Gasoline (MTG) process reported in work funded by the Department of Energy.

The sources of data employed in estimates of the various processes for coal gasification (Texaco, Lurgi, Koppers and Winkler) and those for conversion of synthesis gas to liquids (Fischer-Tropsch, Methanol and Methanol to Gasoline are referenced in the report and Appendix D.

The list of processes or combinations of processes selected for evaluation are tabulated as follows: (Table 3.1)

Table 3.1 Processes Selected for Evaluation

Process Type	Process Description	Case	Appendix D Figure	Table
A. Direct hydrogenation*	H-Coal	A1	D1.1	3.3
	EDS	A2	D1.2	3.3
	SRC-II	A3	D1.3	3.3
	SRC-I	A4	D1.4	3.3
B. Fischer-Tropsch synthesis	Texaco gasification	B1	D1.5	3.4
	Koppers gasification	B2	D1.6	3.4
	Winkler gasification	B4	D1.7	3.4
	Lurgi (Sell Fines) gasification	B4	D1.8	3.4
	Lurgi (Maximum Power) gasification	B5	D1.9	3.4
	Lurgi & Texaco combination	B6	D1.10	3.4
	Lurgi & Koppers combination	B7	D1.11	3.4
C. Methanol synthesis	Texaco gasification	C1	D1.12	3.6
	Koppers gasification	C2	D1.13	3.6
	Winkler gasification	C3	D1.14	3.6
	Lurgi (Sell Fines)	C4	D1.15	3.6
	Lurgi (Maximum Power)	C5	D1.16	3.6
	Lurgi & Texaco combination	C6	D1.17	3.6
	Lurgi & Koppers combination	C7	D1.18	3.6
D. Methanol to Gasoline	Texaco gasification	D1	D1.19	3.7
	Koppers gasification	D2	D1.20	3.7
	Winkler gasification	D3	D1.21	3.7
	Lurgi (Sell Fines)	D4	D1.22	3.7
	Lurgi (Maximum Power)	D5	D1.23	3.7
	Lurgi & Texaco combination	D6	D1.24	3.7
	Lurgi & Koppers combination	D7	D1.25	3.7
E. Production of synthetic natural gas (SNG)	Methanation based on Lurgi gasification	E1	D1.26	3.10

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*Cases A1, A2 and A3 produce liquid fuels and Case A4 (SRC-I) produces mainly clean solid boiler fuel by hydrogenation.

For the cases enumerated under the above studies, A, B, C and D, overall material and energy balances were estimated based on the available information for the various component subprocesses of which they are composed. Block flow diagrams, showing coal, air, water and power requirements and product rates, were drawn. Concise performance diagrams showing yields based on the energy in one tonne of coal were developed from the estimated overall performance data. This data are summarized in Figs. Dl.1 - Dl. 26 (Appendix D) and Tables 3.3 - 3.7 on the basis of coal properties summarized in Table 2.2 corresponding to the "As-Received-Corrected" basis.

In order to permit economic studies to be performed, capital cost estimates were also made for the various cases listed above. Literature and file cost data on the various processes and component subprocesses apply to different capacity plants at different points on the escalation curve. For each case adjusted estimates were prepared applying to a plant capacity of 316.5 TJ/d of products or 50,000 BPD Fuel Oil Equivalent (FOE), in 1980 Canadian Dollars.

Wherever possible, contingencies and safety factors have been eliminated from the estimated data; only basic erected plant costs are estimated. Capital costs are indicated to be "without adjustments" meaning that interest during construction, owners costs, working capital, starting costs and similar items are not included in them. It is intended that the sensitivity analyses be employed to assess the impact of various levels of contingency on the economic feasibility of the project. The addition of commonly assumed contingency levels to the quoted capital cost figures will elevate the costs considerably (See Tables 3.3 - 3.7). Economic assumptions are discussed in detail in Section 5.

3.2 Synthetic Fuel Specifications

The product specifications required on produced product qualities from the various synthetic fuel facilities evaluated are summarized in Tables 3.2A to I, as follows:

Table 3.2A	LPG Quality - Approximate Average
3.2B	Gasoline Specifications
3.2C	Jet Fuel Specifications
3.2D	Light and Heavy Diesel Fuel Specifications
3.2E	Light Fuel Oil Specifications
3.2F	Ethylene Quality
3.2G	Sulfur Quality (byproduct)
3.2H	Alcohols Quality (Fischer Tropsch)
3.2I	Ammonia Quality (byproduct)

In the case of each process studied, considerable flexibility exists to tailor the product slate and product specifications in order to match particular market demands. The product slates made in the various cases studied generally correspond to the "as produced" situation without adjustment by further product conversion and refining operations. Fischer Tropsch processing may, for example, be selected to maximize gasoline, diesel, chemicals or SNG production. Heavier H-Coal products may be converted in high proportion to transportation fuels by additional hydroprocessing. Methanol and SNG can be manufactured in varying proportions in a facility, though for this study methanol production was maximized in the methanol studies.

In some cases, product qualities exceeded Canadian market requirements and could command premium values to the synthetic fuels manufacturer selling to the petroleum refining industry. It was beyond the scope of this study to compare processes taking into account such special premium product values. However, the influence of special credits will be an important consideration in future more detailed studies. The capability of the Exxon Donor Solvent (recycle mode) process to produce high yields of excellent quality gasoline, as an example, should be considered in more comprehensive studies carried out in consultation with the process developers.

Table 3.2A LPG Quality - Approximate Average

Vapor Pressure	38°C	29 KPa (maximum)
Sulfur	wt%	0
Butane and Heavier	vol% max.	0.5 (liquid)
Moisture	wt%	0

Table 3.2B Gasoline Specifications

	Summer	Winter	
Reid VP KPa minimum	-	62	
maximum	76	97	
Lead mg/l	13	13	
Octane Number (ON)			
RON	-	-	Research Method
MON	82	-	Motor Method
(RON + MON)/2	87	89.1	

Table 3.2C Jet Fuel Specification

	Minimum	Normal	Maximum
End Point °C	-	300°C	-
Distillation Residue vol%	-	1.5	-
Flash Point	38°C	-	-
Freezing Point °C	-	-47°C	-
Net Heat of Combustion MJ/kg	-	42.8	-
Sulfur wt%	-	0.2%	-
Compositon			
napthalenes vol%	-	3.0%	-
aromatics vol%	-	22.0%	-
Viscosity -20°C CS	-	8	-

Table 3.2D Light and Heavy Diesel Fuel Specifications

	Light	Heavy
Flash Point °C	40 (min)	40 (min)
Cloud Point °C	-	0°C
Pow Point °C	-	-6°C
Distillation 90% point °C	290 (max)	360 (max)
Sulfur wt%	0.2% max	0.7% max
Cetane Number	40 min	40 min
Viscosity 38°C CS	1.2 min	4.1 max
CS	-	4.1 max

Table 3.2E **Light Fuel Oil Specifications**

Flash Point °C (min)	54°C
Water and Sediment vol% (max	0.5%
Ash wt% max	0.1%
Viscosity min/max CS	5/24 CS

Table 3.2F **Ethylene Quality**

Purity vol% min			99.9
Acetylene	vol.ppm	max	2
Hydrogen	vol.ppm	max	5
Propylene	wt.ppm	max	25
Sulfur	wt.ppm	max	1
Oxygenates	wt.ppm	max	1
Carbon Dioxide	vol.ppm	max	15
Water	vol.ppm	max	5
Nitrogen + Argon	vol.ppm	max	50

Table 3.2G Sulfur Quality (byproduct)

Purity	wt% min	99.0
Color		Bright Yellow

Table 3.2H Alcohols Quality (Fischer Tropsch)

Reid Vapor Pressure	KPa abs	11.0
Research Octane Number		93
Motor Octane Number		90

Table 3.2I Ammonia Quality (byproduct)

Purity	wt% min	99.7

3.3 Projected Process Data for Hat Creek Coal

Approximate projections from existing design data were made in order to evaluate the potential processes for manufacturing synthetic fuels from Hat Creek coal. The cases studied are as enumerated in Section 3.1. The cases for study were selected recognizing the characteristics of Hat Creek coal, as discussed in Section 2.4. In summary, the coal is suitable for gasification to varying degrees by the processes selected; it has been found to be relatively easy to hydrogenate and is therefore probably suitable for liquefaction by direct hydrogenation; and it has been found to produce unsatisfactory tar yields in Fischer Assay studies (Table 2.5) making it an unsatisfactory candidate for pyrolysis. The various processing component operations in the cases studied are described in Appendix D. The methods of linking together the component process units are indicated in Figs. D1.1 to D1.26 of Appendix D. Material requirements and product yields are indicated in Figs. D1.1 to D1.26. Tables 3.3A to 3.10 summarize estimated capital costs and material and energy balances as follows:

Table 3.3A	Direct Liquefaction of Coal Estimates Data for Cases A1 - A4
Table 3.4	Fischer-Tropsch Estimates Data for Cases B1 - B7
Table 3.6	Methanol Estimates Data for Cases C1 - C7
Table 3.7	Methanol to Gasoline Estimates Data for Cases D1 - D7
Table 3.10	SNG Estimates Data for Case E1

Thermal efficiency, an important measure of synfuel plant performance, is indicated on the above-noted set of tables of data.

Operating costs for the various cases defined relating to labor requirements, maintenance costs, catalyst and chemicals and other items are summarized in Section 5 - Economics.

3.3.1 Direct Coal Liquefaction

Table 3.3A summarizes estimated cost and performance data for processes involving hydrogenation of coal. Case A1 is representative of H-Coal process performance based on studies of application of the process to Wyodak coal. Table 3.3B compares Hat Creek and Wyodak coals [Ref. 56]. The newly developed mode of Exxon Donor Solvent (EDS) process operation, in which vacuum tower bottoms slurry (VTBS) is recycled to coal liquefaction, may be expected to resemble Case A1 in thermal efficiency and capital requirements, though it may have a greater value product slate because of higher gasoline production.

This projection of performance for the H-Coal type process producing a synthetic crude oil compares favorable with the indirect coal liquefaction cases discussed in relation to the subsequent B, C, D and E group tables. The H-coal type process is projected to have a thermal efficiency approximately ten points greater than the best indirect liquefaction case for Hat Creek coal. However, attainment of an efficiency of 60 percent is contingent upon being technically able to produce hydrogen at currently projected efficiencies by partial oxidation of VTBS. Schemes which produce hydrogen by steam reforming of the gaseous product fraction tend to be similar to indirect coal liquefaction in thermal efficiency. Schemes which produce hydrogen by steam reforming of product gas introduce the need to dispose of residue VTBS by other methods than in hydrogen manufacture.

Table 3.3A Direct Coal Liquefaction Estimates

Process	Catalytic Hydroliquefaction Processes						Non-Catalytic Hydroliquefaction Processes					
Case	H-Coal & EDS (VTBS Recycle Type Process) A1			EDS (Zero VTBS Recycle Type Process) A2			SRC-II Type Process Liquid Fuels A3			SRC-I Type Process Solid Fuels A4		
Total Plant Investment Capital $ million(CDN)												
Coal preparation and drying	120											
Coal liquefaction	935						Notes 3,4,5,6			Notes 7,8		
Hydrogen production	688											
Offsites	565											
Investment (w/o adjustments) (Note 10)	2,208 (Note 1)			2,760 (Note 2)			2,208			2,208		
Raw Materials												
Coal t/d	38,095			45,714			38,095			36,281		
Power MW	0			0			0			0		
Raw Water m^3/d	28,000			31,540			29,000			25,000		
Product Slate	TJ/d	m^3/d	t/d	TJ/d	m^3/d	t/d	TJ/d	m^3/d	t/d	TJ/d	m^3/d	t/d
Naphtha 54° API	133.1	3805		150.8	4309		}58.5	}1560	(Note 11)			
Turbine fuel 28° API	115.2	2934		104.1	2665							
Boiler fuel 11° API	68.2	1631		61.6	1476		258.0	6480		63.3	1680	
Ammonia	3.6		160	4.3		190	3.6		160	2.8		125
Sulfur	1.6		175	1.9		210	1.6		175	0.9		100
Phenols	0.7		30	0.8		36	0.7		30	0.4		17.0
SRC-Boiler fuel										253.2	6360	
Total energy products	316.5	8370		316.5	8450		316.5	8370		316.5	8040	
Thermal efficiency (Note 9)		60			50			60			63	

Notes

1. H-Coal capital cost is based on study of the process applied to Wyodak coal. Adapted from EPRI Report AF-1297.
2. EDS capital cost is based on study of the process applied to Wyodak coal. Adapted from DOE Report FE 2353-13
3. The SRC-II process is generally reported to be similar in capital requirements to the catalytic type processes. In this case, the capital cost is taken to be the same as H-Coal, since both are based on hydrogen production by the gasification of vacuum residue from coal liquefaction.
4. Reliable yield data from a commercial plant design based on Western sub-bituminous coal is not available for the SRC-II process.
5. Laboratory studies by NAMCQ indicate the feasibility of SRC-II type processing applied to Hat Creek Coal.
6. Under equivalent hydrogen consumption conditions, the SRC-II process is expected to produce a heavier product slate (more boiler fuels) than the H-Coal process.
7. SRC-I process has been found to have similar capital requirements to be catalytic hydroliquefaction processes. The smaller hydrogen manufacturing facility in the SRC-I process tends to be offset as a capital saving by the higher cost of filtration and/or solvent de-ashing facility.
8. SRC-I data is taken from former work for B.C. Hydro and data collected by the contractor for the Northeast Coal Utilization Program (USA).
9. Thermal efficiency exclusive of allowance for sulfur, ammonia and phenols.
10. Adjustments to estimate made subsequently in developing economics include adding construction interest cost and items of a similar nature - See Page 59 for additional explanation.
11. Naphtha and turbine fuel are lumped in Case A3 since insufficient data exists to estimate split of products for Hat Creek coal in the literature. See comments on Page D-10 of Appendix D.

Table 3.3B Comparison of Wydodak (Sub-bituminous) and Hat Creek Coals

Coal	Percent Dry Basis Wyodak	 Hat Creek*
Ash	9.1	46.14
Heating Value MJ/kg.	27.05	13.85
Carbon	66.8	46.14
Hydrogen	5.2	3.66
Oxygen	17.0	15.28

*Performance Blend

The EDS process, Case A2 of Table 3.3A, is illustrative of producing hydrogen from the gas and disposing of VTBS by Flexicoking, a proprietary Exxon process. A lower efficiency is predicted for EDS operation on Hat Creek coal based on information reported on Wyodak coal liquefaction plant designs. Diversion of the product gas from hydrogen manufacture into the product slate leads to an improvement in thermal performance. The Flexicoking process consists of a fluidized bed coking (pyrolysis) operation with an added fluidized bed gasifier disposing of the coke to produce fuel gas or hydrogen. Considerable development work in the application of Flexicoking to VTBS conversion has been carried out by Exxon and it may be regarded as the best available technology for this purpose.

Case A3, Table 3.3A is representative of SRC-II type process performance. The SRC-II type process produces a heavier liquid product slate with a greater gas yield than the H-Coal process. SRC-II performance differs from H-Coal in the major respect of non-use of catalyst. With the provision that the product has a lower value than that from H-Coal, the SRC-II plant appears to closely resemble H-Coal in capital requirements and thermal efficiency. However, the H-Coal process may be more technically advanced than SRC-II, in view of experience gained at the Catlettsburg H-Coal unit. SRC-II has been presented as a process aiming at the boiler (utility) fuel market in view of its heavier, lower hydrogen, product slate.

No reduction in capital is claimed in Case A4, Table 3.3A as a result of producing SRC-I type product. The SRC-I type process is penalized by the high capital cost of the section of the process where ash is separated from the molten solid SRC-I type fuel product. Filtration or solvent precipitation (solvent de-ashing) are the process elements present in SRC-I but not in coal liquefaction types which tend to destroy any advantages of the solid fuel process.

The high oxygen content of Hat Creek coal tends to penalize its conversion by direct hydrogenation. The oxygen is converted to water consuming process hydrogen in the operation. Hat Creek coal may be expected to compete poorly with higher rank coals as feed for coal liquefaction. A mid-continent U.S. bituminous coal such as Illinois No. 6, shows a thermal efficiency about ten percentage points greater than the corresponding application of a process to Hat Creek coal, and correspondingly better economics.

Considerable scope exists for further study of the applicability of direct coal liquefaction to Hat Creek coal. Further work could usefully be undertaken if economic studies supporting the present review support continued interest in the direct route.

A pilot plant test program would be necessary in order to establish fully the applicability of any direct liquefaction process to Hat Creek coal. Laboratory tests performed by NAMCO in 1978 indicated that Hat Creek Coal is reactive and readily liquefied, so that all of the major coal liquefaction process developments involving direct hydrogenation may expect some measure of success in the application.

In view of the low sulfur and nitrogen content of Hat Creek coal, the products of direct liquefaction may be expected to be of excellent quality, requiring minimum additional upgrading. The high ash content and evidence of reactivity may point to non-catalytic processes such as SRC-I and SRC-II and possibly EDS as satisfactory options. Since sulfur content may have a role in establishing reactivity, it may be necessary to employ a catalyst as in the H-Coal process in order to obtain satisfactory high conversion levels.

3.3.2 Fischer-Tropsch Synthesis

Table 3.4 is the result of a screening study examining the use of the Texaco, Koppers and Lurgi gasifiers to produce synthesis gas as feed to Fischer-Tropsch synthesis. All cases in Table 3.4 show lower thermal efficiencies, and higher capital requirements, than the direct liquefaction cases studied in relation to Table 3.4. The limited Fischer-Tropsch data available in the literature applies to use of the Lurgi gasifier. Studies where other gasifiers are used in association with the Synthol or Arge Fischer-Tropsch process are not known. For this work, the gasifiers other than Lurgi were employed to produce a gas of similar composition to Lurgi, as required by Fischer-Tropsch stoichiometry, on a methane-free basis. Adjustments were then made for the methane difference in the subsequent process analysis by reducing the steam-methane reforming process provisions.

The results of the above procedure are approximate and can only be employed as a very rough guide in assessing potential advantages of other systems. In fact, a convincing argument to adopt other than the Lurgi, established, route could not be identified.

A consideration of importance in relation to the Lurgi gasifier is the method of disposing of the ungasifiable fine coal residue. Fines are produced as coal is crushed to the size range permissible as feed to the Lurgi gasifier and in the case of Hat Creek coal amount to 36 percent of the coal supplied at battery limits [Table 2.1]. In this work four methods for utilizing the fines were considered. In the first method, the simple option of selling the fines was assumed to be accessible. However, the process has the need to import a substantial amount of power in the Lurgi case. Therefore, in the second alternative, the

Table 3.4 Fischer-Tropsch Process Estimates

Case	B1	B2	B3	B4	B5	B6	B7
Gasifier Type	Texaco	Koppers	Winkler	Lurgi Sell Fines	Lurgi Max. Power	Lurgi & Texaco	Lurgi & Koppers
Capital $ million CDN (excluding adjustments)	4416	4796	3657	3738	4347	3968	3968
Raw Materials							
Coal (AR) t/d	69890	71247	71791	66395	66395	58600	58710
Power net MW	(100)	100	150	450	(73)	315	360
Raw water m^3/d	52000	51300	41000	45800	51800	47000	47000
Products							
Fuels TJ/d	316.5	316.5	316.5	316.5	316.5	316.5	316.5
Coal fines t/d				10700			
Byproducts							
Sulfur t/d	230	230	230	150	150	175	175
Ammonia t/d				300	300	210	210
Thermal Efficiency							
(a) Including byproducts	35.5	31.5	30.9	36.0	37.3	36.3	35.8
(b) Excluding byproducts	35.2	31.3	30.7	35.1	36.4	35.6	35.1

option selected was to employ fines as boiler fuel for power generation, supplying all gasifier and process power needs and producing a net amount of power for export. In the third option, the fines were disposed of in a Texaco gasifier, and in the fourth option, in a Koppers gasifier. Table 3.4 shows no significant differences between the various cases of fines utilization. The difficulty of applying the Texaco gasifier in service to Hat Creek coal probably excludes that case, unless the tars, phenols and oils produced by the Lurgi units can be made to supplement or reduce the use of water in the coal slurry feed to the Texaco gasifier. The possibility of employing the Lurgi gasifier itself to dispose of a considerable portion of tar, oil and phenolic byproducts, as assumed, should be recognized, though this is sometimes a debated point. The Lurgi/Koppers combination is an all commercial system. The Koppers units could be used to dispose of environmentally dangerous materials such as phenols, etc. since these can accommodate both liquid and finely ground solid feedstocks.

The maximum power Lurgi case, where the fines are employed for power generation, shows the largest contingency-excluded capital requirement, due to the incremental cost of the boilers. The viability of this case will depend on the price of power available from outside sources at the Hat Creek site and the sales price of fines, as sold outside the plant.

One of the more costly features of the Fischer-Tropsch process is the extreme range of products. As a result of the wide product spectrum, the Fischer-Tropsch process involves extensive product separation and refining facilities as part of the complex. More specific Fischer-Tropsch catalysts may make the process more attractive. However, specificity of output has not as yet been established in this synthesis technology. The following Table 3.5 summarizes the product slate selected as the basis for Fischer-Tropsch review in this study.

The thermal efficiency quoted in Table 3.4 corresponds to conversion of part of the light hydrocarbons produced to synthesis gas by steam reforming, thereby reducing its level to that quoted. The availability of a market for gaseous, methane and ethane, products may be expected to lead to higher efficiencies. Reports on Sasol plant performance, based on marketing gaseous products lead to the expectation of thermal efficiencies of on the order of 60 percent based on low rank coal gasified by the Lurgi process. The possibility that the Fischer-Tropsch process may be made to match closely a mixed requirement for gas, diesel fuel, gasoline, olefins and alcohols while maintaining satisfactory, economic, levels of thermal efficiency should continue to be considered.

An important difference between performance and economics of Fischer-Tropsch and direct coal liquefaction processing relates to the type of product. The Fischer-Tropsch process produces light, potentially premium, fuel and chemical products which are refined by well-established oil refining technology.

The same may be claimed for the direct liquefaction processes but there are more unknowns and more difficulties associated with the upgrading of direct liquefaction products, particularly in the heavy range. Carcinogenic hazards may be another significant hurdle for direct coal liquefaction to negotiate before Sasol-scale projects can become commercially realized.

Table 3.5 Product Slate Used for Fischer-Tropsch Estimates

	HHT TJ/d	m^3/d	t/d
C_1 - C_4	6.2		108
Ethylene	34.8		670
Jet fuel*	31.3	850	
Gasoline	167.5	4800	
Diesel (D_1&D_2)	58.1	1574	
Light fuel oil	5.0	119	
Mixed alcohols	13.6	563	

Note: Coal fines production in cases where Lurgi process is used and coal fines are sold (see Table 2.1).

*Included in avaiation gasoline.

3.3.3 Methanol and Methanol-to-Gasoline Synthesis

Tables 3.6 and 3.7 summarize methanol studies and methanol to gasoline (MTG) studies based on the gasifiers considered in relation to previous Fischer-Tropsch related discussions of Table 3.4. Methanol production generally shows higher thermal efficiency and lower capital requirements than the corresponding Fischer-Tropsch cases. MTG is somewhat less attractive on this basis than methanol but is considerably more attractive than the corresponding Fischer-Tropsch case. The MTG process products are more widely generally useful than methanol in present energy economies. However, a widespread swing to methanol-based fuels could lead to the direct use of methanol in transport, obviating the need for MTG.

Many of the observations made in the previous discussion of Fischer-Tropsch technology apply in the case of methanol and MTG processing. The limitations on Texaco gasification, because of the high water and ash content of Hat Creek coal, make it a possibly unsatisfactory process for use in methanol and MTG based processes. However, the operating pressure of the Texaco gasifier is well matched to that of methanol low pressure technology and the results indicate that the Texaco process should be considered in any further venture evaluation studies. The advantage which Lurgi processes seemed to have in Fischer-Tropsch studies appears to be much reduced or not to exist in the cases of the methanol and MTG process routes. Only the Koppers-based process route appeared to border on a significantly lower economic level of attractiveness than Lurgi-based technologies. Environmental constraints on Lurgi processing and the possible difficulty in recycling tars and oils to extinction in that case may make the Koppers process more attractive than Lurgi, irrespective of economic projections.

Table 3.6 Methanol Synthesis Estimates

Case		C1	C2	C3	C4	C5	C6	C7
Gasifier		Texaco	Koppers	Winkler	Lurgi Sell Fines	Lurgi Max. Power	Lurgi & Texaco	Lurgi & Koppers
Capital $ million CDN (excluding adjustments)		2990	3226	2473	2703	3117	2818	2818
Raw Materials								
Coal	t/d	47483	48027	44671	46647	46647	41371	41123
Power net	MW	(75)	350	150	359	0	275	370
Raw water	m^3/d	34600	34600	20600	35450	42000	37230	37000
Products								
Fuels	TJ/d	316.5	316.5	316.5	316.5	316.5	316.5	316.5
Coal fines	t/d				7342			
Byproducts								
Sulfur	t/d	150	150	150	150	150	150	150
Ammonia	t/d				300	300	210	210
Thermal Efficiency								
(a) Including byproducts		51.2	42.3	48.5	50.2	50.2	50.3	48.6
(b) Excluding byproducts		50.9	42.1	48.3	49.0	49.0	49.4	47.6

Table 3.7 Mobil Type (Methanol to Gasoline) Estimates

Case		D1	D2	D3	D4	D5	D6	D7
Gasifier		Texaco	Koppers	Winkler	Lurgi Sell Fines	Lurgi Max. Power	Lurgi & Texaco	Lurgi & Koppers
Capital $ million CDN (excluding adjustments)		3416	3652	2852	3105	3623	3278	3278
Raw Materials								
Coal	t/d	50385	50748	48027	51700	51700	44989	44717
Power-net	MW	(50)	375	175	369	(70)	285	380
Raw water	m^3/d	30000	30000	19200	31000	35000	28800	28600
Products								
Fuels	TJ/d	316.5	316.5	316.5	316.5	316.5	316.5	316.5
Coal fines	t/d				8843			
Byproducts								
Sulfur	t/d	170	170	170	170	170	170	170
Ammonia	t/d				330	330	230	230
Thermal Efficiency								
(a) Including byproducts		47.3	40.0	44.9	46.5	47.9	46.6	45.3
(b) Excluding byproducts		47.1	39.8	44.7	45.2	46.6	45.6	44.4

Reliability and operating experience are important considerations in selecting a gasification process synthesis process combination. Lurgi is reported to have accomplished satisfactory performance at the Sasol installation. Koppers-based methanol (and ammonia) production from coal is reported to have accomplished an operating reliability comparable with that of units employing gas and naphtha as feedstock for steam reforming. This statement is based on experience gained at the unit in Modderfontein, South Africa, which has been operating for five years [Ref. 75].

Table 3.8 Product Slate Used for Methanol Estimates

	TJ/d	m^3/d	t/d
Methanol	316.5	17535	13970

Table 3.9 Product Slate Used for MTG Estimates

	TJ/d	m^3/d
LPG	53.5	1907
Gasoline	263.0	7652

Note: Coal fines production in cases where Lurgi process is used and fines are sold (see Table 2.1).

3.3.4 Synthetic Natural Gas

Since 1977, when the previous study was completed [Ref.1] very large new reserves of natural gas have been discovered in British Columbia. Some indication of the extent of these new reserves is provided by evidence presented to the British Columbia Energy Commission during the Fall of 1980. For example, the submission by Westcoast Transmission Company Limited provided details of the historical growth of proved initial pipeline gas reserves in Northeast British Columbia. For the period since 1968 the results indicate a linear increase, the slope being equivalent to the proving of 13 x $10^9 m^3$ (490 Bcf.) per year, and this trend is expected to continue for a further five years, thereafter declining at 5 percent per year. It is further predicted that the favorable geology of northeastern British Columbia, including the Elmworth trend, Foothills structures, Devonian Reef trends and Fort St. John reefs will provide a reserve of 663 x $10^9 m^3$ (23.4 Tcf.) by 1999, compared with 357 x $10^9 m^3$ (12.6 Tcf.) in 1979 with an ultimate potential of between 570 - 850 x $10^9 m^3$ (20 - 30 Tcf.) [Ref. 84]. This situation is therefore considerably more optimistic than views expressed by the British Columbia Energy Commission in 1976 and referred to in the previous Study. [Ref. 1]

Market studies now indicate that SNG cannot compete economically with natural gas in the area potentially served by the Hat Creek coal deposit. Table 3.10 and Fig. D1.26 indicate cost and performance data associated with facilities for the manufacture of SNG based on use of fines for power generation and the Lurgi gasifier. However, it should be noted that as a synthetic fuel considered on a $/GJ basis, SNG is highly competitive with all other synfuels considerd in this study.

Table 3.10

Comparative SNG Data

Case		E1
Gasifier		Lurgi (Max. Power Production)
Capital $ million (excluding adjustments)		3105
Raw Materials		
Coal	t/d	38100
Power-net	MW	0
Raw water	m^3/d	50000
Products		
Fuels	TJ/d	316.5
Coal fines	t/d	0
Byproducts		
Sulfur		120
Ammonia		250
Thermal Efficiency		
(a) Including byproducts		61
(b) Excluding byproducts		60

Table 3.11 Estimates of Ammonia and Sulfur Production

Process	Gasifier	Texaco	Koppers	Winkler	Lurgi Sell Fines	Lurgi Max. Power	Lurgi & Texaco	Lurgi & Koppers
Fischer-Tropsch	Ammonia t/d				300	300	210	210
	Sulfur t/d	225	230	230	150	150	175	175
Methanol	Ammonia t/d				300	300	210	210
	Sulfur t/d	150	150	150	150	150	150	150
Methanol to Gasoline	Ammonia t/d				330	330	230	230
	Sulfur t/d	170	170	170	170	170	170	170

4. PRODUCT MARKETS

4.1 World Oil Outlook

Oil is projected to remain the largest single energy source at least for the next 20 years. The world's remaining conventional oil resources are estimated to be in the range of 1 to 1½ trillion barrels. This number includes oil which has yet to be discovered.

Growth of world oil consumption in the past has been largely in the major industrialized countries. In 1979, almost 58 percent of the world's total oil consumption was in the United States, Europe and Japan. This figure is projected to drop to 48.6 percent in 1990 and to 42.9 percent in the year 2000.

Table 4.1

Projected World Oil Demand

Area	million m^3/d		
	1979	1990	2000
United States	2.9	2.5	2.4
Europe	2.4	2.1	2.1
Japan	0.8	0.8	0.8
Other Industrial Countries	0.6	0.6	0.6
Developing Countries	1.7	2.7	3.8
Centrally Planned Economies	2.1	2.4	2.5
Total	10.5	11.1	12.2

[Ref.100]

The ability to produce oil is limited by the availability of discovered reserves, reservoir characteristics and the rate at which new reserves are found and developed. The use of secondary and tertiary recovery methods, by substantially increasing the stock of recoverable reserves, is beginning to make a major contribution to existing resources but it is anticipated that more oil will continue to be produced than discovered, so discovered reserves will continue to decline.

There is a growing tendency among many oil exporting countries to limit their oil production in anticipation of higher prices. Limitation of production quotas has been a major feature of the OPEC cartelization. The relatively small current needs of some of these countries for increased revenues and their problems in finding satisfactory investments for excess funds also encourage this tendency.

Oil supply is expected to increase by 10 percent between 1990 and 2000, North American supply is expected to drop by 25 percent from its 1979 level of 12 million b/d to 9 million b/d in 1990. Table 5.2 below shows 1979 actual and projected 1990 and 2000 world oil supply from major sources.

Table 4.2

Projected World Oil Supply

	million m^3/d		
Source	**1979**	**1990**	**2000**
United States and Canada	1.9	1.4	1.4
Europe	0.3	0.6	0.6
Other	1.0	1.6	2.1
OPEC	5.1	4.8	4.6
Centrally Planned Economies	2.2	2.4	2.5
Synthetic and Heavy Oil	-	0.3	1.0
Total	10.5	11.1	12.2

[Ref. 100]

Conventional oil share of world energy supply is expected to shrink from an estimated 54 percent in 1980 to 45 percent in 1990 and only 37 percent by the year 2000 (Table 4.3). Oil exploration and production will become more costly because many of the newer oil reserves are situated in remote locations, in harsh operating environments or are offshore.

Because of longer-range limits on the ability to continue to expand conventional oil and gas development, synthetic fuels are likely to be needed at some time in the future to meet unique liquids and gas requirements in transportation and certain industrial sectors, especially petrochemicals. Because there have been no new confirmed projects to construct commercial synfuels facilities, any projection for the future is highly speculative. Depending upon events in the next few years, a synfuels industry could emerge in several countries. This projection includes a total of some 20 to 30 plants, producing nearly 2 million b/d oil equivalent by the end of the century.

The lead times for a coal-based synthetic fuels project are estimated at 7-10 years. The plants to produce these fuels are technologically complex and many will be situated close to coal mines in remote areas, often requiring a long time for construction. However, as experience is gained and standardization of technology takes place, these lead times could be reduced.

Table 4.3

Projected World Energy Supply

Source	Percent of Total 1980	1990	2000
Oil	54	45	37
Gas	18	18	16
Synthetic fuels	0	2	4
Coal	18	20	24
Nuclear	3	7	10
Hydro & Other	7	8	9
Total	100	100	100

[Ref. 100]

4.1.1 Chemical Feedstock Requirements

Despite the fact that the chemical industry's consumption of feedstock amounts to only 3 percent of the world's total demand for petroleum and natural gas, the uncertainties associated with petroleum and natural gas supplies and their increasing costs have led most manufacturers to begin searches for new raw materials. The interest in coal-derived chemicals has been renewed. Also under study as potential replacements for oil and natural gas are other fossil materials such as oil shale, tar sands and peat as well as non-fossil materials such as organic wastes, land and water vegetation, natural biochemical methods, and the indirect conversion of solar energy to organic chemicals via hydrogen and inorganic carbon resources.

In the next decade, petroleum and natural gas will continue to serve as the largest sources of feedstocks for organic chemicals manufacture. Only small contributions will be made by other materials to the feedstock pool. But beyond the late 1980's increasing use of other materials is expected to occur. Coal is expected to make major contributions in the 1990's, and then renewable non-fossil materials and oil shale and possibly peat are projected to be used on a large scale.

Technology advances in both fuel and feedstocks will be coming at a fast rate, forced by increasing prices for conventional feedstocks and fuels. Chemicals derived from coal may be obtained either as byproducts of coal carbonization which has been their principal source of supply and in which lies the origin of much of the present petrochemicals industry, or as primary products of coal liquefaction, and gasification. All the necessary ingredients for coal-derived chemical products are there, but economically competitive technology and/or petrochemical pricing is not expected to be sufficient to attract large coal chemical production before the 1990's.

4.1.2 Oil Prices

World oil prices are expected to increase during the next 20 years, in real terms after adjustment for inflation, by 50 percent from current levels. From 1975 to 1978, oil prices were essentially flat in real terms, then they almost doubled after the Iranian revolution in 1978. World oil prices have increased by an average rate of 34 percent annually between 1970 and 1980, (Table 4.4)

Table 4.4

World Oil Prices

Year	U.S $/$m^3$	U.S. $/b	Percent Average Annual Increase over 1970
1970	8.74	1.39	-
1973	12.58	2.00	13.0
1974	52.33	8.32	56.4
1977	76.04	12.09	36.2
1979	83.91	13.34	28.6
1980	163.53	26.00	34.0

[Ref. 100]

The projected prices for crude oil, if materialized, could have far-reaching consequences for worldwide energy supply and demand. As a result, the growth in the world gross national product, which has averaged 5 percent during 1965-1973, could drop substantially in the next 20 years.

4.2 Canadian Oil Outlook

According to industry sources, Canada has only a remote chance of becoming independent of imported oil in the 1990's.

Only with maximum production of conventional and heavy oil from Western Canada, acceleration of current plans for frontier production, and completion of tar sands plants would self-sufficiency be possible in the early 1990's. [Based on submissions to the National Energy Board in the matter of Order EHR-1-80]

Table 4.5

Crude Oil Consumption in Canada

million m^3

	1979	1978	1977	1976	1975
Production	78.5	68.1	65.5	56.1	51.0
Imports	30.2	35.8	38.8	42.0	47.9
Total	108.7	103.9	104.3	98.1	98.9

A more realistic outlook is that Canada will require average imports of 31,800 - 47,700 m^3/d (200,000 - 300,000 b/d) throughout the 1990's because it is unlikely that multi-billion dollar projects can be accelerated. Even this level of import dependence would require expansion of existing syncrude tar sand plants by 1990.

This level of oil self-sufficiency also requires construction of three upgrading plants for western Canadian heavy oil production and about 47,700 m^3/d (300,000 b/d) from the frontiers.

Table 4.6

Crude Oil Reserves in Canada

1/1/1980

million cubic meters

Region	Natural Depletion	Enhanced Recovery	Total
Mainland	12.0	13.0	25.0
British Columbia	37.9	40.0	77.9
Alberta	1,316.4	721.1	2,037.5
Saskatchewan	219.6	162.4	382.0
Manitoba	14.2	11.5	25.7
Ontario	8.8	1.3	10.1
Other	0.1	-	0.1
Total	1,609.0	949.3	2,558.3

Table 4.7

Natural Gas Reserves In Canada

1/1/1980

billion cubic meters

Region	10^9 m^3
Mainland	38.9
Mackenzie Delta/Beaufort Sea	240.0
Arctic Islands	524.8
British Columbia	570.4
Alberta	4,116.1
Saskatchewan	160.7
Manitoba	1.7
Ontario	29.0
Other Eastern Canada	1.3
Total	5,682.9

4.2.1 Canadian Energy Policy

During recent months Canada's principal oil producing provinces, Alberta, Saskatechewan and British Columbia have signed separate agreements with the Canadian federal government concerning the future pricing of crude oil and the distribution of tax and royalty revenues. These agreements are generally consistent with the aims of the National Energy Policy announced in October 1980.

Table 4.8

Estimated 1980 Canadian Oil Production

Rank	Province	Percent
1	Alberta	86.0
2	Saskatchewan	10.0
3	British Columbia	3.0
4	Manitoba	0.6
5	Northwest Territories	0.3
6	Ontario	0.1
		100.0

The recent negotiation regarding an energy policy in Canada involved the following issues:

- By the year 1984 Canadian domestic oil prices to equal 85 percent of Chicago prices or the landed cost of imported oil at Montreal, whichever is lower.
- Oil prices to increase by $4 per barrel in 1980 and $4.50 per barrel annually during 1981-1983.

Industry	0.90	0.90
Province	1.90	2.15
Federal	1.20	1.45
	4.00	4.50

- Natural gas price to be based on 85 percent heating value parity with oil.

Future crude oil prices for Canada were based on the National Energy Program. Conventional oil prices are expected to reach $20.70 per barrel in 1985 and $30.20 per barrel in 1990, in 1980 constant dollars. This forecast is based on an annual average increase of 6.1 percent.

Table 4.9

Projected Canadian Crude Oil Prices
1981 Dollars

	1981	1985	1990	Percent Annual Increase 1981-1990
Oil Sands	38.00	38.00	38.00	-
Tertiary Oil	30.00	30.00	30.00	-
Conventional Oil	17.75	20.70	30.20	6.1
Estimated Blended Oil	22.40	27.30	33.75	4.3

4.2.2 Tar Sands

In Canada, tar sands represent a huge resource of heavy oils and represent the principal means for replacing the declining production of conventional oil in the western provinces. Two commercial plants are already on stream and others are planned. The largest of the commercial plants is Syncrude Canada, 130,000 b/d unit near Mildred Lake, north of Fort McMurray in Alberta. It is expected to turn out more than 100,000 b/d of synthetic crude oil for the 25 year economic life of the plant. The Alsand Project Group headed by Shell Canada Ltd. is planning a 140,000 b/d (synthetic crude and LPG) surface mining and bitumen upgrading facility, also north of Fort McMurray. It will use essentially the same processing scheme as Syncrude.

4.2.3 Synfuels in Canada

Liquid fuels are of particular importance to Canada because it does not produce enough of them to meet its needs.

New imports of oil excluding LPG's are currently about 300,000 b/d. This deficit is of great concern to federal and provincial governments in Canada because foreign oil has become very expensive and the supply may not always be secure.

Most of Canada's liquid fuels are produced in the western provinces. The heaviest requirements are in eastern Canada so that long-range transportation becomes a major difficulty. Canada, therefore, exports oil to western United States and imports oil from OPEC countries into the eastern provinces.

Coal liquids produced by direct hydrogenation are highly aromatic and could be refined through hydrotreating to high octane gasoline, heating oil, and boiler fuel. The production of a satisfactory diesel fuel would require more hydrogen and considerable processing, and would be more expensive.

In western Canada, coal liquids could be cheaper because of the availability of natural gas. The use of natural gas to supply the necessary hydrogen, rather than generate it internally in the process, would lower the cost. A recent report by the Canadian government has concluded that coal liquids are likely to remain more expensive than crude oil from tar sands. [Ref. 94]

In British Columbia, the Energy Development Agency of the provincial government is conducting separate feasibility studies into the technology and economics of producing transport fuels in a coal-to-liquids facility located at Hat creek.

In the Maritime Provinces, the formation of a new consortium of Canadian companies, the Scotia Coal Synfuels Corporation, was announced in April 1981. The Corporation proposes to construct and operate, on a site on Cape Breton Island, a plant to produce liquid transportation fuels from bituminous coal.

Among other major proposed synthetic fuels projects complementary or competitive with a possible Hat Creek synfuels project, Petro-Canada, British Columbia Resources Investment Corporation, and Westcoast Transmission, have formed a consortium to develop a U.S. $3 billion to $5 billion coal liquefaction plant in British Columbia. A study group has been set up to do the following:

- Select the most suitable coal deposit in British Columbia for a liquefaction plant.
- Examine environmental and other impacts of the development.
- Select the most suitable liquefaction process that will be used in the plant.

The consortium has already applied to the Canadian government for matching financing to research the project. The consortium has also invited two Japanese companies to participate.

There are also several proposals for manufacture of methanol from natural gas feedstocks at west coast sites. These projects could ultimately produce significant quantities of liquid fuels or chemical intermediates, both for home and export use, and have a major impact on liquid fuel supplies in British Columbia.

4.3 Upgraded Solid Products

Solvent-refined coal (SRC) can be used as boiler fuel, hydrogenated to liquid fuels, or converted to low-ash coke for electrodes. There is good evidence that an acceptable product can be obtained; the solvent-refining technology is being developed in the United States, the United Kingdom, and elsewhere; the delayed coking and coke calcination technologies are already developed. Solvent refining without hydrogenation, followed by coking of the dissolved coal, under development by the British National Coal Board, may prove to be more economical than solvent refining with hydrogenation because the hydrogen and the high extraction pressure required for hydrogenation add considerably to process costs. However, this process is still a long way from the necessary commercial demonstration required for consideration in this study.

4.3.1 Electrode Carbons and Petroleum Cokes

A review of possible alternatives to petroleum coke for the manufacture of carbon electrodes for aluminum production led to the conclusion that production of electrode coke from solvent refined coke in the long run. [Ref. 89]

Coke is produced in Japan, West Germany, Poland,and the Soviet Union by the coking of coal-tar pitch. Production capacities in Japan and West Germany are about 400,000 and 200,000 tonnes per year respectively. This coke is preferred to petroleum coke for aluminum production, especially for Soderberg anodes, largely because of its low sulfur and vanadium content. It commands an appreciably higher price. No pitch coke is produced in North America because it is not economical.

(i) U.S. Market

Present U.S. petroleum coke production is about 40,000 t/d most of which is delayed coke. Only 3,650 t/d of fluid coke is produced. Approximately 56 percent of the total coke produced in the U.S. is fuel grade, and 44 percent is calcined for higher value use.

Table 4.10 Distribution of U.S. Petroleum Coke Market

	Percent Calcined	Percent Fuel
Domestic	19	12
Export	18	44
Electrode	2	-
Other	5	-
Total	44	56

An eventual major shortage of petroleum coke could not be met by increased production of pitch coke because of an insufficient supply of coal tar. In the United States in 1980, an estimated 220 million gallons of tar was burned as fuel. Even if all this tar was diverted to coke production, which is unlikely for economic reasons, it would produce 300,000 tonnes per year of pitch coke. This is only 5 percent of the current consumption of calcined coke by the world aluminum industry outside communist areas which is 6 million tonnes in 1980. [Ref. 101]

(ii) Canadian Market

Alcan Trading Limited of Montreal is presently searching for a long-term source of 300,000 tonnes per year green petroleum coke for the Pacific area. The company requires a coke with 3 percent sulfur maximum and metals generally under 0.03 percent each. There are also sizing and strength characteristics which must be met. The market price for calcined petroleum coke is about $220/tonne.

Great Canadian Oil Sands Ltd. in Alberta produces 2,800 tonnes of delayed coke per stream day of which 2,300 tonnes are burned as fuel and 500 tonnes are stockpiled. Syncrude Canada Ltd., also in Alberta, produces 2,000 tonnes per day of fluid coke which is being stockpiled until an economic use is developed. Both these cokes, however, are unsuitable for electrodes. In addition to high sulfur, vanadium, and nickel content, they are high in silicon and other elements because of residual colloidal clay not removed during the hot water separation process.

Alcan has carried out a bench-scale investigation of the purification of this coke. Leaching with a mixture of hydrochloric and hydrofluoric acids removes significant amounts of the metallic impurities from fine particles,but is rather ineffective on the coarse coke also required for anode manufacture. Treatment with chlorine at 1,400°C removes significant proportions of the metallic impurities, but the silicon and vanadium contents still exceed the maximum values specified for anode coke. They are,therefore,not regarded as potential future sources of high grade electrode carbons.

The posssible production of solvent-refined coal at Hat Creek together with subsequent conversion to electrode carbons as a second process step is therefore a feasible potential utilization. However, only the primary conversion, to solvent-refined coal, is considered further in this study.

4.4 Gaseous Products

4.4.1 Liquefied Petroleum Gas

(i) Propane

Canada produces 20670 m^3/d (130,000 b/d) propane of which only about one-half is required for domestic consumption. The current thinking is to divert excess propane supply into new domestic markets. These markets include:

- Heating oil: as much as 4770 m^3/d (30,000 b/d), 23 percent of current propane supply, could displace heating oil in central and eastern Canada by 1985.

- Vehicle fuels: it is estimated that by 1990 vehicle fleets could consume more than 3980 m^3/d (25,000 b/d) propane. Propane is an excellent vehicle fuel with 10 percent higher combustion efficiency than gasoline. It also lowers service costs and prolongs engine life by between two-to-three times compared with gasoline engines.

- Petrochemicals: propane could be an acceptable feedstock for ethylene and propylene production. A propane-based ethylene and propylene plant could consume 6360 m^3/d (40,000 b/d).

The most important factor limiting the widespread use of propane in Canada is price. Propane price in eastern Canada is higher than the domestic price of oil products it might replace. This, however, is not always true in western Canada because of the higher transportation cost to the east. The new discoveries of natural gas in the west and its competitive price advantage over propane has contributed to the increase in propane surplus. Propane is currently in surplus in Canada and exported to the United States and Japan.

Table 4.11

Propane (Liquid) Supply/Demand Forecast

	$10^3\ m^3/d$		
Canada	**Supply**	**Demand**	**Balance**
1980	20.8	7.1	13.7
1990	22.5	11.3	11.2
2000	20.4	12.9	7.5
British Columbia			
1980	0.2	0.6	(0.4)
1990	0.3	1.0	(0.7)
2000	0.3	1.2	(0.9)

[Ref. 77, 91]

While Canada is expected to have propane surplus over the next two decades, British Columbia is expected to have a deficit. The 1980 estimated deficit of 400 cubic meters per day is expected to reach 700 cubic meters per day in 1990 and 900 cubic meters per day in the year 2000.

The Ministry of Energy, Mines and Petroleum Resources has estimated that annual growth rate in LPG requirements will average 0.5 percent annually over the next fifteen years.

Table 4.12

B.C. LPG Requirements

Petajoules

1978	1981	1986	1991	1996	Average Annual Growth Percent
8.94	9.16	9.41	9.31	9.80	0.5

[Ref. 77]

(ii) Butane

Canada consumes only a small amount of its butane supply. The federal government estimates that in 1980 Canadian demand for butane was 1,500 cubic meters per day, or less than 12 percent of total supply of 12,700 cubic meters per day, with a surplus of 11,200 cubic meters per day.

Table 4.13

Butane (Liquid) Supply/Demand Projection

	10^3 m^3/d		
Canada	**Supply**	**Demand**	**Balance**
1980	12.7	1.5	11.2
1990	14.0	1.9	12.1
2000	12.7	2.1	10.6
British Columbia			
1980	0.3	0.13	0.17
1990	0.4	0.14	0.26
2000	0.4	0.14	0.26

[Ref. 77, 91]

British Columbia butane requirements are far less than supply. The 1980 surplus of 170 cubic meters per day is expected to reach 260 cubic meters per day in 1990. Supply is expected to reach 400 cubic meters per day in 1990, a one-third increase over the estimated figure of 300 cubic meters per day in 1980.

4.4.2 Ethylene

Ethylene is the world's most important petrochemical feedstock in both volume produced and consumed. Future growth of the petrochemical industry will depend largely on having enough ethylene to meet the growth of its plastics, fiber, paint, and solvent derivatives. North American, Western Europe and Japan accounted for 81 percent of the world's ethylene plant capacity in 1980.

Table 4.14

World Ethylene Capacity*
million tonnes/year

Country/Area	1980	1990	Percent 1980-1990 Annual Increase
Canada	1.5	3.5	8.8
United States	16.8	30.2	6.0
Latin America	2.6	4.8	6.5
Western Europe	17.0	33.7	7.1
Japan	6.0	12.1	7.2
Pacific	1.4	3.7	10.2
Eastern Europe	5.0	12.4	9.4
Other	0.7	2.4	13.2
Total	51.0	100.4	7.0

Current world ethylene capacity is more than adequate to meet projected demand for the next few years. The United States ethylene capacity was pushed near its operating limit during 1979 and the early part of 1980 mainly because of the strong export market for ethylene derivatives.

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*Stone & Webster Estimates

However, the new capacity addition which came on-stream during the second half of 1980 coupled with the economic recession have resulted in lowering operating capacity at the end of 1980. As a result, the United States ethylene capacity should be adequate to meet future demand probably through 1985, without any new projects.

Worldwide capacity gained about four million tonnes during 1980. Almost half of the additional capacity was in the United States. This increase in Worldwide capacity is outstripping increases in demand. No new major ethylene projects are planned for Western Europe in the immediate future. Industry sources indicate that Western Europe could achieve a comfortable supply/demand situation by 1990 if those plants already announced are built.

In Japan, many of the projects have been scheduled for a number of years but delayed by the sharp rise in energy and feedstock prices and slow economic conditions.

The Middle East/Africa and Asia/Pacific area total of ethylene projects jumped from a combined 13 to 77 during 1980. The major portion of these are located in Japan (five), China (four), and Saudi Arabia (four).

Over one-third of the world's ethylene production goes into polyethylene, a major factor in the growth demand for ethylene in recent years. About 25 percent is used for ethylene oxide and its derivatives, of which the most important is ethylene glycol used in antifreeze, polyester fiber, and film. About 10 percent of the ethylene production goes into ethanol, while ethylene dichloride, an intermediate for vinyl chloride, and ethylbenzene, an intermediate for styrene monomer, take about 9 percent each.

Table 4.15

Worldwide Ethylene Uses*

Major Uses	Percent of Total
Polyethylene	34
Ethylene Oxide	25
Ethanol	10
Ethylene Dichloride	9
Ethylbenzene	9
Other	13
	100

Future ethylene capacity will depend on several factors chiefly among them are:

- Future availability of petroleum feedstocks
- Alternative sources for hydrocarbons
- Technology
- Economics
- International Developments.

(i) Feedstocks

In North America, traditional feedstocks for ethylene production consisted of ethane and propane. The trend in recent years has been toward increasing use of naphthas and middle distillates. Prospects are that heavy fuel oil may become increasingly available at attractive prices in the future. Pretreatment of this material would significantly enhance its value as a petrochemical feedstock.

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*Stone & Webster Estimates

Table 4.16

Ethylene Feedstock Sources*

Percent

	1980			1985		
	Naphtha	Gas Oil	Gas Liquids	Naphtha	Gas Oil	Gas Liquids
United States	33	34	33	33	43	24
Western Europe	88	9	3	86	11	3
Japan	100	-	-	98	-	2
Other	98	-	2	95	-	5
World Average	77	13	10	77	14	9

(ii) Alternative Sources for Hydrocarbons

Although there is a consensus that petrochemical producers will be able to successfully bid feedstocks away from the fuel market, the ability to use a variety of different materials to produce ethylene will become increasingly important. Coal utilization as a source of hydrocarbon feedstock for ethylene plants seems a logical extension of the trend by the industry to use cheaper feedstocks. Technology already exists for converting coal into other clean forms of energy. Sasol's plant in South Africa is tied in with its coal-based fuels industry, now in the active stages of construction. Sasol II is completed and significant progress has been made on Sasol III. A promising route to making ethylene from coal is found in Mobil's methanol-to-gasoline process that uses a zeolite catalyst. By stopping the reaction

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*Stone & Webster Estimates

at the proper state, the zeolite can also be used to produce light olefins and aromatics.

(iii) Technology Development

Industry sources predict that unit plant size will continue to increase slowly. In the near future, the first olefins plant with a capacity of more than 700,000 tonnes per year will eventually be realized. These will be cautiously larger versions of existing plants without any major inovations.

Regarding the impact of inflation on the petrochemical industry, plant investments per unit of output have more than doubled over the last ten years. Not only has this resulted in substantially higher manufacturing costs despite larger, more efficient plants, but the capital needs of the industry are growing enourmously.

A major factor contributing to increased plant costs in industrialized countries is environmental control. Based on recent estimates for a 500,000 tonnes per year ethylene plant, approximately 12 percent of the inside battery limits plant cost was associated with environmental control systems.

Table 4.17

Environmental Protection Costs*

	Percent of Total Plant Cost
Air Pollution	3
Water Pollution	7
Safety	2
Total	12

(iv) Economics

Economic considerations are forcing ethylene producers to continue to keep older plants in operation because the higher costs of new plants more than offset the savings attributable to technology and size benefits. Thus, to some extent, inflation will slow down rapid progress towards greater energy efficiency by maintaining the competitiveness of older, less efficient plants.

(v) International Developments

As already noted the world's ethylene capacity is concentrated in the industrialized countries of North America, Western Europe and Japan. Developing countries have recognized the potential contribution that a petrochemical industry could make to their industrial development programs. Ambitious plans have been developed and to a considerable

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*Stone & Webster Estimates

extent, are presently being implemented. Mexico and Brazil have more than one million metric tons per year of ethylene production capacity currently in operation or under construction. In the Far East, Korea and Taiwan have already developed a significant petrochemical industry. Oil producing countries, based on an abundant supply of natural-gas liquids, are expected to build large petrochemical complexes.

Development of petrochemical industries in developing countries generally faces numerous obstacles even where raw materials are plentifully available. One of the major problems is the lack of adequate infrastructure to support such developments. Construction and operating personnel must be trained, technology and equipment has to be imported from industrialized countries, and markets for the products must be developed.

Initial cost, even for facilities of modest capacity and complexity usually run over U.S. $1 billion. Financing of such investments is obviously not an easy task.

In many instances, developing countries have sought partnership with established foreign companies to develop a domestic petrochemical industry. Based on actual implementation of projects, this approach has generally been proven to be more successful than attempts by developing countries to establish a petrochemical industry independently. A foreign partner can usually provide technical, managerial and marketing expertise to the project, and his participation can be instrumental in obtaining financing. Joint ventures between developing countries and multinational companies in the development and operation of petrochemical complexes such as those in Brazil and Korea, have demonstrated that this arrangement can be successful.

(vi) Canadian Ethylene Market

The Canadian petrochemical industry dates from the discovery of substantial oil resources in Western Canada in the early 1950's. By 1960, the Canadian oil-refining industry was the third largest in the world (after the United States and the Soviet Union) in terms of crude oil capacity. However, the development of basic petrochemical industry lagged behind the development of the Canadian market for these raw materials. During the 1960's, it was easier for products to move from the United States into Canada because Canadian tariffs were relatively low and the Canadian market was too small to support large chemical facilities.

In the 1970's Canadian ambitions for petrochemical growth were boosted by rising oil and gas prices. Based on substantial oil and natural gas reserves in Alberta, several world-scale chemical complexes were built.

Canada now has two world-scale ethylene plants, Petrosar's Sarnia plant, which produces 455,000 tonnes of ethylene as well as other petrochemical feedstocks annually from naphtha imput, and the Alberta Gas Ethylene (AGE) plant in Alberta, which uses natural gas for ethylene production.

Table 4.18

Canadian Ethylene Producers

Producer	Location	thousand tonnes 1980	 1990
Esso	Sarnia	225	225
Gulf Canada	Quebec	225	225
Petrosar	Sarnia	455	455
Union Carbide	Montreal	75	75
AGE	Joffre	544	1,305
AGE	Alberta	-	600
Dome Petroleum	B.C.	-	275
Petromone	Montreal	-	300
Total		1,524	3,460

[Ref. 91]

In 1980, Western Canada accounted for about 36 percent of Canada's ethylene capacity. Western Canada's share of the country's total ethylene capacity is expected to reach 59 percent in 1985 and 63 percent in 1990

Table 4.19

Canadian Ethylene Capacity
thousand tonnes per year

	1980	1985	1990	Percent Annual Increase 1980-1990
Western Canada	554	1,380	2,180	14.7
Rest of the Country	1,005	965*	1,280	2.5
Total	1,559	2,345	3,460	8.3

[Ref. 91]

Western Canadian production of pretrochemical products is mainly targeted for export markets. Regional market considerations, as a factor in selecting Western Canada for large ethylene plants, have been of minor importance. The critical factor in justifying the very rapid expansion of ethylene capacity has been the advantageous price of hydrocarbon feedstock, mainly natural gas. The availability of such a source provides petrochemical producers with a lower production cost and therefore, a competitive edge in the international market. Petrochemical production from petroleum is a less desirable alternative in Western Canada. This is mainly because of the marginal price advantage over competing areas of production and the lack of a marketable surplus of liquid feedstock. As a result, petroleum products are unlikely to be used as hydrocarbon feedstock for ethylene plants in Western Canada, but rather, they will be used for benzene and toluene production.

*Note: Closing of Canadian Industries reduces capacity of 40,000 tonnes per year.

There are strong indications that the bulk of future Western Canadian ethane-based ethylene production will be upgraded in the producing provinces. These provinces will take advantage of employment opportunities and other economic benefits as a result of the expected expansion in industries based on ethylene production. The realization of these economic benefits, however, will depend, to a great extent, on the future development of a strong regional market.

Table 4.20

Ethylene Uses in Canada

End-Use	Percent of Total
Low-density Polyethylene	25
Ethylene Oxide	20
High-density Polyethylene	18
Ethylene Dichloride	14
Ethylbenzene	9
Ethyl Alcohol	3
Other	11
	100

[Ref. 91]

In the Fall of 1980, a consortium of Canadian Occidental Petroleum, Dome Petroleum of Calgary, Westcoast Transmission of Vancouver, and Mitsubishi of Japan, announced a $2 billion development that will include an upstream plant to strip and liquify ethane and propane from natural gas out of British Columbia's Fort St. John field. This third ethylene plant will have about 60 percent of a "world-scale" 450,000 tonnes per year.

The viability of petrochemical production from Hat Creek coal is likely to be independent of activity in the ethane and gas-based western petrochemical industry. Ethylene from Hat Creek coal would probably not gain significant end use markets within the region, but might find foreign market opportunities. Many factors such as feedstock prices, transportation costs, and market conditions must be carefully investigated before a decision to produce ethylene from Hat Creek coal is made.

4.5 Upgraded Liquid Products

4.5.1 Gasoline

Canadian demand for motor gasoline in 1980 has been estimated at 107,000 cubic meters per day. Over the next two decades Canadian demand for motor gasoline is expected to decrease by an annual average of between 0.5 and 1 percent. This will come as a result of a shift from gasoline to diesel and other fuels, more efficient and smaller cars, change in driving habits, and decrease in distances traveled.

Table 4.21

Demand Projection for Motor Gasoline

	10^3 m^3/d		
	1980	1990	2000
Canada	107.7	101.1	98.4
British Columbia	11.9	13.6	13.2

[Ref. 77, 91]

Demand for motor gasoline in British Columbia in 1980 has been estimated at 11,900 cubic meters per day. Over the next two decades this demand figure is expected to reach 13,600 cubic meters per day in 1990 and then decrease to 13,200 cubic meters per day in the year 2000.

Table 4.22

British Columbia Projected Gasoline Demand

Petajoules

1978	1986	1996	Average Annual Growth Percent
140.99	170.51	189.28	1.6

[Ref. 77]

Unleaded gasoline is required to prevent poisoning of the catalyst in the exhaust gas converter required to meet pollution emission standards for vehicles. Lead, however, is an inexpensive and energy-efficient octane booster for gasoline. The use of leaded gasoline in Canada, instead of unleaded gasoline, could save as much as 25,000 b/d of crude oil. To implement such a change, new car engines using leaded gasoline and meeting emission standards must be developed. [Ref. 94]

Gasoline is obviously among the attractive alternatives for Hat Creek coal utilization. Coal-derived gasoline could be used in British Columbia or exported to the United States, Japan or Latin America.

Table 4.23

Latin American Gasoline Prices*

Country	Cdn$/liter
Uruguay	1.25
Brazil	1.05
Nicaragua	0.86
Chile	0.77
Argentina	0.75
Bolivia	0.64
Colombia	0.49
Ecuador	0.23
Mexico	0.20
Peru	0.19
Venezuela	0.15

[Ref. 99]

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*January 1981

4.5.2 Jet Fuels

Increased cost and reduced availability of jet fuels along with the potential future depletion of worldwide petroleum resources has created an interest in the feasibility of obtaining jet fuel from non-petroleum resources. Crude oils from coal, oil shale and tar sands, alone or in mixtures with petroleum crudes are likely possibilities. Because of basic chemical difference in these crudes, and processing economics, future fuels may have properties which are different from those of current fuels. The end objective is to optimize the factors of availability, cost, aircraft performance, and safety.

New concepts derived from research and development require from 7 to 10 years before actual production occurs. Even longer time periods (up to 20 years) are required if research is necessary to accomplish goals.

Three jet fuel types are currently in wide use thoughout the free world.

1. JP-4 is the fuel used by the air forces of NATO, including the United States
2. Jet B a fuel nearly identical to JP-4, is used by Canadian commercial airlines.
3. Jet A is the kerosene-based fuel used by most of the worlds commercial airlines, including those of the United States.

JP-4 and Jet B fuels can be grossly represented as a blend of kerosene and gasoline. The high volatility of JP-4 results in a vapor pressure of about 0.17 bar (2.5 psia) at 310 K (100°F), and a flash point of approximately -25°C.

Jet A type has a much lower volatility than JP-4 resulting in a flash point of about 52°C.

Because of the reduced probability of post crash fires and the reduction of combat vulnerability, NATO is considering conversion to JP-8, a jet fuel identical to Jet A-1, a commercial fuel similar to Jet A in all respects except freeze point (-50°C versus -40°C for Jet A). As a result of the unique problems associated with shipboard jet fuel use, the U.S. Navy is now using a third fuel type, JP-5, which has an even higher flash point (more than 63°C). [Ref. 95]

(i) Non-Oil Sources

The Boeing company has investigated a number of alternative airplane fuels, considering their production processes, costs, and impacts on the design and operation of airplanes and airports. A number of novel fuels have been suggested as alternatives to conventional jet fuel: ammonia, alcohol, heavy fuel oil, and even powdered coal. Unfortunately, emission restrictions, low energy content, or incompatibility with airplane requirements eliminate these from consideration. The three most promising candidates for the foreseeable future are liquid hydrogen, liquid methane, and synthetic jet fuels produced from oil shale and coal. [Ref. 95]

(ii) Fuel from Tar Sands

A middle distillate cut (kerosene), taken from synthetic crude oil produced by Great Canadian Oil Sands Co., was found to make JP-5 that was indistinguishable from a high quality petroleum derived JP-5. It successfully passed all specification requirements and hardware tests. [Ref. 95]

(iii) Fuel from Oil Shale

The only jet fuels derived from oil shale that have been available for extensive hardware testing were a part of a group of experimental products refined from 10,000 barrels of Paraho shale oil. The jet fuels were intented to approach, as closely as possible specification grade JP-4 and JP-5. The JP-4 met specification after a minor amount of post refining treatment; the JP-5 fell short in several characteristics. The fuel had high gum and contamination values, a high freezing point and poor thermal stability. The freezing point, while important, is not a factor in engine performance. The other characteristics, all related to chemical instability, caused problems during a full scale engine test in the form of blocked filters and deposits in engine fuel injectors. Recent investigations in the area of shale oil refining have indicated that it may be possible to minimize these problems without excessively intensive refining.

(iv) Fuel from Coal

A JP-5 type fuel from crude coal liquids has been tested on a gas turbine engine and a research combustor. Although this fuel received rather intensive hydrogen treatment in both the liquefaction and refining stages, it still failed to meet the specification requirements in several respects. The thermal stability was poor, heat of combustion marginal, density too high and smoke point too low. However, none of these factors were reported to have any noticeable effect on the performance of the engine! Although density is a factor in the operation of the fuel control, it did not deviate enough to cause a problem.

Smoke point and thermal stability can have long range effects on durability which were not apparent in this short test (4 hours).

Boeing studied three methods of manufacturing synthetic jet fuel. Two of the processes produce a synthetic crude oil, one by direct hydrogenation of coal, the other from shale. In these processes synthetic crude is refined into jet fuel, plus an array of other products, and requires little modification to a conventinal refinery. The third method is one in which coal is converted to Syngas from which the derived fuels are catalytically synthesized by process analogous to Fischer-Tropsch.

Of the fuels examined in the Boeing study, synthetic fuels made from coal and oil shale are the only practical alternatives in terms of availability, cost, and efficient use of energy resources. They can be integrated into existing production and distribution systems and are compatible with contemporary airplane engines. Synthetic fuels require the smallest outlay for capital equipment and the already high efficiency of their manufacturing processes would be further improved with advancements in technology - particularly by in-situ techniques. A synthetic fuel industry could be expanded in stages to gradually support a greater portion of energy demand.

Based on the Boeing assessment of alternative fuel production processes, costs, airplane performance, and airplane and airport design, synthetic jet fuels clearly show the greatest potential as alternatives to alleviate problems of supply shortages and the ever increasing cost of petroleum.

(v) Canadian Market

Demand projection for aviation gasoline and turbo fuel shows a substantial growth rate for both Canada and British Columbia. Canadian per day demand was estimated at 13,100 cubic meters in 1980. This figure is expected to reach 16,900 cubic meters per day in 1990 and 22,200 cubic meters per day in the year 2000. Demand for aviation gasoline and turbo fuel in British Columbia was estimated at 2,000 cubic meters per day in 1980. Over the next ten years demand is expected to increase by 50 percent over the 1980 level to reach 3,000 cubic meters per day by 1990. Between 1990 and 2000 demand is projected to reach 4,200 cubic meters per day, a 40 percent increase over 1990.

Table 4.24

Aviation Gasoline and Turbo Fuel Demand Projection

	10^3 m^3/d			Average Annual Growth Rate	
	1980	1990	2000	1980-1990	1990-2000
Canada	13.1	16.9	22.2	2-6	2-8
British Columbia	2.0	3.0	4.2	4.2	3.5

[Ref. 77, 91]

Indirect coal liquefaction, using Fischer-Tropsch, of Hat Creek coal based on 7950 m^3/d (50,000 b/d) could yield about 850 cubic meters per day of jet fuel or about 28 percent of B.C. projected demand in 1990.

The same capacity M-Gasoline plant would produce 7,650 cubic meters per day of motor gasoline which is 45 percent of the 1990 projected demand in Canada. The plant production in this case would have to be tied to export markets.

4.5.3 Diesel Fuel

Demand forecast of fuel requirements in Canada shows a strong growth in diesel fuel relative to gasoline. Diesel engines have some advantages over gasoline engines because of better fuel economy, particularly in cold weather. Particulate emissions from diesel engines, however, could become a health hazard. Indications are that it will be difficult to produce enough acceptable diesel fuel to meet the demand in Canada unless additional middle distillates are freed from present heating uses and diesel quality requirements are relaxed considerably. As more of Canada's crude oil becomes synthetic, from tar sands and coal, the middle distillates from these crudes do not readily yield a diesel that would meet current fuel standards.

Table 4.25

Diesel Fuel Projected Demand

	10^3 m^3/d			Percent Average Annual Growth Rate	
	1980	1990	2000	1980-1990	1990-2000
Canada	39.3	56.5	79.1	3.7	3.4
British Columbia	6.7	9.8	13.1	4.0	2.9

Canadian demand for diesel fuels in 1980 has been estimated at 39,300 cubic meters per day. This demand figure is estimated to reach 56,500 cubic meters per day in 1990 and 79,100 cubic meters per day in the year 2000, or an annual average growth rate of 3.7 percent during the 1980 - 1990 period, and 3.4 percent during the 1990 - 2000 period.

British Columbia's demand for diesel fuel in 1980 has been estimated at 6,700 cubic meters per day or about 17 percent of total Canadian demand.

Over the next ten years the annual average growth in demand for diesel fuel in British Columbia is projected at 4 percent with a total demand projection of 9,800 cubic meters per day in 1990. By the year 2000, demand for diesel fuel in British Columbia is projected to reach 13,100 cubic meters per day, an average annual growth of 2.9 percent over the 1990 projected demand.

Table 4.26

Canadian Imports of Diesel Fuel*

Year	10^3 m^3
1977	53.26
1978	25.65
1979	48.86

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*Statistics Canada

4.5.4 Fuel Oils

Demand for light, medium, and heavy fuel oils is expected to diminish over the next twenty years in both Canada and British Columbia. Canadian demand for light fuel oil in 1980 has been estimated at 44,000 cubic meters per day. By the year 2000 this figure is estimated to drop by 50 percent to about 22,300 cubic meters per day. Heavy fuel oil demand will drop from the 1980 estimated figure of 41,900 cubic meters per day to a projected figure of 28,300 cubic meters per day by the year 2000. A similar demand trend is expected for British Columbia.

Table 4.27

Fuel Oil Demand Projection

	10^3 m^3/d			Percent Average Annual Change	
Canada	1980	1990	2000	1980/1990	1990/2000
Light Fuel Oil	44.0	29.6	22.3	-4.1	-2.9
Heavy Fuel Oil	41.9	29.5	28.3	-3.6	-0.04
British Columbia					
Light Fuel Oil	2.8	2.3	1.9	-2.0	-2.0
Heavy Fuel Oil	3.4	2.8	2.6	-2.0	-1.0

[Ref. 77, 91]

Published trade statistics show that of all crude oil and refined petroleum products in Canada, fuel oil supply exceeded demand and the surplus was exported.

Table 4.28

Canadian Export of Fuel Oil*
thousand m^3

	1977	1978	1979
Medium Fuel Oil	124.9	322.5	374.8
Heavy Fuel Oil	5.2	909.6	623.6

Direct coal liquefaction processes designed to produce mainly heavy fuel oils would not, therefore, be a viable alternative for Hat Creek coal utilization.

4.6 Methanol

At recent growth rates world demand for methanol may reach 20 million tonnes by 1985. The installed capacity needed to produce this quantity will be about 24 million tonnes compared with 1980 capacity of about 16 million tonnes. Thus, on average, two 2,000 t/d plants must be added each year to keep up with the demand. However, much greater quantities could be required if new markets, such as single cell protein and gasoline extenders increase. New uses for methanol are likely to need 20 million tonnes per year by 1985 with a sharp increase in demand between 1985 and 2000. This will require on the average, another five 2,000 t/d units to be built each year to keep up with the demand.

The 1980 world supply of methanol was about 13 million tonnes. It is estimated that 1990 supply of methanol will reach 23.8 million tonnes. Demand for methanol is estimated to reach 25.5 million tonnes by 1990 and therefore, will exceed the projected supply by 1.7 million tonnes. The United States, Western Europe and Japan are

*Statistics Canada

expected to have capacity shortage. Canada is expected to have over 1.6 million tonnes of surplus methanol in 1990. The largest deficit in 1990 is expected in the United States, with demand outstripping supply by two million tonnes.

The two emerging end-uses that will be especially important in stimulating demand are:

1. Increasing production of gasoline octane, boosting methyl tertiary butyl ether (MTBE).
2. More methanol-based acetic acid plants.

According to industry sources, established end-uses for methanol, such as formaldehyde, are expected to absorb about 50 percent of the methanol in this category of established markets. Methanol demand could grow much faster than predicted if it is used as a direct gasoline ingredient or fuel for power plants late in the 1990's.

In a recent report published by the U.S. Engineering Societies Commission on Energy, solar biomass derived fuels are indicated to be at least 80 percent more costly than projected prices for coal-based methanol. Published reports estimate ethanol fuels cost at $27.50/GJ(in 1980 Cdn.dollars), methanol from coal at $15.56.

Methanol can be used as a fuel extender in automobiles, as a chemical feedstock, gasoline additive to improve the octane value, kerosene or diesel oil extender and a utility fuel to meet peak power needs.

At present, methanol, which is virtually all made from natural gas, costs about U.S. $180 per tonne in bulk (Vancouver, B.C.).

Table 4.29

1990 World Methanol Outlook*

thousand tonnes per year

	Demand		Supply		
	Total	Percent	Total	Percent	Surplus/(Deficit)
Canada	480	1.9	2,100	8.8	1,620
United States	8,000	31.4	6,000	25.2	(2,000)
Latin America	1,000	3.9	1,400	5.9	400
Western Europe	6,000	23.5	5,500	23.1	(500)
Japan	2,800	11.0	1,800	7.6	(1,000)
Eastern Europe	4,500	17.6	4,000	16.8	(500)
Australia/Zealand	120	0.5	400	1.7	280
Africa/Middle East	500	2.0	1,500	6.3	1,000
Other	2,100	8.2	1,100	4.6	(1,000)
Total	25,500	100.0	23,800	100.0	(1,700)

*Stone & Webster Estimates

Table 4.30

World Methanol Demand

thousand tonnes

Country/Area	1979	1985*	1990*	1979-1990 Average Growth Percent
Canada	240	320	480	6.5
United States	3,370	5,200	8,000	8.2
Latin America	290	600	1,000	11.9
Western Europe	3,250	4,500	6,000	5.8
Japan	1,100	2,000	2,800	8.9
Eastern Europe	2,320	3,630	4,500	6.2
Mid East/Africa	45	200	500	24.5
Other	530	1,250	2,220	16.6
Total	11,145	17,700	25,500	7.8

*Stone & Webster Estimates

Table 4.31

1980 World Methanol Capacity
thousand tonnes

Area/Country	Capacity	Percent of Total
North & Latin America		
United States	4,080	28.7
Canada	430	3.0
Latin America	346	2.5
Sub-total	4,856	
Western Europe		
West Germany	1,444	10.1
United Kingdom	692	4.9
Netherlands	660	4.7
Italy	450	3.2
France	425	3.0
Spain	250	1.8
Austria	140	1.0
Norway	50	0.4
Finland	35	0.3
Sub-total	4,146	
Eastern Europe		
Soviet Union	1,500	10.5
Rumania	300	2.1
East Germany	250	1.8
Poland	250	1.8
Yugoslavia	180	1.3
Czechoslovakia	100	0.7
Bulgaria	30	0.2
Sub-total	2,610	
Asia		
Japan	1,254	8.9
Korea	395	2.8
China	139	1.0
Taiwan	125	0.9
India	90	0.6
Philippines	28	0.2
Pakistan	3	-
Sub-total	2,034	
Middle East & Africa		
Libya	330	2.3
Algeria	100	0.7
Israel	50	0.4
South Africa	16	0.1
Egypt	10	0.1
Sub-total	506	
WORLD TOTAL	14,152	100.0

4.6.1 Canadian Market

Canada's methanol capacity was 430,000 tonnes in 1980. Future estimates range between as high as 2 million tonnes in 1985 and 3 million tonnes in 1990. Lower estimates put the 1985 capacity at 1.25 million tonnes and 1990 at 2.1 million tonnes. The higher forecast is based on an average growth rate of 21.5 percent annually between 1980 and 1990, and the lower forecast is based on 17.2 percent.

Table 4.32

Canadian Methanol Supply/Demand*

thousand tonnes per year

Higher Forecast

	1980	1985	1990	Average Growth Rate 1980-1990
Capacity	430	2,000	3,000	21.5 percent
Demand	212	320	480	8.4 percent
Surplus	218	1,680	2,520	

Lower Forecast

	1980	1985	1990	Average Growth Rate 1980-1990
Capacity	430	1,250	2,100	17.2 percent
Demand	212	320	480	8.4 percent
Surplus	218	930	1,620	

[Ref. 1, 15]

*Stone & Webster Estimates

Table 4.33

Forecast Canadian Methanol Capacity

thousand tonnes per year

	1975	1980	1985	1990
Firm				
Celanese Canada, Cornwall, Ont.	50	50	-	-
Alberta Gas Chemicals, Medicine Hat	-	380	760	760
Celanese Canada, Edmonton	-	-	700	700
Ocelot Industries, Kitimat, B.C.	-	-	420	420
Sub-total: firm	50	430	1,880	1,880
Potential				
Ashland et al, Alberta	-	-	700	700
Westcoast Transmission, B.C.	-	-	400	400
Sub-total: potential	-0-	-0-	1,100	1,100
Total: firm + potential	50	430	2,980	2,980

[Ref. 91]

In 1975 the Canadian methanol capacity consisted of Celanese 50,000 tonnes per year plant in Cornwall, Ontario. In 1980 Alberta Gas Chemical completed a 380,000 tonnes per year plant at Medicine Hat.

Celanese Canada, Cornwall, Ontario: This is an old naphtha-based plant. While Celanese have not made a definite statement to the effect, it is probable that the plant will be closed when supply becomes available from the company's new Alberta plant in late 1982 or early 1983.

Alberta Gas Chemicals, Medicine Hat, Alberta: The first half of this gas-based plant was commisssioned in the mid-1970's, and the second half is under construction at the present time with completion scheduled for late 1981. Gas feedstock/fuel is produced from the company's own reserves at a cost which must be well below the current utility price to large industrial customers in Alberta.

Celanese Canada, Edmonton: Celanese have completed hearings related to their industrial development permit application to the Alberta Energy Resources Conservation Board. Commissioning of the new plant is proposed for mid-1982. Gas supply will be delivered by Northwestern Utilities; however there may be procedures such as gas purchase contracts between Celanese and producers which will permit deliveries at below normal utility prices to large industrial customers.

Ocelot Industries, Kitimat: Ocelot are proceeding with construction of their plant at Kitimat. The target date for commissioning is 1982. Gas will be delivered from the Pacific Northern Gas system. Apparently a gas price agreement has been reached with the provincial government, but details are not public. Industry sources believe that Ocelot have been exempted from the full force of the province's recently announced gas pricing policy for industrial processing applications as their project was underway prior to announcement of the policy on May 26, 1980.

Ashland, IMC, Alberta: A proposal for a third gas-based methanol plant in Alberta will probably be announced within the next few months. Size will be similar to the Celanese plant, and gas purchases will involve direct contracts between producers and the plant operator. Market interest is believed to focus on the United States.

Westcoast Transmission, British Columbia: Westcoast's methanol plant would appear to be in competition with the projects of Ocelot and Celanese. Ocelot may have an advantage in gas pricing because of their agreement with the B.C. government. Celanese also have a gas price advantage, though this is offset by the cost of transportation to tidewater. While Westcoast continue to pursue their project, it appears the least certain among current proposals to proceed to production.

The Westcoast methanol project will be subject to British Columbia's new gas pricing policy. In essence this policy blends domestic and export gas prices after taking into account the proportion of the product price accounted for by feedstock gas and the proportion of the product exported. As an illustrative example:

- methanol	\$300/tonne FOB factory
- gas feedstock	37 GJ/tonne
- export proportion	90% of production
- domestic gas price	\$2.25/GJ
- export gas price	\$5.20/GJ

$$\text{FEEDSTOCK PRICE} = \$2.25 + 90\% \ \frac{(37 \times \$5.20)}{(\$300.00)} \ (5.20 - 2.25)$$

$$= \$3.95 \ (\text{ per GJ })$$

In the event that the primary gas-based product is upgraded to a first derivative (e.g. formaldehyde, acetic acid) and the derivative is exported, the domestic market price would be applicable.

Demand for methanol in Canada was 212,000 tonnes. Industry sources estimate that by 1985 demand will be 320,000 tonnes and by 1990 will reach 480,000 tonnes. This forecast is based on an average annual demand growth of about 8.4 percent between 1980 and 1990.

The bulk of the methanol is used for the production of formaldehyde. Over 60 percent of the 1980 demand went into formaldehyde production. Its use for dehydrating pipelines was second and accounted for 19 percent.

Table 4.34

1980 Canadian Methanol Demand

Application	10^3 m^3/a	Percent
Formaldehyde	128.5	60.0
Dehydrating Pipelines	40.0	18.9
Deicing Agent	21.0	9.9
Solvent	9.5	4.5
Methyl Amines	5.5	2.6
Other	7.5	3.5
Total	212.0	100.0

[Ref. 91]

For the future, in terms of traditional methanol end-uses, growth of Western markets will be restricted. Formaldehyde requirements could double if the growth in forest industry activity and further penetration of resin-intensive particle wafer boards is substantial. The weight gain associated with upgrading methanol to formaldehyde will prevent Western production from distant markets. Other traditional uses of methanol will continue to offer modest but expanding markets. Methanol requirements for existing uses could be met from existing Alberta Gas Chemicals capacity.

The production of Methyl tertiary butyl ether (MTBE) could create a demand, but it seems certain that it's production would be integrated into a methanol facility.

The production of motor gasoline from methanol could increase the demand substantially.

The move by some provinces, including British Columbia, to raise the natural gas feedstock price to methanol producers who do not upgrade in the province, will discourage further capacity expansion.

Due to the high cost of transporting gas to Toronto, and the increasing spread between crude oil and gas prices, gas-based methanol production in Alberta and British Columbia will continue to be attractive for sometime into the future and provides a competitive yardstick by which the economics of coal-based methanol production can be measured.

4.6.2 Western European Market

Western European demand for chemical grade methanol is estimated to reach six million tonnes by 1990. This conservative estimate is based on an annual growth rate of 5.8 percent over 1979 demand level of 3.25 million tonnes. Several developments could boost demand even higher. One is the direct addition of methanol to gasoline. Another is its use to produce ethylene and related lower olefins and downstream products. The strong demand for methanol in Western Europe in newly developed uses is a good illustration of the dynamic nature of its market. In 1980, more than 10 percent of the methanol used in Western Europe went into outlets that have developed only during the last five years.

Formaldehyde, made by catalytic oxidation of methanol, is expected to continue as the largest single use of chemical grade menthanol. Its average annual growth in the next ten years, however, will be among the lowest rates. Formaldehyde is used largely to make thermosetting resins.

The other conventional methanol outlets include dimethyl terephthalate, methyl methacrylate, methyl halides, and methyl amines. Together they represent about 16 percent of 1980's total demand. By 1990, they will account for about 840,000 tonnes or 14 percent of total chemical grade methanol demand. The projected growth rate over the next 10 years is about 4.2 percent annually.

The newer markets include methyl tertiary-butyl ether (MTBE) which is used as a replacement for lead-based gasoline antiknock agents and as a blending agent. Methanol is also used as a gasoline extender, and in the production of acetic acid and single cell protein.

Among the new uses of methanol currently being examined is the production of olefins. Methanol can be converted to ethylene and propylene in high yield using a zeolite catalyst. The yield, which can be as high as 70 percent, depends on the reaction conditions. At 60 percent yields, the economics of the methanol-to-ethylene process would allow methanol to displace some of the naphtha currently used as feedstock for ethylene production. Another potential use for methanol is in the production of ethylene glycol. Ethylene glycol could be made by reacting formaldehyde, carbon monoxide, and water to form glycolic acid, followed by esterification and reduction to glycol.

4.6.3 Methyl Tertiary-Butyl Ether (MTBE)

Another big potential application for methanol is the manufacture of methyl tertiary-butyl ether (MTBE), an octane improver for gasoline. In the United States, this hinges on government approval of MTBE for this use. Commerical production of MTBE started in Italy and West Germany in 1974.

World capacity of MTBE, which was estimated at 150,000 tonnes per year in 1978, had reached 800,000 tonnes per year by the end of 1980. Some industry sources expect this figure to more than double in the next five years. The demand for MTBE is very strong because of higher worldwide demands for aromatics. Though MTBE is not expected to come even close to matching aromatics in terms of supply, it does have the special attraction of freeing up toluene that might otherwise be added to gasoline pools in order to meet unleaded octane requirements. MTBE is blended in the gasoline pool in amounts of up to 7 percent by volume.

Table 4.35

Methanol Use - Western Europe

thousand tonnes per year

	1979		1985[*]		1990[*]		Percent Annual Growth
	Total	Percent	Total	Percent	Total	Percent	1979-1990
Formaldehyde	1586	48.8	2040	44.3	2315	38.5	3.5
Gasohol	200	6.1	300	6.5	600	10.0	10.5
Dimethyl terephthalate	160	4.9	200	4.3	250	4.2	4.2
Methyl amines	155	4.8	180	3.9	260	4.3	4.8
Methyl halides	110	3.4	150	3.3	180	3.0	4.6
Methyl methacrylate	110	3.4	130	2.8	150	2.5	2.9
Methyl tertiary-butyl ether (MTBE)	70	2.2	180	3.9	240	4.0	11.9
MTBE blending	30	0.9	75	1.6	100	1.7	11.7
Acetic acid	25	0.8	185	4.0	375	6.3	27.9
Single-cell protein (SCP)	2	negl.	160	3.5	250	4.2	55.1
Other[**]	805	24.7	1000	21.9	1290	21.3	4.3
Total	3253	100.0	4600	100.0	6000	100.0	5.8

[*] Stone & Webster Estimates

[**] Includes solvents and methyl esters of acetic acid, fatty acids, and glycols.

Table 4.36

New Methyl Tertiary-Butyl Ether (MTBE) Plants

Location	Producer	Capacity 1000 t/a	Status
United States	Arco	200	near completion
	Petro-Tex	280	near completion
	Phillips Chemical	100	near completion
	Schenectady Chemical	100	near completion
	Sub-total	680	
West Germany		140	completed
Italy		100	completed
	Sub-total	240	

4.7 Benzene, Toluene, Xylenes

4.7.1 Supply

Canada has an excess capacity of benzene, toluene, and xylenes. In 1980 benzene exports amounted to 120,000 tonnes or about 23 percent of supply. Toluene exports during the same year were 175,000 tonnes or 38 percent of supply, and xylenes exports were 290,000 tonnes or 79 percent of supply.

Table 4.37

1980 Canadian BTX Supply/Demand

	1000 tonnes		
	Benzene	Toluene	Xylene
Supply	530	460	365
Demand	410	285	75
Export	120	175	290

[Ref. 91]

Annual growth rates in Canadian BTX supply over the last five years ranged from 7.9 percent for benzene to 21 percent for xylene.

Table 4.38

Canadian BTX Supply Growth

	1000 tonnes		Average
	1976	1980	Annual Growth
Benzene	362	530	7.9
Toluene	280	460	10.5
Xylenes	142	365	21.0

[Ref. 91]

The high growth rates in BTX supply were mainly a result of even higher growth rates in exports.

Table 4.39

Canadian BTX Export Growth

	1000 Tonnes		Average
	1976	1980	Annual Growth
Benzene	101	120	3.5
Toluene	85	175	15.6
Xylenes	86	290	27.5

[Ref. 91]

With the exception of small toluene and xylenes volumes produced in Vancouver, all of the Canadian BTX current capacity is concentrated in the east.

Table 4.40

1980 Canadian BTX Capacity

	1000 tonnes		
	Benzene	Toluene	Xylenes
Sarnia	401	238	202
Montreal	125	190	155
Toronto	22	16	12
British Columbia	-	38	6
Other	-	-	3
Total	548	477	378

[Ref. 91]

Two major projects for benzene production are planned. The Petalta project with a 500,000 tonnes/annum (t/a) capacity and the Shell project with 236,000 t/a capacity.

Petalta Project:

This joint venture between Alberta Energy Company and ESSO Chemical Canada is based on pentane, and higher hydrocarbons, feedstock. It will have a design capacity of 500,000 t/a of benzene, plus a naphtha/raffinate stream which could be upgraded to olefines, but is currently expected to go to the ESSO refinery for fuel production. The plant is scheduled for operation in 1984.

Shell Canada - Strathcona Synthetic Crude Refinery:

Components of Shell's proposed synthetic crude oil refinery include a 1,750 cubic meter per day aromatics extraction unit and a 950 cubic meters per day dealkylation unit. Benzene capacity will be 236,000 t/a. ERCB approval has been granted and operation is scheduled for 1984.

4.7.2 Demand

Current demand for aromatics in western Canada is limited, with the only upgrading operation of consequence being the 32,000 t/a Dow Chemical phenol plant which purchases toluene from the ESSO refinery at Ioco.

The Shell/Nova plant will be a captive user of the full benzene output of the Shell refinery, while the Petalta plant will take over one-half (316,000 t/a) the benzene output from the associated pentanes - plus processing plant. Since announcement of the Petalta styrene plant there has been no statement as to the intended disposition of potential surplus benzene production.

Other than the styrene projects, no new uses for aromatics in western Canada could currently be considered as "probable". Virtually any of the common BTX derivatives might be produced to complement ethylene-related petrochemical activity, but none would appear to offer an export opportunity comparable to that of styrene. Perhaps the most likely potential upgrading project which would provide a significant demand is a benzene to phenol plant; however, construction of such a plant would displace output from Dow's toluene oxidation plant and eliminate the demand for toluene from the ESSO refinery.

4.7.3 Pricing

As aromatics are produced from petroleum, an alternative to their use as chemicals is in manufacture of liquid fuels. The shortage of indigenous crude oil in Canada, the probable move of the Canadian crude price towards world levels, and the fact that Canadian capital costs for petrochemical projects are higher than in competing countries, are factors which discourage expansion of aromatics production in western Canada. In general, production is seen as being limited to situations in which the advantageous gas/ethane price situation provides an opportunity for upgrading aromatics in conjunction with ethylene, with the low ethylene cost permitting the product to be competitive in world markets.

4.7.4 Impact of Production at Hat Creek

The direct liquefaction of Hat Creek coal using H-Coal or the EDS processes will produce substantial amounts of naphtha which could be upgraded to produce BTX or alternatively be utilized as motor benzole, i.e., for blending in gasoline.

Table 4.41

Hat Creek Potential BTX Production

	Annual Production 1000 tonnes		1980 Canadian Production	Percent of 1980 Canadian Production	
	EDS	H-Coal	1000 tonnes	EDS	H-Coal
Benzene	1063	938	530	200.6	177.0
Toluene	123	109	460	26.7	23.7
Xylenes	62	55	367	16.9	15.0

The EDS process would produce double the 1980 Canadian supply of benzene and about 27 percent of the toluene and 17 percent of the xylenes. The introduction of direct coal liquefaction e.g. EDS processes may therefore be anticipated to have a major influence on the international BTX market.

4.8 Ammonia

Ammonia can be produced from Hat Creek coal by the gasification/synthesis route. The yield relationship is such that 1 tonne of ammonia can be produced from approximately 3 tonnes of Performance Blend coal. A production of 1,000,000 tonnes per year of ammonia would, therefore, consume 3 million tonnes of coal or about one-sixth of the requirement for the other processes considered in this Report. However,this also represents 40 percent of total planned Canadian capacity additions scheduled for 1985 and hence, if implemented, would meet severe competition.

4.8.1 International Ammonia and Nitrogen Fertilizer Market

Availability of reliable data is the major limitation of this type of study. China and Russia, for instance, do not make data readily available though they are two of the largest producers and consumers. The degree of accuracy of data from many other countries is unknown.

Data from countries reporting on a calendar year have been combined with those reporting in a fiscal year. Consumption, as reported by the TVA, is estimated as the difference between reported production and export/import data. Thus, consumption estimates can be innaccurate if inventories change significantly from year to year.

Estimating production based on capacity is difficult because of international trade. The data provided by TVA's National Fertilizer Development Center (NFDC) assumes a given operating rate, which may differ from the method utilized by the Food and Agriculture Organization (FAO) of the UN. In addition, the rate may vary from country to country and from year to year for the same country.

Thus, the above limitations may cause a difference in the figures for any one region (e.g. Tables 4.42 and 4.45). Nevertheless, the two most widely recognized sources - FAO and the NFDC - are used in this study for comparison purposes.

The following sections are divided by world regions each discussing the nitrogen fertilizer market, including consumption/production and demand/supply capabilities. In addition, the ammonia market is detailed by region whenever possible.

4.8.2 World Nitrogen Fertilizer Market

During 1977 the world nitrogen production reached 46 million metric tons (Table 4.42, Table 4.43), with an estimated 57 percent increase expected by 1985. During 1978 and 1979, world production has averaged slightly more than 60 percent of the world's ammonia capacity. There are two main factors which influence demand for nitrogen (Table 4.43). First, modern crop systems need large amounts of nitrogen fertilizers to reach their full yield potential. Secondly, there is an increased demand for meat throughout the world requiring more feed grains. In addition, livestock expansion means more intensive use of pasture and forage crops, creating a greater need for nitrogen. This shift in diets should help to increase demand

Table 4.42 **World Nitrogen Fertilizer Production**

Region	1976	1977	1978^{E}	1979^{E}	1980^{E}	1981^{E}	1982^{E}	1983^{E}	1984^{E}	1985^{E}	Rate Of Change 1976-85 %
	(million tonnes)										
North America	10.4	10.8	11.4	11.2	11.2	11.2	11.3	11.3	11.3	11.3	9
Latin America	1.2	1.3	1.8	2.0	2.5	2.5	2.7	3.0	3.5	3.8	217
Western Europe	9.0	9.2	9.6	9.7	10.1	10.4	10.4	10.5	10.6	10.6	18
Eastern Europe	5.3	5.8	6.2	6.5	6.9	7.2	7.6	7.9	8.0	8.0	51
USSR	8.5	8.5	9.3	10.5	12.2	13.9	15.7	17.2	17.8	17.8	109
Africa	0.6	0.7	0.8	0.8	1.1	1.4	1.8	1.9	2.0	2.0	233
Asia	8.7	9.4	9.6	11.1	12.5	13.8	15.0	16.0	17.2	17.8	105
Oceania	0.2	0.2	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	50
World	43.9	45.9	49.0	52.1	56.8	60.7	64.8	68.1	70.7	71.6	63

Note: E = Estimates and Projections, based on 90 percent rate in developed countries and a 70 percent operating rate in developing countries.

Source: TVA, National Fertilizer Development Center

Table 4.43 Nitrogen Fertilizer Consumption, Trade and Production - 1976 to 1979

	North America	Latin America	Europe	USSR	Africa	Asia	Oceania	World
				(million tonnes)				
1976								
Consumption	10.3	2.2	12.5	7.3	1.3	12.5	0.2	46.3
Net Trade[1]	-0.1	-1.0	2.6	0.4	-0.6	-1.8	-0.03	-0.5
Total	10.2	1.2	15.1	7.7	0.7	10.7	0.17	45.8
Production	11.0	1.3	15.1	8.5	0.6	10.8	0.2	47.5
Difference[2]	0.8	0.1	-0-	0.8	-0.1	0.1	0.03	1.7
1977								
Consumption	9.7	2.5	13.0	7.5	1.3	15.5	0.3	49.8
Net Trade[1]	0.7	-1.2	2.6	0.6	-0.5	-2.1	-0.9	-0.8
Total	10.4	1.3	15.6	8.1	0.8	13.4	-0.6	49.0
Production	11.1	1.3	15.8	9.5	0.8	13.2	0.2	51.9
Difference[2]	0.7	-0-	0.2	1.4	-0-	-0.2	0.8	2.9
1978								
Consumption	10.5	2.5	14.0	7.7	1.4	17.4	0.3	53.8
Net Trade[1]	1.1	-1.1	2.7	0.7	-0.6	-2.6	-0.1	0.1
Total	11.6	1.4	16.7	8.4	0.8	14.8	0.2	53.9
Production	11.7	1.4	17.2	9.2	0.8	15.4	0.2	55.9
Difference[2]	0.1	-0-	0.5	0.8	-0-	0.6	-0-	2.0
1979								
Consumption	11.1	2.7	14.5	7.5	1.5	19.6	0.3	57.2
Net Trade[1]	0.9	-1.2	2.5	0.7	-0.6	-2.7	-0.1	-0.5
Total	12.0	1.5	17.0	8.2	0.9	16.9	0.2	56.7
Production	12.9	1.4	18.0	9.0	0.8	17.5	0.2	59.8
Difference[2]	0.9	-0.1	1.0	0.8	-0.1	0.6	-0-	3.1

Notes:

[1] Net Trade = Exports less Imports

[2] Difference = Production less (Consumption and Net Trade)

Source: Food & Agriculture Organization of the U.N.

in the years ahead. Increased demand for meat is partially offset by the increasing areas of land used for soybeans which fix atmospheric nitrogen, requiring little nitrogen fertilizer.

World demand for nitrogen from 1978 through 1984 is expected to grow at an average annual compound rate of 4.8 percent (Table 4.44).

On the supply side, nitrogen is expected to increase at a compound rate of 4.8 percent for the same period, reaching almost 72 million tonnes by 1984 (Table 4.45).

The world ammonia capacity is expected to increase by 60 percent through 1985, reaching 113 million tonnes (Table 4.46). Natural gas is the preferred feedstock, which is the basis for 68 percent of the world's supply of nitrogen. Thus, the location of cheap natural gas will be a major determinant in the location of future ammonia production (Table 4.47).

The most current sources indicate that by December 1978 the number of plants closed totalled 6 million tonnes of ammonia capacity.

Table 4.47 World Natural Gas Reserve

	Reserves	
	10^{12} m^3	10^{12} ft^3
North America	7.59	268.0
Latin America	3.07	108.5
Western Europe	3.87	136.8
Eastern Europe	0.32	11.4
USSR	26.05	920.0
Africa	5.88	207.5
Asia	23.49	829.4
Oceania	1.08	38.0
World	71.35	2,519.6

Source: TVA, Natural Fertilizer Development Center

Table 4.44 World Nitrogen Fertilizer Consumption

Region	1976	1977	1978^E	1979^E	1980^E	1981^E	1982	1983	1984	1985	Rate Of Change 1976-85 %
					(million tonnes)						
North America	10.0	10.3	9.7	10.8	11.2	11.6	11.9	12.3	12.7	13.1	31
Latin America	2.0	2.3	2.5	2.7	2.9	3.1	3.3	3.6	3.8	4.1	105
Western Europe	7.7	8.0	8.2	8.5	8.8	9.0	9.3	9.6	9.8	10.0	30
Eastern Europe	4.5	4.4	4.8	5.1	5.4	5.7	5.9	6.2	6.4	6.7	49
USSR	7.3	7.3	8.1	8.6	9.0	9.6	10.1	10.6	11.1	11.6	59
Africa	1.3	1.4	1.5	1.6	1.7	1.8	1.9	2.0	2.2	2.3	77
Asia	10.3	11.2	12.4	13.3	14.1	15.0	15.9	16.8	17.7	18.5	80
Oceania	0.2	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.4	100
World	43.3	45.2	47.5	50.9	53.4	56.1	58.6	61.4	64.0	66.7	54

NOTE: E = Estimates and Projections, based on a 90 percent rate in developed countries and a 70 percent operating rate in developing countries.

Source: TVA, National Fertilizer Development Center

Table 4.45 World Nitrogen Fertilizer Demand/Supply Capabilities

	1976	1977	1978	1979^{E}	1980^{E}	1981^{E}	1982	1983	1984
				(million tonnes)					
NORTH AMERICA									
Supply Capability	10.75	11.28	11.83	10.85	12.97	12.92	12.83	12.82	12.79
Consumption	10.26	9.69	10.49	10.53	11.09	11.66	12.22	12.70	13.18
Surplus (- Deficit)	0.49	1.59	1.34	0.32	1.88	1.26	0.61	0.12	-0.39
LATIN AMERICA									
Supply Capability	1.32	1.36	1.34	2.78	3.04	3.54	3.94	4.11	4.34
Consumption	2.27	2.58	2.49	2.72	2.96	3.16	3.37	3.58	3.81
Surplus (- Deficit)	-0.95	-1.22	-1.15	0.06	0.08	0.38	0.57	0.53	0.53
WESTERN EUROPE									
Supply Capability	9.59	10.09	11.14	9.54	9.26	9.52	9.47	9.46	9.52
Consumption	8.42	8.75	9.42	9.62	9.84	10.09	10.35	10.61	10.87
Surplus (- Deficit)	1.17	1.34	1.72	-0.08	-0.58	-0.57	-0.88	-1.15	-1.35
EASTERN EUROPE/USSR									
Supply Capability	13.95	14.71	15.28	17.03	19.38	21.59	21.94	22.18	22.20
Consumption	11.28	11.69	12.24	12.72	13.44	14.18	14.91	15.65	16.39
Surplus (- Deficit)	2.67	3.02	3.04	4.31	5.94	7.41	7.03	6.53	5.81
AFRICA									
Supply Capability	0.16	0.15	0.17	0.12	0.23	0.35	0.42	0.46	0.55
Consumption	0.52	0.47	0.49	0.55	0.60	0.65	0.69	0.74	0.79
Surplus (- Deficit)	-0.36	-0.32	-0.32	-0.43	-0.37	-0.30	-0.27	-0.28	-0.24
ASIA									
Supply Capability	8.37	9.90	11.84	14.47	16.40	17.35	18.52	19.88	20.87
Consumption	10.39	13.27	14.92	16.26	17.19	18.17	19.51	20.21	21.24
Surplus (- Deficit)	-2.02	-3.37	-3.08	-1.79	-0.79	-0.82	-0.99	-0.33	-0.37
OCEANIA									
Supply Capability	0.22	0.22	0.20	0.17	0.17	0.19	0.20	0.20	0.20
Consumption	0.23	0.23	0.24	0.27	0.27	0.29	0.30	0.32	0.33
Surplus (- Deficit)	-0.01	-0.01	-0.04	-0.10	-0.10	-0.10	-0.10	-0.12	-0.13
OTHER									
Supply Capability	1.53	1.91	1.98	1.85	1.65	1.60	1.55	1.49	1.42
Consumption	1.08	1.07	1.14	1.16	1.18	1.22	1.26	1.29	1.33
Surplus (- Deficit)	0.45	0.84	0.84	0.69	0.47	0.38	0.29	0.20	0.09
WORLD TOTAL									
Supply Capability	45.89	49.62	53.78	56.81	63.10	67.06	68.87	70.60	71.89
Consumption	44.45	47.75	51.43	53.83	56.57	59.42	62.61	65.10	67.94
Surplus (- Deficit)	1.44	1.87	2.35	2.98	6.53	7.64	6.62	5.50	3.95

Source: U.S. Department of Agriculture.

Table 4.46 World Ammonia Capacity

Region	1976	1977	1987^{E}	1979^{E}	1980^{E}	1981^{E}	1982^{E}	1983^{E}	1984^{E}	1985^{E}	Rate Of Change 1976-85 %
	(million tonnes)										
North America	15.3	18.2	17.4	17.3	17.4	17.4	17.4	17.4	17.4	17.4	14
Latin America	2.8	2.8	4.3	4.3	4.4	4.4	5.6	6.6	6.8	6.8	143
Western Europe	14.8	14.8	14.9	15.8	16.1	16.1	16.1	16.4	16.4	16.4	11
Eastern Europe	9.2	9.4	9.9	10.8	10.8	11.6	12.4	12.4	12.4	12.4	35
USSR	11.4	12.6	14.2	16.5	19.4	20.9	24.7	24.7	24.7	24.7	117
Africa	1.2	1.2	1.5	1.6	2.9	2.9	3.5	3.5	3.5	3.5	192
Asia	15.3	17.2	19.3	22.3	24.5	25.6	28.5	30.7	31.3	31.3	105
Oceania	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	-
World	70.4	76.6	81.9	89.0	95.9	99.3	108.6	112.1	112.9	112.9	60

NOTE: E = Estimates and Projections, based on a 90 percent rate in developed countries and a 70 percent operating rate in developing countries.

Source: TVA, National Fertilizer Development Center

4.8.3 Asia consumes over 19 million tonnes of nitrogen, which is about 2 million more than it produces. Consumption for selected countries is further broken down below:

Table 4.48 Nitrogen Consumption - Pacific Rim 1976 to 1977 and 1978 to 1979

	million tonnes	
Canada	0.6	0.8
United States	9.7	10.3
Japan	0.7	0.8
Philippines	0.2	0.2
Indonesia	0.3	0.6
World	46.3	57.2

Source: Food and Agriculture Organization of UN.

China also consumes a considerable amount of nitrogen, totalling about 7 million tonnes in 1978. The Japanese industry is reducing its fertilizer capacity, since its export market is declining while the cost of importing feedstock is increasing. Japan has closed 1.3 million tonnes of ammonia capacity since 1978 and may close 20 percent of the remaining plants. Despite Japan's decreasing ammonia capacity, Asia is expected to have one of the largest increases in ammonia capacity in the world. However, a net increase of 12 million tonnes in Asian ammonia capacity is expected between 1978 and 1985. Total nitrogen production in Asia could increase 90 percent between 1977 and 1985, while consumption will increase by 66 percent. Asia is expected to be a net importer of nitrogen through 1985, the largest consumers being China and India.

Large increases in ammonia capacity are planned in India, China, Turkey, Indonesia and Pakistan. Planned capacity in the Middle East market is uncertain. Though the demand for fertilizers is low in this area, it has 29 percent of the world's natural gas reserves at a low cost, which encourages construction of processing facilities.

4.8.4 Europe

During 1977 Western Europe produced about 9.2 million tonnes of nitrogen and consumed only 8.0 million tonnes (Table 4.42 and 4.44). Production is expected to increase to 10.6 million tonnes and consumption to 10.0 million tonnes by 1985. Capacity expansions in Europe are expected to slow down.

Ammonia capacity is expected to increase by only 11 percent between 1977 and 1985 (Table 4.46). Some plants have recently closed because of poor profits. Consumption is expected to increase by 1985, the nitrogen surplus is expected to continue. France, West Germany and the United Kingdom are the largest consumers in Western Europe.

4.8.5 Developing Countries*

Consumption of nitrogen fertilizer is expected to grow fastest in developing economies since populations are increasing most rapidly and government policies are aimed at augmenting food production. In 1977 these countries consumed about 14 million tonnes of nitrogen, totalling about 31 percent of the world's consumption (Table 4.42). The region's production of nitrogen fertilizer reached

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NOTE: Developing countries generally include Africa, Latin America, Asia (except Japan and Israel) and other developing economies.

10 million tonnes in 1977. Ammonia capacity is expected to grow at a rate of 27 percent through 1985 (Table 4.46). The major importers of nitrogen will continue to be the developing countries through 1985.

4.8.6 North America

Historically North America (Canada and the United States) produces more nitrogen than is consumed. In 1977 production was 10.8 million tonnes, while consumption was 10.3. The surplus in this region is expected to continue through 1981 (Tables 4.42 and 4.43). During 1977, however, more nitrogen was imported than was exported since ammonia producers found imports cheaper.

Ammonia capacity is expected to be relatively constant at 17.4 million tonnes through 1985 (Table 4.46). However, it is not known whether plants which had been closed due to poor market conditions will be brought back when the market improves. This is significant since idle capacity in the US is estimated to be over 4 million tonnes of ammonia.

4.8.7 Canada

Ammonia plants in Canada have been operating at about 90 percent of design capacity, or 1.9 million tonnes a year (Table 4.49). The latest sources report almost equal levels of consumption and production for 1978 and 1979 at 2.35 and 2.45 million tonnes, respectively. This is a one percent increase in production between 1978 and 1979.

The proposed new ammonia plants in Canada are expected to reach a capacity of 1.46 million tonnes by 1985 (Table 4.49), and total capacity should reach about 2.5 million tonnes. However, the three plants planned by Union Oil are still uncertain. Most of this production is

Table 4.49 Ammonia Plants in Canada

Company	Location	Status	Capacity Tonnes/Year (000)
Operating Units[a]			
Canadian Fertilizer, Ltd.	Medicine Hat, AB	1976-1980	726
Cominco, Ltd.	Calgary, AB	1976-1980	113
	Trail, BC	1976-1980	63
	Carseland, AB	1977-1980	363
Cyanamid of Canada	Welland, ON	1976-1980	227
Genstar Chemical	Maitland, ON	1976-1980	80
Sherritt-Gordon Mines	Ft. Saskatchewan, AB	1976-1980	145
J. R. Simplot Co.	Brandon, MB	1976-1980	100
Western Coop. Fertilizer	Calgary, AB	1976-1980	63
	Medicine Hat, AB	1976-1980	60
Total Operating Units		1976-1980	1,940
Proposed Units[b]			
Canadian Industries, Ltd.	Eastern Canada (Courtwright)	1985	1.1
Ethyl Chemicals (Esso)	Western Canada	1983-1984	544
Sherritt-Gordon Mine	Western Canada	1983-1984	374
Union Oil (3 plants)	- - - - - - - - - - Planned- - - - - - - -		544
Total Proposed Units		1983-1985	1,463.1
TOTAL	(Approximate) -	1976-1985	3.403.1

Sources:

(a) National Fertilizer Development Center, TVA.

(b) The Canadian Fertilizer Institute.

expected to take place in Western Canada where gas is available at low cost. About 50 percent of Canada's production in now exported to the U.S. This is expected to continue since most of the new production will be for exports.

Table 4.50 Summary - World Nitrogen Fertilizer Consumption 1970 to 1985

	Annual Growth(a) %		1980(a)	Annual Growth(a) %	1985E(a)	Consumption Per Cap. % Increase
	1970-75	1976-80	million tonnes	1980-85E	million tonnes	1968-1978
North America	3.4	2.3	11.2	3.4	13.0	47
Latin America	10.3	9.3	2.9	8.1	4.1	73
Western Europe	3.8	2.8	8.8	2.8	10.0	48
Eastern Europe	5.5	3.8	5.4	4.8	6.7	86
USSR	12.2	4.6	9.1	5.5	11.6	
Africa	8.4	7.2	1.7	6.6	2.3	100
Asia	7.3	7.6	14.1	6.2	18.5	148
Oceania	2.0	9.6	0.3	4.5	0.3	11
World	-	-	53.5	5.0	66.6	-

Sources: (a) TVA, National Fertilizer Development Center.

(b) U.N., Food & Agriculture Organization.

NOTE: E = Estimates and Projections, based on a 90 percent rate in developed countries and a 70 percent rate in developing countries.

5 ECONOMIC ANALYSES

The method and requirements of the economic analysis were outlined in the Terms of Reference (Appendix A) and the Financial/Economic Criteria (Appendix B).

5.1 Conversion Processes - Basis of Selection

The processes selected for economic evaluation in this Study are listed in Table 3.1. This list is the result Of a systematic elimination of a large number of potential processes for converting Hat Creek coal to upgraded solid, liquid or gaseous products. The reasons for elimination have been discussed separately in Sections 2, 3 and 4 of this report but are summarized as follows:

A. Coal-related factors

Any conversion process considered must be compatible with or adaptable to the particular properties of Hat Creek coal. As was discussed in detail in Section 2 these are:

(i) a high inerts content, moisture and ash accounting for more than half the weight of coal in the as received conditions.

(ii) low rank, being borderline sub-bituminous/lignite. Additional tests demonstrated that the coal hydrogenated readily but gave low yields of liquid hydrocarbons on pyrolysis, hence eliminating pyrolysis from further consideration.

(iii) unsuitability for beneficiation by currently available commercial methods for removing mineral matter. This virtually rules out all processes for conversion to upgraded solid products except where the ash is removed at some later process stage, e.g. solvent refined coal.

(iv) a strong tendency for the dried coal to reabsorb water without measurable volume change. This is a particular disadvantage in cases where a slurried coal feedstock is required e.g. Coal/oil mixtures (COM), coal/water mixtures (CARBOGEL), Texaco gasification.

B. Process-related factors

The time frame of the study requires that any processes considered must be already commercially proven or at such an advanced stage of demonstration that engineering design of a commercial plant could reasonably be undertaken at the present time.

(i) Solid Products

No processes for direct conversion to upgraded solid products are considered available. Conversion to liquids or gaseous products were considered as follows:

(ii) Coal liquids

(a) Solvent Extraction

The liquid solvent extraction process (LSE) and the supercritical gas solvent extraction process (SGE) under development by the NCB (UK) are not commercially demonstrated

The hydrogenated liquid (donor) solvent extraction process under development by EXXON (EDS Process) is presently operating at the 250 tons of coal per day scale and was considered to be sufficiently advanced to be considered in the Study.

Other processes in this category were not considered.

(b) Uncatalyzed Direct Hydrogenation

The solvent refined coal processes (SRC-1 and SRC-2) under development by the Gulf Oil Company are considered to be sufficiently demonstrated and are included.

No other variations of the Pott-Broche process are near commercial demonstration.

(c) Catalyzed Hydrogenation

The H-Coal Process (Hydrocarbon Research Inc.) is considered to be sufficiently demonstrated for inclusion, following operation of the 200/600 ton per day plant at Catlettsburg, Kentucky in 1980/1981. The development of other catalyzed processes e.g. the Consol Process (CRESAP) are not being pursued to commercial demonstration at the present time and are not considered.

(d) Indirect Liquefaction

All these processes commence with a synthesis gas produced by total gasification of a coal followed by shift reaction to adjust the carbon monoxide/hydrogen ratios according to the product slate required.

The following synthetic processes are considered to be fully commercial:

a) Methanol Sythesis

This is the most widely demonstrated process in commercial use. Several processes are available.

b) Fischer-Tropsch (Arge Process and Synthol Process)

Have been full demonstrated on large commercial scale at SASOL.

c) Mobil Methanol-to-Gasoline Process (Mobil - MTG)

Has been demonstrated at continuous pilot plant stage and the New Zealand Government is funding a detailed design of 2070 m^3/d (13,000 b/d) plant. At least four other major projects are under development and the process is considered to be commercially available in this Study.

No other processes for synthesis of liquids have been considered.

(iii) Gaseous Products

The process options available are conversion to fuel gases (low BTU, medium BTU); syhthetic natural gas (SNG) by synthesis from medium BTU gas; or ammonia by synthesis from medium BTU gas.

Low BTU fuel gases have not been considered for resons discussed under marketing factors.

A number of processes capable of producing medium BTU gases by total coal gasification have been commercially demonstrated, the best known of which are the Lurgi (pressurized fixed bed), Winkler (atmospheric fluidized bed) Koppers (atmospheric entrained flow) and Texaco (pressurized entrained flow) processes. Other, so-called, second generation processes are under development none of which, however, having been commercially proven to the same degree. This Study therefore considers only these four primary processes. But since there is no discernible market for medium BTU fuel gases (see Paragraph C, below) these processes are utilized principally for the preparation of a synthesis gas for subsequent conversion to SNG, ammonia or synthetic coal liquids. Economic evaluation of SNG and synthetic coal liquids has been carried out. Ammonia has not been evaluated for reasons stated in Section 4.8.

C. Market-related factors

The primary questions addressed in considering market opportunities for Hat Creek coal conversion products have been to determine whether a demand presently exists or will develop in the foreseeable future, the size and location of that demand and the potential revenues. These considerations have been influenced by the extremely large reserves of natural gas that have been discovered in the Province, the small and diminishing market for heavy fuel oils, the excess capacity situation in the world nitrogenous fertilizer market and the continuing concern over the future supplies of transportation fuels. Other factors have been the absence of a developed industrial market in the immediate vicinity of Hat Creek and the necessity for transportation the products to the Lower Mainland. In general, it has appeared clear that the primary thrust for conversion processes, as an alternative to electric power generation, should be directed to production of liquid fuels.

Table 5.0 presents a summarized breakdown of the procedures employed for process elimination and selection.

Table 3.1 lists 26 processes or process combinations, of which four are processes for direct liquefaction by hydrogenation. The remaining 22 processes cover the routes

Table 5.0 Process Selection - Utilization of Hat Creek Coal

Type of Process	Related Factors	Decision
I Solid Application		
a) Direct Combustion	Studied in separate evaluation	Study has objective of evaluation options which are competitive with combustion of Hat Creek coal.
b) Beneficiation	Coal unsuitable for beneficiation	Beneficiation studies not included.
c) Active Carbon	Application of coal to water treatment taking advantage of ion exchange capability.	Study excluded because of lack of local market for application.
d) SRC	Solid solvent refined coal (SRC) is produced by the hydrogenation of coal at low severity.	Rejected in favor of more severe hydrogenation processes which make liquid products.
II Gaseous Fuel Production		
a) Low BTU Gas	Lack of local industrial market for products rules against production from Hat Creek coal.	Rejected LBG as method of utilization.
b) High BTU Gas		Rejected HBG as method of utilization
c) Substitute Natural Gas	Appears to be an unattractive option in view of abundant supplies in B.C. at present.	Evaluate comparative economics of SNG vs. coal liquefaction. Base study on Lurgi gasification process.

Table 5.0 (continued)

Type of Process	Related Factors	Decision
II Gaseous Fuel Production (continued)		
Gasification Processes		
i) Lurgi Process	Hat Creek coal is reported, on the basis of laboratory tests, to be a satisfactory feedstock for Lurgi Fixed Bed gasifiers but this is contingent on the removal of fines smaller than 13 mm from the feed. These constitute a substantial proportion of total coal to plant and economic factors are critically dependent on satisfactory utilization of the fines.	Evaluate indirect coal liquefaction based on: a) combustion of excess fine coal and export of any excess power b) sale of excess fine coal. c) gasification of excess fine coal by the Koppers and Texaco processes. Evaluate SNG production based on Lurgi gasification process.
ii) Koppers Process	Koppers entrained bed process would be satisfactory for the gasification of Hat Creek coal. However, low pressure gasification is expected to be uneconomical as basis for indirect coal liquefaction.	The Koppers Process could be used as a feasible alternative to the Lurgi process. Perform economic evaluation based on best available data for indirect coal liquefaction.
iii) Texaco Process	The Texaco, entrained bed, high pressure process based on water slurry feed is in the developmental phase. Application to Hat Creek coal is a possibility but this could only be proved based on a pilot test comparing it with other gasifiers.	Estimate economics of indirect coal liquefaction using the best available data. Perform evaluation in order to recognize if this second generation technology offers potential future advantages to B.C. Hydro.
iv) Winkler Process	Hat Creek coal is potentially gasifiable by a fluidized bed process and closely related data based on lignite gasification is available.	Develop order of magnitude comparative economics based on Winkler gasification applied to indirect coal liquefaction.

Table 5.0 (continued)

Type of Process	Related Factors	Decision
II Gaseous Fuel Production (continued)		
Gasification Processes		
v) Other Processes		
British Gas/Lurgi Slagging Gasifier Cogas Shell Koppers		Excluded from consideration because these processes are not available for application to Hat Creek coal in the near future.
III Liquid Fuel Production		
i) Direct coal Liquefaction	Hat Creek coal has been shown to liquefy readily and with high carbon conversion efficiencies in laboratory tests of direct hydrogenation. It is a potential feedstock for processes derived from Potte Broche (uncatalyzed) and Bergius (catalyzed) processes. However this is contingent on satisfactory removal of the mineral residues (ash).	Include direct hydrogenation by SRC-I, SRC-II, H-Coal and EXXON Donor Solvent processes. Exclude long-term prospects such as the DOW coal liquefaction process and the EPRI short residence time approach. Though it could be applicable, the Consol (Conoco) process was excluded from consideration because of the lack of current activity in its development.
NCB Processes Super Critical Gas Extraction and Liquid Solvent Extraction	Preliminary investigations have been carried out of applicability to Hat Creek coal. Only the supercritical extraction results appeared to be unsatisfactory as a basis for commercial application.	Economic study not carried out because processes are not available for commercial application in the near future.

Table 5.0 (continued)

Type of Process	Related Factors	Decision
ii) Indirect Coal Liquefaction		
Processes for production of liquid fuels based on synthesis gas feedstocks.		
a) Fischer-Tropsch	Process is well established in commercial service based on low grade coal (South Africa). Process lacks capability to satisfy specific synthetic fuel needs. However, it could assume national importance as a source of synthetic fuels.	Evaluate order of magnitude economics of Fischer-Tropsch technology applied in association with above selected gasifiers viz. Lurgi, Koppers, Texaco and Winkler.
b) Methanol	Methanol production processes are well established. For all practical purposes methanol production based on coal equals Fisher-Tropsch technology in commercial feasibility.	Evaluate order of magnitude economics of methanol technology with above selected gasifiers (as for Fisher-Tropsch).
c) Methanol Conversion to Gasoline	In view of the lack of a general methanol based economy, the conversion of methanol to gasoline may be the best near term policy for utilizing coal gasification. Process has excellent specifications for producing motor fuels.	Evaluate order of magnitude impact of producing gasoline rather than methanol. Use best available published data on the Mobil process.
IV Pyrolysis	Poor yield of liquid products obtained in Fisher Assay test.	Rejected pyrolysis as method of utilization of Hat Creek coal.

to indirect liquefaction products - Fisher Tropsch Synthesis; Methanol Synthesis; Methanol to M-Gasoline Synthesis - based on seven processes or combinations of processes for producing the necessary synthesis gas. Four of the seven processes are proprietary, single processes - Koppers, Lurgi, Texaco and Winkler gasification processes. Of these, the Lurgi and Koppers processes have found the greatest commercial application.

The Lurgi process, however, presents an additional problem because the gasifiers are unable to accept coal fines (See Paragraph 2.1.7) which, following a special crushing and screening operation are expected to comprise 36 percent of the plant feedstock at Hat Creek. In the absence of a thermal electric power generating plant at Hat Creek no other external demand for the coal fines is expected to develop adjacent to the Hat Creek site. The high inerts content of the fines, moisture and ash together totalling more than 54 percent, militates strongly against rail transportation to the coast.

Alternative cases were therefore considered in which the fines produced could be totally consumed with the battery limits of the conversion facility. In one set of cases the fines are used for co-generation, the boiler and turbogenerator costs being included in the overall capital costs. In a few cases this results in a small excess of electricity available for export with corresponding adjustment of the plant revenues. In another set of cases the fines rejected by the Lurgi gasifier are first used to raise the required process steam, and to provide a minimum emergency power generation supply (150 MW_e). The excess of fines is then passed to other types of gasifiers capable of handling pulverized coal feeds viz. Texaco or Koppers. The gases produced by two sets of gasifiers are combined, after preliminary purification, to provide a single syngas feed to the Synthesis units. (See Paragraph 2.4.3 and Appendix D, Table D1.1, Cases B6, B7, C6, C7, D6, D7.)

5.2 Financial Criteria and Assumptions

(i) Exchange Ratio:

U.S. $1.00 = Cdn. $1.15

(ii) Capital Costs:

Include costs of permitting, design, engineering procurement and construction of the battery limits coal conversion plant. It is assumed that coal, electric power, water and other support facilities are available at battery limits. No allowance has been made for infrastructure development needs at Hat Creek.

(iii) Contingency:

No allowance for contingency has been made in this work. So far as possible economics have all been developed on a contingency-free basis.

The influence of various levels of contingency can then be seen by considering various levels of capital investment in the sensitivity studies.

(iv) Import Duties and Taxes:

Not included. The sources and countries of origin of major machinery and equipment items for a major conversion plant at Hat Creek cannot be predicted at present.

(v) Project Schedule

The period from date of filing of project applications to initial plant start-up is taken to be eight years. It is assumed that certain tasks such as permitting, design, engineering and preliminary construction can be overlapped to minimize the time required.

The assumed schedule of investment is as shown in Table 5.1.

Table 5.1 Project Investment Schedule

Year	Investment Percent	Total to Date Percent
1	2	2
2	5	7
3	15	22
4	20	42
5	22	64
6	20	84
7	10	94
8	6	100

(vi) Interest Rate:

The rates specified (Appendix B) are 3, 6 and 10 percent per annum. Similar rates are used to discount the net cash flows of the project in calculating the Net Present Value (NPV) of each investment.

(vii) Plant Life:

The economic life for each process considered is taken as 30 years from start up operations. No residual (book value), after 30 years is realized.

(viii) Plant Operations:

330 days per year; 3 shifts per day; 8 hours per shift.

(ix) Operating Costs

(a) <u>Coal</u>

B.C. Hydro provided estimated costs of $8, $10 and $12 per tonne as received. A base case assuming $10 per tonne has been used.

(b) <u>Electricity</u>

B.C. Hydro provided a formula for electric power costs based on demand charges and load factors. However this provides a cost close to ¢2.5 per KWh, with little significant effect of variations on overall costs, and this fixed cost has been used.

(c) <u>Water</u>

A delivered cost of $0.38/m^3 for both process and domestic water has been assumed.

(d) Catalysts and Chemicals

Costs calculated according to individual process requirements for start-up and normal operations.

(e) Labor

It is assumed that the plant is fully manned from commissioning.

- 4 shift crews covering 3-shift operation.
- Crew make-up is 40 percent operational, 60 percent maintenance.*
- Average hourly rate is $15.00
- Payroll burden is 30 percent of average hourly rate.
- Hourly wages = $19.50/hour ($39,000/year).
- Overhead costs are 100 percent of wages.
- Total labor costs = $78,000 per plant worker.

(x) Replacement/Maintenance

Assumed to be 3 percent of capital investment annually over 30 year economic life of plant. (Note effect of capital cost escalation).

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* An allowance for contracted maintenance has also been made in estimates of Replacement/Maintenance costs. This reduces the number of directly engaged maintenance workers required.

(xi) Start-up Costs

Assummed to be 25 percent of one month's productive supplies (coal, power, water) plus initial catalyst costs plus one month capital equivalent (Total capital ÷ 360).

(xii) Working Capital

Assumed to be cost of coal for one month plus value of products for one month plus one percent of capital.

Change in working capital - 50 percent in the first year of operation and 100 percent in second and succeeding years.

(xiii) Depreciation

Not included

(xiv) Taxes

Federal and provincial taxes are not included.

(xv) Inflation

Is not included.

(xvi) Revenues

(a) It is assumed that plant products are fully sold over 30 year life of plant.

(b) Product prices reflect those existing as at end of 1980.

(c) Because of the existing differential in oil prices between Canada and world-posted prices, an adjustment was made to reflect future revenues from product sales in international markets. At the end of 1980, this price differential amounted to $20.25 per barrel.

The product prices and annual sales revenues for the four principal types of coal conversion processes included in the economic analyses are summarized in Tables 5.2 - 5.5.

(d) The SNG case has been calculated assuming an export price of 0.159 $\$/m^3$ ($4.50 per 1,000 cu. ft.)

All costs expressed in 4th Quarter 1980 are Cdn.$.

5.2.2 Financial Results

The financial results of applying these assumptions to the twenty-five cases are summarized in Tables 5.6 - 5.9A. For each case condisered there is a surplus of revenues over operating costs (gross margin), these being generally higher in the following classes of cases.

Process Type	Process Description
Direct Liquefaction	H-Coal
Indirect Liquefaction:	
Fischer-Tropsch	Lurgi (Make Power)
Methanol	Lurgi (Make Power)
M-Gasoline	Lurgi (Make Power)

Such screening makes no allowance for assessment of project risk, which requires that the margin on revenues be assessed in relation to capital employed. However,this

Table 5.2

PRODUCT PRICES AND ANNUAL SALES REVENUES

DIRECT LIQUEFACTION PROCESSES

Products	Unit Price	SRC-I		SRC-II		H-COAL		EDS	
		Daily Production	Annual Sales $ million	Daily Production	Annual Sales $ million	Daily Production	Annual Sales $ million	Daily Production	Annual Sales $ million
Naphtha	$251.6/$m^3$	-	-	1,560 m^3	130	3,105 m^3	316	4,309 m^3	358
Turbine fuel	$220.1/$m^3$	-	-	-	-	2,934 m^3	213	2,665 m^3	194
Boiler fuel	$188.7/$m^3$	1,680 m^3	105	6,480 m^3	404	1,631 m^3	102	1,471 m^3	92
Ammonia	$200.0/t	125 t	8	160 t	11	160 t	11	190 t	13
Sulfur	$105.0/t	100 t	3	175 t	6	175 t	6	210 t	7
Phenols	$940.0/t	17 t	5	30 t	9	30 t	9	36 t	11
SRC solids	$ 24.0/m^3	40,000 m^3	317	-	-	-	-	-	-
Sub-total			438		560		657		675
Adjustment*			334		334		334		334
Total Revenues			772		894		991		909

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*Adjustment for differential between Canadian and international market prices.

Table 5.3 Product Prices and Annual Sales Revenues

FISCHER-TROPSCH SYNTHESIS

(a) Principal Products (all cases)

Products	m^3/d	t/d	Unit Price $	Annual Volume $m^3 x 10^6$	Annual $ million
Gasoline	4886	-	251.73	1.613	406
Jet Fuel	840	-	220.26	0.277	61
Diesel	1574	-	201.38	0.519	105
Fuel Oil	120	-	157.33	0.040	6
Mixed Alcohols	563	-	207.68	0.186	39
C_1-C_4	-	108	144	-	5
Ethylene	-	670	300	-	66
Ammonia	-	see (b) below	200	-	see (b) below
Sulfur	-	see (b) below	105	-	see (b) below
					688

	Texaco	Koppers	Winkler	Lurgi Sell Fines	Lurgi Make Power	Lurgi & Texaco	Lurgi & Koppers
(b) Byproducts							
Production:							
Ammonia (t/d)	-	-	-	300	300	210	210
Sulfur (t/d)	225	230	230	150	150	175	175
Coal Fines (t/d)	-	-	-	10,700	-	-	-
Electricity (MW)	100	-	-	-	73	-	-
Revenues:							
Ammonia	-	-	-	20	20	14	14
Sulfur	8	8	8	4	4	5	5
Coal Fines	-	-	-	31	-	-	-
Electricity	20	-	-	-	14	-	-
Total	28	8	8	55	38	19	19
(c) Total Revenues							
Products	688	688	688	688	688	688	688
Byproducts	28	8	8	55	38	19	19
Adjustment*	334	334	334	334	334	334	334
Total	1,050	1,030	1,030	1,077	1,060	1,041	1,041

*Adjustment for differential between Canadian and international market prices.

Table 5.4 Product Prices and Annual Sales Revenue

METHANOL SYNTHESIS

(a) Principal Products (all cases)

			Annual	
	t/d	$/t	million tonnes	$ million
Methanol	13,979	207	4.6	954
Ammonia	(a)	200	-	-
Sulfur	(b)	105	-	-

	Texaco	Koppers	Winkler	Lurgi Sell Fines	Lurgi Make Power	Lurgi & Texaco	Lurgi & Koppers
(b) Byproducts							
Production:							
(a) Ammonia (t/d)	-	-	-	300	300	210	210
(b) Sulfur (t/d)	150	150	150	150	150	150	150
Coal Fine (t/d)	-	-	-	7,342	-	-	-
Electricity (MW)	75	-	-	-	-	-	-
Revenues:							
Ammonia	-	-	-	20	20	14	14
Sulfur	5	5	5	5	5		
Coal Fines	-	-	-	21	-	-	-
Electricity	15	-	-	-	-	-	-
Total	20	5	5	46	25	19	19
(c) Total Revenues							
Methanol	954	954	954	954	954	954	954
Byproducts	20	5	5	46	25	19	19
Total	974	964	964	1,000	979	973	973

Table 5.5 Product Prices and Annual Sales Revenues

M - GASOLINE SYNTHESIS

(a) Principal Products (all cases)

	m^3/d	$\$/m^3$	Annual million m^3	Annual $ million
Gasoline	7648	251.73	2.523	635
LPG	1318	125.87	0.435	55

	Texaco	Koppers	Winkler	Lurgi Sell Fines	Lurgi Make Power	Lurgi & Texaco	Lurgi Koppers
(b) Byproducts							
Production:							
Ammonia (t/d)	-	-	-	330	330	231	231
Sulfur (t/d)	170	170	170	170	170	170	170
Coal Fines (t/d)	-	-	-	8,843	-	-	-
Electricity (MW)	50	-	-	-	70	-	-
Revenues:							
Ammonia	-	-	-	22	22	15	15
Sulfur	6	6	6	6	6	6	6
Coal Fines	-	-	-	25	-	-	-
Electricity	10	-	-	-	14	-	-
Total	16	6	6	53	42	21	21
(c) Total Revenues							
Gasoline	635	635	635	635	635	635	635
LPG	55	55	55	55	55	55	55
Byproducts	16	6	6	53	42	21	21
Adjustment*	334	334	334	334	334	334	334
Total	1,040	1,030	1,030	1,077	1,066	1,045	1,045

*Adjustment for differential between Canadian and international market prices.

Table 5.6

FINANCIAL SUMMARY

DIRECT COAL LIQUEFACTION

	$ million			
	SRC-I	SRC-II	H-COAL	EDS
Capital	2208	2208	2208	2760
Revenues	772	894	991	909
Operating Costs				
Raw Materials				
Coal	126	126	126	151
Water	4	4	4	5
Chemicals & Catalysts	12	15	28	28
Sub-total	142	145	158	184
Labor & Overhead	55	55	55	55
Replacement	66	66	66	83
Total	263	266	279	322
Gross Margin	509	628	712	587
Start-Up Costs	20	36	48	50
Working Capital	50	50	51	51

Table 5.7

FINANCIAL SUMMARY

FISCHER - TROPSCH SYNTHESIS

Base Case:	Texaco	Koppers	Winkler	Lurgi Sell Fines	Lurgi Max Power	Lurgi & Texaco	Lurgi & Koppers
				$ million			
Capital (1)	4,416	4,796	3,657	3,738	4,347	3,968	3,968
Revenues (2)	1,050	1,030	1,030	1,077	1,060	1,041	1,041
Operating Cost:							
Raw Materials:							
Coal (3)	231	235	237	219	219	193	194
Electricity (4)	-	20	30	89	-	62	71
Chemicals and Catalyst	13	13	15	10	10	12	12
Water (5)	8	8	7	7	8	7	7
Sub-total	252	276	289	325	237	274	284
Labor and Overhead (6)	70	70	70	70	70	70	70
Replacement (7)	132	144	110	112	130	119	119
Total Operating Costs	454	490	469	507	437	463	473
Gross Margin	596	540	561	570	623	578	568
Start-Up Cost	40	41	41	34	34	39	39
Working Capital	69	73	62	62	67	62	62

(1) Base case no escalation.
(2) Include $334 million adjustment to reflect crude oil price increase from $17.75/b to $38.00/b.
(3) Transfer price, as received at plant gate, is $10/t.
(4) Electric power cost supplied to the plant is ¢2.5/kwh. Texaco and Lurgi (make power) export power at same rate.
(5) Water cost supplied at plant gate ¢37.4/m^3 ($1.70 per thousand gallons).
(6) Based on plant Labor of 900 workers. Overhead is 100% of labor cost.
(7) Based on 3% of Capital and includes an element for contract maintenance labor.

Table 5.8

FINANCIAL SUMMARY

METHANOL SYNTHESIS

	Texaco	Koppers	Winkler	Lurgi Sell Fines	Lurgi Max Power	Lurgi & Texaco	Lurgi & Koppers
	$ million						
Capital	2,990	3,226	2,473	2,703	3,117	2,818	2,818
Revenues	974	964	964	1,000	979	973	973
Operating Costs:							
Raw Materials:							
Coal	157	158	147	154	154	137	136
Electricity	-	69	30	71	-	54	73
Chemicals & Catalyst	18	23	24	16	16	20	20
Water	5	5	5	5	6	6	6
Sub-total	180	255	353	246	176	217	235
Labor and Overhead	55	55	55	55	55	55	55
Replacement	90	97	74	81	94	85	85
Total	325	407	482	382	325	357	375
Gross Margin	649	557	482	612	654	616	598
Start-Up Cost	43	51	54	38	37	44	45
Working Capital	58	64	66	60	59	58	59

Table 5.9

FINANCIAL SUMMARY

M - GASOLINE SYNTHESIS

	Texaco	Koppers	Winkler	Lurgi Sell Fines	Lurgi Max Power	Lurgi & Texaco	Lurgi & Koppers
	$ million						
Capital	3,416	3,652	2,852	3,105	3,623	3,278	3,278
Revenues	1,040	1,030	1,030	1,077	1,066	1,045	1,045
Operating Costs:							
Raw Materials:							
Coal	166	168	159	171	171	149	148
Electricity	-	74	35	73	-	56	75
Chemicals & Catalyst	23	29	30	21	21	24	24
Water	4	4	4	4	5	4	4
Sub-total	193	275	228	269	197	382	251
Labor & Overhead	62	62	62	62	62	62	62
Replacement	103	110	86	93	109	98	98
Total	358	447	376	424	368	542	411
Gross Margin	682	583	654	653	698	503	634
Start-Up Cost	52	62	63	46	46	57	54
Working Capital	64	73	61	68	67	77	66

Table 5.9A Financial Summary
SNG Synthesis

Case	Lurgi Max. Power
Capital (1)	$ 3105 million
	$ million/year
Revenue (2) @ $4.2654/GJ	455
Operating Costs:	
Raw Materials	
Coal	126
Electricity	0
Catalyst and Chemicals	16
Water	4
Sub-total	146
Labor and Overhead	55
Replacement (3)	93
Total Operating Costs	294
Gross Margin	161
Start-up Cost	36
Working Capital	62

NOTES: 1) Base case, no contingency.
2) Based on plant labor of 700 workers, overhead is 100 percent of labor costs.
3) Based on 3 percent of capital.

becomes included in estimations of net present values. Given the future needs of the Province of British Columbia for light liquid fuels the H-Coal process has an advantage over the other direct liquefaction processes because it gives higher yields of light hydrocarbons than the SRC-I and SRC-II processes and requires less capital for a given yield than the EDS process.

In the case of the Indirect Liquefaction, the Lurgi process is undoubtedly that having the greatest commercial acceptance at the present time and has lower capital requirements. The cases in which the fines are totally consumed within the battery limits to produce electric power also provide the easiest solution to what would otherwise be a very difficult and expensive disposal problem. It was therefore determined that these four processes, each representing a different product slate of light liquid fuels, would be selected for further analysis to determine the sensitivity of these results to variation in certain important variables on net present values. A computer program was developed to examine the various combinations of changes in the different variables. (See Appendix C)

The economics of manufacturing SNG were also estimated for comparison with the data related to the liquefaction of coal.

5.3 Sensitivity Analyses

These analyses examine the effect of variation in capital costs, sales revenues and interest rates on net present value. The results are expressed in a series of tables (Tables 5.10 - 5.14) and figures (Figs. 5.1 - 5.9). Inspection of the tables generates the following comments.

Tables 5.10 - 5.14 show the effects of increase in capital costs and decrease in revenues on NPV at different interest rates of 3, 6 and 10 percent. B.C. Hydro indicated that

Table 5.10 Effect of Variation in Capital Costs, Sales Revenues and Interest Rates on Net Present Value

DIRECT LIQUEFACTION - H-COAL
(in millions of dollars)

CAPITAL COSTS		SALES REVENUES		NET PRESENT VALUE OF GROSS MARGIN		
Total	Change(1)	Total	Change(1)	3%*	6%*	10%*
2208	-	991	-	8487	3741	1068
2650	+ 20%	991	-	7882	3270	693
3312	+ 50%	991	-	6976	2558	139
4416	+100%	991	-	5469	1381	- 785
2208	-	892	-10%	7009	2930	666
2208	-	793	-20%	5535	2117	258
2208	-	595	-40%	2583	490	- 551
2650	+ 20%	892	-10%	6412	2457	288
2650	+ 20%	793	-20%	4934	1643	- 114
2650	+ 20%	595	-40%	1981	23	- 922
3312	+ 50%	892	-10%	5499	1750	- 266
3312	+ 50%	793	-20%	4027	934	- 670
3312	+ 50%	595	-40%	1076	- 691	-1476
4416	+100%	892	-10%	3994	564	-1191
4416	+100%	793	-20%	2519	- 244	-1596
4416	+100%	595	-40%	- 433	-1867	-2406

(1) from base case

* coal price is: $10/t at 3 percent interest rate
: $12/t at 6 percent interest rate
: $14/t at 10 percent interest rate

Table 5.11 Effect of Variation in Capital Costs, Sales Revenues and Interest Rates on Net Present Value

INDIRECT LIQUEFACTION-FISHER-TROPSCH
(in millions of dollars)

CAPITAL COSTS		SALES REVENUES		NET PRESENT VALUE OF GROSS MARGIN		
Total	Change(1)	Total	Change(1)	3%*	6%*	10%*
4347	-	1060	-	5177	1087	- 992
5216	+ 20%	1060	-	3991	151	-1722
6521	+ 50%	1060	-	2193	-1252	-2818
8694	+100%	1060	-	- 778	-3575	-4642
4347	-	954	-10%	3597	218	-1426
4347	-	848	-20%	2019	- 652	-1857
4347	-	636	-40%	-1142	-2389	-2724
5216	+ 20%	954	-10%	2413	- 715	-2154
5216	+ 20%	848	-20%	833	-1586	-2585
5216	+ 20%	636	-40%	-2327	-3325	-3453
6521	+ 50%	954	-10%	615	-2118	-3254
6521	+ 50%	848	-20%	- 968	-2987	-3685
6521	+ 50%	636	-40%	-4126	-4724	-4551
8694	+100%	954	-10%	-2361	-4444	-5077
8694	+100%	848	-20%	-3940	-5315	-5507
8694	+100%	636	-40%	-7099	-7052	-6347

(1) from base case

* coal price is: $10/t at 3 percent interest rate
: $12/t at 6 percent interest rate
: $14/t at 10 percent interest rate

Table 5.12 Effect of Variation in Capital Costs, Sales Revenues and Interest Rates on Net Present Value

INDIRECT LIQUEFACTION-METHANOL
(in millions of dollars)

CAPITAL COSTS		SALES REVENUES		NET PRESENT VALUE OF GROSS MARGIN		
Total	Change(1)	Total	Change(1)	3%*	6%*	10%*
3117	-	979	-	6786	2469	130
3740	+ 20%	979	-	5945	1804	- 390
4676	+ 50%	979	-	4665	805	-1175
6234	+100%	979	-	2527	- 864	-2487
3117	-	881	-10%	5324	1663	- 268
3117	-	783	-20%	3867	861	- 670
3117	-	587	-40%	944	- 747	-1473
3740	+ 20%	881	-10%	4485	1004	- 792
3740	+ 20%	783	-20%	3024	201	-1191
3740	+ 20%	587	-40%	107	-1408	-1993
4676	+ 50%	881	-10%	3203	2	-1574
4676	+ 50%	783	-20%	1743	- 803	-1973
4676	+ 50%	587	-40%	-1175	-2407	-2780
6234	+100%	881	-10%	1068	-1665	-2887
6234	+100%	783	-20%	- 393	-2471	-3288
6234	+100%	587	-40%	-3314	-4075	-4087

(1) from base case

* coal price is: $10/t at 3 percent interest rate
: $12/t at 6 percent interest rate
: $14/t at 10 percent interest rate

Table 5.13 Effect of Variation in Capital Costs, Sales Revenues and Interest Rates on Net Present Value

INDIRECT LIQUEFACTION - M-GASOLINE
(in millions of dollars)

CAPITAL COSTS		SALES REVENUES		NET PRESENT VALUE OF GROSS MARGIN		
Total	Change(1)	Total	Change(1)	3%*	6%*	10%*
3623	-	1066	-	6970	2375	- 85
4348	+ 20%	1066	-	5980	1609	- 689
5435	+ 50%	1066	-	4490	439	-1608
7246	+100%	1066	-	2013	-1497	-3123
3623	-	959	-10%	5372	1496	- 523
3623	-	848	-20%	3715	590	- 974
3623	-	640	-40%	615	-1116	-1826
4348	+ 20%	959	-10%	4383	731	-1128
4348	+ 20%	848	-20%	2732	- 179	-1582
4348	+ 20%	640	-40%	- 368	-1885	-2433
5435	+ 50%	959	-10%	2893	- 442	-2043
5435	+ 50%	848	-20%	1238	-1347	-2497
5435	+ 50%	640	-40%	-1857	-3054	-3348
7246	+100%	959	-10%	419	-2373	-3558
7246	+100%	848	-20%	-1234	-3282	-4012
7246	+100%	640	-40%	-4331	-4988	-4864

(1) from base case

* coal price is: $10/t at 3 percent interest rate
: $12/t at 6 percent interest rate
: $14/t at 10 percent interest rate

Table 5.14 Effect of Variation in Capital Costs, Sales Revenues and Interest Rates on Net Present Value - SNG Synthesis Export Selling Price of $4.2654/GJ

(millions of Dollars)

Capital Costs		Sales Revenue		Net Present Value		
Total	Charge(1)	Total	Charge(1)	3%*	6%*	10%*
3105	-	455	-	- 543	-1513	-1825
3726	+ 20%	455	-	-1403	-2179	-2346
4658	+ 50%	455	-	-2674	-3177	-3130
6210	+100%	455	-	-4792	-4621	-4429
3105	-	410	-10%	-1213	-1881	-2005
3105	-	364	-20%	-1905	-2258	-2195
3105	-	273	-40%	-3226	-2977	-2537
3726	+ 20%	410	-10%	-2609	-2550	-2533
3726	+ 20%	364	-20%	-2757	-2923	-2710
3726	+ 20%	273	-40%	-4083	-3645	-3066
4658	+ 50%	410	-10%	-3345	-3547	-3314
4658	+ 50%	364	-20%	-4030	-3896	-3473
4658	+ 50%	273	-40%	-5338	-4642	-3848
6210	+100%	410	-10%	-5463	-5175	-4585
6210	+100%	364	-20%	-6120	-5550	-4773
6210	+100%	273	-40%	-7477	-6300	-5147

---

(1) from base case

* coal price is $10/t at 3% interest rate
$12/t at 6% interest rate
$14/t at 10% interest rate

these changes in interest rates would also affect the cost of coal delivered to the battery limits, as detailed at the base of each table.

Capital costs were increased by 20, 50 and 100 percent. Revenues were decreased by 10, 20 and 40 percent. No account has been taken of changes more favorable than the base case.

The predominant effects of interest rates on overall project economics is summarized in Table 5.15 which shows the effect of higher interest rates on NPV for the four processes at base capital costs and sales revenues; and for the assumed worst cases in which estimated capital costs are doubled and sales revenues reduced by 40 percent.

Table 5.15 Effect of Interest Rate on Net Present Values of Gross Margin

(millions of dollars)

Process	Case	3 Percent	6 Percent	10 Percent
H-Coal	Base	8487	3741	1068
	Worst	- 433	-1867	-2408
M-Gasoline	Base	6970	2375	- 85
	Worst	-4331	-4988	-4864
Methanol	Base	6786	2469	130
	Worst	-3314	-4075	-4087
Fischer-Tropsch	Base	5177	1087	- 992
	Worst	-7099	-7052	-6347
SNG (Export)	Base	- 543	-1513	-1825

Figures 5.1 - 5.4 show NPV at a fixed 3 percent interest rate for variable capital costs and sales revenues.

Figures 5.5 - 5.8 show NPV at fixed sales revenues but variable interest rates and capital costs.

Figure 5.9 shows the effects of interest rate variations alone on NPV, capital costs and sales revenues being fixed at the base case for all four selected processes.

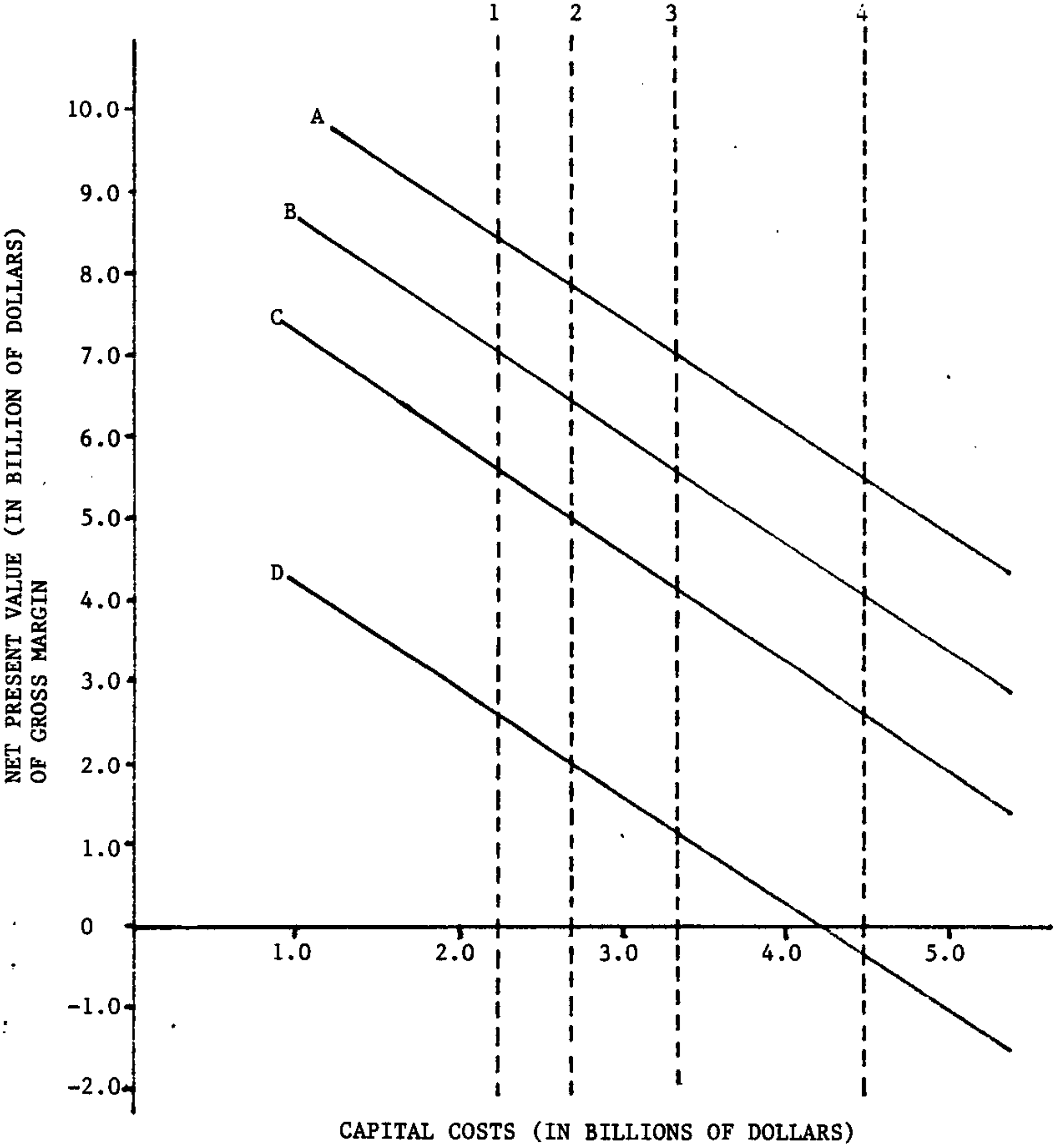
For each figure and set of conditions the break-even point is represented by the abscissa at zero NPV. In Figures 5.1 - 5.4 the various levels of capital costs are represented by vertical dotted lines. In Figure 5.9 the vertical dotted lines represent the interest rates set at 3, 6 and 10 percent.

> N.B. In interpreting the results in these tables and figures it must be borne in mind that the financial analysis employed does not make provisions for taxation of rates applicable to Canadian corporations and does not allow for depreciation.

With this proviso, the results indicate that the financial viability of the selected processes may be ranked

H-Coal
Methanol
M-Gasoline
Fischer-Tropsch.

In the case of SNG the results show that, even based on current natural gas export prices, an investment in SNG would incur substantial negative net present values over the whole range of parameters considered. If current domestic gas prices are used, even larger losses would result. The manufacture of SNG is not an economically viable option at the present time. This situation could change in the future if the prices of natural gas escalated at a rate greater than future escalations in oil prices.



CAPITAL COSTS
1. Base case ($2208 million)
2. 20 percent over base case ($2650 million)
3. 50 percent over base case ($3312 million)
4. 100 percent over base case ($4416 million)

ANNUAL SALES REVENUES
A. Base case ($991 million)
B. 90 percent of base case ($892 million)
C. 80 percent of base case ($793 million)
D. 60 percent of base case ($595 million)

B. C. HYDRO

FIG. 5.1

DIRECT LIQUEFACTION
H-COAL
COAL AT $10/t
INTEREST RATE - 3 PERCENT

NJ 1024

NORTH AMERICAN MINING
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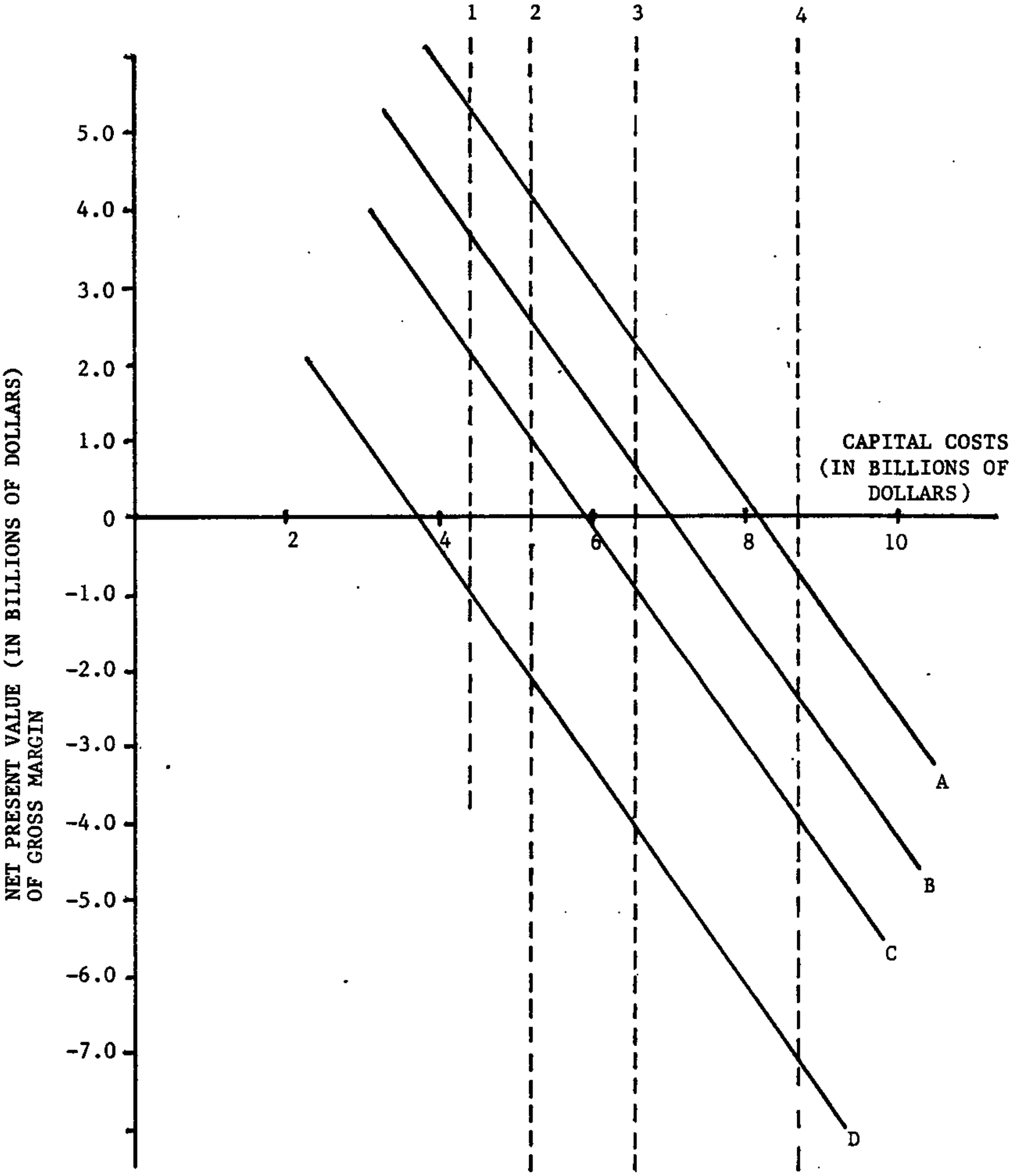


CAPITAL COSTS

1. Base case ($4347 million)
2. 20 percent over base case ($5216 million)
3. 50 percent over base case ($6521 million)
4. 100 percent over base case ($8694 million)

ANNUAL SALES REVENUES

A. Base case ($1060 million)
B. 90 percent of base case ($954 million)
C. 80 percent of base case ($848 million)
D. 60 percent of base case ($636 million)

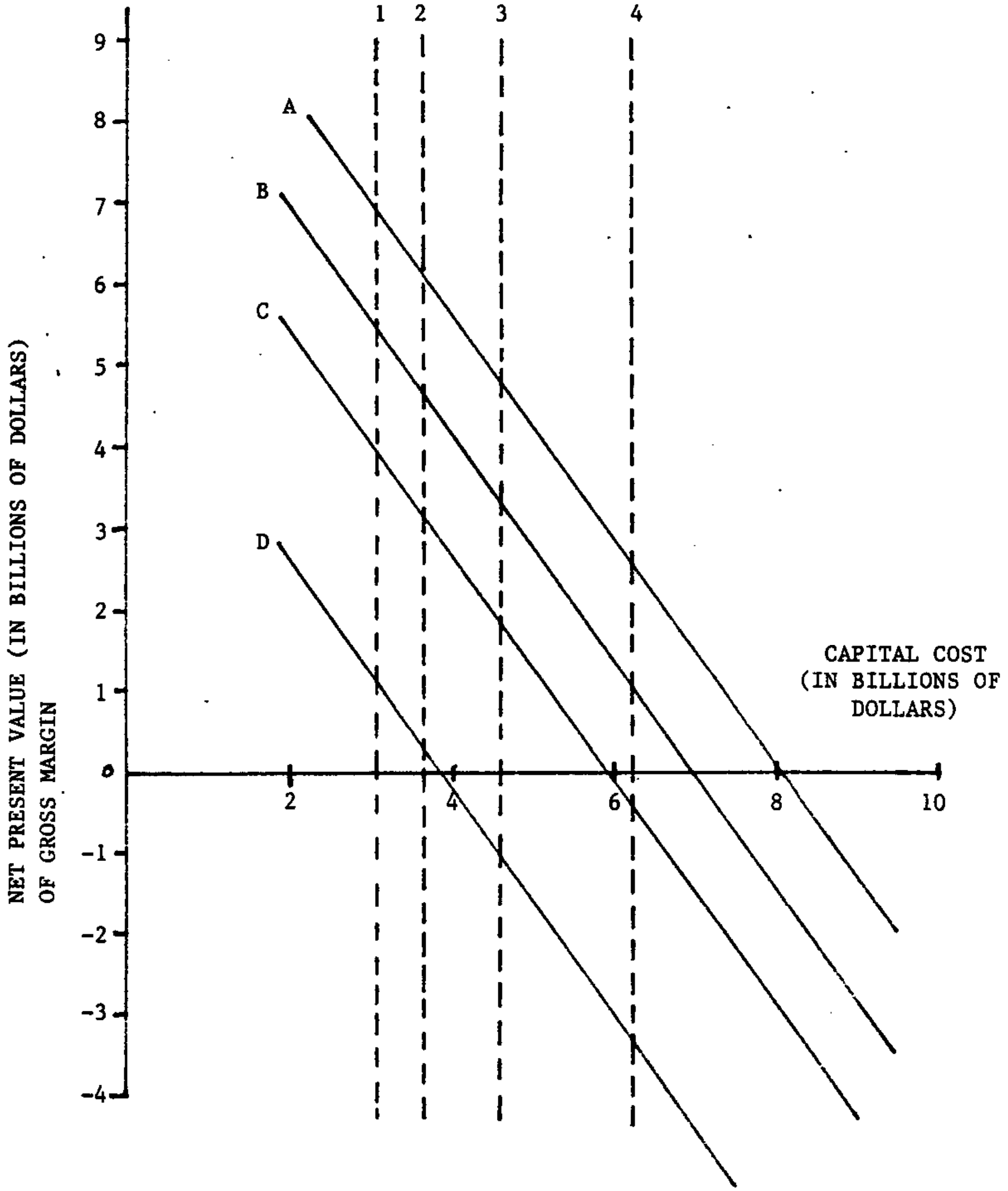
B. C. HYDRO

FIG. 5.2

INDIRECT LIQUEFACTION
FISCHER-TROPSCH
LURGI (MAX POWER)
COAL AT $10/t
INTEREST RATE - 3 PERCENT

NJ 1024

NORTH AMERICAN MINING
CONSULTANTS, INC.



CAPITAL COSTS
1. Base case ($3117 million)
2. 20 percent over base case ($3740 million)
3. 50 percent over base case ($4676 million)
4. 100 percent over base case ($6234 million)

ANNUAL SALES REVENUES
A. Base case ($979 million)
B. 90 percent of base case ($881 million)
C. 80 percent of base case ($783 million)
D. 60 percent of base case ($587 million)

B. C. HYDRO

FIG. 5.3

INDIRECT LIQUEFACTION
METHANOL
LURGI (MAX POWER)
COAL AT $10/t
INTEREST RATE - 3 PERCENT

NJ 1024

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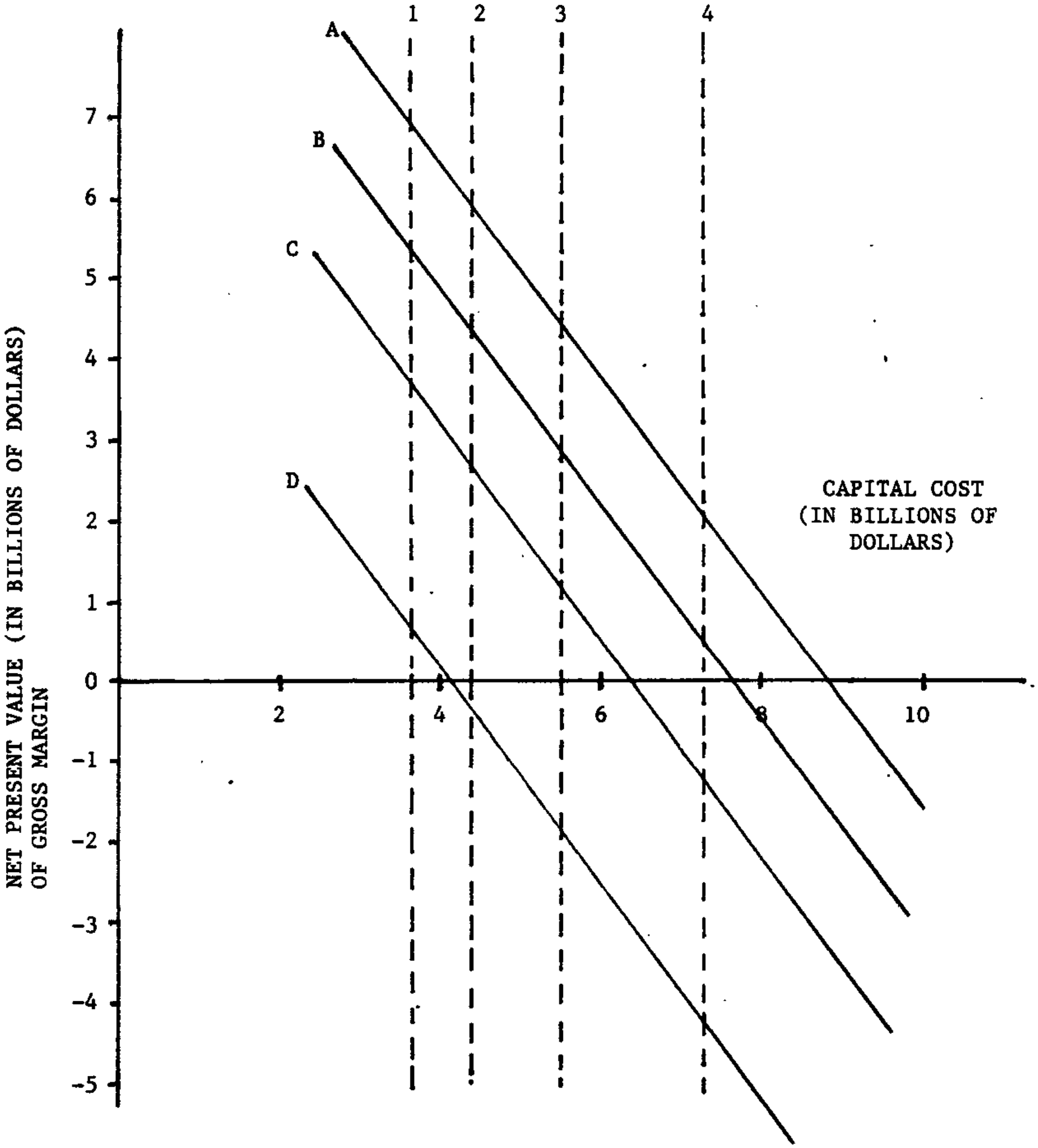
CONSULTANTS, INC.



CAPITAL COSTS
1. Base case ($3623 million)
2. 20 percent over base case ($4348)
3. 50 percent over base case ($5435)
4. 100 percent over base case ($7246)

ANNUAL SALES REVENUES
A. Base case ($1066 million)
B. 90 percent of base case ($959 million)
C. 80 percent of base case ($848 million)
D. 60 percent of base case ($640 million)

B. C. HYDRO

FIG. 5.4

INDIRECT LIQUEFACTION
M-GASOLINE
LURGI (MAX POWER)
COAL AT $10/t
INTEREST RATE - 3 PERCENT

NJ 1024

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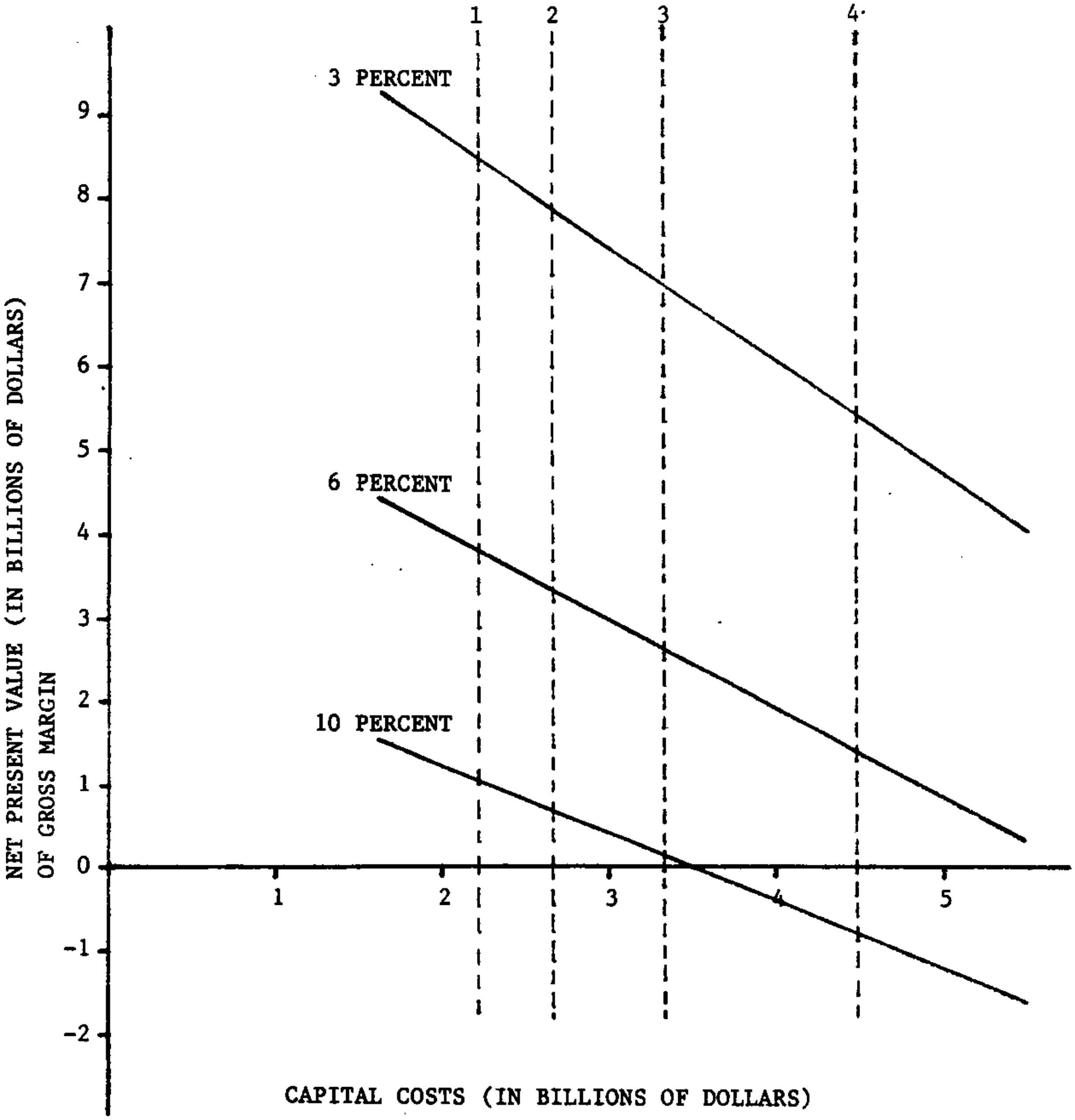
CONSULTANTS, INC.



CAPITAL COSTS

1. Base case ($2208 million)
2. 20 percent over base case ($2650 million)
3. 50 percent over base case ($3312 million)
4. 100 percent over base case ($4416 million)

ANNUAL SALES REVENUES

Base case ($991 million) fixed.

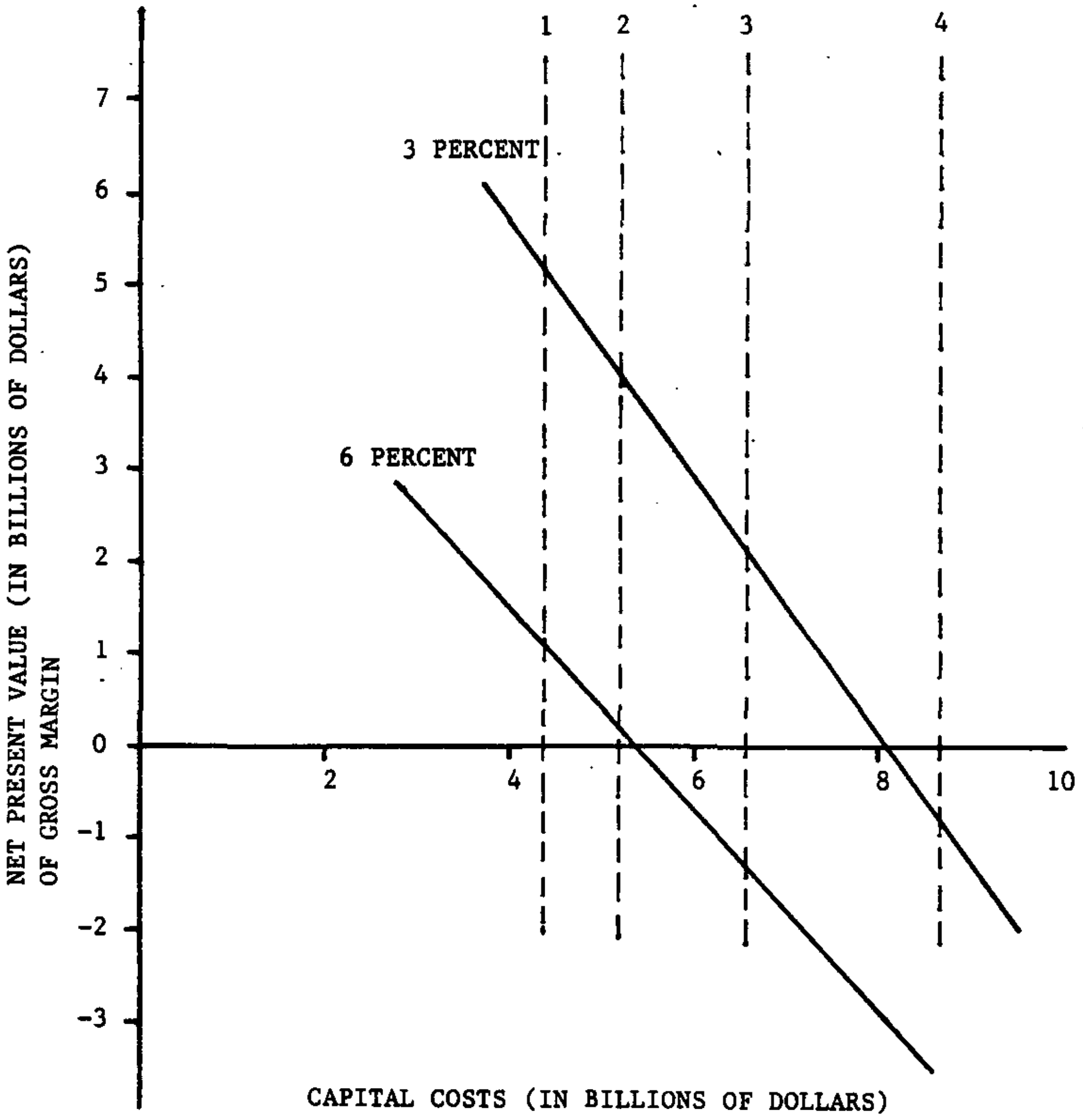
B. C. HYDRO

FIG. 5.5

H-COAL
N.P.V. AT DIFFERENT INTEREST
RATES AND CAPITAL COSTS
FIXED SALES REVENUES

NJ 1024

NORTH AMERICAN MINING
CONSULTANTS, INC.



CAPITAL COSTS

1. Base case ($4347 million)
2. 20 percent over base case ($5216 million)
3. 50 percent over base case ($6521 million)
4. 100 percent over base case ($8694 million)

ANNUAL SALES REVENUES

Base case ($1060 million) fixed

NOTE: AT 10 PERCENT INTEREST RATE ALL NET PRESENT VALUES ARE NEGATIVE.

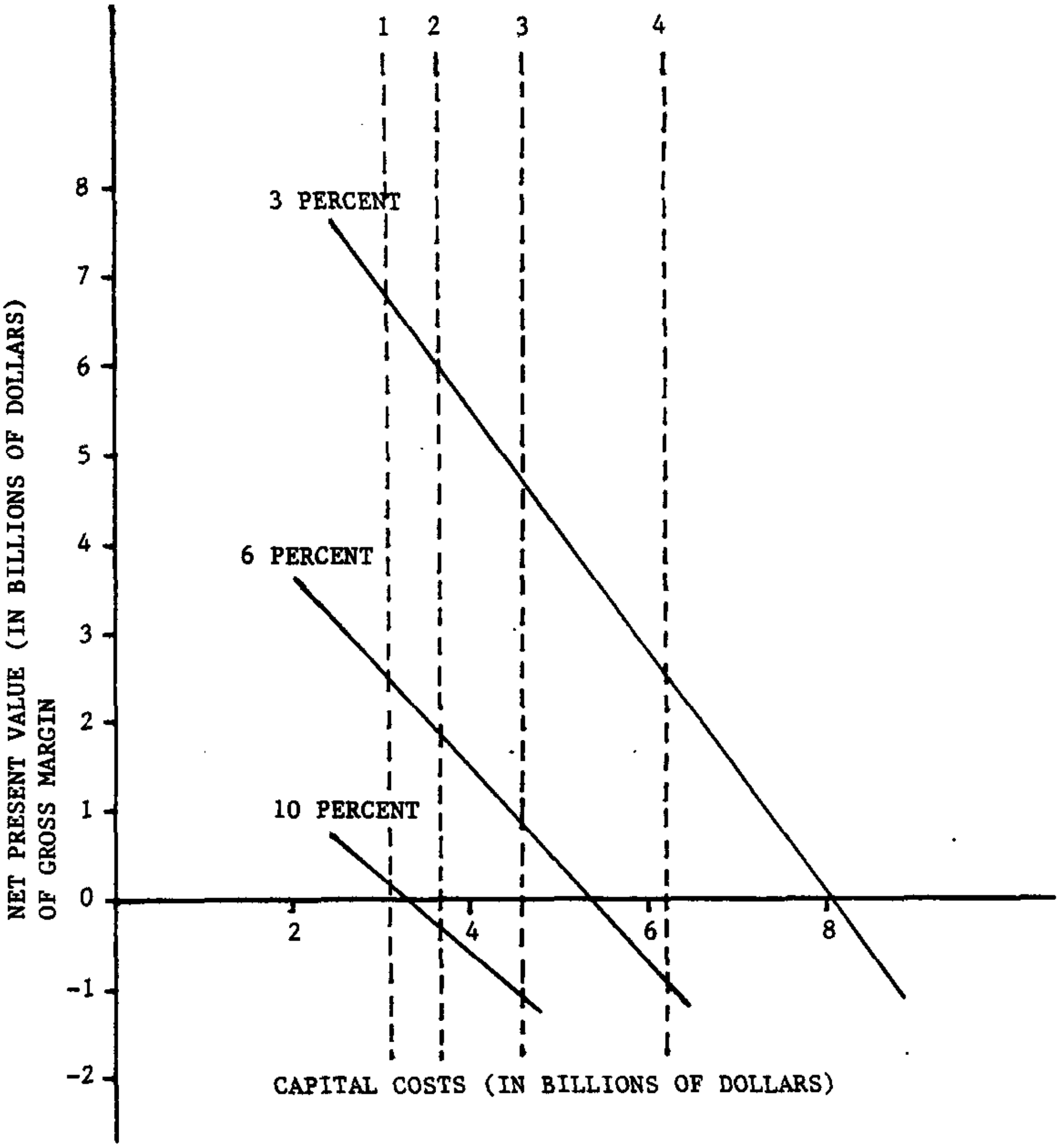
B. C. HYDRO

FIG. 5.6

FISCHER-TROPSCH
N.P.V. AT DIFFERENT INTEREST RATES
AND CAPITAL COSTS
FIXED SALES REVENUES

NJ 1024

NORTH AMERICAN MINING
CONSULTANTS, INC.



CAPITAL COSTS

1. Base case ($3117 million)
2. 20 percent over base case ($3740 million)
3. 50 percent over base case ($4676 million)
4. 100 percent over base case ($6234 million)

ANNUAL SALES REVENUES

Base case ($979 million) fixed

B. C. HYDRO

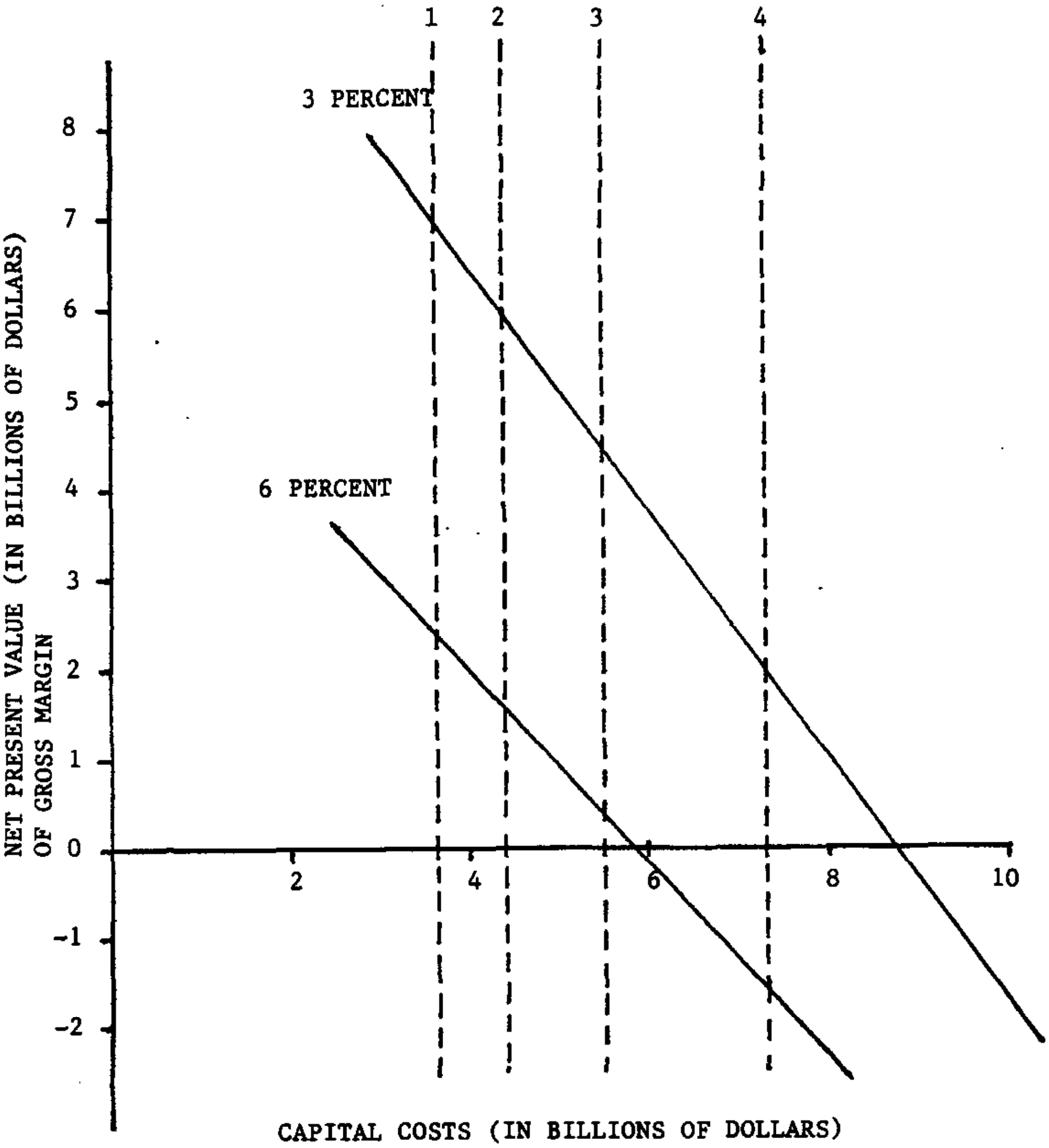
FIG. 5.7

METHANOL

N.P.V. AT DIFFERENT INTEREST RATES AND CAPITAL COSTS

NJ 1024

NORTH AMERICAN MINING CONSULTANTS, INC.



CAPITAL COSTS

1. Base case ($3623 million)
2. 20 percent over base case ($4348 million)
3. 50 percent over base case ($5435 million)
4. 100 percent over base case ($7246 million)

ANNUAL SALES REVENUES

Base case ($1066 million) fixed

NOTE: AT 10 PERCENT INTEREST RATE ALL NET PRESENT VALUES ARE NEGATIVE.

B. C. HYDRO

FIG. 5.8

M-GASOLINE

N.P.V. AT DIFFERENT INTEREST RATES AND CAPITAL COSTS

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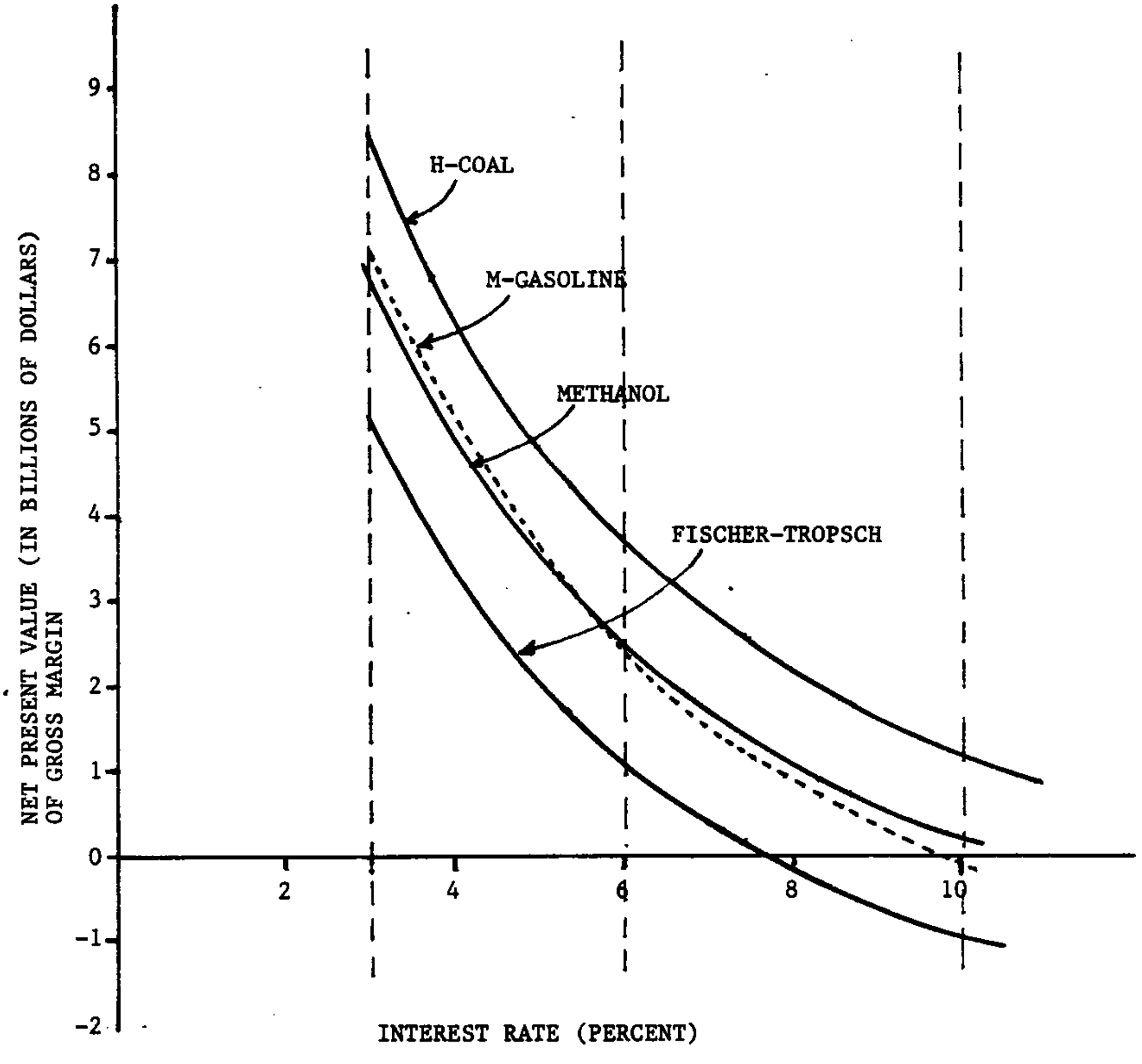

PROCESS	CAPITAL COST	ANNUAL SALES
H-Coal	$2208 million	$ 991 million
Fischer-Tropsch	$4347 million	$1060 million
Methanol	$3117 million	$ 979 million
M-Gasoline	$3623 million	$1066 million

ALL INDIRECT LIQUEFACTION CASES ARE BASED ON LURGI (MAKE POWER) PROCESS ECONOMICS.

B. C. HYDRO

FIG. 5.9

EFFECT OF INTEREST RATE ON NET PRESENT VALUE
BASE CASE CAPITAL COSTS AND SALES REVENUE

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6. ENVIRONMENTAL CONSIDERATIONS

6.1 General Remarks

Coal conversion processes are faced with potential pollution problems that are common to coal burning power plants and with possible pollution problems which are peculiar to the conversion processes. Due to the relative lack of experimental data and full scale coal conversion operating experience, relative environmental impact of such processes compared with the impact produced by a coal fired power plant is difficult to assess. The very nature of most coal conversion systems, including the utilization of sulfur recovery systems and the implementation of zero-discharge liquid waste treatment systems, assures the probability that some potential effluent streams will be reduced in size when compared to a power plant; however, the large numbers of chemical products and wastes produced by these systems provide potential for a variety of undesirable impacts which are not fully understood at this time.

For the assessment of environmental considerations, data availability and process similarities resulted in the categorization of coal conversion processes into three general groups; Low Temperature Gasification, High Temperature Gasification and Direct Liquefaction. The division of gasifiers into two groups is primarily related to the production of phenols, oils and tars during reaction. The raw product gas from Low Temperature Gasifiers contains significant quantities of these constituents as well as other organics and these components eventually appear in process waste streams. The presence of these compounds adds complexity to the water treatment facilities required. High Temperature Gasifiers produce few organic contaminants and water treatment requirements are considerably simplified.

Due to operational experience and data availability, the assessment of Low Temperature Gasification was based entirely on the Lurgi process while the High Temperature Gasifier analysis was based primarily on Koppers-Totzek data. The results, however, should also be generally applicable to the Texaco and Winkler processes. Process differences which are significant relative to environmental assessment are identified and discussed.

The assessment of Direct Liquefaction processes is related to solvent hydrogenation technologies since pyrolysis/hydrocarbonization technologies were not considered viable (See Section 2). Solvent hydrogenation technologies are similar, from a process viewpoint, and waste stream compositions should be comparable although there may be variations in concentration and quantity. Consequently, the conceptual control technologies can basically be established generically for solvent hydrogenation processes. Because of data availability, the Direct Liquefaction assessment is based on the H-Coal and SRC-II processes. The results should, within the accuracy of available information, be also applicable to the Exxon Donor Solvent process.

In general, for the indirect production of coal liquids utilizing either high or low temperature gasification processes, the "add-on" synthesis facilities are environmentally clean in comparison with the gasification processes which preceed them. Vented gaseous emissions and contaminated bottoms and/or condensate streams would be relatively small in quantity and could be combined with comparable streams generated in the gasification process for subsequent treatment. A lack of available data precludes definitive assessment.

In this Study, and in order to facilitate technical and economic comparisons, each coal conversion process considered is based on the production of 316.5TJ/d of useful conversion products, i.e., equivalent to approximately 50,000 barrels/day of liquid products or 3663 MW_t. The particular coal feed rates will then reflect process efficiencies; and for the three categories of conversion processes considered, are summarized in Table 6.1.

All references to plant size which follow are referable to this common base.

An assessment summary is provided in Section 6.6 which includes a comparison with a conventional coal-fired power plant using 40,500 t/d of coal, as described in the Hat Creek Project Environmental Impact Statement [Ref. 102]. Although the comparison provides an instructive view of relative impacts in the vicinity of the plant, it is important to point out that consumption of the products of a coal-conversion plant will occur in a much larger geographical area and will provide environmental impacts which will be significantly different from those associated with the consumption of electric power.

6.2 Applicable Standards

At the time of report preparation, British Columbia had not formulated specific environmental standards directly applicable to coal-conversion technologies. Discussions held with Ministry of Environment personnel in October 1980 indicated that the development of such standards has just recently been considered and that it would be several years before promulgation.

Table 6.1 Coal Feed Summary

Coal Feed (t/d)	Koppers-Methanol	Lurgi-Fischer-Tropsch	Direct Liquefaction
Process	42,960	41,890	33,455
Boilers	3,000	23,560	3,000
Dryers	2,070	945	1,640
Total	48,930	66,395	38,095
Residues	-	-	9,900

It was consequently decided that the "Pollution Control Objectives for the Chemical and Petroleum Industries of British Columbia" [Ref. 72] would be used to determine control levels and for compliance assessments. Although none of the coal conversion processes are specifically described by the Pollution Control Objective guidelines of British Columbia, coal conversion processes are sufficiently similar to the chemical and petroleum industry that extrapolation of the objectives for those industries to the coal conversion industry is a reasonable approach. The Director of Pollution Control Branch may establish other minimum requirements, if it is determined that extrapolation is not suitable.

6.2.1 Air Emissions

Table 6.2 presents the air emission objectives which may be applicable to coal conversion technologies and associated steam generation facilities. Level A objectives are applicable to new facilities. The Pollution Control Objectives [Ref. 72], with respect to specification of control technologies, require the use of smokeless flares and the following emission reduction facilities associated with petroleum product storage vessels having capacities greater than 54,000 imperial gallons (~250 m^3):

Vapor Pressure	Requirements
< 10.5 kPa (1.52 psia)	Conservation vent
10.5 - 76.5 kPa	Floating roof
> 76.5 kPa (11.1 psia)	Vapor recovery system

Other general standards require minimization of cooling tower hydrocarbon emissions, elimination of nuisance odors, and control of sulfur dioxide emissions during sulfur plant catalyst bed regeneration.

Table 6.2 Level A Objectives for Air Emissions Applicable to Coal Conversion Processes

Sulfur Plant	
Sulfur recovery, % (a)	99+
Sulfur dioxide, mg/m^3 (ppm) (b)	830 (300)
Overall Refinery	
Sulfur trioxide, mg/m^3 (gr/SCF) (b)	25 (0.011)
Fluid Catalytic Cracking Unit	
Particulate solids, mg/m^3 (gr/SCF) (b)	115 (0.050)
Hydrocarbons (as Hexane) -	
mg/m^3 (ppm) (b)	90 (25)
g/m^3 cracking feed	57
Carbon monoxide, mg/m^3 (ppm) (b)	2,400 (2,000)
Sulfur dioxide, mg/m^3 (ppm) (b)	830 (300)
Steam Plant	
Particulate solids, mg/m^3 (gr/SCF) (b) (c)	150 (0.065)
Sulfur dioxide, mg/m^3 (ppm) (b)	830 (300)
Sulfur Recovery - Percent	
Acid gas CO_2/H_2S ratio lower than 10	99
Acid gas CO_2/H_2S ratio higher than 10 (d)	95

(a) Total sulfur recovered from refinery fuel gases.
(b) mg/m^3 at 20°C, 760 mm Hg, dry basis.
(c) Corrected to 12 percent carbon dioxide.
(d) Individual assessment may be required.

Table 6.2 (cont'd)

Substance	Level A (a)
Acetaldehyde	3.8 (2.1)
Acrolein	2.5 (1.0)
Alkanolamines	60
Ammonia	180 (250)
Benzene	800 (250)
Carbon monoxide	2,400 (2,000)
Cobalt	7.0 (0.003) (b)
Copper	7.0 (0.003) (b)
Cresol	220 (50)
Dimethyl ether	190 (100)
Diphenyl	10 (2)
Formaldehyde	30 (20)
Formic acid	90 (50)
Hydrogen sulfide	7 (4.7)
Maleic anhydride	10 (2.5)
Mercaptans	(c)
Methanol	2,600 (2,000)
Methyl ethyl ketone (2-butanone)	900 (300)
Methyl Isobutyl Ketone	2,050 (500)
Organic disulfides	4
Particulate solids -	
Total	230 (0.100) (b)
Organic chemical dust	115 (0.050) (b)
Phenol	100 (26)
Phthalic anhydride	120 (20)
Styrene	200 (47)
Toluene	3,750 (1,000)
Vinyl acetate	30 (10)

(a) Concentrations given in milligrams per cubic meter (20 C, 760 mm Hg, dry basis) and in parentheses, parts per million by volume.

(b) Concentrations in parentheses are given in grains per standard cubic foot.

Source: [Ref. 72].

6.2.2 Liquid Effluents

Table 6.3 presents the applicable effluent-quality objectives. Level A objectives would be applicable. These effluent standards are based upon maximum control of losses and reduction of wastes through recovery and recycling. Separate sewer systems should be maintained for uncontaminated and waste water streams. Cooling towers or air-fan coolers are strongly recommended for thermal pollution control.

Where plausible, effluent streams should be combined so that only a single discharge point is necessary. Overall, the waste treatment system should be designed to eliminate toxicity and reduce gross emissions. The utilization of the "zero discharge" concept for the development of environmental controls essentially eliminates the need for liquid effluent compliance determinations. The reuse of liquid wastes, however, will result in an increased concern associated with the handling and disposal of various solid waste streams such as ash/slag which come in contact with the recycled liquids and evaporator residues.

6.2.3 Solid Wastes

As stated in the Pollution Control Objectives [Ref. 72], "Progressive objective levels have not been specified for the disposal of industrial refuse" due to the dependence of disposal requirements on plant and site specific conditions. General guidelines require the protection of ground and surface waters. Sludges should be neutralized and dewatered prior to disposal. Thixotropic sludges require stabilization prior to landfill. Soil cultivation or ground spraying is permitted for biological sludges, however. According to Ministry of Environment personnel, the disposal of potentially toxic and hazardous wastes would be handled under

Table 6.3

Effluent-quality Objectives for Chemical Industries Other Than Petroleum Refineries

	Discharges to Marine Waters			Discharges to Fresh Waters			Monitoring
	Level A	Level B	Level C	Level A	Level B	Level C	
Oil, nonvolatile, mg/l (a)	-	10	15	-	10	15	Daily composite, once per week
Oil, total, mg/l	10	-	-	5	-	-	Daily composite, once per week
BOD, five-day, 20°C, mg/l	20	45	130	20	45	130	COD or TOC once per week, BOD checked quarterly
Ammonia, as N, mg/l	10	15	15	10	15	15	Daily composite, once per week
Nitrates, as N, mg/l	10	50	50	10	50	50	Daily composite, once per week
Total nitrogen, mg/l (kjeldahl)	15	25	25	15	25	25	Daily composite, once per week
Chlorate, mg/l	75	150	150	50	100	100	Daily composite, once per week
Chloride ion, mg/l (b)	-	-	-	-	-	-	Daily composite, once per week
Chlorine, residual, mg/l	0.2	0.5	1.0	0.2	0.5	1.0	Continuous
Fluoride, mg/l	2.5	10	15	2.5	10	15	Daily composite, once per week
Formaldehyde, mg/l	5	5	10	5	5	10	Daily composite, once per week
Metals (total)							
Arsenic, trivalent, mg/l	0.05	0.05	0.08	0.05	0.05	0.05	Daily composite, once per month
Barium, mg/l	1.0	1.0	1.5	1.0	1.0	1.0	Daily composite, once per month
Boron, mg/l	15	15	15	10	10	10	Daily composite, once per month
Chromium, mg/l	0.2	0.2	0.2	0.2	0.2	0.2	Daily composite, once per month
Copper, mg/l	0.1	0.1	0.1	0.1	0.1	0.1	Daily composite, once per month
Lead, mg/l	0.2	0.2	0.2	0.2	0.2	0.2	Daily composite, once per month
Mercury, mg/l	0.002	0.050	0.050	0.002	0.050	0.050	Daily composite, once per month
Nickel, mg/l	0.2	0.2	0.2	0.2	0.2	0.2	Daily composite, once per month
Zinc, mg/l	0.2	0.2	0.3	0.2	0.2	0.2	Daily composite, once per month
Phenols, mg/l	0.2	0.3	1.0	0.2	0.3	1.0	Weekly grab
Phosphate, as P, mg/l	-	-	-	1.0	10	30	Daily composite, once per week
Sulphate, mg/l (b)	-	-	-	-	-	-	Daily composite, once per week
Urea, mg/l	2.0	2.0	2.0	1.0	1.0	1.0	Daily composite, once per week
Sulphides, mg/l	0.10	0.10	1.0	0.10	0.10	1.0	Weekly grab
Cyanide, mg/l	0.10	0.10	0.20	0.10	0.10	0.20	Weekly grab
Suspended solids, mg/l (c)	20	20	30	20	20	30	Daily composite, once per week
Settleable solids, mg/l (c)	<0.5	<0.5	<0.5	<0.5	<0.5	<0.5	Daily composite, once per week
Floatable solids	(d)	(d)	(d)	(d)	(d)	(d)	Daily observation
Total solids, mg/l (e)	3,000	3,000	3,000	1,500	1,500	1,500	Daily composite, once per week
Colour, Pt Co Units at pH 7	20	20	30	15	15	20	
Turbidity, JTU	15	15	25	10	10	15	
Temperature, °C maximum	32	32	32	32	32	32	Continuous
pH	6.5-8.5	6.95-9.0	6.5-9.0	6.5-8.5	6.5-8.5	6.5-9.0	Continuous
Toxicity (f)	50	45	25	100	90	50	Quarterly

(a) For discharge of once-through cooling water used for indirect cooling (heat exchangers, bearings, etc.) the maximum permissible oil concentration is 2 mg/l above background.

(b) While the importance of these characteristics is recognized, no limits have been established at this time.

(c) Not applicable to discharges to exfiltration ponds.

(d) Negligible

(e) Depends upon the nature of solids other than normal marine composition.

(f) 96 hour Tlm Bioassay on salmonid species, expressed as per cent by volume of effluent in receiving-water which is required to give 50 per cent survival over 96 hours.

Source: [Ref. 72]

normal permitting procedures, but would require assessment on a plant/site specific basis.

6.3 Steam Generation

All coal conversion processes require boilers for the generation of steam to meet process requirements and for the production of power (if not imported). For the purposes of assessment, coal fired boilers have been assumed although it is recognized that more detailed study may show distinct advantages associated with the utilization of certain byproduct streams as boiler fuels. For the assessment of Lurgi-based systems, the screened fines will be utilized as boiler fuel. All other analyses assume utilization of Performance Blend coal.

The boiler "package," including makeup water treatment and gaseous/liquid/solid waste treatment systems, can be environmentally investigated almost independently of the associated coal conversion facility. Except for the provision of makeup water and the combination of solid waste streams prior to transport to the disposal facility, very little waste stream integration is required. Integration would become more complex, however, with the utilization of coal conversion byproducts as fuel or if sulfur removal/ recovery facilities were to be combined with systems used in the coal conversion process.

Table 6.4 presents, for the Lurgi-Fischer-Tropsch (maximum power) alternative case which requires the largest boiler capacity, the estimated coal feed rate required for steam generation, a breakdown of makeup water requirements and solid wastes generated, and a comparison of gaseous emissions with applicable standards. The basis for derivation of the various projections is provided in following discussions.

Table 6.4 Example of Steam Generation Operating Characteristics for the Lurgi-Fischer-Tropsch Case using Hat Creek Coal

Coal Feed (t/d)	23,560
Steam (t/h)	2,940
Makeup Water (t/d)	
Bottom Ash	1,560
Fly Ash	1,400
Scrubber	7,070
Total	10,030
Particulate Emission - (mg/m^3)*	
Standard	150
Uncontrolled	60,000
Percent Removal Required	99.75
SO_2 Emissions - (mg/m^3)*	
Standard	830
Uncontrolled	1,750
Percent Removal Required	53
Solid Wastes (Wet) - (t/d)	
Fly Ash	6,990
Bottom Ash	2,500
Scrubber Sludge	480
Total	9,970

*mg/m^3 at 20°C, 760 mm Hg, dry basis

6.3.1 Gaseous Emissions

Untreated boiler flue gas is potentially the most significant gaseous emission associated with conversion processes. However, conventional control technology is well proven and capable of reducing controlled pollutants to levels which are considered environmentally safe. For assessment purposes, electrostatic precipitators with a removal efficiency of 99.75 percent have been assumed for removal of particulates. Tests conducted for British Columbia Hydro and Power Authority by Southern Research Institute, which investigated fly ash emission concentrations and rates, particle size, and resistivity, determined that electrostatic precipitation was viable for Hat Creek Coal fly ash [Ref. 69]. Precipitator costs would vary from $125 million (1980) for the Lurgi maximum power alternative to approximately $20 million (1980) for boilers sized just for generation of on-site steam requirements and emergency power.

Wet nonregenerative scrubbers have been incorporated into the boiler design for sulfur dioxide removal and sized on the basis of meeting Provincial standards. For emission calculations, air flow rates were based on an assumed boiler rating of 0.4 kg/MJ (includes 15 percent excess air). It was also assumed that gases exiting the scrubber were heated to 80°C to achieve better plume rise and dispersion. Scrubber costs would vary from $60 million (1980) for the Lurgi maximum power alternative to $10 million (1980) for the no-export power boilers.

6.3.2 Liquid Wastes

Major potential waste streams associated with boiler operation are bottom ash quench and sluice water, the recirculation lime or limestone solution used for sulfur removal in the wet nonregenerative scrubber, and boiler blowdown.

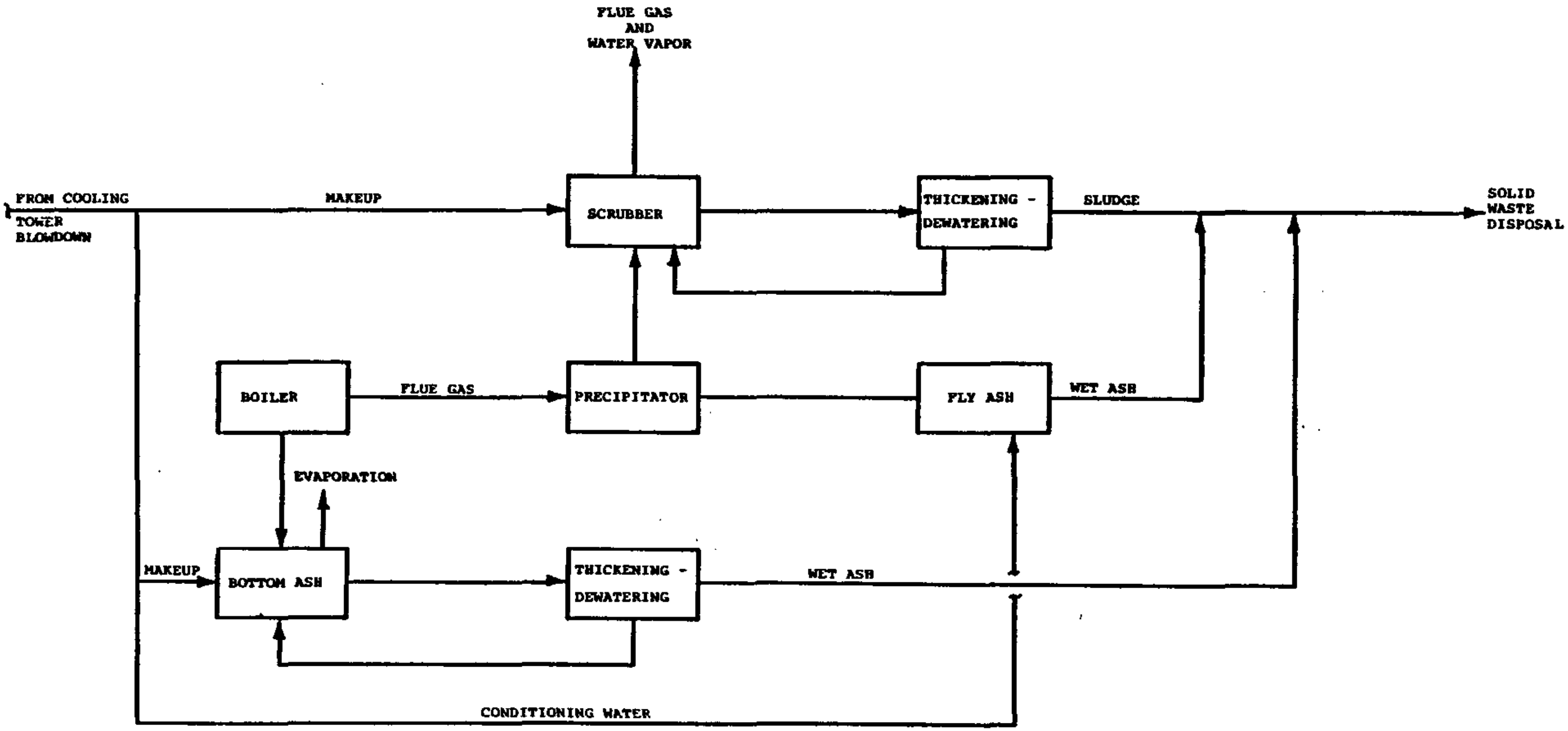
Consistent with the maximum reuse philosophy utilized with the coal conversion process facilities, waste stream recycle is thoroughly implemented resulting in a zero discharge system as indicated on Figure 6.1. The following discussions describe the liquid waste controls shown.

Scrubber solution, following reaction, is directed to conventional solids separation and sludge dewatering facilities. Recovered water will be returned to the scrubber system. Some water will be lost with flue gas emissions and with the sludge sent to disposal which would contain approximately 50 percent water by weight. Based on water-loss-estimating procedures contained in Ref. 64, it is projected that total makeup water requirements for boiler operation will be about 0.3 kg/kg coal.

The bottom ash quench and sluicing system will also be a recycle system. Water will be required for ash quenching, boiler seals, and for intermittently sluicing ash from the hoppers to solids separation and sludge thickening/dewatering facilities. Recovered water would be recycled. Water losses include quench evaporation, which is estimated to be 0.5 kg water/kg ash [Ref. 64], and moisture entrained in the dewatered ash which will contain about 25 percent (by weight) water.

Boiler blowdown, required to maintain dissolved and suspended solids concentration at levels required for reliable boiler operation, will be recycled to the boiler makeup water demineralizer reducing raw water requirements. It is conservatively estimated that boiler blowdown will be about 1 percent (by weight) of steam flow.

Miscellaneous waste streams include regeneration wastes from the boiler makeup water demineralizer and the condensate polisher. These streams, following neutralization, will be



B. C. HYDRO

FIG. 6.1

STEAM GENERATION
MAJOR WASTE STREAMS
WATER UTILIZATION

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stored in a wastewater holding pond and could be used as scrubber makeup water. Other waste streams, generally intermittant in nature, include boiler cleaning/equipment wash waters, localized area runoff, and floor/equipment drains. These streams, following oil separation where appropriate, will be collected in a sump. Following treatment in a sedimentation basin, the streams will be neutralized and combined with the neutralized water treatment regeneration wastes in a holding pond. Sludge collected in the sedimentation basin will be dewatered with the liquid effluent recycled back to the basin.

6.3.3 Solid Wastes

Solid wastes generated as a result of boiler operation include scrubber sludge, fly ash, bottom ash, and relatively small quantities of sedimentation basin sludge.

For the purpose of assessment, it has been assumed that the use of Hat Creek coal will result in a bottom ash/fly ash split of 25 percent/75 percent respectively. For the determination of solid waste quantities associated with ash collection, no allowance has been made for combustibles collected with the ash. Typical ash analyses for Hat Creek coal are presented on Tables 6.5 and 6.6. Examination of these tables show that the elements silicon, aluminum and iron comprise approximately 90 percent of the total ash. Calcium, magnesium, titanium, manganese, sodium, potassium, phosphorus and sulfur comprise an additional 6-8 percent. Based on the trace elements analyzed, they would comprise less than 1 percent of the total ash.

Fly ash will be conditioned with water for dust control during transport and to achieve higher compaction densities. The moistened fly ash will contain about 20 percent water. As previously stated, the dewatered bottom ash will contain about 25% water.

Table 6.5 Hat Creek Coal - Ash Analyses

Fly Ash Analyses	Percent	Bottom Ash Analyses	Percent
Silicon dioxide	55.09	Silicon dioxide	56.54
Aluminum oxide	26.48	Aluminum oxide	23.02
Iron oxide	6.60	Iron oxide	10.21
Calcium oxide	1.85	Calcium oxide	2.14
Magnesium oxide	1.40	Magnesium oxide	1.59
Titanium oxide	1.21	Titanium oxide	1.18
Manganese oxide	0.40	Manganese oxide	0.13
Molybdenum trioxide	-	Molybdenum trioxide	-
Chromium oxide	0.01	Chromium oxide	0.02
Nickel oxide	-	Nickel oxide	-
Vanadium pentoxide	-	Vanadium pentoxide	-
Sodium oxide	0.36	Sodium oxide	0.47
Potassium oxide	1.12	Potassium oxide	0.91
Phosphorus pentoxide	0.18	Phosphorus pentoxide	0.17
Sulfur trioxide	0.02	Sulfur trioxide	1.06
Others	5.28	Others	2.56
Total	100.00	Total	100.00

Source: [Ref. 69]

Table 6.6 Hat Creek Coal - Ash Trace Element Analyses

Trace Element	Mean Concentration (mg/kg)	
	Bottom Ash	Fly Ash
Hg	0.14	0.06
F	73	104
As	5.4	99
Se	13.2	19.1
Be	0.96	1.9
B	29	88
Cd	0.74	1.1
Cr	164	251
Cu	>678	>948
Pb	12.6	57
Mn	>854	>692
Mo	8.2	13.8
Ni	55	107
Sr	380	>503
U	9.6	16.5
V	>564	>703
Zn	58	>364

Source: [Ref. 69]

It is estimated that for wet non-regenerative scrubbers utilizing lime or limestone, solids generation will be approximately 5.5 kilograms of solid per kilogram of sulfur removed [Ref. 64]. Dewatering facilities will reduce the sludge moisture content to approximately 50 percent. It may be desirable to mix fly ash with scrubber sludge in a stabilization process prior to disposal in order to achieve increased landfill stability. All solid wastes will be disposed of in a contained solid waste disposal facility as discussed in Section 6.4.3 (iv).

6.4 Coal Conversion Processes

6.4.1 Liquid Wastes

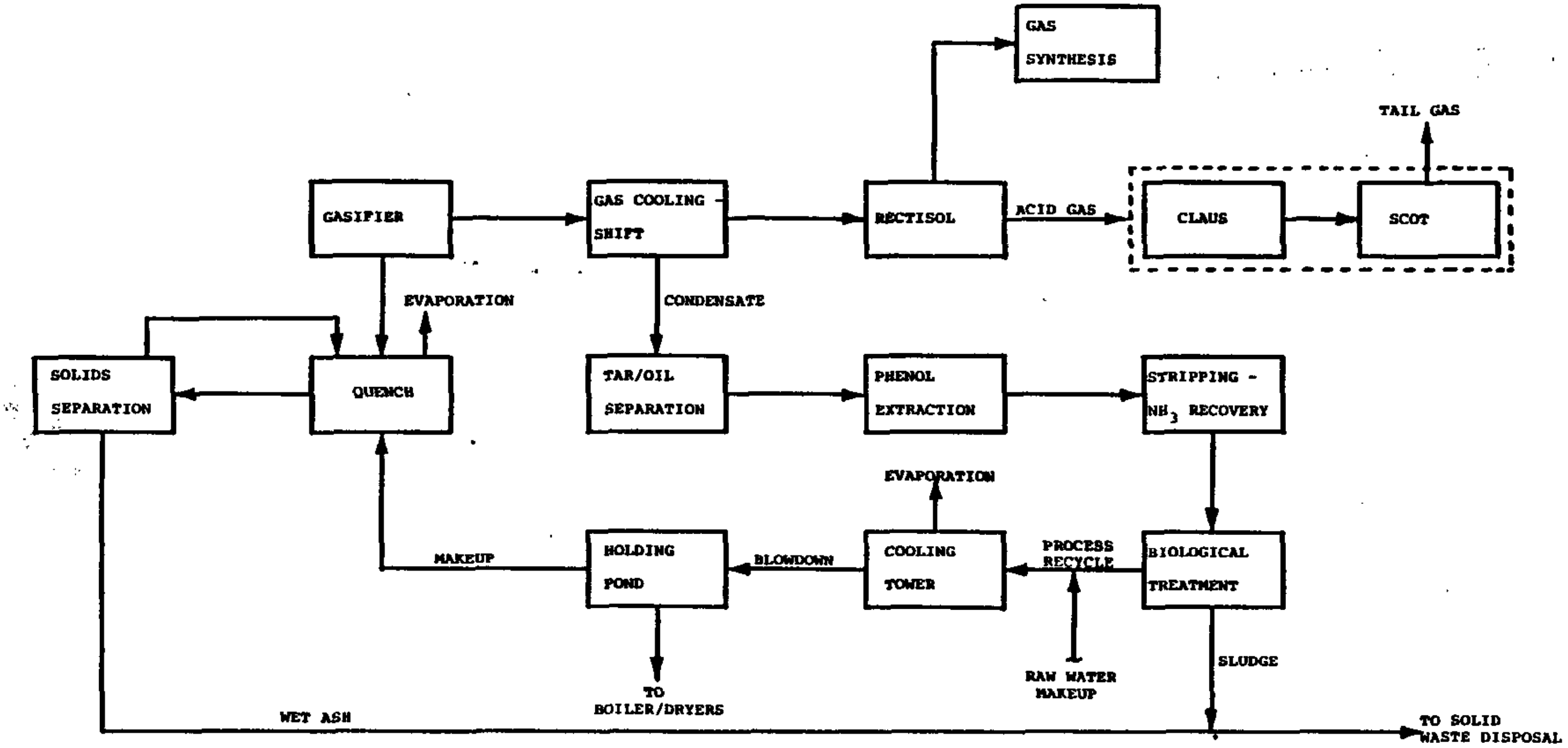
Because of the complexities associated with handling coal conversion wastewaters and anticipated licensing difficulties in permitting discharges, integrated wastewater treatment schemes which maximize reuse and minimize effluents have numerous advantages over "treatment for discharge" schemes. These advantages include reduced water treatment costs (it is more economical to treat for process recycle than to treat for discharge to the environment), smaller raw water requirements, and avoidance of licensing problems associated with meeting present and future environmental standards.

For the development of integrated treatment schemes, liquid waste streams should be combined, where possible, prior to treatment in accordance with their pollutant characteristics. For coal conversion waste waters, three general categories can be established consisting of "clean" streams (boiler blowdown, uncontaminated condensates); organic wastewaters (gas cooling and scrubbing liquids, foul condensates, oily runoff, and sanitary wastes); and low quality waters characterized by high solids' concentrations (quenches/slurries, brines, and sludges).

It is noted that several waste streams, with respect to composition, are basically conventional and would be found in all conversion processes or in a coal-fired power plant. These waste streams include raw water treatment sludges, boiler blowdown, runoff, and sanitary treatment wastes. Cooling tower blowdown is not included in the categorization since its composition is dependent upon the quality of makeup and the cycles of concentration associated with the cooling system.

In order to economically implement a zero-discharge system, it is imperative that major waste streams be integrated. For coal conversion plants, the major streams are process-originated wastewaters (foul condensates and gas cooling/scrubbing waters), cooling tower blowdown, and ash quench/slurry streams.

Integration of these streams requires that process-originated wastes be treated to a level suitable for use as cooling tower makeup, which is required to replace evaporation, drift and blowdown losses. Cooling tower blowdown could be used to replace ash or slag quench evaporation losses or for slurrying gasifier solid residuals, fly ash conditioning, scrubber makeup, and direct contact gas cooling. The ash/slag quench and slurry waters would be recycled following solids separation and dewatering. Conceptual diagrams showing primary system integration and principal systems for the three coal conversion categories are shown in Figures 6.2, 6.3 and 6.4.



B. C. HYDRO

FIG. 6.2

LOW TEMPERATURE GASIFICATION
SYSTEM INTEGRATION FOR
WATER REUSE

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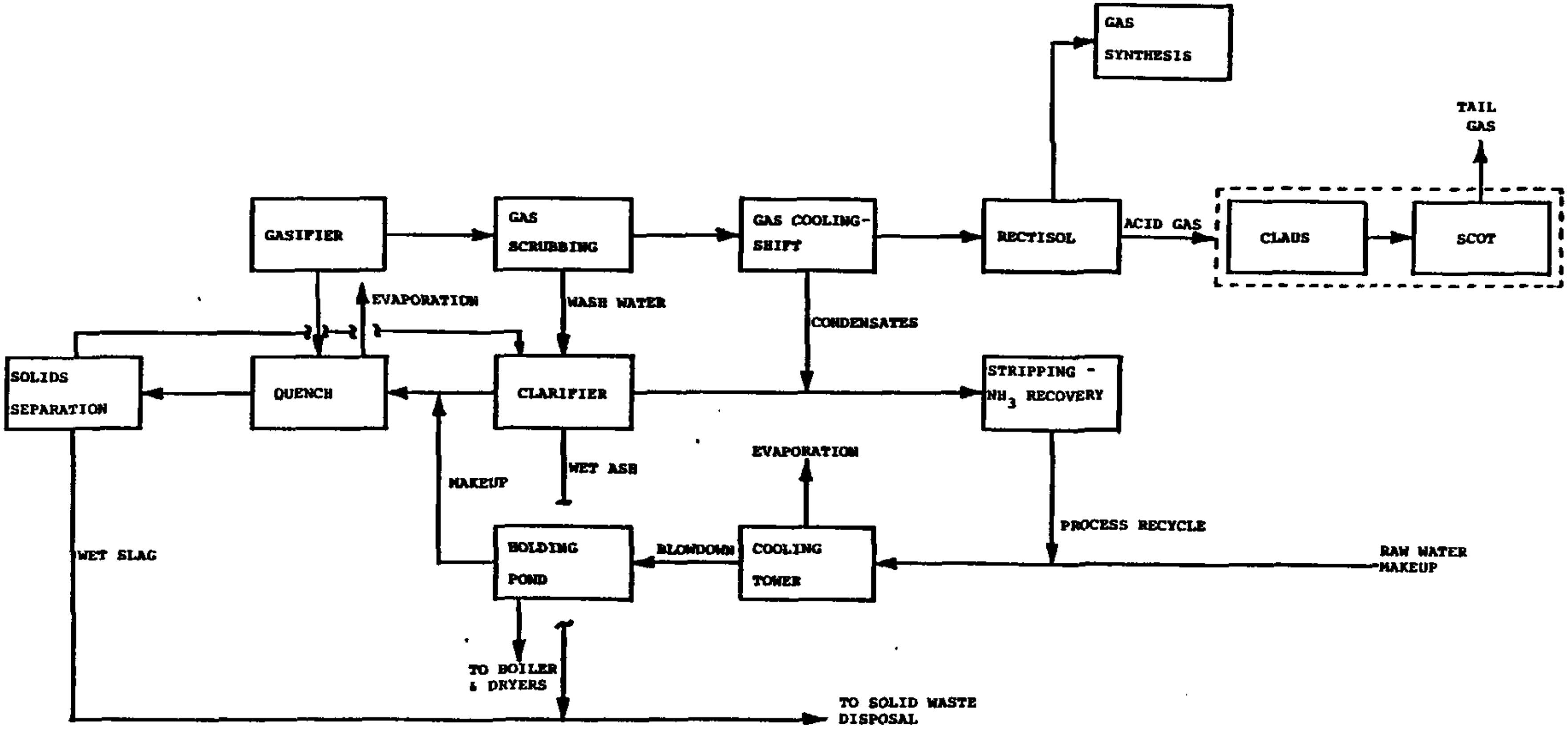

B. C. HYDRO

FIG. 6.3

HIGH TEMPERATURE GASIFICATION SYSTEM INTEGRATION FOR WATER REUSE

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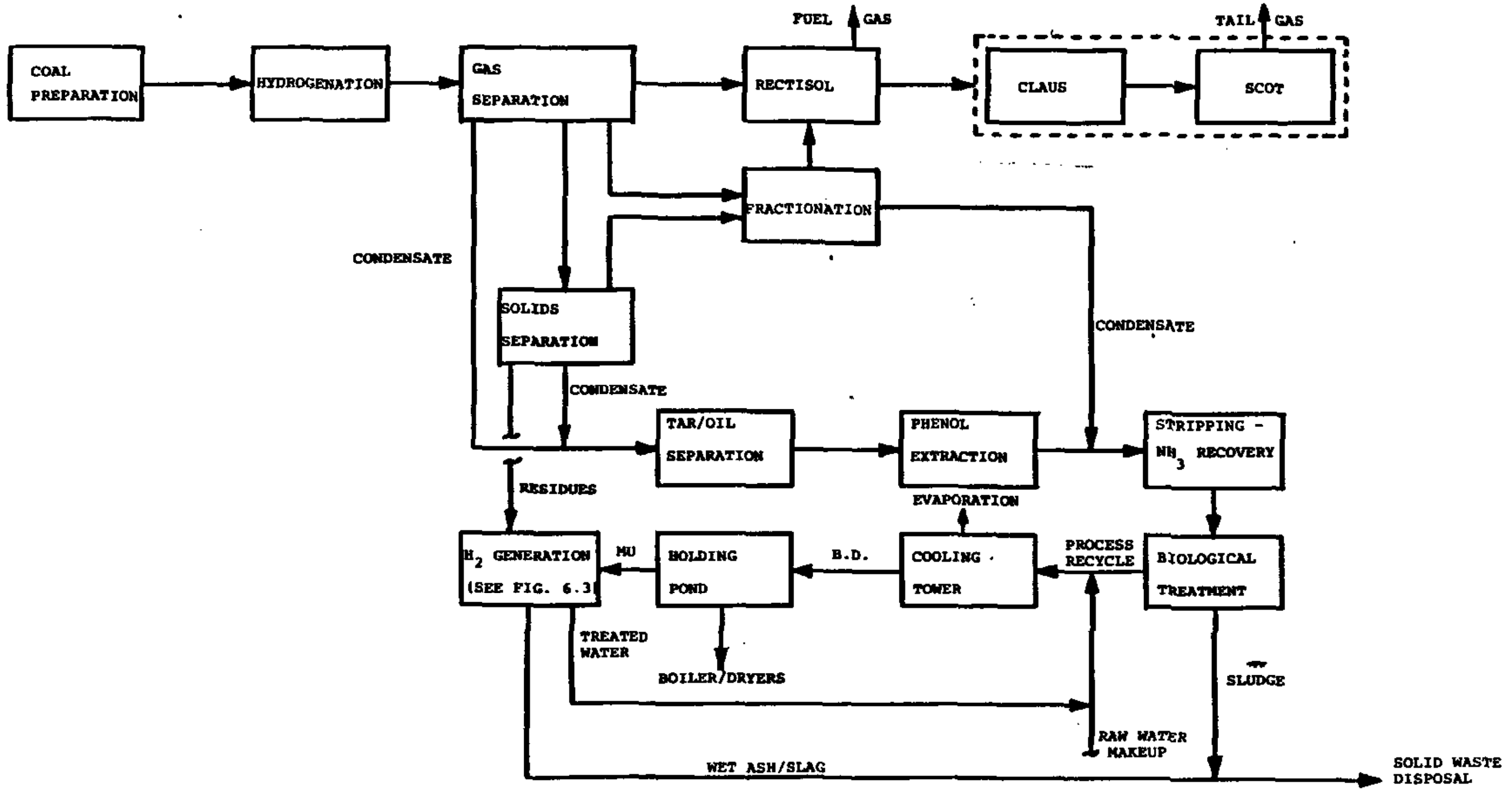

B. C. HYDRO

FIG. 6.4

DIRECT LIQUEFACTION
SYSTEM INTEGRATION
FOR WATER REUSE

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The following discussions identify, for each process category, major and minor waste streams. Stream integration and associated treatment requirements are also described and conceptual water balances associated with major stream reuse and recycle are discussed.

(i) Low Temperature Gasification (see Figure 6.2)

As previously stated, Low Temperature Gasification is characterized by the production of phenols, oils and tars. These contaminants are removed from the gas product stream during cooling and shift processes and are contained in the contaminated gas liquor stream. This stream is characterized by high BOD and COD concentrations due to high levels of dissolved and suspended organics. The inorganic components of the gas liquor consist primarily of ammonia and bicarbonate with smaller quantities of sulfur compounds, thiocyanate, cyanide, chlorides and trace elements. [Ref. 57, p. 119-121] Contaminated gas liquor is the major stream originating in the process and extensive treatment is required if this waste is to be reused as cooling tower makeup. Projected water quality characteristics are shown on Table 6.7.

Primary treatment of the contaminated gas liquor would consist of tar and oil separation with the treated effluent going to a phenol recovery system such as the Lurgi-proprietary Phenosolvan process. Extraction recoveries for coal gasification liquors of 99.5 percent for monohydric phenols, 60 percent for polyhydric phenols, and 15 percent for other organics are reported [Ref. 58, p. 255].

Table 6.7 Lurgi Process Water Analysis

Constituent	Location* 1	2	3
BOD	7200	2070	8
COD	13000	5220	400
TDS	1884	-	-
TSS	4676	125	5
Phenol	3100	410	1
Cyanide	8	5	0.6
Thiocyanate	260	260	5
Ammonia	13600	80	0.9
Sulfide	506	10	0.06
Oil	21000	500	1

*Location 1 - prior to oil separation/phenol extraction
Location 2 - prior to biological treatment
Location 3 - after biological treatment

All values in mg/l

Source: [Ref. 58]

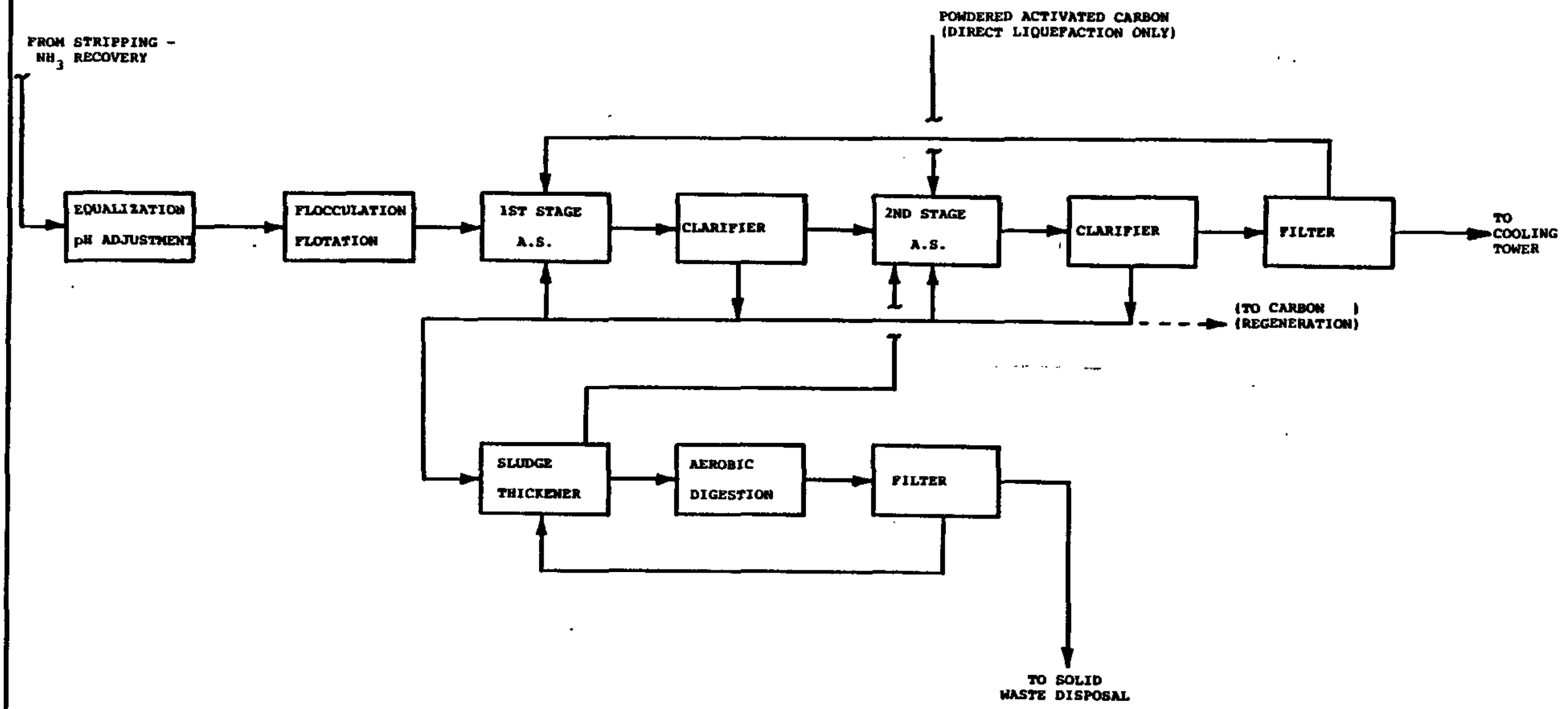
Following phenol extraction, the gas liquor would be steam stripped for the removal of dissolved gases; principally NH_3, H_2S, CO_2 and HCN. The ammonia would be recovered separately and the residual gases would be sent to sulfur recovery facilities. Because the liquor feed is high in CO_2, either the Lurgi-proprietary (Linz-Lurgi) or the Chevron WTT stripping process could be utilized. Characteristics of clean gas liquor are shown on Table 6.7. Residual levels as low as 50 mg/l and 5 mg/l for ammonia and hydrogen sulfide, respectively, are reported as achievable [Ref. 57, p. 177]. However, stripping must be controlled in order to leave sufficient nitrogen relative to biological treatment nutrient requirements. The combined extraction/stripping processes could result in about 90 percent removal of COD, 96 percent removal of total phenols, 95-98 percent removal of ammonia and 95 percent removal of hydrogen sulfide depending on system operation and optimization.

It is estimated, based on operational experience and considerations of coal moisture and steam requirements, that the quantity of clean gas liquor produced would approximate 0.85 kg/kg coal [Ref. 57, p. 119]. As indicated on Table 6.7, this stream still contains a high organic loading and would require biological oxidation prior to reuse as cooling tower makeup. Little data exists with respect to the biodegradability of clean gas liquors. Experience with refinery and coke plant wastewaters indicates that BOD/COD removal efficiencies of 65-95 percent are achieved, [Ref. 57, p. 180].

For conceptual purposes, a two-stage activated sludge treatment process has been assumed as shown on Figure 6.5. The biological sludges would be thickened and dewatered, stabilized by aerobic digestion, and filtered prior to transport to solid waste disposal facilities. The treated gas liquor would be used as makeup to the cooling system significantly reducing raw water requirements. Anticipated water quality following treatment is shown on Table 6.7. The total cost (1980 $) for treatment of the gas liquor waste stream from tar/oil separation through biological treatment, including sludge dewatering and stabilization, is estimated to be approximately $90 million.

Integration of the cooling system into the waste treatment scheme provides for system control and balance. Both the quality and quantity of cooling tower blowdown are controllable variables. Blowdown quantity can be reduced by increasing the cycles of concentration. If increased concentration factors result in scaling, precipitation, or materials' problems, additional treatment can be provided as required to alleviate such situations.

Additional treatment may also be required to prevent the buildup of trace elements and nonbiodegradeable organic constituents to levels considered toxic or hazardous. Additional treatment steps could include polishing of the biologically treated process waste waters with powdered activated carbon and raw water softening.



B. C. HYDRO

FIG. 6.5

CONCEPTUAL ACTIVATED SLUDGE BIOLOGICAL TREATMENT FACILITY

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Cooling tower blowdown would optimally be used to replace gasifier, boiler and dryer ash system water losses and for scrubber makeup. Steam generation system water requirements relative to ash handling and scrubber systems were previously discussed (Section 6.3.2). Coal dryer water utilization relative to ash handling and scrubber systems would be proportional. The quantity of water required to quench the gasifier hot ash is highly dependent upon gasifier operation, coal characteristics, and ultimately on the resulting ash properties. It is anticipated that approximately 0.5 kg water/kg ash will be evaporated during ash quench. The ash slurry would be dewatered to a moisture content of 25 percent. Moist ash would be sent to the solid waste disposal facility while extracted water would be recycled. Consequently, overall water loss associated with ash handling would approximate 0.83 kg/kg ash. A summary of water utilization including system water losses and treated wastewater available for process recycle is provided on Table 6.8.

Assuming a rejected heat duty of 158×10^6 MJ/d (1829 MW_t) for the Lurgi-Fischer-Tropsch maximum power alternative, the annual average cooling tower evaporation would be approximately 46,000 t/d (wet bulb = 13.9°C). It is believed that a cooling tower water balance which equates annual average blowdown to system losses would result in maximum concentration factors over 20. Consequently, depending on cooling tower makeup water quality resulting from the combination of treated wastewater with raw water makeup, a water balance may be required which would result in excess blowdown relative to ash handling losses and scrubber makeup.

Table 6.8 Water Utilization Summary*

	Koppers-Methanol	Lurgi-Fischer-Tropsch	Direct Liquefaction
Process recycle	38,660**	35,610**	22,510
Boiler losses			
Ash	300	2,960	300
Scrubber	900	7,070	900
Dryer losses			
Ash	210	90	160
Scrubber	620	280	490
Gasifier losses			
Ash	10,230	7,730	7,960
Total losses	12,260	18,130	9,810

*) All quantities in t/d

**) Does not include waste streams from gas synthesis

The excess blowdown could probably be used for process quench water or boiler makeup although tertiary treatment may be required for such uses.

Conventional waste streams include runoff and sanitary waste treatment sludges and effluents. Raw water supplied to the plant will already have been clarified so raw water treatment sludges and filter backwashes do not have to be considered. Runoff may require biological treatment, oil separation, or only suspended solids removal depending upon its origination and degree of contamination. Sanitary wastes have been assumed to be treated in separate facilities rather than combined with the gas liquor stream prior to biological treatment. Although there are some advantages to stream combination (e.g. provision of nutrients), adverse synergistic reactions may occur and consequently stream combination is not recommended at this time. Sanitary waste treatment effluents may be used for scrubber makeup while sanitary sludges could be disposed of in the solid waste disposal facility.

Little information is available regarding the quantity or quality of waste streams from synthesis processes. It is believed that the Fischer-Tropsch process would generate larger quantities of wastewater than either the Methanol or M-Gas processes. It is projected that approximately 5000 - 6000 t/d of wastewater would be produced, most of it originating in the Fischer-Tropsch synthesis section. Synthesis wastewaters would be routed to alcohol recovery facilities and then combined with smaller waste streams from hydrotreating, fractionation, and carbon monoxide removal facilities. The combined waste stream would be contaminated with 1 - 2 percent acetic and other organic acids. Small quantities of hydrocarbons may also be present. Definitive treatment requirements for this waste stream are not known. It is estimated that about 30 percent of this waste stream, following treatment, could be recycled back to the synthesis process while the remaining 70 percent could be used for cooling tower makeup [Ref. 65 and 67].

Miscellaneous process waste streams include Rectisol still bottoms and condensates from ash quench vent cooling and the SCOT plant. If the condensates are clean, they can be readily used as boiler makeup. Contaminated condensates and the Rectisol still bottoms would require combination with the gas liquor stream prior to phenol extraction or stripping.

(ii) High Temperature Gasification (see Figure 6.3)

Water treatment requirements for High Temperature Gasification processes are significantly reduced relative to Low Temperature Gasification processes due to the thermal cracking of organics in the gasifier. The high temperature generation of product gases results in negligible formation of heavy organics [Ref. 60, pp. 61-63]. Consequently, biological treatment of process-generated wastewaters would probably not be required prior to water reuse. It is possible that the Winkler gasifier, with a maximum reaction bed temperature of 1000°C, could produce small quantities of organics such as phenols and oil. Biological treatment may be required for these wastewaters.

The primary contaminants of process-originated wastewaters resulting from gas cleaning, cooling, and shift processes would be inorganic in nature and similar in identity to those inorganics associated with Low Temperature Gasification wastewaters (see Section 6.4.1 (i)). Major contaminants would include hydrogen sulfide, ammonia, hydrogen cyanide, thiocyanates, and sulfites.

The process-originated wastewaters will also contain relatively high concentrations of certain trace elements in comparison to Lurgi process wastewaters. High Temperature Gasifiers are characterized by a significant carryover of fly ash (50 - 75 percent of total ash) in the raw product gas from the gasifier due to the utilization of entrained or fluidized beds. The wash waters associated with gas scrubbing would subsequently contain concentrations of those trace elements such as antimony, arsenic, cadmium, lead, selenium, and thallium which tend to volatilize and condense on fine particulates. In a Lurgi-type gasifier, these elements would tend to appear in the ash quench waters rather than in wastewaters associated with gas cleaning and quench processes.

Clarifiers would be provided for the removal of suspended solids from wash waters and solids' separation recycle streams. Clarifier sludge, after dewatering to approximately 50 percent water (by weight), would be combined with dewatered slag (10 percent water by weight) and sent to solid waste disposal. Clarifier overflow would be used for gas scrubbing and for slag quenching and slurrying. Net clarifier overflow, which occurs as a result of condensation of water vapor entrained in the raw product gas, would be combined with contaminated cooling water and condensates from gas cooling and shift processes and sent to water treatment. It is estimated that the quantity of wastewater requiring treatment is approximately 0.90 kg/kg coal (as received). A water analysis for Koppers-Totzek waste water is presented in Table 6.9.

Process-originated wastewaters would require steam stripping for the removal of gaseous components prior to utilization as cooling tower makeup. Consistent with the stripping process utilized for Lurgi gasification, ammonia would be recovered separately and the residual gases sent to sulfur recovery facilities. The removal of ammonia is more critical, however, with respect to treatment of high temperature gasifier wastewaters since residual levels will not be reduced during biological oxidation. Ammonia fixation, for example as ammonium chloride, could result in high residual ammonia levels. Consequently, two stage stripping with lime treatment between stages may be required to ensure residual ammonia levels of less than 50 mg/l. It is anticipated that the utilization of lime treatment for pH adjustment prior to second stage stripping would also remove significant concentrations of the trace metals and other inorganic compounds which otherwise could lead to problems during subsequent water reuse. The anticipated water quality of the treated wastewater is provided in Table 6.9. The cost of water treatment is estimated to be approximately $30 million (1980).

Following treatment, the water would be used as cooling tower makeup as discussed in Section 6.4.1 (i). Cooling tower blowdown would also be utilized as discussed in that section. It is estimated that coal conversion processes of the magnitude contemplated, with steam generation facilities sized to meet on-site steam and emergency

Table 6.9 Koppers-Totzek Process Water Analysis

Constituent	1	2
COD	128	52
TDS	831	475
TSS	5081	5
Phenol	0.01	0.01
Cyanide	13	6
Oil	-	-
Ammonia	184	20
Sulfide	7	7
Hardness ($CaCO_3$)	630	80
Alkalinity ($CaCO_3$)	650	10

Location 1 - prior to stripping
Location 2 - after stripping/pH adjustment

All values in mg/l

Source: [Ref. 58]

power requirements, would have a cooling tower heat duty of approximately 105 x 10^6 MJ/d (1215 MW_t). Corresponding annual average and design point evaporation rates would be 30,000 and 40,000 t/d, respectively. High concentration factors would again probably dictate the generation of excess blowdown relative to ash handling and scrubber water losses.

A combined clarifier overflow/cooling tower blowdown stream would be used for slag quenching and transporting the slag to dewatering facilities. It is again estimated that evaporative losses would be about 0.5 kg/kg ash. The combined stream consisting of dewatered slag and ash would contain approximately 30 percent water (by weight). Liquid effluents associated with solids separation facilities would be recycled as required to the clarifier prior to reuse. Overall water loss associated with ash handling would be 0.93 kg/kg ash.

Conventional and miscellaneous process waste streams would be similar to those associated with low temperature gasification and would be handled as discussed previously (see Section 6.4.1 (i)). Water utilization is summarized in Table 6.8.

(iii) Direct Liquefaction (see Figure 6.4)

The use of solvent hydrogenation processes for direct liquefaction of coal and production of hydrocarbon liquids inevitably results in wastewaters which are significantly contaminated with a variety of organic and inorganic constituents. Although relatively smaller quantities of wastewater are produced than with gasification technologies, the variety of organic contaminants reported present appears to be significantly greater. Potential organic contaminants include phenols, significant amounts of acetic acid and smaller amounts of carboxylic acids; light hydrocarbons including benzene, toluene and xylene; naphthalene; fatty acids; cresols; alkylated phenols, naphthols, and benzene; tars; and oil. Inorganic contaminants are similar to those present in gasification waste streams and include sulfur compounds, nitrates, cyanide, thiocyanates, and chlorides. The trace elements present in the coal will be present in liquid waste streams in various forms. Porphyrins, carbonyls, metal alkyls and metal chelates may all form during the liquefaction process and ultimately appear in wastewaters. Dissolved gases requiring removal include hydrogen sulfide, ammonia, carbon dioxide, carbonyl sulfude and hydrogen cyanide [Ref. 61, pp. 245-257].

The major waste stream produced during the liquefaction of coal would be from the gas separation process. This stream would contain high concentrations of dissolved gases and hydrocarbons including phenols. The waste stream from liquids/solids separation would also contain quantities of phenols and other hydrocarbons. A sour condensate is also produced in the fractionation stage. This stream would have small quantities of dissolved gases and hydrocarbons, but little, if any, phenols. Table 6.10 shows the primary constituents and quantities associated with Liquefaction sour waters and condensates while Table 6.11 shows the composition of combined wastewaters prior to treatment.

In addition to the waste streams discussed above, significant quantities of wastewater would be generated in association with hydrogen generation. It has been assumed that a high temperature gasifier would be utilized and consequently, wastewater characteristics would be as discussed in Section 6.4.1 (ii).

Table 6.10 Direct Liquefaction Process Water Characteristics

Constituent	Waste Stream (mg/l) Gas Separation	Liquid/Solids Separation	Fractionation
CO_2	37,700	-	1,900
H_2S	55,600	-	900
NH_3	29,200	-	-
Hydrocarbons	5,000	5,300	6,600

Source: [Ref. 61]

Table 6.11 H-Coal Process Water Analysis

	Location*		
Constituent	1	2	3
BOD	52,700	9,100	40
COD	88,600	14,200	950
TDS	5,300	-	-
TSS	2	2	5
Phenol	6,800	410	1
Cyanide	10	7	0.05
Thiocyanate	350	350	35
Ammonia	14,400	45	1
Sulfide	29,300	10	0.06
Oil	608	50	1
Hardness ($CaCO_3$)	80	80	80
Alkalinity ($CaCO_3$)	80,000	700	700
pH	9.5	7.5	7.5

*Location 1 - prior to oil separation/phenol extraction
Location 2 - prior to biological treatment
Location 3 - after biological treatment

All values in mg/l

Source: [Ref. 58, p. 288, 365, 491]

The gas separation/liquids-solids separation waste streams would be combined with the fractionation condensates and sent to stripping facilities for the recovery of ammonia and removal of dissolved gases as previously described (see Section 6.4.1(ii)).If the liquefaction wastewaters are low in carbon dioxide concentrations, the United States Steel Phosam process would have to be used rather than the Chevron WTT, Linz-Lurgi, or two-stage steam stripping.

Table 6.11 shows anticipated waste stream water quality prior to biological treatment which would be required to reduce the high residual organic loading. It is estimated that approximately 0.35 kg water/kg coal would require biological treatment. The biological treatment of organically contaminated wastewaters was previously discussed in Section 6.4.1 (i). Because of the high organic loading, powdered activated carbon would be added to the second stage of the treatment to improve removal efficiencies (see Figure 6.5). The carbon could be recovered from the final clarifier, regenerated, and recycled. Recent experimental studies on H-coal wastewaters have shown that low COD, BOD and phenol residuals can be achieved [Ref. 58, pp. 316-317]. Table 6.11 shows anticipated water quality following biological treatment. The cost of liquefaction wastewater treatment is estimated to be about $65 million (1980).

Wastewaters from the high temperature gasifier used for hydrogen generation should be treated separately from the liquefaction process waste streams since they contain little, if any, organic contaminants. If the waste streams were combined, the hydraulic loading on the biological treatment plant would be significantly increased. Gasifier wastewaters would be treated as described in Section 6.4.1 (ii). The associated cost would be approximately $14 million (1980).

Following treatment, the gasifier liquids would be combined with the other treated wastewaters and used for cooling tower makeup. The use of treated wastewaters for cooling was previously discussed in Section 6.4.1 (i). Problems with the optimum utilization of cooling tower blowdown would be similar to those discussed in Sections 6.4.1 (i) and 6.4.1 (ii).

Conventional and miscellaneous waste streams would be similar to those associated with low temperature gasification except that gas synthesis waste streams would not be present. The handling of these waste streams would be as previously discussed (Section 6.1.1 (i)). Water Utilization is summarized on Table 6.8.

6.4.2 Gaseous Emissions

Pollution control requirements for gaseous emissions are relatively simple in comparison with water treatment requirements discussed in the following section. Commercially proven technologies can be readily utilized to reduce residual pollutant levels to criteria requirements. Table 6.12 provides a summary of particulate and sulfur emissions associated with major onsite sources, including the steam boilers (Section 6.3). NOx emissions have not been quantified since they are not addressed in the section dealing with applicable standards (Section 6.2). If required, control could be readily provided through boiler/burner design. Hydrocarbon emissions have also

Table 6.12 Summary of Particulate and Sulfur Emissions

	Koppers-Methanol	Lurgi-Fischer-Tropsch	Direct Liquefaction
Particulate Emissions (t/d)			
Boilers	2.0	14.0	2.0
Dryers	1.4	0.6	1.1
Total	3.4	14.6	3.1
Sulfur Emissions (t/d)			
Boilers	5.8	38.8	5.8
Dryers	4.0	1.9*	3.2*
SCOT	1.6	1.6	1.1
Total	11.4	42.3	10.1

*Sulfur recovery may not be required if emissions are less than 5 tonnes sulfur/day provided ambient air quality guidelines are not exceeded [Ref. 72, Section 2.2.2].

not been quantified since they would be primarily associated with leaks and spills. Control would be as discussed in the "Fugitive Emissions" portion of this section.

With respect to volume and concentration of potential air pollutants, the waste stream of most significance associated with the coal conversion technologies addressed would be the concentrated acid gases from the acid gas removal facilities. The concentrated acid gas waste streams would be sent to sulfur recovery facilities. Sulfur recovery tail gas would require additional treatment prior to release to the atmosphere. For indirect liquefaction processes, the raw product gas would require purification prior to synthesis. For direct liquefaction technologies, product gas from the hydrogen generation system would require clean-up prior to process utilization. In addition, gaseous waste streams from gas separation and fractionation would require treatment for recovery of heating values and sulfur recovery prior to release. Pollutants of major concern in the concentrated acid gases would be reduced sulfur compounds, hydrogen cyanide, and hydrocarbons. Essentially all the sulfur contained in the coal feed would appear in the concentrated acid gases. Allowing for some sulfur in ash/slag, byproduct tars, and liquid products, approximately 90 - 95 percent of the feed coal sulfur would be contained in the acid gas removal facilities' waste streams.

It has been assumed that the Rectisol process would be used for acid gas removal although a number of process options are available. All such processes produce a product gas relatively free of sulfur compounds and carbon dioxide. The Rectisol process produces a number of gaseous waste streams, all of which would be fed to a sulfur recovery plant except for a rich carbon dioxide stream which also contains most of the wasted hydrocarbons. This stream would require incineration prior to release if combination with the product gas is not feasible.

There are a large number of process options available for sulfur recovery and tail gas cleanup. For purposes of analysis, it has been assumed that a Claus-SCOT treatment train will be utilized and it is anticipated that an overall sulfur removal efficiency of greater than 99 percent can be achieved. It is estimated that the cost of these facilities would be approximately $15 million (1980). The Claus plant will remove approximately 90 - 97 percent of the feed gas sulfur with higher removals achieved with higher feed concentrations. Lean feed streams would also require the provision of supplemental fuel since reaction heat would be insufficient to maintain process temperature. Concentration processes such as Shell ADIP could be used to provide the Claus plant with a rich hydrogen sulfide stream relatively free of hydrocarbons. If the feed gas contains over 3 percent ammonia, scrubbing may be required for ammonia removal, especially if the feed gas also contains high levels (greater than 30 percent) of carbon dioxide.

The SCOT process will convert all residual sulfur species in the Claus tail gas to H_2S. A rich H_2S stream is recycled back to the Claus plant. Emissions from the SCOT plant should contain less than 300 ppm (v) of sulfur. Residual levels of 100 ppm (v) are reported achievable [Ref. 59]. A typical analysis of SCOT tail gas is shown on Table 6.13.

Gaseous waste streams sent to the Claus-SCOT treatment system in addition to those from acid gas removal include the stripped gases from the ammonia recovery facilities associated with water treatment. In the Lurgi process, expansion gases are

Table 6.13 Typical SCOT Tail-Gas Analysis

Component	Volume (percent)
H_2S	.03
SO_2	-
S	-
COS	10 ppm
CS_2	1 ppm
CO	-
CO_2	3.05
H_2	0.96
H_2O	7.0
N_2	89.0

Source: [Ref. 76]

generated in the gas liquor separator which contain sufficient quantities of H_2S to warrant Claus plant treatment.

Miscellanous waste streams associated with all coal conversion processes are described in the following discussions:

(i) Fugitive Dust

Fugitive dust would occur as a result of coal preparation including thermal drying and handling. Conventional controls such as water sprays at dump hoppers, transfer points, screens and crushers and the use of exhausted ducts, dust collectors, and fabric filters should maintain fugitive dust emissions at acceptable levels.

(ii) Nitrogen

Nitrogen from the air liquefaction plant is available for use as transport gas for dry pulverized coal and for process purging requirements. Excess nitrogen from the oxygen plant should be vented from an elevated stack to enhance dispersion.

(iii) Coal Lock Hopper Vents

The use of raw product gas for lock hopper pressurization and recovery via compression is recommended. Nitrogen could be used to displace residual gases prior to recharging the hopper after a feed cycle. The residual gases purged by nitrogen would probably require incineration.

(iv) Ash Lock Hopper/Quench Vents

These vents are comprised mostly of steam contaminated with particulates and volatile components of the wastewater used for quenching. The steam could be sent through a wet cyclone for particulates removal and then condensed. Residual noncondensables should be incinerated.

(v) Coal Dryer Flue Gases

Coal drying is required for most conversion processes investigated as discussed in Section 2. Thermal drying, using Performance Blend coal, is proposed. Since drying is accompanied by contacting the feed coal with combustion gases, dryer emissions will contain particulates, volatile compounds, and conventional combustion products. Because of the high moisture content of the coal and the large quantity of coal being dried for most processes, dryer emissions could be environmentally significant. It is probable that pollution-control requirements similar to those associated with steam generation will be necessary (see Section 6.3).

(vi) Preheater Flue Gases

Both direct and indirect liquefaction processes require a number of furnaces or heaters which generate combustion flue gases. If byproduct fuel gas, treated for acid gas removal, is utilized to supply process heat requirements, airborne pollutants associated with flue gas emissions should be minimal.

(vii) Fugitive Emissions

Because of potential hazards associated with coal conversion products (see Section 6.5), the most difficult gaseous emission problem may be control of fugitive releases associated with product production, handling and storage. Escapes could be minimized by the adequate design and maintenance of valve stems, pump packing or mechanical seals, flanges/gaskets, relief valves, instrument/piping connections and compressor seals. Specific control methods include the use of positive pressure sealing systems on rotating equipment and closed relief valve systems.

Storage tank emissions would be controlled through implementation of the requirements dictated by Provincial standards (Section 6.2.1).

6.4.3 Solid Wastes

(i) Low Temperature Gasification

The major solid waste produced as a result of gasifier operation would be ash removed from the bottom of the gasifier through lockhoppers. The ash would be quenched; slurried to solids separation, thickening and dewatering facilities; and subsequently sent to a contained landfill for disposal with a moisture content of approximately 25 percent by weight.

In a Lurgi gasifier, essentially all ash contained in the feed coal exits from the bottom of the reactor. Very little ash is carried over with the raw product gas to the gas cooling and shift facilities. It has been subsequently assumed that the quantity of ash requiring disposal is equivalent to the ash contained in the feed coal.

The ash would contain most of the inorganic content of the coal including trace elements not volatilized during gasification. It is believed that the ash constituency would not vary significantly from that generated during conventional coal combustion. The characteristics of Hat Creek coal ash is presented in Section 6.3.3.

In addition to ash, other solid wastes include organic sludges from biological and sanitary waste treatment, and byproduct storage; inorganic sludges generated as a result of runoff treatment; evaporator residues; and spent catalysts and filter media.

The organic sludges associated with biological treatment and byproduct storage may be hazardous in nature. These wastes could be disposed of with the ash in a secure landfill. Incineration may be preferable due to potential odor problems. Sanitary waste treatment sludges and inorganic sludges could also be landfilled or incinerated.

If metals' recovery is not viable, the spent catalysts will require handling as a solid waste. The spent catalysts, as well as filter media and evaporator residues, could be very hazardous and require special handling and disposal. Chemical fixation and/or containerization may be necessary before landfill disposal.

(ii) High Temperature Gasification

Ash (or slag) would be the major solid waste produced as a result of gasifier operations. As discussed in Section 6.4.1 (ii), High Temperature gasifiers are characterized by an entrainment in the raw product gas exiting the gasifiers of 50 - 75 percent of the ash contained in the coal fed to the gasifier. The remaining ash would be discharged from the bottom of the gasifier as a molten slag.

The slag, following quenching, would be granular in nature and less susceptible to leaching than Lurgi ash or fly ash. Because of the slag's granular character, it would be relatively easy to dewater. Following solids separation, the slag solid waste would contain only

5 - 10 percent water. The slag produced by Koppers-Totzek and Texaco gasifiers would be inorganic in constituency and would resemble ash produced during conventional coal combustion (see Section 6.3.3). Winkler bottom ash would contain relatively high concentration of unburned carbon. Unless this ash is fed to a fluidized bed combustor for recovery of the residual energy values, this material would constitute a solid waste.

Fly ash would be removed from the raw product gas through utilization of wet scrubbers, cyclones and/or electrostatic precipitators. The fly ash collected in association with Winkler and Texaco processes has sufficient carbon to warrant recovery. The Winkler fly ash could be dewatered as required and combined with the slag for use as fuel while Texaco fly ash would be recycled back to the coal slurry feed. Koppers-Totzek fly ash would have little, if any, organic content. Washwaters would be clarified and the resulting sludge would be filtered to a solids content of approximately 50 percent. The filtered sludge would be combined with the slag and sent to the solid waste disposal facility for burial.

In summary, all the ash contained in the coal fed to any of the high temperature gasifiers considered, including the high-carbon Winkler residues, would ultimately end up as a solid waste requiring disposal.

Miscellaneous solid wastes associated with High Temperature gasification processes would be similar to those identified in Section 6.4.3 (i) except that biological treatment sludges would not be present.

(iii) Direct Liquefaction

Since the char and heavy oil residuals produced during the liquefaction process will be utilized for gasifier fuel, they will not constitute a solid waste. However, essentially all ash in the coal feed will become concentrated in these byproducts and will ultimately end up as gasifier bottom ash or fly ash. The collection and handling of these waste products was discussed in Section 6.4.3 (ii).

Miscellaneous solid wastes would be similar to those discussed in Section 6.4.3 (i) and would be handled accordingly.

(iv) Solid Waste Disposal Facility

As discussed in the preceding section, coal conversion processes generate large quantities of solid wastes of varying characteristics. It is estimated that the total quantity of solid wastes produced would be approximately 4.9 - 7.5 million t/year (wet). As indicated on Table 6.14, the majority of these wastes would be ash and/or slag. The only viable method of disposing of the projected quantities of solid wastes would be in a landfill facility such as that proposed in association with the 2000 MW conventional coal-fired power generation facility [Ref.102] which is designed to handle up to 4 million t/year of ash over the 35-year life of the facility. The cost of facility development, including transfer equipment, has been estimated to be $50-60 million (1980).

The primary environmental concern associated with landfill operation would be ground water and surface water contamination caused by runoff and leachate. As discussed in the preceding sections, it is anticipated that the composition of the gasifier ash or slag would resemble ash produced by the conventional burning of coal. The projected quality of leachate for various trace

Table 6.14 Solid Waste Summary

Solid Waste (t/d-dry)	Koppers-Methanol	Lurgi-Fischer-Tropsch	Direct Liquefaction
Gasifier Ash	11,000	9,310	8,560
Boiler/Dryer Ash	1,300	7,730	1,190
Scrubber Sludge	50	250	50
Total Water	6,510	5,440	5,130
Total	18,860	22,730	14,930

elements from a 75/25 percent combination of Hat Creek coal fly ash/bottom ash is shown on Table 6.15. It is believed that a waste stream of this quality would not be considered hazardous by regulatory agencies. However, the use of treated process waters for ash slurrying, which may contain concentrations of the volatilized trace elements as well as organic contaminants, could significantly change ash leachate characteristics. These waters may contain residual concentrations of organic substances, such as phenols, fatty acids, and polycyclic aromatics. In addition, concentrations of the volatilized trace elements and inorganic ions,such as SCN^-, CN^- and $S^=$,may also be present. The presence of some of these substances may enhance the solubility of certain metals contained in the ash while the alkaline ash may absorb or cause some of the quench water contaminants to precipitate out of solution [Ref. 57, Page 119].

Although it is impossible to project leachate quality at this time, it is probably safe to assume that regulatory agencies will consider the leachate as potentially hazardous and will require contained disposal with associated leachate/runoff collection systems and groundwater monitoring. The inclusion of organic sludges and spent catalysts would also contribute to the suspect nature of the solid wastes requiring disposal.

6.5 Toxic and Hazardous Considerations

Conventional pollutants produced during coal conversion such as sulfur dioxide, nitrogen oxide, carbon monoxide, hydrogen sulfide, particulates, and gross hydrocarbons, acting as individual agents or as a result of synergistic or environmental reactions, have the potential to cause acute respiratory

Table 6.15 Projected Combined Ash Leachate Quality for Trace Elements*

Trace Element	Concentration (mg/l)
Arsenic	0.6 - 2.4
Boron	3.0 - 3.6
Cadmium	0.10
Chromium	0.12 - 0.20
Copper	0.23 - 0.33
Fluoride	3.3 - 4.9
Lead	0.05
Mercury	0.0013 - 0.0023
Vanadium	0.18 - 0.22
Zinc	0.82 - 2.5

*Based on fly ash to bottom ash ratio of 75/25, conditioned and wetted with recycled powerplant waste waters to 20 percent and 40 percent moisture respectively.

Source: [Ref. 68]

illness or chronic lung and skin diseases. Control technologies are commercially available to limit the emissions of these pollutants at levels for which adverse health effects are not anticipated. However, waste streams and coal conversion products will potentially contain unknown concentrations of complex organic constituents as well as trace metals, a number of which are known to be carcinogenic or otherwise hazardous. The following sections address concerns associated with the presence of these pollutants in coal conversion products and waste streams.

6.5.1 Waste Treatment Concerns

(i) Trace Elements

An unknown factor regarding waste treatment in coal conversion facilities is the ultimate fate of trace elements. Some trace elements, either through bioaccumulation or on a direct dose/response basis, are known to have the potential to cause both chronic and acute toxic effects.

During direct coal liquefaction, trace elements would tend to concentrate in the residues with less than 1 ppm of any element distributed in the coal products [Ref. 61, pp. 297-299]. Since it is proposed to gasify the residues, the ultimate fate of trace elements in either direct or indirect liquefaction technologies hinges on gasification reactions. Trace elements tend to partition and concentrate themselves during combustion relative to their volatility. Nonvolatile elements would tend to concentrate in the ash or slag. Volatile elements would tend to concentrate in the raw product gas where a secondary partitioning would occur since most of the volatilized elements would condense or adsorb on fly ash particles as the gas cools.

An analysis of trace element redistribution associated with the conventional combustion of Hat Creek coal identified 23 elements of environmental concern [Ref. 68]. The primary criteria for selection included presence in Hat Creek coal and potential toxicity. Table 6.16 lists the identified elements and their combustion partitioning characteristics. It is believed that the partitioning characteristics shown on the table would be representative of initial trace element distribution during gasification.

It should be recognized that the trace element characteristics shown on Table 6.16 represent general trends and should not be interpreted as absolutes. For example, arsenic may be found in bottom ash and in the cleaned product gas as well as being a fly ash constitutent. The use of water to quench ash, slag, and raw product gas; water reuse and system interactions; metals' solubility; and the removal efficiency of water treatment systems all contribute to the unknown fate of trace elements.

Trace elements would primarily be returned to the environment as components of ash/slag and biological treatment sludge which are proposed to be landfilled; with stack emissions associated with coal combustion in the plant boilers and thermal dryers; or with cooling tower drift due to the utilization of treated water for make-up and the trace metals contained in the raw water. The previously referenced report addressing trace element redistribution associated with the utilization of Hat Creek coal in a 42,000 t/d conventional coal-fired power plant concluded that no significant environmental impacts would occur [Ref. 68]. The environmental

Table 6.16 Trace Element Partitioning Characteristics

		Volatile	
Element	Nonvolatile	Condense/Adsorb	Remain Volatile
Silver (Ag)	X		
Arsenic (As)		X	X
Boron (B)	X	X	
Beryllium (Be)	X	X	X
Cadmium (Cd)		X	X
Cobalt (Co)	X	X	
Chromium (Cr)	X	X	
Copper (Cu)	X	X	X
Fluorine (F)		X	X
Mercury (Hg)			X
Manganese (Mn)	X		
Molybdenum (Mo)		X	
Nickel (Ni)		X	X
Lead (Pb)		X	X
Antimony (Sb)		X	
Selenium (Se)		X	X
Tin (Sn)	X		
Thorium (Th)	X		
Thallium (Tl)	X	X	X
Uranium (U)	X		
Vanadium (V)	X	X	
Tungsten (W)	X		
Zinc (Zn)	X	X	

Source: [Ref. 68]

pathways associated with conventional coal combustion are essentially identical to those of coal conversion although the relative distribution of metals with respect to the pathways would be different. Conventional coal plants would have relatively larger stack emissions while coal conversion facilities would have larger cooling tower emissions. Although it is anticipated that there would be no major adverse environmental impacts associated with trace element releases from a coal conversion facility, insufficient data exists to allow a definitive conclusion at this time.

(ii) Water Treatment Efficiency

Low temperature gasification and direct liquefaction facilities produce waste waters which are contaminated with organics as previously discussed in this report. Biological treatment would be required to reduce the organic loading to acceptable levels and it is believed that some metals removal would also be achieved. Most of the organic compounds would be phenols which, at the inlet concentrations anticipated, would be readily biodegradable. The residual phenols would probably be complex polyhydric or polyaromatic compounds which resist degradation. Other organic compounds which resist degradation and are potentially present would include aromatics such as benzene, toluene, and xylene as well as related sulfur/nitrogen substituted compounds. It is likely that polynuclear aromatic hydrocarbons would also be present in small concentrations. Many of the complex compounds which would potentially be present in the treated wastewaters are known to be carcinogenic or otherwise toxic to various body organs and systems. [Ref. 61, pp. 243 <u>et seq</u>; Ref. 74, pp. 2-32 to 2-35].

Since there is a general lack of data regarding the biotreatability of coal conversion wastewaters relative to complex organics and trace metals, as well as a lack of specific health effects data, an assessment of water treatment adequacy relative to these contaminants is not possible. The primary environmental pathways associated with residual pollutants in the treated wastewaters would be via cooling tower drift or as part of the moisture component in ash/slag or biological sludge. Drift emissions from cooling towers would be widely dispersed and, because of atmospheric dilution, it is anticipated that no significant impacts would occur. The lack of data regarding residual pollutant levels and dose/response information for many of the contaminants precludes definitive determinations at this time. Definitive analyses would be complicated by the reuse of water in evaporative systems such as the cooling water system and the ash quench system. With the implementation of the zero discharge concept, system blowdown would consist primarily of water removed from the site with solid wastes. It is possible that nonvolatile contaminants could build up to unsafe levels and supplemental blowdown, most likely from the ash quench systems, would be required. Pollutants associated with the landfilled solid waste would be controlled through containment. Leachate would be collected and disposed of as discussed in the following section.

(iii) Ultimate Disposal of Wastewater

With the use of zero-discharge water treatment systems, it is probable that some form of ultimate wastewater disposal will be required. Potential streams requiring such disposal would include leachate collected from the solid waste disposal area and the supplemental blowdown streams identified in the prior section. Forced evaporation is the most likely treatment system although natural evaporation may be viable in the proposed site area. Other ultimate disposal technologies such as ion exchange, reverse osmosis, and electrodialysis are subject to fouling by organics and may not be practical considering anticipated wastewater characteristics from most of the coal conversion processes investigated. It is estimated that the cost of forced evaporation, conservatively assuming a waste flow of 3000 t/d, would be about $5 - 6 million (1980).

Forced evaporation of wastewaters would result in a flash gas, waste brine, and product water suitable for reuse. The flash gas may require treatment for ammonia or sulfur removal prior to release to the atmosphere. The waste brine would require dewatering/drying prior to burial in the solid waste disposal facility.

6.5.2 Coal Conversion Products and Byproducts

Products and byproducts from coal conversion facilities contain numerous substances known to be toxic or otherwise present hazards for occupational or public exposure. A hazard potential assessment of 216 specific substances supected or known to be present in product streams resulted in 37 judged to be hazardous, 24 - very hazardous and 15 - most hazardous [Ref. 61, p. 19]. It is estimated that this assessment addressed less than 10 percent of the possible compounds in liquefaction products [Ref. 70, p. 510]. It is estimated that coal liquefaction products contain over 70 percent (by weight) aromatic and heterocyclic hydrocarbons. Some of those compounds, which would include polynuclear aromatic hydrocarbons, are known to be carcinogenic or mutagenic.

Therefore, leaks, spills, handling, transportation and storage will require special attention during all phases of a facility's life from conceptual design through decommissioning. It is possible that toxic and hazardous considerations may restrict product utilization although the extent of such restrictions are unknown at this time.

6.6 Relative Impact Comparison between Coal Conversion and Power Generation Facilities

The local environmental significance of a coal conversion complex or power generating station can be investigated on a relative basis through comparison of various parameters associated with facility construction and operational characteristics. It must be recognized, however, that such comparisons are only indicative of the potential for impact. More detailed analyses beyond the scope of this report would be required to identify and subsequently compare the absolute effects of facility construction and operation on the local environment. Table 6.17 summarizes important impact indicators which are presented elsewhere in this report for selected representative technologies and provides a tabular comparison with similar data associated with a 2000 MW coal-fired power generating station, as described in Ref.102.

Examination of Table 6.17 indicates that while solid waste production is proportionate to coal feed rates due to ash being the major component, atmospheric emissions are not. With the use of acid gas and tail gas treatment systems and the control of fugitive emissions, the atmospheric emission of pollutants associated with either coal conversion or power generation is primarily related quantitatively to the combustion of coal. It is inherently obvious that coal-fired power generation facilities will consequently have larger emission levels of combustion products than conversion facilities with equivalent total coal feed rates. Correspondingly, the indicated Lurgi-Fischer Tropsch alternative,which utilizes coal fines for the production of export power, has higher emission levels than the other two conversion alternatives which have boilers sized to meet only onsite steam and power requirements.

It is recognized that the utilization of liquid products produced by coal conversion facilities will probably result in additional pollutant emissions to the atmosphere. Because of the potentially hazardous nature of coal liquids (see Section 6.5.2), such emissions could be environmentally significant. However, the variety of possible products and end-use diversity precludes detailed assessment.

Water consumption is primarily dictated by cooling water system losses in facilities designed for maximum reuse of waste water. Power generation facilities consequently require much more water than conversion facilities having similar total coal feed rates since so much low-level waste heat is rejected by the thermal production of electricity. The Lurgi-Fischer Tropsch maximum power alternative uses more water than the other two conversion alternatives due to large quantities of coal being utilized to generate electricity.

Coal conversion technologies, because of their complex petrochemical nature, require the preemption of much more land for the construction of onsite facilities. It is estimated that approximately 200-350 ha would be required for siting a coal conversion unit of the size contemplated [Ref.1], while only about 100 ha would be required for the power generation facilities [Ref.102]. Area requirements do not include allowances for coal preparation or solid waste disposal facilities.

Facility complexity would also be reflected in construction work-force requirements. The peak work-force requirement for power plant construction is about 3,150 [Ref. 76], while it is estimated that two to three times that number of workers would be required to construct a conversion facility of equivalent size. Since the work force required for power plant construction exceeds the required supply of unionized construction workers [Ref. 102],it can be projected that each additional worker required for construction of the conversion facility would probably be an inmigrating worker. Secondary employment induced by increased work-force inmigration would probably also be supplied through inmigration. The incremental socioeconomic impacts associated with an influx of thousands of people into the Hat Creek Valley region beyond that required for power plant construction would be significant but impossible to assess without detailed study. "Boom-bust" cycle effects associated with coal conversion facility construction could be greatly magnified relative to power plant construction since the permanent operating staff for both types of facilities are projected to be about equal (around 1000 personnel).

Table 6.17 Comparison of Coal Conversion and Power Generation Facilites' Operating Parameters.

Parameter (t/d)	Koppers-Methanol	Lurgi-Fischer Tropsch (Max Power)	Direct Liquefaction	Power Generation
Coal Requirement	48,030	66,395	38,095	40,500
Water Requirement	34,600	51,600	28,800	100,200
Particulate Emission	3.4	14.6	3.1	17
Sulphur Emission (as S)	11.4	42.3	10.1	75
Solid Wastes (dry)	12,350	17,290	9,800	10,830

7. CONCLUSIONS

7.1 This study finds that the conversion of Hat Creek Coal to light liquid fuels -

(i) is best suited to future energy and petrochemical market needs in British Columbia.

(ii) is technically viable by several alternative process routes.

(iii) is economically viable within the Terms of Reference and economic/financial criteria imposed on the Study.

This utilization offers a technically and economically viable alternative to the use of the coal for thermal electric power generation.

N.B. It should be noted, however, that the provision of a thermal electic power generation plant and/or a coal conversion plant are not mutually exclusive concepts for utilization of the huge coal deposits at Hat Creek. It is considered feasible, given suitably expanded mining plans, to produce sufficient coal to supply both such facilities simultaneously. And while these may be best operated under separate managements there could, by suitable integration of services, be synergistic effects tending to improve the overall utility and economics of the complex. In this context it may be noted that some of the alternative processes considered in this report require local generation of about 700 MWe of electric power.

7.2 The most attractive of these alternate uses and the economic limits within which they are viable, are illustrated in Figures 5.1 - 5.9. These processes, ranked in decending order are -

- (i) H-Coal Process
- (ii) Lurgi/Methanol Synthesis Process
- (iii) Mobil Methanol-to-Gasoline Process
- (iv) Lurgi/Fischer Tropsch Process.

However, the H-Coal Process has not yet been commercially demonstrated to the same degree as the Methanol and Fisher-Tropsch Processes, and its heavier oil products are not as readily marketable.

7.3 The limited market in Western Canada for heavy fuel oils renders the Solvent Refined Coal processes unattractive.

7.4 The conversion of Hat Creek coal to Synthetic Natural Gas (SNG) is not economically viable at current gas export prices.

7.5 There appear to be no opportunities for usefully converting Hat Creek coal to upgraded solid products.

7.6 The low yields of liquid products on pyrolysis of Hat Creek coal preclude its use for combined pyrolysis/thermal electric power generation applications.

7.7 The processes for conversion of Hat Creek coal to light liquid fuels listed in 7.2 can be operated within current environmental control regulations and this study has included the best available control technology procedures, with attendant costs.

7.8 There have been no developments in coal cleaning techniques during the past five years which are encouraging for the cleaning and beneficiation of Hat Creek coal in the future. All uses of the coal must therefore continue to be predicted on a low-rank, low-grade quality feedstock.

7.9 Although Hat Creek coal exhibits adsorptive and ion exchange properties, its use as a medium for effluent treatment and wastewater purification is not recommended for the following reasons: -

(i) these properties are not extraordinary as compared with other, commercially available water treatment materials.

(ii) the presence of swelling clays in the coal will, themselves, create a serious water treatment problem through dispersion in the water being treated.

(iii) the coal cannot be economically regenerated and it is a low grade fuel.

(iv) the coal cannot be prepared and transported to industrial regions for use in water treatment at costs competitive with existing materials and methods.

7.10 The chemically and structurally bound water in the minerals present in Hat Creek coal, by interfering with the Standard method of analyses, may cause misleading proximate and ultimate analyses.

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Appendix A

TERMS OF REFERENCE

1. Identify and evaluate major alternative uses of Hat Creek coal. The study is to include but not be limited to a comparative analysis of the following areas:

    (a) Principally solid products

    - for combustion for process heat generation
    - from solvent refining

    (b) Principally liquid products

    - from pyrolysis
    - from solution and hydrogenation of coal and tar
    - from synthesis gas

    (c) Principally gaseous products

    - from coal gasification

    The study will evaluate the major uses of the Hat Creek coal as alternatives to the 2000 MW (net) thermal electrical power generating plant at Hat Creek.

    The coal consumption and specifications for the proposed power plant will be advised by B.C. Hydro. Coal quality tentative specifications for a process plant adjacent to the 2000 MW power plant will also be advised by B.C. Hydro.

2. For each of the selected process applications, material and energy balances per unit of feed material

(one tonne) shall be developed. Material and energy flows per unit time (one hour) shall be presented on flow diagrams showing the thermodynamic states of reactants and products.

Power, steam, land and water requirements associated with various plants shall be clearly identified.

The manpower requirements for construction and operation of the various processes will be identified.

3. Capital investment and operating costs for each selected process will be identified. The plant availability will be specified. Capital and operating cost data should be specified in constant 1980 dollars and displayed on a cash flow basis.

4. In considering alternate uses of the resource, a market forecast within the time frame 1990 - 2010 will be developed for these uses based on:

   (a) products (identified under 1.) manufactured at Hat Creek and supplied to meet market demands

   (b) potential development of secondary industry in the Province using the products from (a).

   Data relating to existing or anticipated future productive capacity, supply and demand for individual products will be accumulated and forecasts of probable future markets will be prepared. The evaluations will include estimates of probable selling prices at selected locations, the costs involved in delivering products from Hat Creek to those areas, an indication of the profitability of serving the principal markets from Hat Creek, and a resulting evaluation of whether a given product will be economically viable.

5. The study will derive opportunity costs (or values in alternative uses) based on alternate uses of Hat Creek coal. A framework within which the opportunity costs will be evaluated shall be agreed upon by B.C. Hydro.

6. The economic evaluation will develop cash flow projections showing costs, potential sales dollars and resulting net income and cash generation.

7. Environmental considerations associated with various processes will be described. The best practical technology available to reduce the emissions to stringent levels will be outlined including costs.

8. Economic criteria for the study will be provided by B.C. Hydro.

9. The work will be carried out including reference to the related studies supplied by B.C. Hydro.

   The consultant will provide, in questionnaire form, at the start of the study, the list of information he requires from B.C. Hydro.

10. The consultant will prepare minutes of all joint meetings held with the study coordinator or others contacted during the course of the study and submit monthly progress reports which include a summary of expenditures to date.

11. The study is to be controlled and coordinated on behalf of B.C. Hydro by the Vice-President, Engineering Group or his appointee.

12. Draft report shall be submitted to B.C. Hydro for review within 24 weeks of the date of commencement of the study.

Note:

(i) International system of units (S.I. units) should be used throughout the report. Conventional American or English units should be put in brackets following the S.I. units.

(ii) All calculations and use of formulas should be clearly presented for easy reading.

(iii) Sources of information used in report should be documented.

(iv) All tables and figures in the report should have descriptive titles.

(v) The report should have a table of contents and an index of tables and figures.

Appendix B

FINANCIAL/ECONOMIC CRITERIA

1. Cost and revenue data to be presented on an annual basis in 1980 Canadian dollars.

2. Exchange Rate Cdn. $ per U.S. $: range of 1.11 to 1.17

3. Project life for a thermal station is 35 years. Project life for other facilites to be based on industry experience or estimated separately. Data on the thermal plant will be provided by B.C. Hydro for inclusion in the analysis.

4. For comparative purposes an in-service date of 1989 should be assumed, if practical. However, cost and revenue estimates should also be presented independently of an in-service date, e.g. year -6, -5, -4, -3, -2, -1, 0, 1, 2, 3, etc. (0 = in-service year).

5. Complete cash flows are to be provided for cost and revenue estimates over the life of each development. Operating costs should be broken down into major components (BCH will provide power costs) on an annual basis.

6. The analysis should be prepared to show the net return to the fixed resource, i.e. the coal, on a present worth basis (present discounted value of revenues minus costs). The estimation and reporting of net present values will be related to the investment levels that pertain to the particular coal use considered in order to provide a more useful comparison.

These net present values can then be converted to the net value of the resource under alternative uses on a $/tonne basis. The opportunity cost of coal in alternative uses will be an output and should not be included in the costs of any single process.

7. Discount rates to be used in the analysis: real rates of 3, 6 and 10 percent.

8. Coal costs will be provided on a $/tonne basis for each discount rate.

9. The analysis should be in constant dollars. Escalation of certain relative prices should be incorporated if experience suggests these will be different from general inflation.

10. If a program is not already in place one should be developed to allow any recalculations based on subsequent revisions of estimates.

11. Sensitivity analysis should be performed on all major cost and revenue estimates with a most likely scenario bounded by high and low scenarios. These will be established in consultation with the client and will include such things as ranges for future product prices, pollution control requirements and capital costs.

12. Taxes should not be considered in looking at project costs and benefits. Tax regimes which will affect final demand, e.g. methanol/gasoline, will have to be taken into account. (cf. "Liquid Fuels from Renewable Resources: Feasibility Study", Inter-Group Consultants).

13. Depreciation should be taken into account only in estimates of O&M operating costs.

Appendix C

COMPUTER PROGRAM FOR ESTIMATION OF NET PRESENT VALUES

1. INTRODUCTION

The program computes Net Present Values (NPV in accordance with the Financial/Economic Criteria (Appendix B.) and certain other assumptions listed in Section 6.

The program has general utility but,in this instance, is arranged to provide specific outputs for the following table of coal conversion cases.

The following description is intended to provide the necessary information to use the program.

2. PROGRAM

The program has been written in Fortran IV and is operational on IBM-3033 computer system. It consists of a main program and a subroutine by the name REPORT.

Input is provided through Device No. 5 and output through Device No. 6.

3. CAPABILITY

Given total capital costs of the four processes, sales revenues generated, interest rate and coal price, the program computes various economic parameters and the net present value for each of the four processes and presents the results in the manner described under Output Section.

SELECTED COAL CONVERSION CASES

Process Type	Process Description	Case	Fig.	Table
Direct Hydrogenation	H-Coal	A1	D1.1	3.3
Indirect Liquefaction	Lurgi (Max. Power)	B5	D1.9	3.4
Indirect Liquefaction	Methanol (Lurgi)	C5	D1.16	3.6
Indirect Liquefaction	M-Gasoline (Lurgi)	D5	D1.23	3.7

Each "case" thus consists of input for capital costs and sales revenues associated with the four processes, interest rate and coal price. User can run as many cases as desired (up to 25) at one time.

User can also exercise the option of printing either the detailed yearly projection for 38 years or only the net present value and corresponding input economic parameters. There is a "switch" provided for this purpose, which is set either on or off as described under input.

4. NOMENCLATURE

	Variable	Mode	Explanation
1.	TOTCAP	Integer	DIMENSION (4)
	TOTCAP (1)		Total Capital Cost for H-Coal Process, $ million
	TOTCAP (2)		Total Capital Cost for F.T. Process, $ million
	TOTCAP (3)		Total Capital Cost for Methanol Process, $ million
	TOTCAP (4)		Total Capital Cost for M-Gasoline Process, $ million
2.	SALES	Integer	DIMENSION (4)
	SALES (1)		Sales Revenue for H-Coal Process, $ million/year
	SALES (2)		Sales Revenue for F.T. Process, $ million/year
	SALES (3)		Sales Revenue to Methanol Process, $ million/year
	SALES (4)		Sales Revenue for M-Gasoline Process, $ million/year

3. RATE	Real	Interest Rate, %
4. COALPR	Real	Coal Price, $/tonne
5. ISW	Integer	Switch: = 1 : only net present value and corresponding input economic parameters printed
6. ISEQ	Integer	Sequence Number of the Case

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 0001               COMMON/PRSENT/CAPTAL,CONINT,TOTCST,REVENU,COAL,POWER,CHEM,
                   1  WATER,SUBTOT,OHDLAB,REPLAC,OPTOT,GRMRGN,START,WORK,CASH,
                   2  DCF,VALUE
 0002               INTEGER CAPTAL,CONINT,TOTCST,REVENU,COAL,POWER,CHEM,WATER,
                   1  SUBTOT,OHDLAB,REPLAC,OPTOT,GRMRGN,START,WORK,CASH,DCF,VALUE
 0003               INTEGER TOTCAP,SALES
 0004               REAL LABOR,INCATL
 0005               DIMENSION CAPTAL(38,4),CONINT(8,4),TOTCST(8,4),REVENU(30,4),
                   1  COAL(30,4),POWER(30,4),CHEM(30,4),WATER(30,4),SUBTOT(30,4),
                   2  OHDLAB(30,4),REPLAC(30,4),OPTOT(38,4),GRMRGN(38,4),START(4),
                   3  WORK(2,4),CASH(38,4),DCF(38,4),VALUE(4)
 0006               DIMENSION TOTCAP(4),SALES(4),COALFD(4),PERCAP(8),PERSAL(30),
                   1  CCOP(4),AGUA(4),LABOR(4),INCATL(4)
 0007               DIMENSION NPV(25,4),KAPTAL(25,4),IVEND(25,4),RINT(25,4),COLL(25,4)
 0008               DATA PERCAP/2.,5.,15.,20.,22.,20.,10.,6./
 0009               DATA PERSAL/50.,75.,28*100./
 0010               DATA COALFD/38100.,66395.,46647.,51700./
 0011               DATA CCOP/28.,10.,16.,21./
 0012               DATA AGUA/9.5,17.,13.7,11.4/
 0013               DATA LABOR/700.,900.,700.,800./
 0014               DATA INCATL/35.,17.,25.,32./
 0015               COMMON/PRSENT/IVARS
 0016               DIMENSION IVARS(1800)
 0017               COMMON/SWITCH/ISW
              C *** INITIALIZE
 0018               DO 500 I =1,25
 0019               DO 503 J = 1,4
 0020               NPV(I,J) = 0
 0021               KAPTAL(I,J) = 0
 0022               IVEND(I,J) = 0
 0023               RINT(I,J) = 0.
 0024               COLL(I,J) = 0.
 0025           503 CONTINUE
 0026           500 CONTINUE
              C *** READ THE PARAMETERS
 0027            10 READ (5,100) TOTCAP,SALES
 0028               IF (TOTCAP(1).EQ.0) GO TO 900
 0029               READ (5,103) RATE,COALPR,ISW,ISEQ
 0030           100 FORMAT (6I6)
 0031           103 FORMAT (2E10.2,2I5)
 0032               DO 300 I =1,4
              C *** INITIALIZE COMMON
 0033               DO 400 IN = 1,1800
 0034               IVARS(IN) = 0
 0035           400 CONTINUE
 0036               DO 305 J =1,8
 0037               CAPTAL(J,I) = TOTCAP(I)*PERCAP(J)/100. + .5
 0038               CONINT(J,I) = CAPTAL(J,I)*RATE/100. + .5
 0039               TOTCST(J,I) = CAPTAL(J,I) + CONINT(J,I)
 0040               GRMRGN(J,I) = 0
 0041               OPTOT(J,I) = 0
 0042           305 CONTINUE
 0043               DO 307 J = 9,38
```

```
      CAPTAL(J,I) = 0
  307 CONTINUE
      DO 310 J = 1,30
      REVENU(J,I) = SALES(I)*PERSAL(J)/100. + .5
      COAL(J,I) = COALPR*COALFD(I)*330.*PERSAL(J)/10.**8 + .5
      POWER(J,I) = 0
      CHEM(J,I) = CCOP(I)*PERSAL(J)/100. + .5
      WATER(J,I) = AGUA(I)*.449092489*PERSAL(J)/100. + .5
      SUBTOT(J,I) = COAL(J,I) + POWER(J,I) + CHEM(J,I) + WATER(J,I)
      OHDLAB(J,I) = LABOR(I)*.078  + .5
      REPLAC(J,I) = .03*TOTCAP(I) + .5
      JX = J + 6
      OPTOT(JX,I) = SUBTOT(J,I) + OHDLAB(J,I) + REPLAC(J,I)
      GRMRGN(JX,I) = REVENU(J,I) - OPTOT(JX,I)
  310 CONTINUE
      START(I) = (COAL(3,I) + POWER(3,I) + WATER(3,I))/48. +
     1   TOTCAP(I)/360. + INCATL(I) + .5
      WORK(1,I) = .5*((2.*COAL(3,I) + POWER(3,I) + CHEM(3,I) +
     1 WATER(3,I) + REPLAC(3,I))/12. + .01*TOTCAP(I)) + .5
      WORK(2,I) = WORK(1,I)
      VALUE(I) = 0
      DO 320 J = 1,36
      IF (J.LE.8) CASH(J,I) = -(CAPTAL(J,I) + CONINT(J,I))
      IF (J.EQ.9) CASH(J,I) = GRMRGN(J,I)-START(I)-WORK(1,I)
      IF (J.EQ.10) CASH(J,I) = GRMRGN(J,I) - WORK(1,I)
      IF (J.GE.11) CASH(J,I) = GRMRGN(J,I)
      DCF(J,I) = CASH(J,I)/(1. + RATE/100.)**J
      VALUE(I) = VALUE(I) + DCF(J,I)
  320 CONTINUE
      NPV(ISEQ,I) = VALUE(I)
      KAPTAL(ISEQ,I) = TOTCAP(I)
      IVEND(ISEQ,I) = SALES(I)
      RINT(ISEQ,I) = RATE
      COLL(ISEQ,I) = COALPR
  300 CONTINUE
      IF (ISW.EQ.1) CALL REPORT
      GO TO 10
  900 IF (ISW.EQ.1) GO TO 950
      DO 407 N = 1,4
      GO TO (601,602,603,604), N
  601 WRITE (6,701)
      GO TO 610
  602 WRITE (6,702)
      GO TO 610
  603 WRITE (6,703)
      GO TO 610
  604 WRITE (6,704)
  610 WRITE (6,705)
      DO 410 M = 1,25
      IF (NPV(M,N).EQ.0) GO TO 410
      WRITE (6,710) KAPTAL(M,N),IVEND(M,N),RINT(M,N),COLL(M,N),NPV(M,N)
  410 CONTINUE
  407 CONTINUE
```

```
 0095            701 FORMAT (1H1//51X,'DIRECT LIQUEFACTION : H-COAL'//)
 0096            702 FORMAT (1H1//32X,'INDIRECT LIQUEFACTION : FISCHER TROPSCH        LU
                    1RGI (MAKE POWER)'//)
 0097            703 FORMAT (1H1//37X,'INDIRECT LIQUEFACTION : METHANOL       LURGI (MAK
                    1E POWER)'//)
 0098            704 FORMAT (1H1//36X,'INDIRECT LIQUEFACTION : M-GASOLINE      LURGI (M
                    1AKE POWER)'//)
 0099            705 FORMAT (T10,'CAPITAL',T30,'REVENUE',T50,'INTEREST',T70,
                    1  'COAL PRICE',T90,'NET PRESENT VALUE'/T12,'$ MM',T32,'$ MM',T53,
                    2  '%',T71,'$ PER TON',T95,'$ MM'//)
 0100            710 FORMAT (T12,I5,T32,I5,T52,F5.2,T73,F5.2,T95,I5/)
 0101            950 STOP
 0102                END
```

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0001                SUBROUTINE REPORT
0002                COMMON/PRSENT/CAPITAL,CONINT,TOTCST,REVENU,COAL,POWER,CHEM,
                  1  WATER,SUBTOT,OHDLAB,REPLAC,OPTOT,GRMRGN,START,WORK,CASH,
                  2  DCF,VALUE
0003                COMMON/SWITCH/ISW
0004                INTEGER CAPITAL,CONINT,TOTCST,REVENU,COAL,POWER,CHEM,WATER,
                  1  SUBTOT,OHDLAB,REPLAC,OPTOT,GRMRGN,START,WORK,CASH,DCF,VALUE
0005                DIMENSION CAPITAL(38,4),CONINT(8,4),TOTCST(8,4),REVENU(30,4),
                  1  COAL(30,4),POWER(30,4),CHEM(30,4),WATER(30,4),SUBTOT(30,4),
                  2  OHDLAB(30,4),REPLAC(30,4),OPTOT(38,4),GRMRGN(38,4),START(4),
                  3  WORK(2,4),CASH(38,4),DCF(38,4),VALUE(4)
0006                DO 500 I = 1,4
0007                DO 520 IPG = 1,3
0008                WRITE (6,100) IPG
0009                GO TO (501,502,503,504),I
0010            501 WRITE (6,101)
0011                GO TO 510
0012            502 WRITE (6,102)
0013                GO TO 510
0014            503 WRITE (6,103)
0015                GO TO 510
0016            504 WRITE (6,104)
0017            510 IF (IPG.EQ.3) GO TO 512
0018                WRITE (6,107)
0019                GO TO 515
0020            512 WRITE (6,109)
0021            515 IX = (IPG-1)*13 + 1
0022                JX = IX + 12
0023                IF (IPG.EQ.3) JX = 38
0024                WRITE (6,111) (K), K = IX,JX)
0025                IF (IPG.GT.1) GO TO 600
0026                WRITE (6,115) (CAPITAL(K,I),K=1,8)
0027                WRITE (6,117) (CONINT(K,I),K=1,8)
0028                WRITE (6,119) (TOTCST(K,I),K=1,8)
0029            600 IF (IPG-2) 601,602,603
0030            601 WRITE (6,121) (REVENU(K,I),K=1,5)
0031                WRITE (6,150)
0032                WRITE (6,123) (COAL(K,I),K=1,5)
0033                WRITE (6,125) (POWER(K,I),K=1,5)
0034                WRITE (6,127) (CHEM(K,I),K=1,5)
0035                WRITE (6,129) (WATER(K,I),K=1,5)
0036                WRITE (6,131) (SUBTOT(K,I),K=1,5)
0037                WRITE (6,133) (OHDLAB(K,I),K=1,5)
0038                WRITE (6,135) (REPLAC(K,I),K=1,5)
0039                WRITE (6,161) (OPTOT(K,I),K =9,13)
0040                WRITE (6,163) (GRMRGN(K,I),K=9,13)
0041                GO TO 612
0042            602 WRITE (6,215)
0043                WRITE (6,217)
0044                WRITE (6,219)
0045                WRITE (6,221) (REVENU(K,I),K=6,18)
0046                WRITE (6,150)
0047                WRITE (6,223) (COAL(K,I),K=6,18)
```

```
          WRITE (6,225) (POWER(K,I),K=6,18)
          WRITE (6,227) (CHEM(K,I),K=6,18)
          WRITE (6,229) (WATER(K,I),K=6,18)
          WRITE (6,231) (SUBTOT(K,I),K=6,18)
          WRITE (6,233) (OHDLAB(K,I),K=6,18)
          WRITE (6,235) (REPLAC(K,I),K=6,18)
          GO TO 610
      603 WRITE (6,215)
          WRITE (6,217)
          WRITE (6,219)
          WRITE (6,221) (REVENU(K,I),K=19,30)
          WRITE (6,150)
          WRITE (6,223) (COAL(K,I),K=19,30)
          WRITE (6,225) (POWER(K,I),K=19,30)
          WRITE (6,227) (CHEM(K,I),K=19,30)
          WRITE (6,229) (WATER(K,I),K=19,30)
          WRITE (6,331) (SUBTOT(K,I),K=19,30)
          WRITE (6,233) (OHDLAB(K,I),K=19,30)
          WRITE (6,335) (REPLAC(K,I),K=19,30)
          WRITE (6,341) (OPTOT(K,I),K=IX,JX)
          WRITE (6,343) (GRMRGN(K,I),K=IX,JX)
          WRITE (6,349) (CASH(K,I),K=IX,JX)
      610 WRITE (6,141) (OPTOT(K,I),K=IX,JX)
          WRITE (6,143) (GRMRGN(K,I),K=IX,JX)
      612 IF (IPG.EQ.1) GO TO 615
          WRITE (6,333)
          IF (IPG.EQ.3) GO TO 619
          GO TO 617
      615 WRITE (6,145) START(I)
          WRITE (6,147) (WORK(K,I),K=1,2)
      617 WRITE (6,149) (CASH(K,I),K=IX,JX)
      619 WRITE (6,151) (DCF(K,I),K=IX,JX)
          IF (IPG.NE.1) GO TO 520
          WRITE (6,153) VALUE(I)
      520 CONTINUE
      500 CONTINUE
      100 FORMAT (1H1////45X,'BRITISH COLUMBIA HYDRO & POWER AUTHORITY',
         1  36X,'PAGE = ',I2//49X,'HAT CREEK COAL UTILIZATION STUDY'
         2  //56X,'ECONOMIC ANALYSIS'/T54,'(IN MILLIONS OF DOLLARS)'/)
      101 FORMAT (51X,'DIRECT LIQUEFACTION : H-COAL'//)
      102 FORMAT (32X,'INDIRECT LIQUEFACTION : FISCHER-TROPSCH       LURGI (
         1MAKE POWER)'//)
      103 FORMAT (37X,'INDIRECT LIQUEFACTION : METHANOL       LURGI (MAKE PO
         1WER)'//)
      104 FORMAT (36X,'INDIRECT LIQUEFACTION : M-GASOLINE       LURGI (MAKE
         1POWER)'//)
      107 FORMAT (39X,13(3X,'YEAR'))
      109 FORMAT (39X,12(3X,'YEAR'))
      111 FORMAT (43X,12(I2,5X),I2)
      115 FORMAT (//T6,'CAPITAL COSTS',T40,8(2X,I5)/)
      117 FORMAT (T6,'INTEREST CHARGES',T40,8(2X,I5)/)
      119 FORMAT (T10,'TOTAL COSTS',T40,8(2X,I5)//)
      150 FORMAT (/T6,'OPERATING COSTS'//T10,'RAW MATERIALS'/)
```

```
0096          160 FORMAT (/T125,'.....',I1)
0097          121 FORMAT (T6,'SALES REVENUES',T96,5(2X,I5))
0098          123 FORMAT (T10,'COAL',T96,5(2X,I5))
0099          125 FORMAT (T10,'ELECTRICITY',T96,5(2X,I5))
0100          127 FORMAT (T10,'CHEMICALS',T96,5(2X,I5))
0101          129 FORMAT (T10,'WATER',T96,5(2X,I5))
0102          131 FORMAT (T12,'SUB-TOTAL',T96,5(2X,I5)//)
0103          133 FORMAT (T10,'LABOR & OVERHEAD',T96,5(2X,I5))
0104          135 FORMAT (T10,'REPLACEMENT',T96,5(2X,I5)//)
0105          215 FORMAT (//T6,'CAPITAL COSTS'/)
0106          217 FORMAT (T6,'INTEREST CHARGES'/)
0107          219 FORMAT (T10,'TOTAL COSTS'//)
0108          221 FORMAT (T6,'SALES REVENUES',T40,13(2X,I5))
0109          223 FORMAT (T10,'COAL',T40,13(2X,I5))
0110          225 FORMAT (T10,'ELECTRICITY',T40,13(2X,I5))
0111          227 FORMAT (T10,'CHEMICALS',T40,13(2X,I5))
0112          229 FORMAT (T10,'WATER',T40,13(2X,I5))
0113          231 FORMAT (T12,'SUB-TOTAL',T40,13(2X,I5)//)
0114          233 FORMAT (T10,'LABOR & OVERHEAD',T40,13(2X,I5))
0115          235 FORMAT (T10,'REPLACEMENT',T40,13(2X,I5)//)
0116          141 FORMAT (T6,'TOTAL OPERATING COST',T40,13(2X,I5)//)
0117          143 FORMAT (T6,'GROSS MARGIN',T40,13(2X,I5)//)
0118          145 FORMAT (T6,'START-UP COST',T96,I5)
0119          147 FORMAT (T6,'WORKING CAPITAL',T96,2(2X,I5)///)
0120          149 FORMAT (T6,'NET CASH FLOW',T40,13(2X,I5)/)
0121          151 FORMAT (T6,'DISCOUNTED CASH FLOW',T40,13(2X,I5))
0122          153 FORMAT (//T6,'NET PRESENT VALUE',T42,I5)
0123          161 FORMAT (T6,'TOTAL OPERATING COST',T96,5(2X,I5)//)
0124          163 FORMAT (T6,'GROSS MARGIN',T96,5(2X,I5)//)
0125          331 FORMAT (T12,'SUB-TOTAL',T40,12(2X,I5)//)
0126          335 FORMAT (T10,'REPLACEMENT',T40,12(2X,I5)//)
0127          341 FORMAT (T6,'TOTAL OPERATING COST',T40,12(2X,I5)//)
0128          343 FORMAT (T6,'GROSS MARGIN',T40,12(2X,I5)//)
0129          349 FORMAT (T6,'NET CASH FLOW',T40,12(2X,I5)//)
0130          333 FORMAT (T6,'START-UP COST'/T6,'WORKING CAPITAL'///)
0131          750 RETURN
0132              END
```

6. INPUT

Variable	Format	Columns
Card No. 1		
TOTCAP (1)	I8	1-8
TOTCAP (2)	I8	9-16
TOTCAP (3)	I8	17-24
TOTCAP (4)	I8	25-32
SALES (1)	I8	33-40
SALES (2)	I8	41-48
SALES (3)	I8	49-56
SALES (4)	I8	57-64
Card No. 2		
RATE	F10.2	1-10
COALPR	F10.2	11-20
ISW	I5	21-25
ISEQ	I5	26-30

Listing of example inputs for 16 cases with "switch" on and off follow:

```
 2208   4347   3117   3623    991   1060    979   1066
   3.    10.      0      1
 2650   5216   3740   4348    991   1060    979   1066
   3.    10.      0      2
 3312   6521   4676   5435    991   1060    979   1066
   3.    10.      0      3
 4416   8694   6234   7246    991   1060    979   1066
   3.    10.      0      4
 2208   4347   3117   3623    892    954    881    959
   3.    10.      0      5
 2208   4347   3117   3623    793    846    783    848
   3.    10.      0      6
 2208   4347   3117   3623    595    636    587    640
   3.    10.      0      7
 2650   5216   3740   4348    892    954    881    959
   3.    10.      0      8
 2650   5216   3740   4348    793    848    783    846
   3.    10.      0      9
 2650   5216   3740   4348    595    636    587    640
   3.    10.      0     10
 3312   6521   4676   5435    892    954    881    959
   3.    10.      0     11
 3312   6521   4676   5435    793    846    783    846
   3.    10.      0     12
 3312   6521   4676   5435    595    636    587    640
   3.    10.      0     13
 4416   8694   6234   7246    892    954    881    959
   3.    10.      0     14
 4416   8694   6234   7246    793    846    783    846
   3.    10.      0     15
 4416   8694   6234   7246    595    636    587    640
   3.    10.      0     16
    0
```

INPUT WITH SWITCH OFF

```
2208    4347    3117    3623     991    1060     979    1066
  3.     10.       1       1
2650    5216    3740    4348     991    1060     979    1066
  3.     10.       1       2
3312    6521    4676    5435     991    1060     979    1066
  3.     10.       1       3
4416    8694    6234    7246     991    1060     979    1066
  3.     10.       1       4
2209    4347    3117    3623     892     954     881     959
  3.     10.       1       5
2208    4347    3117    3623     793     848     783     848
  3.     10.       1       6
2208    4347    3117    3623     595     636     587     640
  3.     10.       1       7
2650    5216    3740    4348     892     954     881     959
  3.     10.       1       8
2650    5216    3740    4348     793     848     783     848
  3.     10.       1       9
2650    5216    3740    4348     595     636     587     640
  3.     10.       1      10
3312    6521    4676    5435     892     954     881     959
  3.     10.       1      11
3312    6521    4676    5435     793     848     783     848
  3.     10.       1      12
3312    6521    4676    5435     595     636     587     640
  3.     10.       1      13
4416    8694    6234    7246     892     954     881     959
  3.     10.       1      14
4416    8694    6234    7246     793     848     783     848
  3.     10.       1      15
4416    8694    6234    7246     595     636     587     640
  3.     10.       1      16
     0
```

INPUT WITH SWITCH ON

7. OUTPUT

Output corresponding to the first of the 16 cases of the example input with the "switch" on and that corresponding to all 16 cases with the "switch" off follow:

```
                              DIRECT LIQUEFACTION : H-COAL

    CAPITAL          REVENUE          INTEREST          COAL PRICE          NET PRESENT VALUE
     $ MM             $ MM               %              $ PER TON                $ MM

      2208             991              3.00              10.00                  6467
      2650             991              3.00              10.00                  7882
      3312             991              3.00              10.00                  6976
      4416             991              3.00              10.00                  5469
      2208             892              3.00              10.00                  7009
      2208             793              3.00              10.00                  5535
      2208             595              3.00              10.00                  2563
      2650             892              3.00              10.00                  6412
      2650             793              3.00              10.00                  4934
      2650             595              3.00              10.00                  1961
      3312             892              3.00              10.00                  5499
      3312             793              3.00              10.00                  4027
      3312             595              3.00              10.00                  1076
      4416             892              3.00              10.00                  3994
      4416             793              3.00              10.00                  2519
      4416             595              3.00              10.00                  -433
```

~~OU~~TPUT WITH SWITCH OFF

INDIRECT LIQUEFACTION : FISCHER TROPSCH LURGI (MAKE POWER)

CAPITAL $ MM	REVENUE $ MM	INTEREST %	COAL PRICE $ PER TON	NET PRESENT VALUE $ MM
4347	1060	3.00	10.00	5177
5216	1060	3.00	10.00	3991
6521	1060	3.00	10.00	2193
8694	1060	3.00	10.00	-778
4347	954	3.00	10.00	3597
4347	846	3.00	10.00	2019
4347	636	3.00	10.00	-1142
5216	954	3.00	10.00	2413
5216	848	3.00	10.00	833
5216	636	3.00	10.00	-2327
6521	954	3.00	10.00	615
6521	848	3.00	10.00	-968
6521	636	3.00	10.00	-4126
8694	954	3.00	10.00	-2361
8694	848	3.00	10.00	-3940
8694	636	3.00	10.00	-7049

INDIRECT LIQUEFACTION : METHANOL LURGI (MAKE POWER)

CAPITAL $ MM	REVENUE $ MM	INTEREST %	COAL PRICE $ PER TON	NET PRESENT VALUE $ MM
3117	979	3.00	10.00	6786
3740	979	3.00	10.00	5945
4676	979	3.00	10.00	4665
6234	979	3.00	10.00	2527
3117	881	3.00	10.00	5324
3117	783	3.00	10.00	3867
3117	587	3.00	10.00	944
3740	881	3.00	10.00	4485
3740	783	3.00	10.00	3024
3740	587	3.00	10.00	107
4676	881	3.00	10.00	3203
4676	783	3.00	10.00	1743
4676	587	3.00	10.00	-1175
6234	881	3.00	10.00	1068
6234	783	3.00	10.00	-393
6234	587	3.00	10.00	-3314

INDIRECT LIQUEFACTION : M-GASOLINE LURGI (MAKE POWER)

CAPITAL $ MM	REVENUE $ MM	INTEREST %	COAL PRICE $ PER TON	NET PRESENT VALUE $ MM
3623	1066	3.00	10.00	6970
4348	1066	3.00	10.00	5980
5435	1066	3.00	10.00	4490
7246	1066	3.00	10.00	2013
3623	959	3.00	10.00	5372
3623	848	3.00	10.00	3715
3623	640	3.00	10.00	615
4348	959	3.00	10.00	4383
4348	848	3.00	10.00	2732
4348	640	3.00	10.00	-368
5435	959	3.00	10.00	2893
5435	848	3.00	10.00	1238
5435	640	3.00	10.00	-1857
7246	959	3.00	10.00	419
7246	848	3.00	10.00	-1234
7246	640	3.00	10.00	-4331

OUTPUT WITH SWITCH ON

```
                        BRITISH COLUMBIA HYDRO & POWER AUTHORITY                            PAGE =  1

                               HAT CREEK COAL UTILIZATION STUDY

                                       ECONOMIC ANALYSIS
                                   (IN MILLIONS OF DOLLARS)

                                 DIRECT LIQUEFACTION : H-COAL

                          YEAR  YEAR  YEAR  YEAR  YEAR  YEAR  YEAR  YEAR  YEAR  YEAR  YEAR  YEAR  YEAR
                            1     2     3     4     5     6     7     8     9    10    11    12    13

CAPITAL COSTS               44   110   331   442   486   442   221   132

INTEREST CHARGES             1     3    10    13    15    13     7     4

   TOTAL COSTS              45   113   341   455   501   455   228   136

SALES REVENUES                                                            496   743   991   991   991

OPERATING COSTS

   RAW MATERIALS

   COAL                                                                    63    94   126   126   126
   ELECTRICITY                                                              0     0     0     0     0
   CHEMICALS                                                               14    21    28    28    28
   WATER                                                                    2     3     4     4     4
     SUB-TOTAL                                                             79   118   158   158   158

   LABOR & OVERHEAD                                                        55    55    55    55    55
   REPLACEMENT                                                             66    66    66    66    66

TOTAL OPERATING COST                                                      200   239   279   279   279

GROSS MARGIN                                                              296   504   712   712   712

START-UP COST                                                              44
WORKING CAPITAL                                                            26    26

NET CASH FLOW              -45  -113  -341  -455  -501  -455  -228  -136   226   478   712   712   712

DISCOUNTED CASH FLOW       -43  -106  -322  -404  -432  -381  -185  -107   173   355   514   499   484

NET PRESENT VALUE         0487
```

OUTPUT WITH SWITCH ON

BRITISH COLUMBIA HYDRO & POWER AUTHORITY

HAT CREEK COAL UTILIZATION STUDY

ECONOMIC ANALYSIS
(IN MILLIONS OF DOLLARS)

DIRECT LIQUEFACTION : H-COAL

	YEAR 14	YEAR 15	YEAR 16	YEAR 17	YEAR 18	YEAR 19	YEAR 20	YEAR 21	YEAR 22	YEAR 23	YEAR 24	YEAR 25	YEAR 26
CAPITAL COSTS													
INTEREST CHARGES													
TOTAL COSTS													
SALES REVENUES	991	991	991	991	991	991	991	991	991	991	991	991	991
OPERATING COSTS													
RAW MATERIALS													
COAL	126	126	126	126	126	126	126	126	126	126	126	126	126
ELECTRICITY	0	0	0	0	0	0	0	0	0	0	0	0	0
CHEMICALS	28	28	28	28	28	28	28	28	28	28	28	28	28
WATER	4	4	4	4	4	4	4	4	4	4	4	4	4
SUB-TOTAL	158	158	158	158	158	158	158	158	158	158	158	158	158
LABOR & OVERHEAD	55	55	55	55	55	55	55	55	55	55	55	55	55
REPLACEMENT	66	66	66	66	66	66	66	66	66	66	66	66	66
TOTAL OPERATING COST	279	279	279	279	279	279	279	279	279	279	279	279	279
GROSS MARGIN	712	712	712	712	712	712	712	712	712	712	712	712	712
START-UP COST													
WORKING CAPITAL													
NET CASH FLOW	712	712	712	712	712	712	712	712	712	712	712	712	712
DISCOUNTED CASH FLOW	470	457	443	430	418	406	394	382	371	360	350	340	330

BRITISH COLUMBIA HYDRO & POWER AUTHORITY

HAT CREEK COAL UTILIZATION STUDY

ECONOMIC ANALYSIS
(IN MILLIONS OF DOLLARS)

DIRECT LIQUEFACTION : H-COAL

	YEAR 27	YEAR 28	YEAR 29	YEAR 30	YEAR 31	YEAR 32	YEAR 33	YEAR 34	YEAR 35	YEAR 36	YEAR 37	YEAR 38
CAPITAL COSTS												
INTEREST CHARGES												
TOTAL COSTS												
SALES REVENUES	991	991	991	991	991	991	991	991	991	991	991	991
OPERATING COSTS												
RAW MATERIALS												
COAL	126	126	126	126	126	126	126	126	126	126	126	126
ELECTRICITY	0	0	0	0	0	0	0	0	0	0	0	0
CHEMICALS	28	28	28	28	28	28	28	28	28	28	28	28
WATER	4	4	4	4	4	4	4	4	4	4	4	4
SUB-TOTAL	158	158	158	158	158	158	158	158	158	158	158	158
LABOR & OVERHEAD	55	55	55	55	55	55	55	55	55	55	55	55
REPLACEMENT	66	66	66	66	66	66	66	66	66	66	66	66
TOTAL OPERATING COST	279	279	279	279	279	279	279	279	279	279	279	279
GROSS MARGIN	712	712	712	712	712	712	712	712	712	712	712	712
NET CASH FLOW	712	712	712	712	712	712	712	712	712	712	712	712
TOTAL OPERATING COST	279	279	279	279	279	279	279	279	279	279	279	279
GROSS MARGIN	712	712	712	712	712	712	712	712	712	712	712	712
START-UP COST												
WORKING CAPITAL												
DISCOUNTED CASH FLOW	320	311	302	293	284	276	268	260	253	245	236	231

BRITISH COLUMBIA HYDRO & POWER AUTHORITY

HAT CREEK COAL UTILIZATION STUDY

ECONOMIC ANALYSIS
(IN MILLIONS OF DOLLARS)

INDIRECT LIQUEFACTION : FISCHER-TROPSCH LURGI (MAKE POWER)

	YEAR 1	YEAR 2	YEAR 3	YEAR 4	YEAR 5	YEAR 6	YEAR 7	YEAR 8	YEAR 9	YEAR 10	YEAR 11	YEAR 12	YEAR 13
CAPITAL COSTS	87	217	652	869	956	869	435	261					
INTEREST CHARGES	3	7	20	26	29	26	13	8					
TOTAL COSTS	90	224	672	895	985	895	448	269					
SALES REVENUES									530	795	1060	1060	1060
OPERATING COSTS													
RAW MATERIALS													
COAL									110	164	219	219	219
ELECTRICITY									0	0	0	0	0
CHEMICALS									5	8	10	10	10
WATER									4	6	6	6	6
SUB-TOTAL									119	178	237	237	237
LABOR & OVERHEAD									70	70	70	70	70
REPLACEMENT									130	130	130	130	130
TOTAL OPERATING COST									319	378	437	437	437
GROSS MARGIN									211	417	623	623	623
START-UP COST									34				
WORKING CAPITAL									46	46			
NET CASH FLOW	-90	-224	-672	-895	-985	-895	-448	-269	131	371	623	623	623
DISCOUNTED CASH FLOW	-87	-211	-614	-795	-849	-749	-364	-212	100	270	450	436	424
NET PRESENT VALUE	5177												

BRITISH COLUMBIA HYDRO & POWER AUTHORITY

HAT CREEK COAL UTILIZATION STUDY

ECONOMIC ANALYSIS
(IN MILLIONS OF DOLLARS)

INDIRECT LIQUEFACTION : FISCHER-TROPSCH   LURGI (MAKE POWER)

	YEAR 14	YEAR 15	YEAR 16	YEAR 17	YEAR 18	YEAR 19	YEAR 20	YEAR 21	YEAR 22	YEAR 23	YEAR 24	YEAR 25	YEAR 26
CAPITAL COSTS													
INTEREST CHARGES													
TOTAL COSTS													
SALES REVENUES	1060	1060	1060	1060	1060	1060	1060	1060	1060	1060	1060	1060	1060
OPERATING COSTS													
RAW MATERIALS													
COAL	219	219	219	219	219	219	219	219	219	219	219	219	219
ELECTRICITY	0	0	0	0	0	0	0	0	0	0	0	0	0
CHEMICALS	10	10	10	10	10	10	10	10	10	10	10	10	10
WATER	8	8	8	8	8	8	8	8	8	8	8	8	8
SUB-TOTAL	237	237	237	237	237	237	237	237	237	237	237	237	237
LABOR & OVERHEAD	70	70	70	70	70	70	70	70	70	70	70	70	70
REPLACEMENT	130	130	130	130	130	130	130	130	130	130	130	130	130
TOTAL OPERATING COST	437	437	437	437	437	437	437	437	437	437	437	437	437
GROSS MARGIN	623	623	623	623	623	623	623	623	623	623	623	623	623
START-UP COST													
WORKING CAPITAL													
NET CASH FLOW	623	623	623	623	623	623	623	623	623	623	623	623	623
DISCOUNTED CASH FLOW	411	399	388	376	365	355	344	334	325	315	306	297	288

HAT CREEK COAL UTILIZATION STUDY

ECONOMIC ANALYSIS
(IN MILLIONS OF DOLLARS)

INDIRECT LIQUEFACTION : FISCHER-TROPSCH LURGI (MAKE POWER)

	YEAR 27	YEAR 28	YEAR 29	YEAR 30	YEAR 31	YEAR 32	YEAR 33	YEAR 34	YEAR 35	YEAR 36	YEAR 37	YEAR 38
CAPITAL COSTS												
INTEREST CHARGES												
TOTAL COSTS												
SALES REVENUES	1060	1060	1060	1060	1060	1060	1060	1060	1060	1060	1060	1060
OPERATING COSTS												
RAW MATERIALS												
COAL	219	219	219	219	219	219	219	219	219	219	219	219
ELECTRICITY	0	0	0	0	0	0	0	0	0	0	0	0
CHEMICALS	10	10	10	10	10	10	10	10	10	10	10	10
WATER	8	8	8	8	8	8	8	8	8	8	8	8
SUB-TOTAL	237	237	237	237	237	237	237	237	237	237	237	237
LABOR & OVERHEAD	70	70	70	70	70	70	70	70	70	70	70	70
REPLACEMENT	130	130	130	130	130	130	130	130	130	130	130	130
TOTAL OPERATING COST	437	437	437	437	437	437	437	437	437	437	437	437
GROSS MARGIN	623	623	623	623	623	623	623	623	623	623	623	623
NET CASH FLOW	623	623	623	623	623	623	623	623	623	623	623	623
TOTAL OPERATING COST	437	437	437	437	437	437	437	437	437	437	437	437
GROSS MARGIN	623	623	623	623	623	623	623	623	623	623	623	623
START-UP COST												
WORKING CAPITAL												
DISCOUNTED CASH FLOW	280	272	264	256	249	241	234	228	221	214	208	202

BRITISH COLUMBIA HYDRO & POWER AUTHORITY PAGE = 1

HAT CREEK COAL UTILIZATION STUDY

ECONOMIC ANALYSIS
(IN MILLIONS OF DOLLARS)

INDIRECT LIQUEFACTION : METHANOL LURGI (MAKE POWER)

	YEAR 1	YEAR 2	YEAR 3	YEAR 4	YEAR 5	YEAR 6	YEAR 7	YEAR 8	YEAR 9	YEAR 10	YEAR 11	YEAR 12	YEAR 13
CAPITAL COSTS	62	156	468	623	686	623	312	187					
INTEREST CHARGES	2	5	14	19	21	19	9	6					
TOTAL COSTS	64	161	482	642	707	642	321	193					
SALES REVENUES									490	734	979	979	979
OPERATING COSTS													
RAW MATERIALS													
COAL									77	115	154	154	154
ELECTRICITY									0	0	0	0	0
CHEMICALS									8	12	16	16	16
WATER									3	5	6	6	6
SUB-TOTAL									88	132	176	176	176
LABOR & OVERHEAD									55	55	55	55	55
REPLACEMENT									94	94	94	94	94
TOTAL OPERATING COST									237	281	325	325	325
GROSS MARGIN									253	453	654	654	654
START-UP COST									37				
WORKING CAPITAL									33	33			
NET CASH FLOW	-64	-161	-482	-642	-707	-642	-321	-193	183	426	654	654	654
DISCOUNTED CASH FLOW	-62	-151	-441	-570	-609	-537	-261	-152	140	312	472	450	445
NET PRESENT VALUE	6786												

HAT CREEK COAL UTILIZATION STUDY

ECONOMIC ANALYSIS
(IN MILLIONS OF DOLLARS)

INDIRECT LIQUEFACTION : METHANOL LURGI (MAKE POWER)

	YEAR 14	YEAR 15	YEAR 16	YEAR 17	YEAR 18	YEAR 19	YEAR 20	YEAR 21	YEAR 22	YEAR 23	YEAR 24	YEAR 25	YEAR 26
CAPITAL COSTS													
INTEREST CHARGES													
TOTAL COSTS													
SALES REVENUES	979	979	979	979	979	979	979	979	979	979	979	979	979
OPERATING COSTS													
RAW MATERIALS													
COAL	154	154	154	154	154	154	154	154	154	154	154	154	154
ELECTRICITY	0	0	0	0	0	0	0	0	0	0	0	0	0
CHEMICALS	16	16	16	16	16	16	16	16	16	16	16	16	16
WATER	6	6	6	6	6	6	6	6	6	6	6	6	6
SUB-TOTAL	176	176	176	176	176	176	176	176	176	176	176	176	176
LABOR & OVERHEAD	55	55	55	55	55	55	55	55	55	55	55	55	55
REPLACEMENT	94	94	94	94	94	94	94	94	94	94	94	94	94
TOTAL OPERATING COST	325	325	325	325	325	325	325	325	325	325	325	325	325
GROSS MARGIN	654	654	654	654	654	654	654	654	654	654	654	654	654
START-UP COST													
WORKING CAPITAL													
NET CASH FLOW	654	654	654	654	654	654	654	654	654	654	654	654	654
DISCOUNTED CASH FLOW	432	419	407	395	384	372	362	351	341	331	321	312	303

BRITISH COLUMBIA HYDRO & POWER AUTHORITY

HAT CREEK COAL UTILIZATION STUDY

ECONOMIC ANALYSIS
(IN MILLIONS OF DOLLARS)

INDIRECT LIQUEFACTION : METHANOL LURGI (MAKE POWER)

	YEAR 27	YEAR 28	YEAR 29	YEAR 30	YEAR 31	YEAR 32	YEAR 33	YEAR 34	YEAR 35	YEAR 36	YEAR 37	YEAR 38
CAPITAL COSTS												
INTEREST CHARGES												
TOTAL COSTS												
SALES REVENUES	979	979	979	979	979	979	979	979	979	979	979	979
OPERATING COSTS												
RAW MATERIALS												
COAL	154	154	154	154	154	154	154	154	154	154	154	154
ELECTRICITY	0	0	0	0	0	0	0	0	0	0	0	0
CHEMICALS	16	16	16	16	16	16	16	16	16	16	16	16
WATER	6	6	6	6	6	6	6	6	6	6	6	6
SUB-TOTAL	176	176	176	176	176	176	176	176	176	176	176	176
LABOR & OVERHEAD	55	55	55	55	55	55	55	55	55	55	55	55
REPLACEMENT	94	94	94	94	94	94	94	94	94	94	94	94
TOTAL OPERATING COST	325	325	325	325	325	325	325	325	325	325	325	325
GROSS MARGIN	654	654	654	654	654	654	654	654	654	654	654	654
NET CASH FLOW	654	654	654	654	654	654	654	654	654	654	654	654
TOTAL OPERATING COST	325	325	325	325	325	325	325	325	325	325	325	325
GROSS MARGIN	654	654	654	654	654	654	654	654	654	654	654	654
START-UP COST												
WORKING CAPITAL												
DISCOUNTED CASH FLOW	294	285	277	269	261	253	246	239	232	225	219	212

BRITISH COLUMBIA HYDRO & POWER AUTHORITY

HAT CREEK COAL UTILIZATION STUDY

ECONOMIC ANALYSIS
(IN MILLIONS OF DOLLARS)

INDIRECT LIQUEFACTION : M-GASOLINE LURGI (MAKE POWER)

	YEAR 1	YEAR 2	YEAR 3	YEAR 4	YEAR 5	YEAR 6	YEAR 7	YEAR 8	YEAR 9	YEAR 10	YEAR 11	YEAR 12	YEAR 13
CAPITAL COSTS	72	181	543	725	797	725	362	217					
INTEREST CHARGES	2	5	16	22	24	22	11	7					
TOTAL COSTS	74	186	559	747	821	747	373	224					
SALES REVENUES									533	800	1066	1066	1066
OPERATING COSTS													
RAW MATERIALS													
COAL									65	126	171	171	171
ELECTRICITY									0	0	0	0	0
CHEMICALS									11	16	21	21	21
WATER									3	4	5	5	5
SUB-TOTAL									99	148	197	197	197
LABOR & OVERHEAD									62	62	62	62	62
REPLACEMENT									109	109	109	109	109
TOTAL OPERATING COST									270	319	368	368	368
GROSS MARGIN									263	481	698	698	698
START-UP COST									46				
WORKING CAPITAL									30	30			
NET CASH FLOW	-74	-186	-559	-747	-821	-747	-373	-224	179	443	696	698	698
DISCOUNTED CASH FLOW	-71	-175	-511	-663	-706	-625	-303	-176	137	329	506	469	475
NET PRESENT VALUE	6970												

BRITISH COLUMBIA HYDRO & POWER AUTHORITY

HAT CREEK COAL UTILIZATION STUDY

ECONOMIC ANALYSIS
(IN MILLIONS OF DOLLARS)

INDIRECT LIQUEFACTION : M-GASOLINE LURGI (MAKE POWER)

	YEAR 14	YEAR 15	YEAR 16	YEAR 17	YEAR 18	YEAR 19	YEAR 20	YEAR 21	YEAR 22	YEAR 23	YEAR 24	YEAR 25	YEAR 26
CAPITAL COSTS													
INTEREST CHARGES													
TOTAL COSTS													
SALES REVENUES	1066	1066	1066	1066	1066	1066	1066	1066	1066	1066	1066	1066	1066
OPERATING COSTS													
RAW MATERIALS													
COAL	171	171	171	171	171	171	171	171	171	171	171	171	171
ELECTRICITY	0	0	0	0	0	0	0	0	0	0	0	0	0
CHEMICALS	21	21	21	21	21	21	21	21	21	21	21	21	21
WATER	5	5	5	5	5	5	5	5	5	5	5	5	5
SUB-TOTAL	197	197	197	197	197	197	197	197	197	197	197	197	197
LABOR & OVERHEAD	62	62	62	62	62	62	62	62	62	62	62	62	62
REPLACEMENT	109	109	109	109	109	109	109	109	109	109	109	109	109
TOTAL OPERATING COST	368	368	368	368	368	368	368	368	368	368	368	368	368
GROSS MARGIN	698	698	698	698	698	698	698	698	698	698	698	698	698
START-UP COST													
WORKING CAPITAL													
NET CASH FLOW	698	698	698	698	698	698	698	698	698	698	698	698	698
DISCOUNTED CASH FLOW	461	446	434	422	410	398	386	375	364	353	343	333	323

BRITISH COLUMBIA HYDRO & POWER AUTHORITY

HAT CREEK COAL UTILIZATION STUDY

ECONOMIC ANALYSIS
(IN MILLIONS OF DOLLARS)

INDIRECT LIQUEFACTION : M-GASOLINE LURGI (MAKE POWER)

	YEAR 27	YEAR 28	YEAR 29	YEAR 30	YEAR 31	YEAR 32	YEAR 33	YEAR 34	YEAR 35	YEAR 36	YEAR 37	YEAR 38
CAPITAL COSTS												
INTEREST CHARGES												
TOTAL COSTS												
SALES REVENUES	1066	1066	1066	1066	1066	1066	1066	1066	1066	1066	1066	1066
OPERATING COSTS												
RAW MATERIALS												
COAL	171	171	171	171	171	171	171	171	171	171	171	171
ELECTRICITY	0	0	0	0	0	0	0	0	0	0	0	0
CHEMICALS	21	21	21	21	21	21	21	21	21	21	21	21
WATER	5	5	5	5	5	5	5	5	5	5	5	5
SUB-TOTAL	197	197	197	197	197	197	197	197	197	197	197	197
LABOR & OVERHEAD	62	62	62	62	62	62	62	62	62	62	62	62
REPLACEMENT	109	109	109	109	109	109	109	109	109	109	109	109
TOTAL OPERATING COST	368	368	368	368	368	368	368	368	368	368	368	368
GROSS MARGIN	698	698	698	698	698	698	698	698	698	698	698	698
NET CASH FLOW	698	698	698	698	698	698	698	698	698	698	698	698
TOTAL OPERATING COST	368	368	368	368	368	368	368	368	368	368	368	368
GROSS MARGIN	698	698	698	698	698	698	698	698	698	698	698	698
START-UP COST												
WORKING CAPITAL												
DISCOUNTED CASH FLOW	314	305	296	287	279	271	263	255	248	240	233	227

Appendix D

COAL PROCESSING PLANT - OVERALL PERFORMANCE DATA

1. COAL PROCESSING

1.1 Basis of Process Selection:

The discussion of general technical considerations in the preceding sections indicated that the available methods of utilizing Hat Creek coal included manufacturing the following:

a) Principally solid products.

- Combustion fuel for thermal power generation.
- Conversion to solvent refined coal by hydrogenation under high severity conditions.

b) Principally liquid products including:

- Conversion to liquid forms by hydrogenation under suitable conditions of high severity.
- Conversion of synthesis gas produced from coal to liquid forms by Fischer-Tropsch and methanol process technologies.
- Pyrolysis of coal to produce tars, oils and residual char for subsequent use.

c) Principally gaseous products including:

- Conversion to "low BTU gas" based on gasification with air.
- Conversion to "high BTU gas" based on gasification with oxygen.
- Conversion to Substitute Natural Gas (SNG) by methanation of high BTU gas.

The use of the coal as combustion fuel for thermal power generation is outside ths scope of this study. For the other selected processes in the above categories material and energy balances have been estimated based on producing approximately 316.5 TJ/d (equivalent to 3663 MW_t or 50,000 BPD petroleum derived fuel oil) of energy products, exclusive of energy value credit for byproducts sulfur and ammonia.

The major focus of attention in this study is coal conversion to liquids and solids by direct hydrogenation and indirect methods in which the coal is first gasified. Sources of data related to the direct liquefaction of Hat Creek Coal were as follows:

i) H-Coal process data is based on a recent study of that process published by EPRI* and based on data provided by the developer, Hydrocarbon Research Incorporated [Ref. 29]. That work has recently been amplified in additional studies funded by EPRI [30-33].

ii) Hat Creek data estimates for the Exxon Donor Solvent Process are based on application of the process to Wyodak coal, a western USA sub-bituminous coal, as published in reports on work funded by the Department of Energy (USA). [Ref. 37-38]

iii) Data on the SRC-II (liquid fuel product) and SRC-I (solid fuel product) are based on studies performed by the contractor in connection with the Northeast Coal Utilization Program NECUP. [Ref. 34-36]

The SRC-I and SRC-II processes are not generally considered to be suitable for processing coals of the Hat Creek type. However studies by NAMCO indicate that Hat Creek coal is a reactive variety which is possibly suitable for conversion by the SRC type processes [Ref. 16]. Recent discussions with the process developers indicate that SRC type technology is applicable to subbituminous coals of Wyodak type.

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*EPRI - The Electrical Power Research Insititute, Palo Alto, California.

The various processes referred to above related to direct coal hydrogenation are described in the following report material.

The sources of data related to indirect coal liquefaction were as follows: [Ref. 39-56]

i) Private files and estimates of cost and performance data for methanol manufacture and Texaco gasification.

ii) Published sources and previous studies for B. C. Hydro on Fischer-Tropsch technology and Lurgi gasification.

iii) Recently published data on use of Winkler gasification applied to lignites in the manufacture of methanol.

iv) Data on the gasification of sub-bituminous coals by the Koppers process as published by Koppers.

v) Concepts from the literature such as combination methods where the Lurgi process is used as the primary method and coal fines are disposed of in Texaco or Koppers type gasification units.

vi) Data on the Methanol to Gasoline (MTG) process reported in work funded by the Department of Energy.

The sources of data employed in estimates of the various processes for coal gasification (Texaco, Lurgi, Koppers and Winkler) and those for conversion of synthesis gas to liquids (Fischer-Tropsch, Methanol and Methanol to Gasoline are referenced in the report and Appendix D.

The list of processes or combinations of processes selected for evaluation are tabulated as follows: (Table D1.1)

Table D1.1 Processes Selected for Evaluation

			Appendix D	
Process Type	Process Description	Case	Figure	Table
A. Direct hydrogenation*	H-Coal	A1	D1.1	3.3
	EDS	A2	D1.2	3.3
	SRC-II	A3	D1.3	3.3
	SRC-I	A4	D1.4	3.3
B. Fischer-Tropsch synthesis	Texaco gasification	B1	D1.5	3.4
	Koppers gasification	B2	D1.6	3.4
	Winkler gasification	B4	D1.7	3.4
	Lurgi (Sell Fines) gasification	B4	D1.8	3.4
	Lurgi (Maximum Power) gasification	B5	D1.9	3.4
	Lurgi & Texaco combination	B6	D1.10	3.4
	Lurgi & Koppers combination	B7	D1.11	3.4
C. Methanol synthesis	Texaco gasification	C1	D1.12	3.5
	Koppers gasification	C2	D1.13	3.5
	Winkler gasification	C3	D1.14	3.5
	Lurgi (Sell Fines)	C4	D1.15	3.5
	Lurgi (Maximum Power)	C5	D1.16	3.5
	Lurgi & Texaco combination	C6	D1.17	3.5
	Lurgi & Koppers combination	C7	D1.18	3.5
D. Methanol to Gasoline	Texaco gasification	D1	D1.19	3.6
	Koppers gasification	D2	D1.20	3.6
	Winkler gasification	D3	D1.21	3.6
	Lurgi (Sell Fines)	D4	D1.22	3.6
	Lurgi (Maximum Power)	D5	D1.23	3.6
	Lurgi & Texaco combination	D6	D1.24	3.6
	Lurgi & Koppers combination	D7	D1.25	3.6
E. Production of synthetic natural gas (SNG)	Methanation based on Lurgi gasification	E1	D1.26	3.7

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*Cases A1, A2 and A3 produce liquid fuels and Case A4 (SRC-I) produces mainly clean solid boiler fuel by hydrogenation.

For the cases enumerated under the above studies, A, B, C and D, overall material and energy balances were estimated bsed on the avilable information for the various component subprocesses of which they are composed. Block flow diagrams, showing coal, air, water and power requirements and product rates, were drawn. Concise performance diagrams showing yields based on the energy in one tonne of coal were developed from the estimated overall performance data. This data is summarized in Figs. D1.1 - D1.26. and Tables 3.3 - 3.7* of the Report on the basis of coal properties summarized in Table 2.2* of the Report corresponding to the "As-received-corrected" basis.

In order to permit economic studies to be performed, capital cost estimates were also made for the various cases listed above. Literature and file cost data on the various processes and component subprocesses apply to different capacity plants at different points on the escalation curve. For each case adjusted estimates were prepared applying to a plant capacity of 316.5 TJ/d of products or 50,000 BPD Fuel Oil Equivalent (FOE), in 1980 Canadian Dollars.

Wherever possible, contingencies and safety factors have been eliminated from the estimated data; only basic erected plant costs are estimated. Capital costs are indicated to be "without adjustments" meaning that interest during construction, owners costs, working capital, starting costs and similar items are not included in them. It is intended that the sensistivity analyses be employed to assess the impact of various levels of contingency on the economic feasibility of the project. The addition of commonly assumed contingency levels to the quoted cpaital cost figures will elevate the costs considerably (See Tables 3.3 - 3.7*). Economic assumptions are discussed in detail in Section 5.

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*See Chapters 2 and 3 of the Report

1.2 Direct Coal Liquefaction

1.2.1 The H-Coal Process

Hydrocarbon Research Incorporated - Developer

Fig. D1.1 is a diagram of a feasible H-Coal process configuration. Other configurations may be preferred for use in liquefaction of Hat Creek coal. Feedstock coal is pulverized and dried in the coal preparation area. From coal preparation, the coal enters the coal hydrogenation (liquefaction) area.

Coal liquefaction typically comprises facilities for slurrying the coal with recycled oil, pumping coal-oil slurry to about 13.8 MPa, mixing it with hydrogen, heating it to about 460°C and hydrogenating the coal in an HRI-designed ebullated bed reactor to produce coal liquids. The ebullated bed reactor contains a turbulent fluidized bed of catalyst pellets which makes it suited for use in coal liquefaction since it has less tendency to plug than fixed bed reactors. Product from the coal liquefaction area is separated into gaseous and liquid fractions.

The gaseous fraction of the product stream passes through a sweetening and hydrogen recovery unit. From the gas treating area, desulfurized product gas passes into the fuel gas system.

The liquid fraction of the product stream passes into a product separation area where lighter liquids vaporize and the liquid stream is divided into two parts in hydroclones. Hydroclone overflow containing the lower solids concentration is recycled to the coal liquefaction operation (hydroclone system not shown in Fig. D1.1).

Hydroclone underflow with higher solids content, containing the net mineral residue and undissolved coal fed to the reactor, pass into the solids separation area of the plant.

If hydrogenation is severe and the plant produces synthetic crude oil, it will be feasible to distill product liquids and produce a solids rich residue for feed to the hydrogen unit referred to as Vacuum Tower Bottoms Slurry (VTBS). Product distillation under vacuum may not be feasible in the case where the boiler fuel production is to be maximized; where use of a process such as solvent de-ashing will be necessary due to the higher boiling nature of the product. Filtration would be a candidate process for application in solids separation but because of its associated mechanical problems and high cost, filtration is avoided in the H-Coal process.

In solvent de-ashing, a paraffine hydrocarbon called anti-solvent, is added to the hydroclone underflow where it causes the precipitation of a small quantity of "sticky" asphaltenic material. The precipitated material assists in agglomeration of the solids which are separated in settling tanks. It is anticipated that the liquid decanted from the top of the settling tanks will contain approximately 0.1 percent (weight) solids after anti-solvent separation from it.

The feed to the hydrogen plant is the solids containing stream from vacuum distillation or solvent de-ashing depending on type of product and residue separation. The Texaco Partial Oxidation Process is widely assumed to be used for this service.

For this study of Hat Creek coal, severe hydrogenation leading to the production of upgraded synthetic fuels is the basis of study.

1.2.2 Exxon Donor Solvent (EDS) Process
Exxon - Developer

The Exxon Donor Solvent Process, illustrated in Figure D1.2A employs technology which in areas other than that of coal liquefaction is used in petroleum refining. The processing sequence provides flexibility for liquefaction of different coals and control of product distribution based on market demand. It is not possible to draw up a detailed comparison of EDS and H-Coal operations on Hat Creek coal at the time of writing using data which is in the public domain.

In liquefaction, coal, solvent and hydrogen are reacted to produce gas, coal liquids, and a bottom stream containing the unreacted coal and mineral matter. The liquefaction liquid product is separated by distillation and the separated recycle solvent is catalytically hydrogenated in the solvent hydrogenation block to produce a specially active coal liquefaction solvent. Bottoms for distillation are subjected to coking to produce additional liquids. Hydrogen or fuel gas are produced by gasification of the coke produced as the other product of coking.

The EDS process uses a hydrogenated recycle solvent called the donor solvent. An example of the donor solvent molecule is tetralin. Figure D1.2B illustrates how tetralin donates hydrogen to free radicals formed by the disintegration of the coal during liquefaction and is converted to naphthalene. Naphthalene converts back to tetralin in the solvent hydrogenation step. Tetralin is one of many molecules which donate hydrogen. Ability to donate hydrogen is not the only criteria of coal liquefaction solvent characterization. The solvent transports coal into the liquefaction reactor and promotes dissolving of the coal particles. The composition of the recycle solvent is an important variable in the process. Exxon has characterized the quality and defined the molecular composition of preferred solvents as part of the foundation pilot work supporting current proposals to demonstrate the process.

Current EDS developments include introduction of a residue recycle step which diverts material from coker feed back to the coal liquefaction reactors, thereby increasing conversion and improving product quality. Residue may be gasified rather than coked in schemes more closely analogous to that described in the H-Coal process description. Vacuum bottoms recycle is indicated as a dashed (future) feature of the EDS process. With the use of vacuum bottoms recycle, the EDS process will be competitive with other processes and exhibit similar or superior efficiency, capital cost, hydrogen consumption and product quality.

1.2.3 SRC-II Type Technology
Gulf Oil - Developer

Figure D1.3 is a diagram of a more recent solvent refined coal process developed by Gulf Oil and designated SRC-II. The primary product of the SRC-II process is a low-sulfur distillate boiler fuel oil. Detailed comparisons of this processes applied to Hat Creek coal is not possible at this time.

SRC-II process technology is founded upon the recognition that mineral matter inherent in coal is capable of catalysis of coal liquefaction. The SRC-II process has the same range of flow schemes accessible to it as those of the other direct liquefaction processes. For Hat Creek coal, assuming satisfactory indigenous mineral catalyst properties, a scheme like that shown for the H-Coal process will probably be feasible.

The SRC-II process does not employ manufactured catalysts in coal liquefaction or hydrogenation of solvent recycles. Some differences are therefore seen in its performance with various coals due to variations in the catalytic properties of indigenous minerals. Preliminary tests by NAMCO indicated that SRC-II technology may be applicable in processing Hat Creek coal.

1.2.4 SRC-1 Type Technology

Gulf Oil, Southern Services, Department of Energy, Electric Power Research Institute - Process Developers.

Figure D1.4 is a block flow diagram of a process of the solvent refined coal (SRC-I) type. Coal is pulverized and slurried with coal derived solvent in the coal preparation and drying area. The slurry flows to the coal hydrogenation and gas recovery area where it is mixed with hydrogen and in excess of 90 percent of the coal is dissolved and hydrogenated. The process is accomplished without the aid of catalysts.

A slurry of solvent, mineral matter, unconverted coal and dissolved coal is separated in the coal hydrogenation area from a gaseous product fraction which may be internally used as fuel. The slurry then passes to a filtration or solvent de-ashing area where solid material is separated from it. The filtrate or de-ashed liquid solvent is heated and passes to a vacuum distillation section in the product and solvent recovery area where the solvent is separated and recycled for use in slurry preparation in the coal liquefaction area. Bottom product from vacuum distillation is molten solvent refined coal which may melt at about 200°C and is a major product of the process.

The process also produces a light fuel oil product which may be further separated into naphtha and medium fuel oil fractions in a commercial facility. However insufficient data are available to predict the naphtha/fuel oil split for Hat Creek coal with any precision.

Filter cake, if filters are used, is washed on the filters and sent to a drying operation in order to recover the wash solvent. The dried material is about 40 percent undissolved coal and the remainder ash. The dried material is a source of hydrogen or fuel gas following gasification by a Texaco or Koppers type of process.

In the solids separation area, the objective is to separate fine (1.0 to 40.0 micron) particles from the hot liquid solvent-refined coal. The molten solvent refined coal must be maintained at an operating temperature of approximately 290°C to establish a viscosity suitable for separation of solid material from it. At present, filtration is used in the SRC-I service. Its use requires the use of diatomaceous earth filter aid (precoat or body coat) because of the small filtered particle size. Because of the non-porous nature of the filter cake on the diatomaceous earth, continuous fresh filter aid surface regeneration is required or frequent cake removal by washing is required.

The type of filter which provides for continuous filter surface regeneration is the rotary precoat type. This filter consists of a rotating drum partially and horizontally immersed in a reservoir of product slurry. The surface of the drum is a mesh screen carrying a layer of filter aid. As the drum rotates, a knife blade pares off a skin of precoat together with a filter cake of solids deposited on the drum surface.

A wash solvent is sprayed onto the filter cake as it rotates to displace trapped coal liquids. Wash solvent is subsequently "dried" from the cake. Wash solvent passing into the filtrate is subsequently recovered in the vacuum distillation operation. Pressure in the filter is maintained by circulating an inert gas.

A rotary filter is used at the Tacoma (U.S.) pilot facility. At the Wilsonville pilot facility, a pressure leaf filter is used. The leaf filter uses a fixed filter surface and has a shorter operating cycle as a result of quick blinding of the precoat.

The drive toward SRC-II development is largely motivated by a desire to avoid problems encountered in filtration, drying, and solidification. However, SRC-I development is continuing, based on use of solvent de-ashing technology.

The current development program in the United States involves hydrogenation of the SRC further to produce coal liquids, rather than solid fuel, using the LC Fining (Lummus) process to hydrogenate the SRC to liquid products.

1.3 Gasification Processes

Table D1.2 summarizes data describing the performance of the various gasifiers studied in this project. The gasifier types are described in more detail in the following paragraphs:

1.3.1 Texaco Gasification Process

Texaco has considerable experience in partial-oxidation of liquid hydrocarbon residues, tars, and petroleum cokes. The Texaco SYN Gas process has been licensed since 1953 and there are 75 plants, with 160 gasifiers operating in 22 countries. The synthesis gas produced is used in ammonia, methanol, hydrogen, and chemicals production.

Figures D1.5, D1.12 and D1.19 illustrate applications of the Texaco coal gasification process to Hat Creek Coal. The Texaco coal gasification process operates at pressures between 2.1 - 8.3 MPa and incorporates a single stage, slagging,

Table D1.2

Estimated Data For Hat Creek Coal

Gasification Processes

Type of Gasifier Studied	Texaco	Koppers	Winkler	Lurgi
Applicability to Hat Creek Coal	doubtful	satisfactory	satisfactory	satisfactory
Operating Pressure Bar	20-80	1.0	1.0	20-30
Exit Temperature °C	1500-1600°C	1400-1600°C	300-1050°C	300-600°C
Oxygen t/t (coal substance)	0.31	0.25	0.20	0.15
Steam Oxygen Ratio t/t	-	0.2	2.0	7.0
Fuel Supply Form	water slurry	dry ground	small grains	dumped lumps
Fuel Supply condition	-	< 0.1 mm	2.8 mm	13.0 mm
Counter or Cocurrent	cocurrent	cocurrent	cocurrent	counter current
Classification Fixed/Fluidized/ Entrained Bed	entrained	entrained	fluidized	fixed, moving or descending bed*
Requirements on fuel	constraints on ash and moisture	none	reactive	non-caking, and non-swelling coals
Method of Heat Recovery	WHB prior to quench	WHB prior to quench	WHB prior to quench	quench prior to WHB
Gas yield Nm^3 $CO+H_2$/1000kcal HHV** of fuel charged (Hat Creek coal)	0.17	0.20	0.22	0.27
Byproducts	sulfur	sulfur	sulfur	tar, oil, naphtha, phenol sulfur ammonia. (hydrocarbons can be recycled to extinction)

*Equivalent terms

**In cases where gas yield is low, additional waste heat is generated which must be supplied by boilers in cases where gas yield is low.

entrained bed down-flow gasifier. The Texaco gasifier utilizes feed consisting of a concentrated slurry of coal particles in water fed in mixture with oxygen. A refractory-lined gasifier is used which is similar to those employed in Texaco's oil gasification process, except that provision is made to remove quenched slag through a water-sealed lock-hopper system.

The process is capable of gasifying a wide range of caking and non-caking bituminous and sub-bituminous coals, as well as petroleum coke. However, coal of the Hat Creek type is not an attractive feed since it requires a relatively large amount of energy to heat and vapourize the slurry water as well as the intrinsic moisture which does not contribute to the slurrying medium; and this coal is of high ash content.

The raw coal is first ground either wet or dry to a carefully controlled size distribution. Control of the size distribution is important to maximize the coal concentration in the resultant slurry made in the slurry preparation tank.

The slurry is pumped to a burner where it is mixed with oxygen and partially oxidized to a synthesis gas rich in carbon monoxide and hydrogen at temperatures of 1300 - 1480°C. As the reactants pass through the reaction zone, the ash is slagged and drops into a water quench pot at the bottom of the gasifier. The slag, which contains a small concentration of unreacted carbon, is removed through a pressurized lock hopper.

The hot synthesis gas passes through an outlet located below the gasification zone but above the slag quench pot. It then passes through a waste heat boiler and is contacted with water in a scrubber operating at gasifier pressures to remove entrained particulates. Water removed from the scrubbing system and the slag quench pot is recovered through a settler where the particulates are extracted and recycled to the gasifier. The water scrubbed synthesis gas is subjected to further cooling and purification.

The Texaco coal gasifier is in development in two pilot gasifiers located at Texaco's Montebello Refinery, California which are designed to operate at:

2.4 MPa	15-20 tonnes per day solids feed, and
8.3 MPa	15-20 tonnes per day solids feed.

There is also a 150 tonnes per day demonstration plant at Ruhrchemie in Oberhausen, West Germany, which has thousands of hours of operation at pressures ranging from 2.1 - 8.3 MPa

The high pressure operation (2.1 - 8.3 MPa) of the Texaco coal gasifier is an advantage in that compression can be minimized or eliminated in conveying synthesis gas to a medium or high pressure downstream loop for synthesis of methanol. Use of a slurry feed eliminates the use of coal lock-hoppers, which require a considerable amount of maintenance.

An important advantage of this process is the high temperature operation which produces a clean synthesis gas, free of tars, oils, and phenols. The high carbon monoxide and hydrogen content of the gas, with negligible methane, make the gas a good candidate for methanol production.

The use of a water slurry feed system makes this process unsuitable for application with Hat Creek type coals. The oxygen requirements for combustion of a slurry of Hat Creek coal are about twice those for a slurry of low moisture bituminous coals. The gasifier is suited for use with bituminous and sub-bituminous coals.

1.3.2 Koppers Gasification Process

Figures D1.6, D1.13, and D1.20 illustrate processes using Koppers gasification. The Koppers gasifier operates at near atmospheric pressure. Gasification is carried out in a refractory lined vessel which is water cooled. Each gasifier has two or four coal injection heads (burners) through which steam, oxygen, and pulverized coal enter the unit via screw feeders. Velocities in the mixers are maintained high to avoid "backflashing" out of the gasifier.

The temperature at the burners is in the region of 1900°C and the gas, cooled by radiation to cooled studs in the ceiling, leaves the radiant chamber at about 1480°C.

Ash is melted to form a slag in the gasifier. About half of the slag produced drops into a quench water reservoir in the base of the gasifier. The remainder passes, entrained, overhead. The entrained material is frozen by injecting steam into the upper part of the gasification chamber and use of additional radiation cooling surface in the gasifier ceiling. The gas and non-sticky cooled particles pass into a waste heat boiler where high pressure steam is produced. Solids are removed from the crude cooled gas using cyclone and electrostatic methods and by water scrubbing. The clean gas is subsequently compressed to the process conditions required for shift conversion, acid gas removal and subsequent processing.

In shift conversion, the water gas reaction is employed to adjust the carbon monoxide ratio to the value required by the subsequent processing steps. Differing ratios are called for according to whether SNG, methanol or Fischer-Tropsch processing is employed.

The acid gas removal operation involves the separation of carbon dioxide and hydrogen sulfide from the gasifier product stream. The Rectisol process has been employed for acid gas removal in all studies on which this study is based. The Rectisol process has the capability to reduce sulfur contaminants, hydrogen sulfide and carbonyl sulfide, to the low levels (below one tenth part per million) required for avoiding catalyst poisoning in methanol, Fischer-Tropsch and SNG manufacture.

The Koppers gasifier product is low in methane content and contains no heavier hydrocarbon contaminants. It is well suited for methanol manufacture where contaminants are minimized in the product. The Koppers system has the capability to process a wide range of coals and may be considered to be fully commercially demonstrated. The need to compress gas produced in the Koppers gasifier leads to the necessity to cool the gas prior to compression and to reheat it prior to shift conversion. A substantial amount of steam is added to the gas prior to shift conversion, this represents a thermal burden tending to reduce the thermal efficiency of the process.

Factors to be investigated in connection with a proposed venture involving Koppers gasification include:

- Solid gas separation requirements
- Oxygen consumption and relative level compared to other gasifiers
- Requirements for coal pulverizing and drying
- Cost impact, if any, of low pressure operation
- Need to use flux to control ash fusion properties
- Expected waste heat boiler life under erosive and corrosive action of entrained particulates.

1.3.3 Winkler Gasification Process

Figures D1.7, D1.14 and D1.21 illustrate processes using Winkler gasification. The Winkler gasification process employs a fluidized bed. The height-to-diameter ratio of the bed is selected to permit the near completion of gasification in the bed. Secondary oxygen and steam are added above the bed in order to maximize carbon conversion to synthesis gas. Carbon remains in the fly ash taken overhead from the gasifier. Ash also is removed from the bottom of the gasifier and this also tends to carry carbon from the unit.

Crushed, dried, and sized coal is screw fed into a base of the gasifier from a feed bunker system. The gasifier operates at near atmospheric pressure at an average temperature in the region of 800°C to 900°C.

Gases leaving the gasifier pass through a waste heat boiler, cyclones for carbon recovery and ash removal, and water scrubbing to achieve a high degree of solids removal.

The Winkler gasifier is ideal for gasification of reactive lignites of the Hat Creek type. It should have a high ash fusion temperature to avoid the formation of clinkers in the bed.

The scrubbed cooled gas is compressed to the process conditions required for shift conversion, acid gas removal and subsequent processing. In shift conversion, the water gas reaction is employed to adjust the carbon monoxide ratio to the value required by the subsequent processing steps. Differing ratios are called for according to whether SNG, methanol or Fischer-Tropsch processing is subsequently employed. The acid gas removal operation involves the separation of carbon dioxide and hydrogen sulfide from the gasifier product stream. The Rectisol process has been employed in the studies on which this study is based. The Rectisol process, which employes methanol as the solvent, has the proven capability to reduce sulfur contaminants to the low levels required to insure long catalyst life in downstream processing.

1.3.4 Lurgi Dry Bottom Gasification Process

Figures D1.8, D1.10, D1.15-D1.18, D1.22-D1.26 illustrate process flow schemes employing Lurgi (dry bottom gasification). Lurgi experience with the fixed bed (also variously referred to as "moving bed" and "descending bed") gasifier dates from the 1930's. A total of 16 plants have been built using 65 gasifiers, excluding the Sasol II plant but including Sasol I. The Sasol II complex, with an additional 36 gasifiers, is currently in its early operating phases. A duplicate of Sasol II is currently under construction.

The Lurgi gasifier has undergone extensive testing and may be considered to be commercially available for certain lignites and sub-bituminous coals. The gasifier is a water-jacketed vessel system which operates at up to about 2.8 MPa and temperatures below the ash softening point in the bottom section. Temperature at the gas exit at the top of the gasifier will approximate 300°C with Hat Creek coal. Sized coal (50 x 6 mm) is introduced to the top of the coal bed in the gasifier through a pressurized lock hopper and mechanically distributed by a gear driven arm. Steam and oxygen are introduced at the bottom of the gasifier through a revolving grate. Dry ash falls through the grate into an ash lock chamber for discharge into an ash disposal system.

As the coal moves down through the gasifier bed, drying, devolatilization, gasification and partial combustion occur sequentially. The partial combustion zone at the base of the gasifier supplies the heat required by the endothermic gasification reactions going on in the upper part of the bed. Gasification activity is evident in the bed at temperatures down to about 650°C. Temperature is further reduced below 650°C to about 300°C by the process of vaporizing water and volatile matter from the coal.

Crude Lurgi gas contains tar, oil, naphtha, phenols and ammonia as a result of the low temperature experience of the coal in the final stages of the process (top of bed). Part of the tars and any entrained coal dust are removed in a scrubber and recycled to the gasifier. The cleaned gas, or part of it, is then subject to shift conversion following steam addition in order to achieve proper reactant ratios for the water gas reaction.

Some opportunities are taken to recover heat from the Lurgi gasifier effluent gas in waste heat boilers. Tar fouling and related problems tend to reduce the opportunities for waste heat recovery. The advantages of Lurgi gasification include an attractively low oxygen consumption and high operating pressure, which reduces or eliminates synthesis gas compression requirements. With coals of the Hat Creek type, methane production tends to be high, making the gasifier the best suited for SNG manufacture.

The unattractive features of the Lurgi fixed bed gasifier are the byproduct tar, oils, phenols and ammonia, which lead to high environmental protection costs. In this study all undesirable hydrocarbons products have been recycled to extinction to the gasifiers. The need to feed sized coal to the Lurgi gasifier leads to an associated fines disposal problem.

In this study, fines which cannot be gasified are subject to various utilization strategies:

i) Sale of fines, as in the Lurgi (Sell Fines), L(SF), cases.

ii) Combustion of fines in boilers as in the Lurgi (Max. Power), L(MP), cases.

iii) Gasification of the fines in a gasifier suitable for the purpose.
Lurgi-Koppers and Lurgi-Texaco combination cases are considered.

1.3.5 British Gas Council - Lurgi Slagging Process

In association with British Gas Corporation, Lurgi has developed a slagging gasifier of the fixed bed type. This gasifier is now offered commercially. The gasifier is similar to the dry bottom Lurgi type except that ash is tapped from the base as a slag and the grate in the dry bottom unit is eliminated. The gasifier uses low oxygen and low coal to steam ratios and as a result exhibits good coal throughput characteristics.

1.3.6 Gas Purification Processes

The raw gas produced by coal gasifiers must be cleaned of impurities and the ratio of hydrogen and carbon monoxide must be adjusted prior to synthesis or methanation in order to attain the most effective production of the desired energy products. Particulates and hydrocarbons are typically removed in water quench systems that function as effective scrubbers, and are usually an integrated part of the gasification system. Other impurities which must be removed include hydrogen sulfide and carbonyl sulfide, to prevent catalyst poisoning and environmental releases, and carbon dioxide which acts largely as a process and product diluent.

The technology for removal of these acid gases has been widely demonstrated in a variety of industrial applications. A number of process systems exist which can be utilized for acid gas removal; however, operational characteristics vary somewhat and the selection of the most appropriate gas cleanup system is dependent upon the desired product and the operating conditions of the various process elements in the overall system. The selection of the gas cleanup system is influenced by a number of factors including potential for contaminating the synthesis gas, utility requirements, chemical makeup needs, gas contaminant levels in cleaned synthesis gas, and maintenance requirements.

The Rectisol process, developed by Lurgi, is particularly well suited to the Lurgi gasification process because it removes and isolates the naphthas produced in gasification, sulfur compounds to the very low concentration levels required to protect methanation and synthesis catalysts, and carbon dioxide. The process uses methanol as the solvent.

1.4 Indirect Liquefaction Processes

The following processes produce coal-derived liquid hydrocarbons and other liquids by hydrogenation of carbon monoxide in suitably produced synthesis gases.

1.4.1 Fischer-Tropsch Process (FT)

Carbon monoxide is reduced by hydrogen in FT to produce a range of oxygenated and regular hydrocarbon products. Iron catalysts are used in current commercial projects. Synthesis gas can be derived by steam hydrocarbon reforming for this process; Exxon practiced FT based on steam reforming at Bayway, New Jersey, but the process was not economical and was abandoned.

The FT process is carried out in part in a fluidized bed reactor in the Sasol (SA) applications of the Synthol process. These reactors have been developed over a period of 20 years since startup of Sasol I. The reactors are about 36 m high by 2.2 m diameter, and cooling loops remove the high level of heat of reaction which is evolved. About 25 percent of the heating value of the synthesis gas is liberated in the FT synthesis reactors.

A generally referenced problem in FT synthesis is its lack of specificity of product type or catalyst selectivity. The range of products possible is illustrated by the range quoted in the FT cases of this work. The requirements to separate the many products and fractions leads to the provision of extensive separation and refining facilities at FT sites.

The primary FT reaction is (C&E News, Feb. 23rd 81):

$$nCO + 2nH_2 = (-CH_2-)_n + nH_2O + n(201\ KJ)$$

Fixed bed reactors (Arge synthesis) are also employed in FT synthesis. The fixed bed process produces less transportation fuels than the fluidized bed process but more heavy compounds and waxes.

1.4.2 Methanol Synthesis Processes

Two process routes exist for methanol synthesis which are commercially available. One operates at a high pressure range of 21 - 35 MPa, the other operates at a low pressure range of 4.8 - 10.3 MPa. Most modern methanol plants use a low pressure methanol synthesis process employing highly active copper based catalysts.

The leading low pressure methanol processes are the Lurgi Low Pressure Methanol Process, and the Imperial Chemical Industries (ICI) Low Pressure Methanol Process.

The major difference in the two processes, besides that in the catalyst used, is the method employed to remove the heat of reaction in the reactor. Lurgi incorporates a tubular reactor in which the synthesis occurs at 5.2 - 8.3 MPa in the catalyst-packed tubes and the heat of reaction is absorbed by boiling water on the shell side. This results in a uniform catalyst operating temperature of 230°C - 260°C along the length of the reactor, and minimizes byproduct formation. Most of the exothermic heat of reaction is converted directly into medium pressure steam.

ICI utilizes a direct quench system wherein the reaction takes place in a packed bed of catalyst at 5.2 - 10.3 KPa in which the gas temperature rises as the methanol synthesis occurs. The gas is cooled at several points along the bed by injecting cold synthesis gas in order to avoid excess temperature rise beyond the 200 - 300°C operating range.

Both the Lurgi and ICI processes are used in commercial methanol plants operating mainly on natural-gas or petroleum-liquid derived synthesis gas.

1.4.3 Methanol to Gasoline Process (MTG) (Mobil)

The Mobil Corporation has developed a zeolite catalyst (ZSM-5) which is capable of catalysis of the conversion of methanol to high octane gasoline and lighter hydrocarbons. Gasoline is the dominant product of the reaction. Propane and butane products are potentially convertible to additional gasoline by alkylation. Compounds heavier than decane tend not to be formed due to the molecular sieve action of the catalyst. The hydrocarbon product consists of highly branched paraffins and olefins, naphthenes and aromatics.

The MTG process, like Fischer-Tropsch, is exothermic, though not to such a large extent. The process can be based on fixed, tubular or fluidized bed reactors. Fluidized bed reactors have been found to offer the best gasoline yields and development is focused in this direction. A pilot unit in West Germany is operating at a conversion capacity of 100 t/d.

In the MTG type process, fresh and recycled feed methanol enter the reactor where gasoline and other hydrocarbons are produced together with by-product water. Gasoline, water other products and methanol are separated from the reacter effluent stream and recycled or removed from the plant as products. In the situation where methanol is converted to gasoline, water from the conversion step can be recycled for use in the gasification of coal, thereby conserving water in the neighborhood of the plant. Water recovered in the conversion of methanol to gasoline would in effect, be exported from the site in situations where methanol is the single main coal conversion product.

1.5 Manufacture of Synthetic Natural Gas (SNG)

The synthesis of methane is accomplished by conversion of carbon monoxide and hydrogen to methane and water in the presence of a nickel-based catalyst. Extensive small scale and semi-commercial experience exists for two methanation processes, one developed by Lurgi, the other by Conoco, both suitable for use with coal-derived synthesis gas.

1.5.1 Lurgi Methanation Process

The Lurgi methanation process is a hot gas recycle method which was demonstrated successfully at Sasol's coal gasification facility in Sasolburg, South Africa in 1973 - 1974.

The system involves two packed catalyst beds in series, each followed by a heat exchanger to remove some of the heat of reaction. Acid gas removal and carbon monoxide shift to hydrogen to obtain a hydrogen/carbon monoxide ratio of about 3 are required for the process. An SNG product with less than 1.0 percent by volume of hydrogen and a higher heating value of about 36,200 KJ/m^3 after drying and carbon dioxide removal has been produced by the process.

1.5.2 Conoco Methanation Process

The Conoco methanation process was demonstrated at the British Gas Corporation (BGC) gasification testing unit at Westfield, Scotland in 1974. The source of the synthesis gas was a Lurgi Gasifier, and the maximum throughput yielded up to 71,000 cubic meters per day of high BTU SNG which was fed into the local gas grid.

The process consists of three primary methanation reactors in series. Cooled product gas from the third reactor is recycled and combined with synthesis gas for each reactor for temperature control. Recycled product gas acts as a heat sink and limits the temperature rise in each reactor.

The process is designed to maximize the recovery of the heat of methanation by generating steam in waste heat boilers located after the reactors.

1.6 Coal Pyrolysis

During recent years at least seven pyrolysis processes have been under development -

Clean Coke Process - U.S. Steel Corporation
COED - FMC
COGAS - Cogas Development Co.
Garrett's Coal Pyrolysis - Occidental Petroleum Corp.
Lurgi-Ruhrgas - Lurgi GmbH/Ruhrgas AG
Project Seacoke - ARCO Chemical Corp.
Toscoal - The Oil Shale Corporation

In 1978 plans to construct a major demonstration plant were first announced by the U.S. Department of Energy and subsequently withdrawn.

The only fully commercial process is the Lurgi-Ruhrgas, a first plant having gone into operation in 1965 (Yugoslavia) since which time a large number of plants, usually operating with low rank coals, have been erected in various countries.

Development of the other processes listed is not being vigorously pursued at the present time.

1.6.1 The Lurgi-Ruhrgas Process

Lurgi GmbH, Frankfurt, and Ruhrgas AG are the developers of this process.

Feed coal and a heat carrier consisting of hot char are continuously supplied to a mechanical mixer which ensures a uniform mixing of the two components as well as a very rapid equalization of temperature between the char and coal so that a major part of the carbonization occurs at the end of the mixer. The resultant pyrolysis gas and vapors are withdrawn at the end of the mixer, passed through a cyclone for dust removal, and then sent to a condensing unit.

The tar is subjected to dust removal and hydrogenated in the hydro-treatment section to produce a range of liquid products. The gas after cleaning, has a heating value of 26,100-31,670 KJ/kg. This gas may be used as a source of hydrogen for the tar-hydrogenation steps or methanated to pipeline quality.

The char which has been used as a heat carrier and newly-formed char, fall into the carbonizer shaft where additional temperature equalization between the heat carrier and fresh distillation residue takes place so that a subsequent degasification can occur. The char leaves the carbonizer shaft at the lower end and flows to a lift pipe where it is raised by combustion gases and heated simultaneously.

The combustion gases are produced in the lift pipe itself, into which preheated air is blown to cause partial combustion of the char. Char and combustion gases are separated and the gases, after cleaning, are exhausted.

The hot char is collected in a bin and then recirculated to the mixer to complete the cycle. The continuous production of fresh char results in a surplus of circulating char. This surplus is continuously withdrawn and used for steam and electricity production.

The close intermixing of coal and hot char in the mixer avoids the formaton of agglomerates,so that caking coals can be treated.

Operating conditions in the carbonizer are as follows:

Reactor	Temp. °C	Pressure	Reactants	Products
Carbonizer	595	Atmospheric	Coal-Hot Char	Char,Tar,Gas

A materials balance and calculated overall thermal efficiency for raw Hat Creek coal is approximately as follows (see Fig. D1.27), assuming the most favorable yield of liquids:

Liquid product yield is 6.5 percent. Gas yield is 12 percent.

Electric power as coal equivalent (gross calorific value gases) is 21.6 percent of coal feed to plant.

Thermal efficiency is 40.1 percent.

A commercial production plant would have the following performance.

The coal feed for the plant should be 19.3 million t/a run-of-mine coal at an on-stream factor of 90 percent.

For this capacity the major primary units would be:

- Coal-drying unit
- Lurgi-Ruhrgas carbonization-unit with 12 carbonizers including quench- and waste-heat systems
- Hydrocarbons recovery with gas treatment, tar treatment and carbonization-water treatment.

Secondary process units:

- Power- and steam plant
- Flue-gas treatment
- Make-up water unit
- Cooling water unit

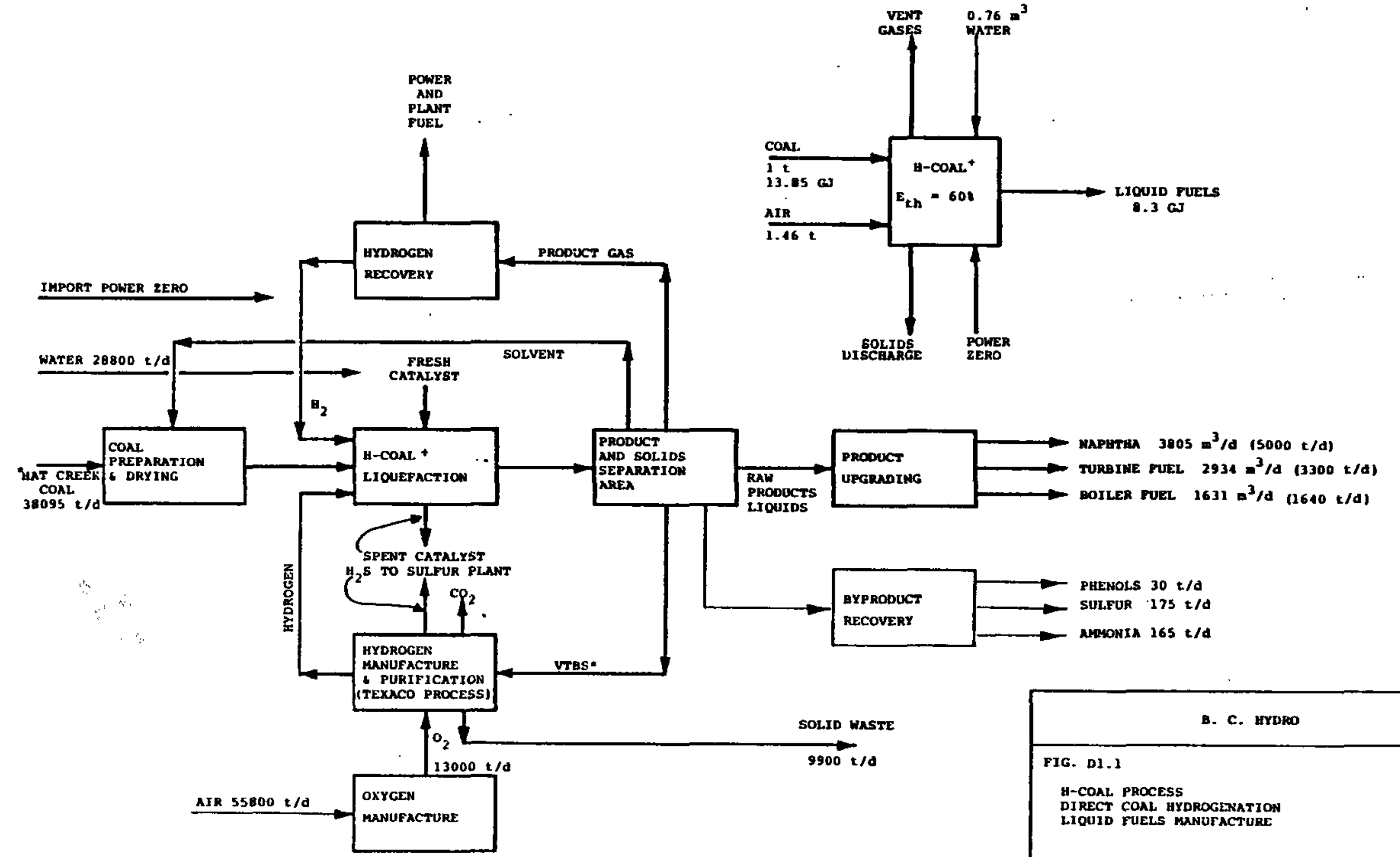
Feeds

Coal for cabonization	19.3	10^6 t/a
Make-up-water	19.3	10^6 t/a

Products

Gas	1424.4	10^6 m^3/a
Tar and gas oil	326.8	10^3 t/a
Gasoline	60.9	10^3 t/a
Phenols	28.9	10^3 t/a
Electric Power	14.3	10^6 KWH/a

Total Products 316.5 TJ/d



*SEE NOTES FIGS. D-1 TO D1-26 (FOLLOWING FIG. D1.26)

+H-COAL IS THE TRADE NAME REG. BY HYDROCARBON RESEARCH INC.

B. C. HYDRO

FIG. D1.1

H-COAL PROCESS
DIRECT COAL HYDROGENATION
LIQUID FUELS MANUFACTURE

NJ 1024

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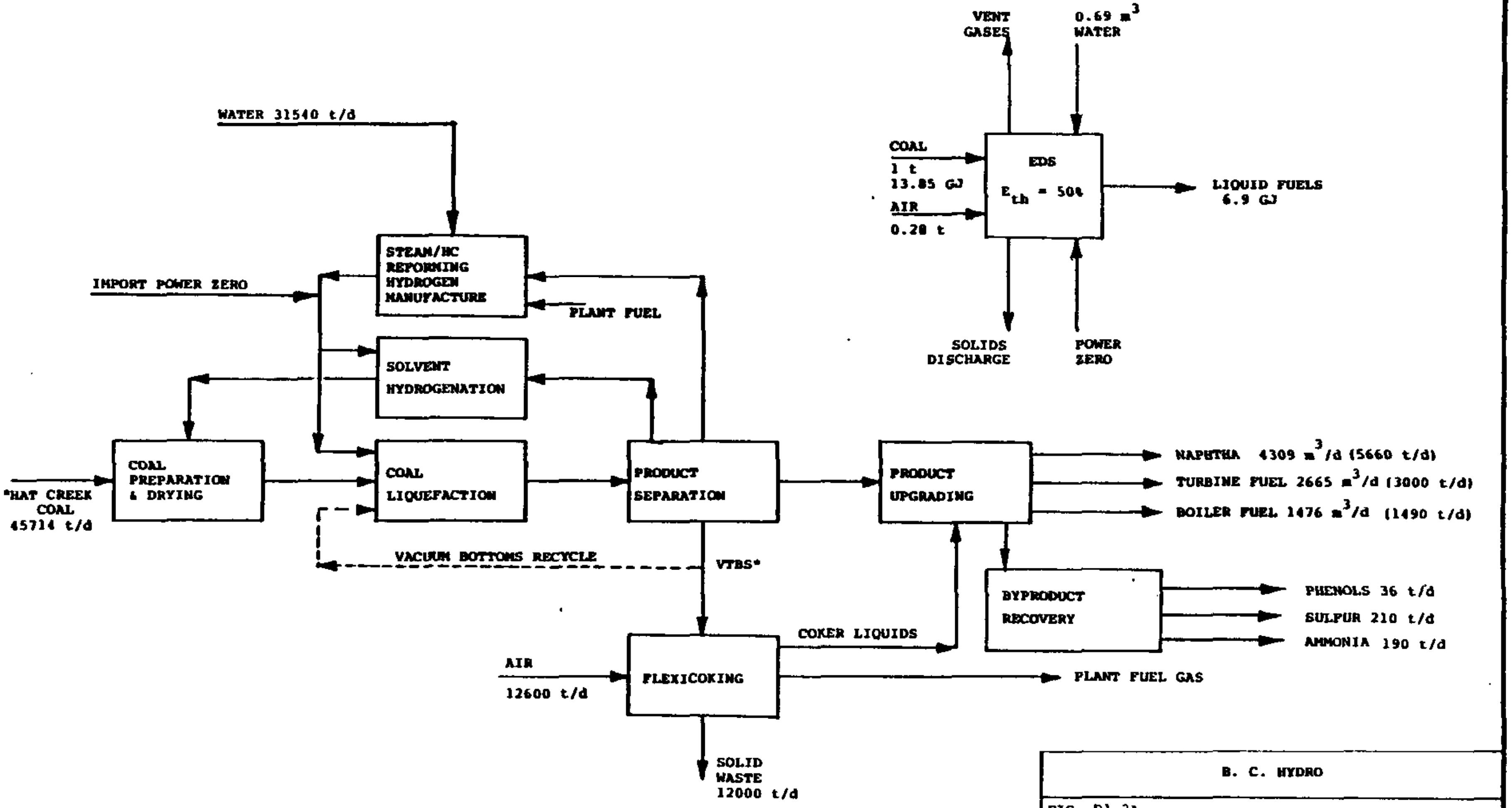

*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.2A

EXXON DONOR SOLVENT PROCESS TYPE
DIRECT COAL HYDROGENATION
LIQUID FUELS MANUFACTURE

NJ 1024

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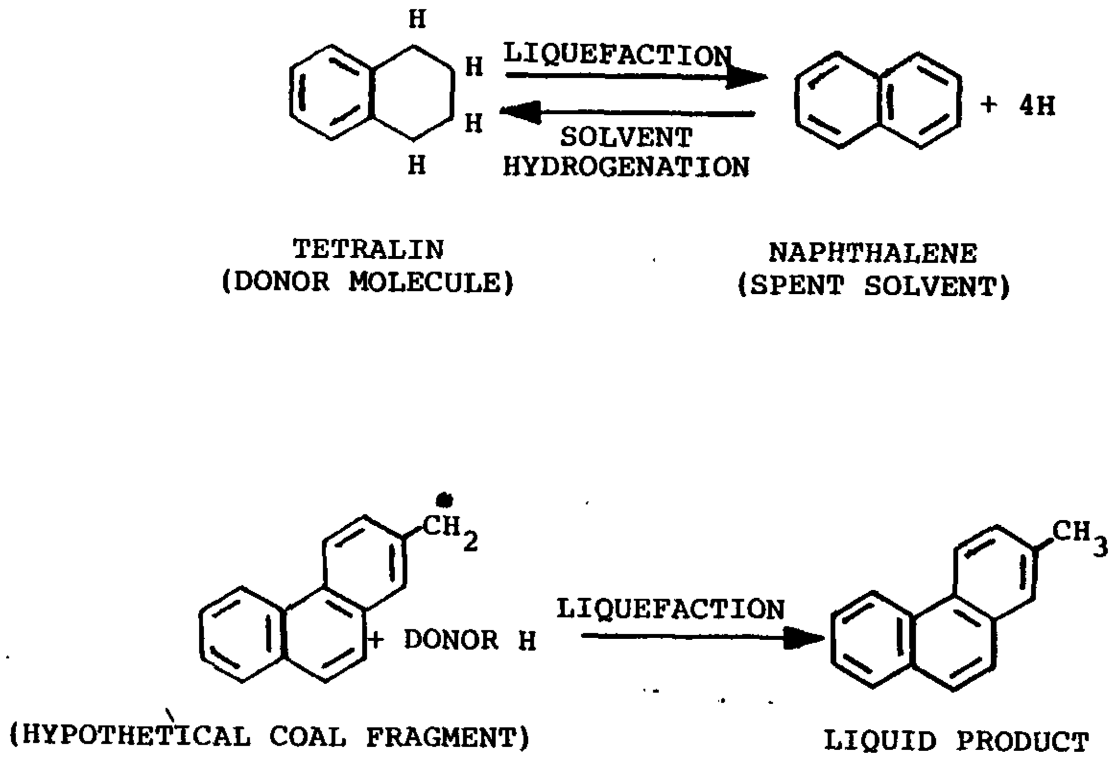

B. C. HYDRO

FIG. D1.2B

MECHANISM OF EXXON DONOR SOLVENT PROCESS -- EXAMPLE

NJ 1024

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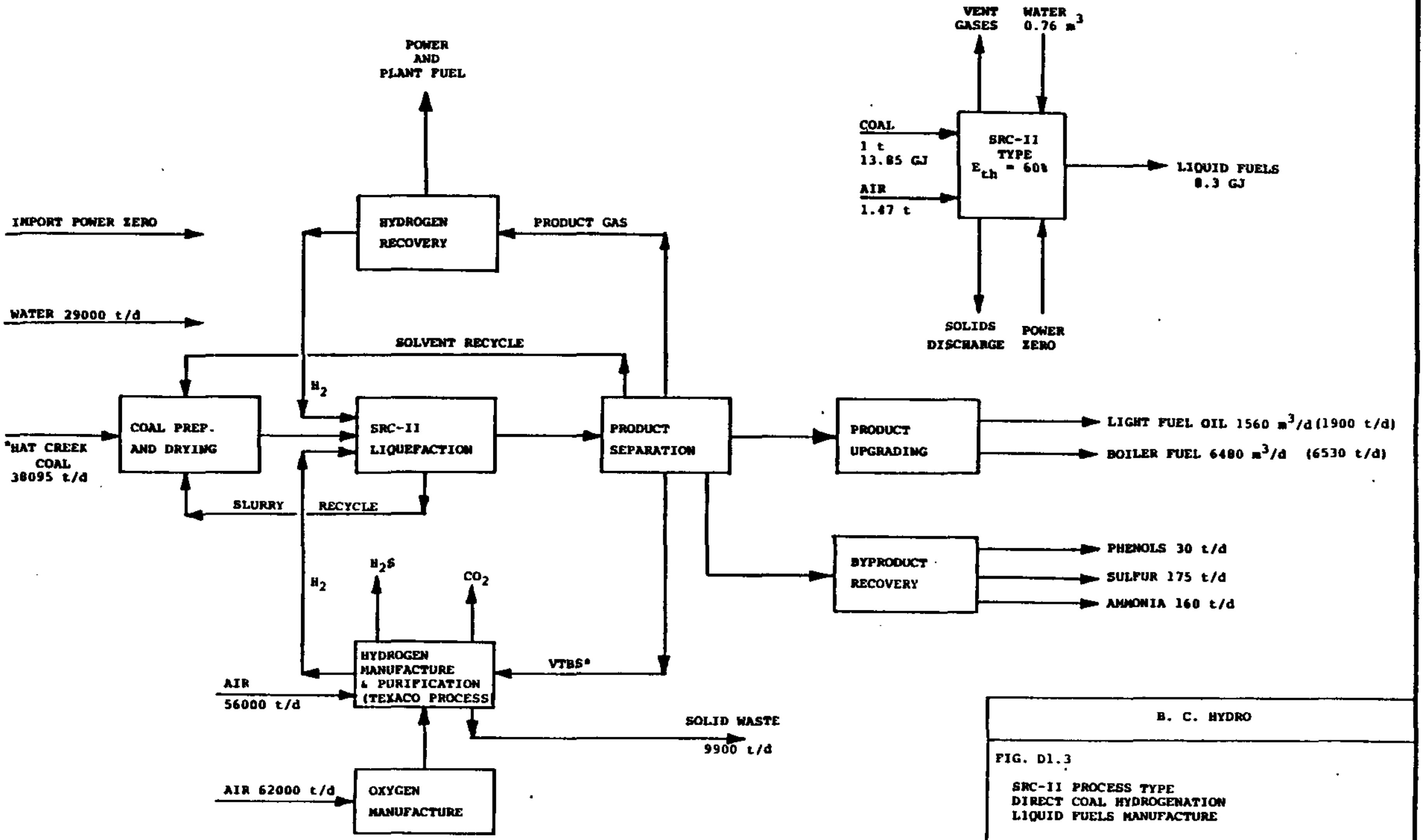

*SEE NOTES D1.1 to D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.3

SRC-II PROCESS TYPE
DIRECT COAL HYDROGENATION
LIQUID FUELS MANUFACTURE

NJ 1024

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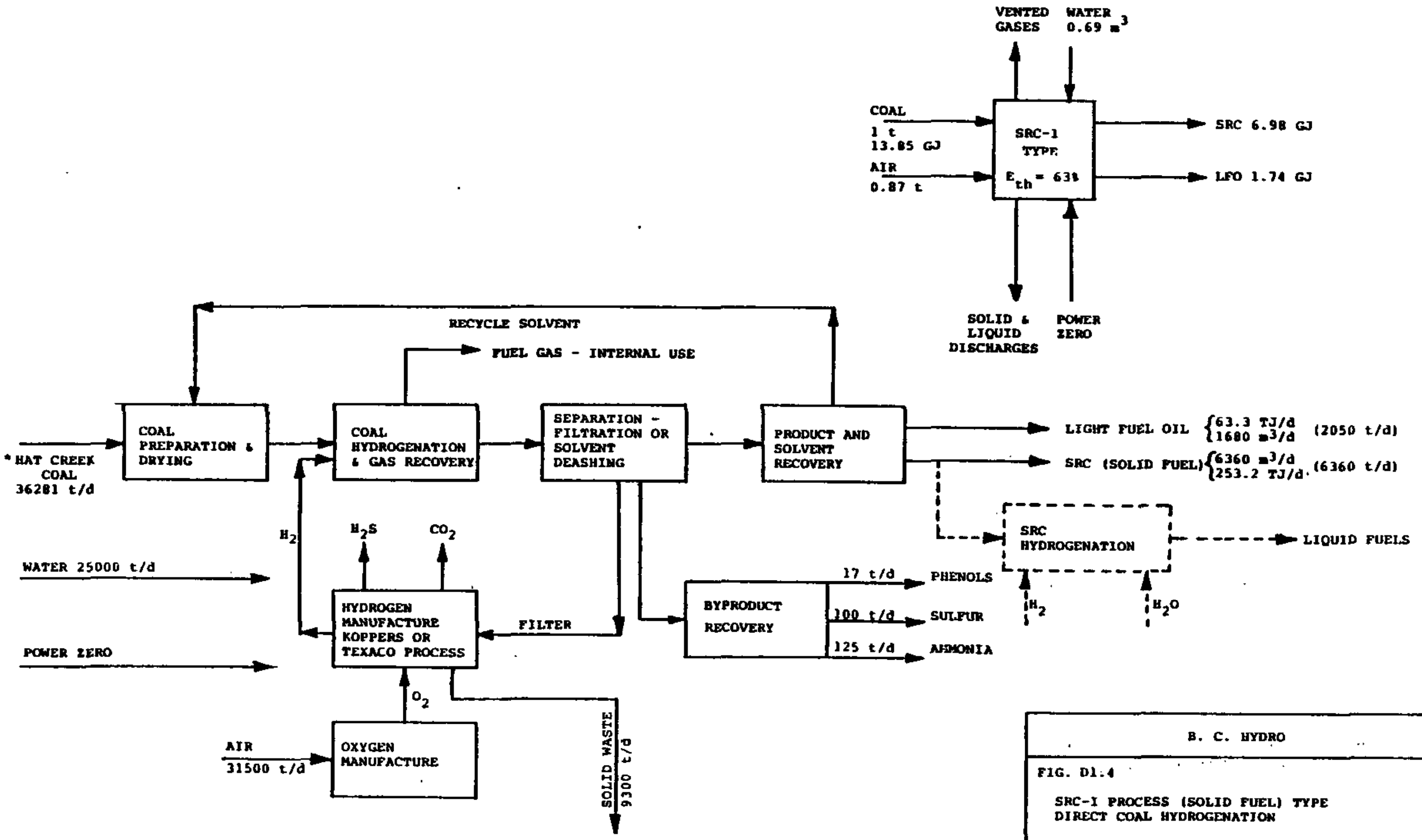

*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.4

SRC-I PROCESS (SOLID FUEL) TYPE
DIRECT COAL HYDROGENATION

NJ 1024

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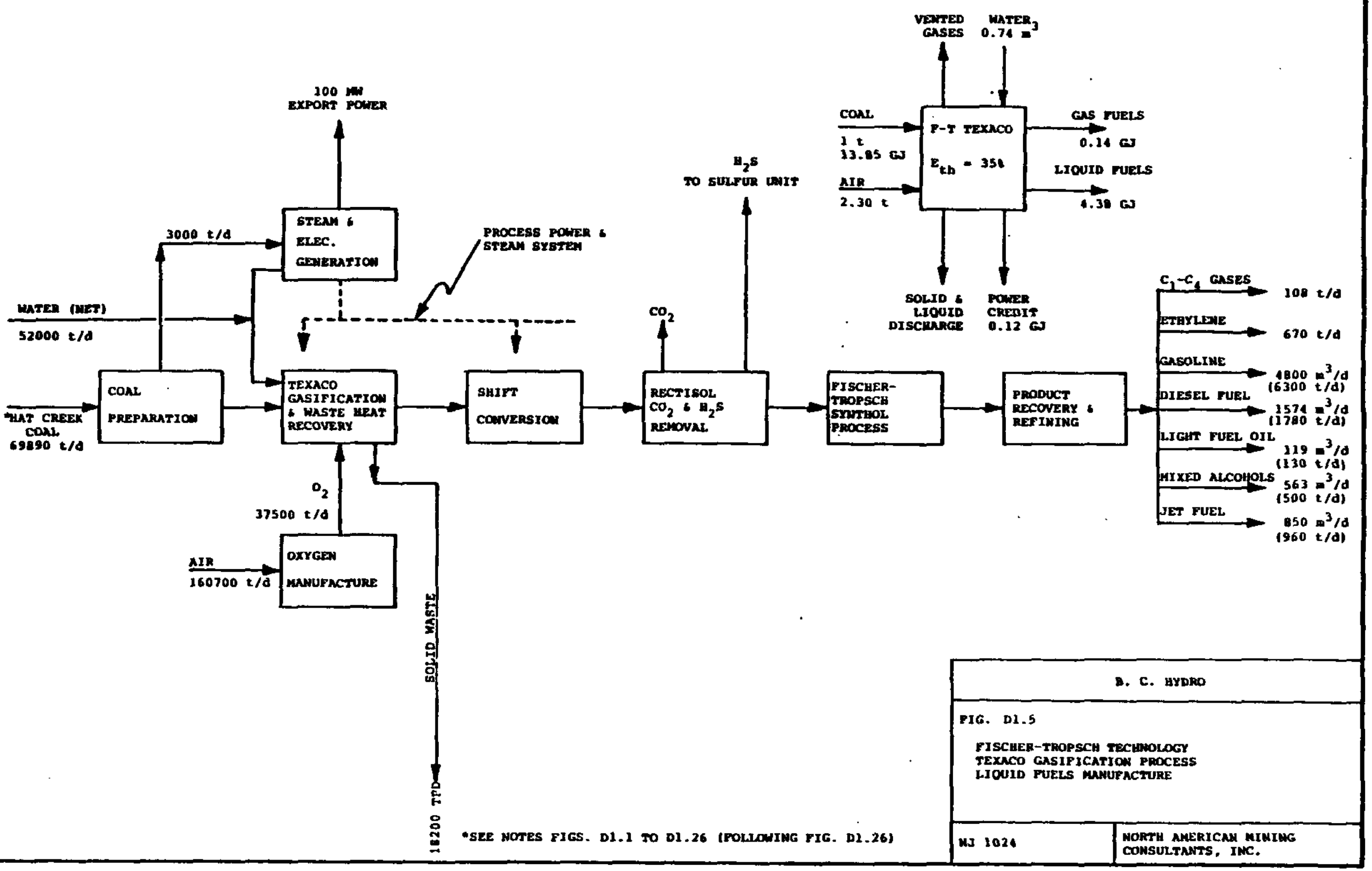


*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

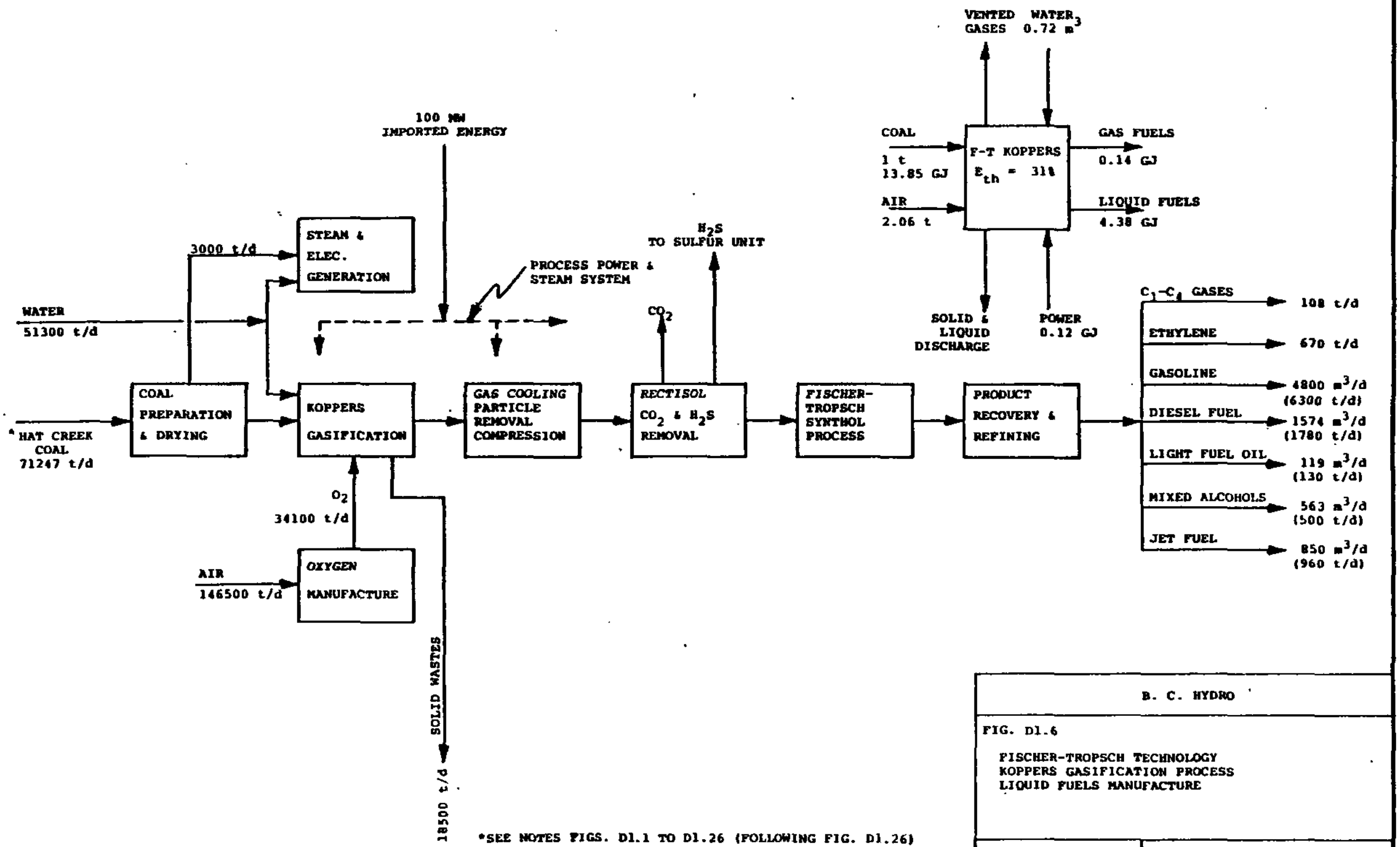
B. C. HYDRO

FIG. D1.5

FISCHER-TROPSCH TECHNOLOGY
TEXACO GASIFICATION PROCESS
LIQUID FUELS MANUFACTURE

NJ 1024

NORTH AMERICAN MINING
CONSULTANTS, INC.



*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.6

FISCHER-TROPSCH TECHNOLOGY
KOPPERS GASIFICATION PROCESS
LIQUID FUELS MANUFACTURE

NJ 1024

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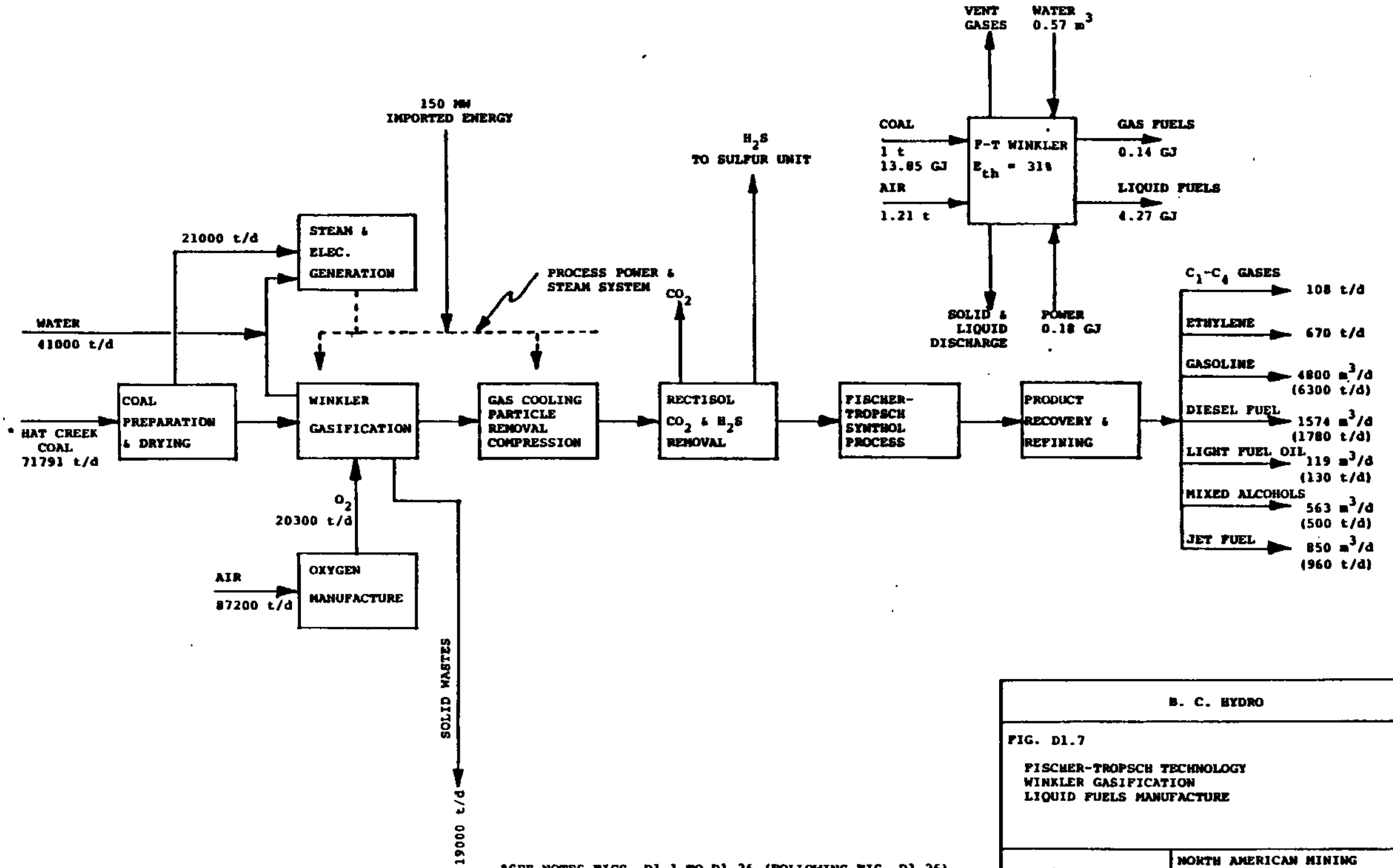

*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.7

FISCHER-TROPSCH TECHNOLOGY
WINKLER GASIFICATION
LIQUID FUELS MANUFACTURE

MJ 1024

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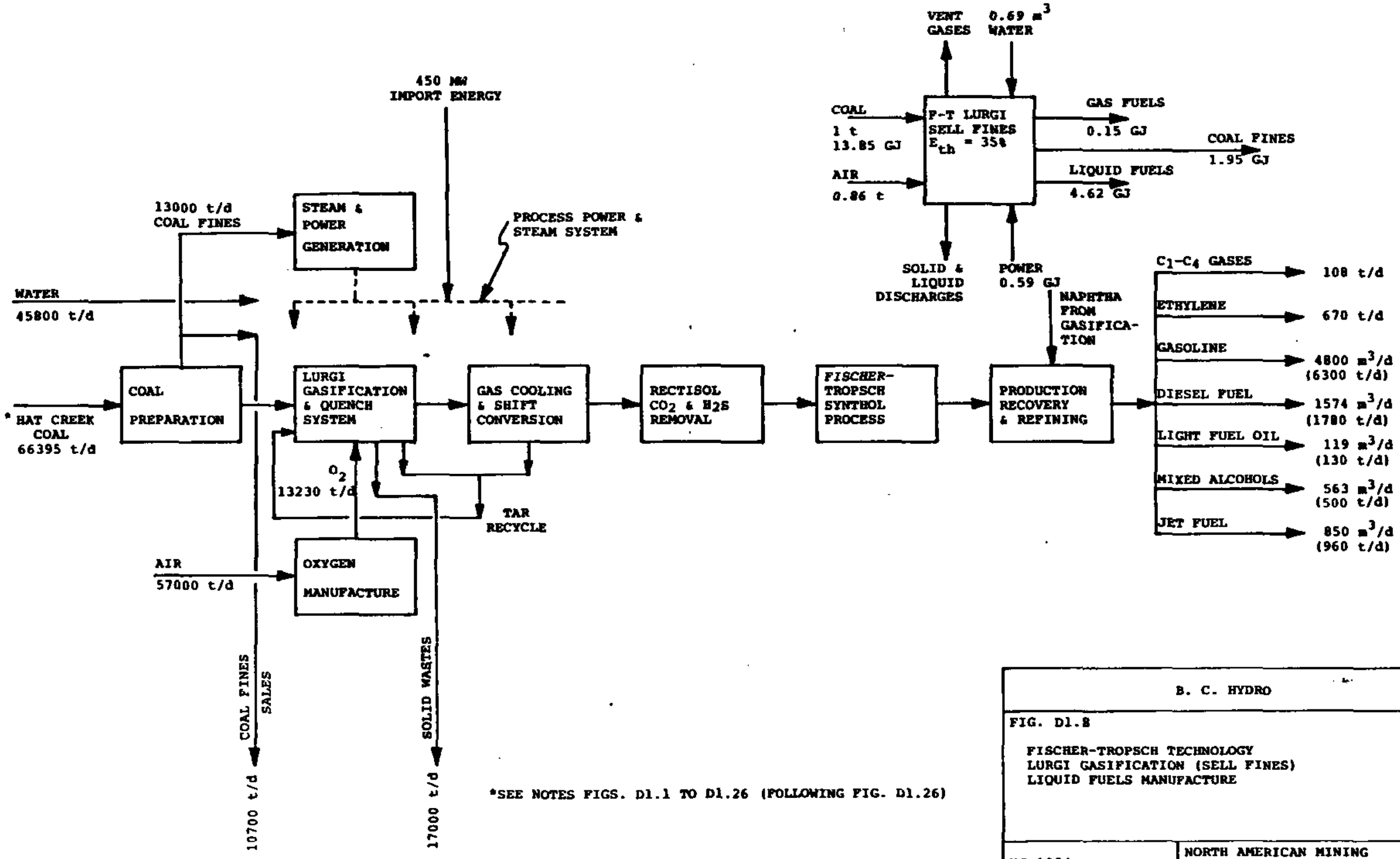
CONSULTANTS, INC.



*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

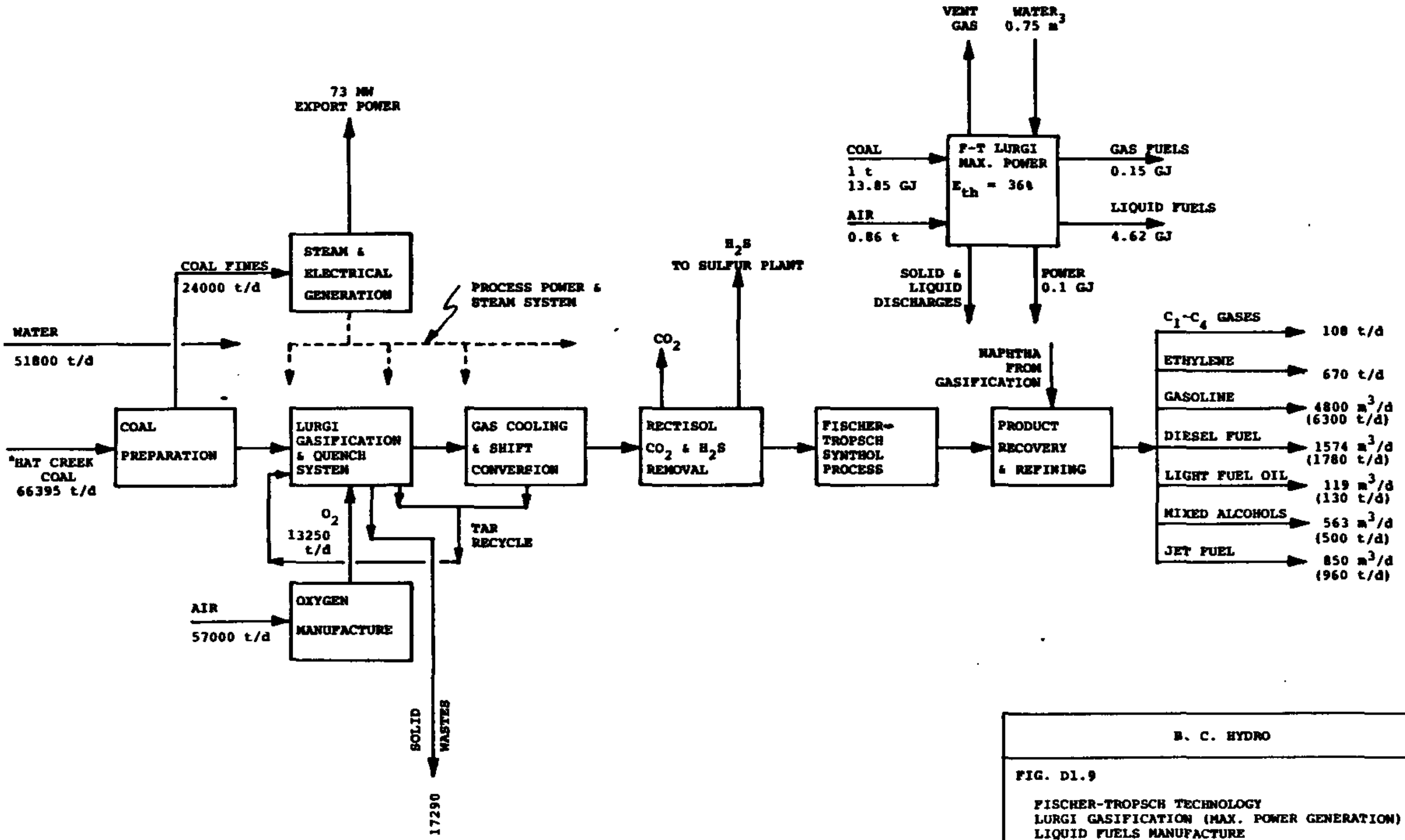
B. C. HYDRO

FIG. D1.8

FISCHER-TROPSCH TECHNOLOGY
LURGI GASIFICATION (SELL FINES)
LIQUID FUELS MANUFACTURE

NJ 1024

NORTH AMERICAN MINING
CONSULTANTS, INC.



*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

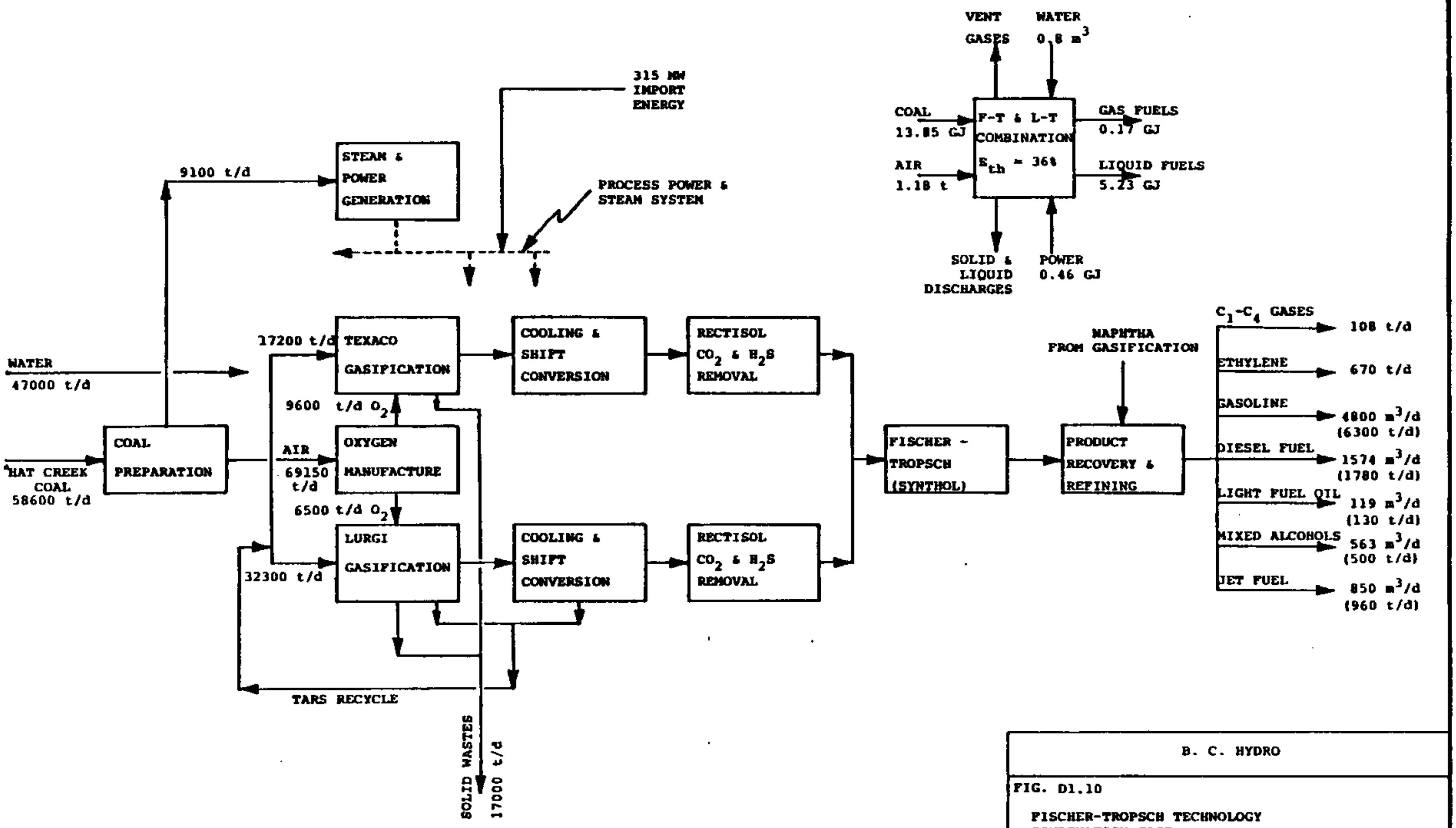
B. C. HYDRO

FIG. D1.9

FISCHER-TROPSCH TECHNOLOGY
LURGI GASIFICATION (MAX. POWER GENERATION)
LIQUID FUELS MANUFACTURE

NJ 1024

NORTH AMERICAN MINING
CONSULTANTS, INC.



*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.10

FISCHER-TROPSCH TECHNOLOGY
COMBINATION CASE
LURGI & TEXACO GASIFICATION PROCESSES
LIQUID FUELS MANUFACTURE

NJ 1024

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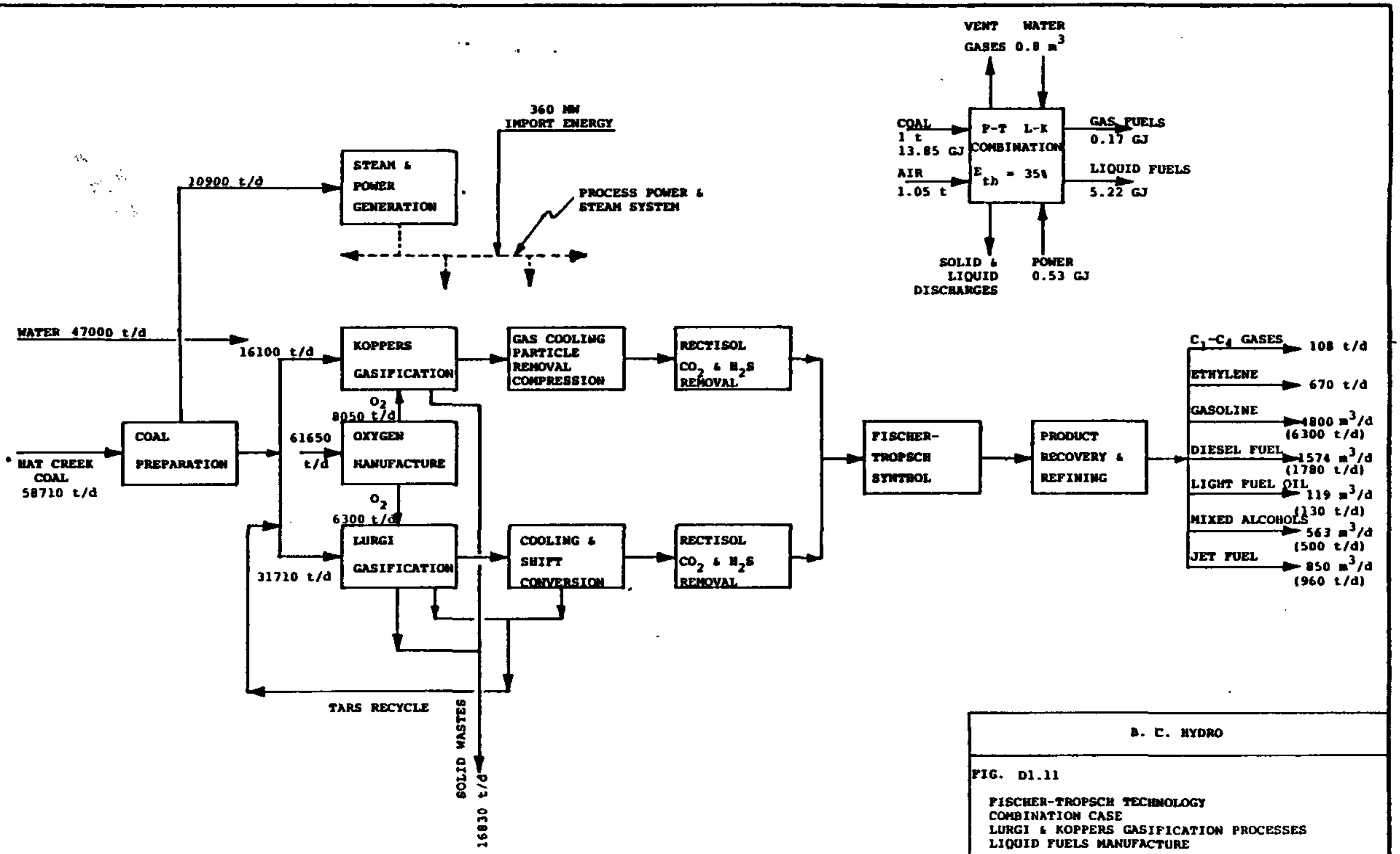

*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.11

FISCHER-TROPSCH TECHNOLOGY
COMBINATION CASE
LURGI & KOPPERS GASIFICATION PROCESSES
LIQUID FUELS MANUFACTURE

NJ 1024

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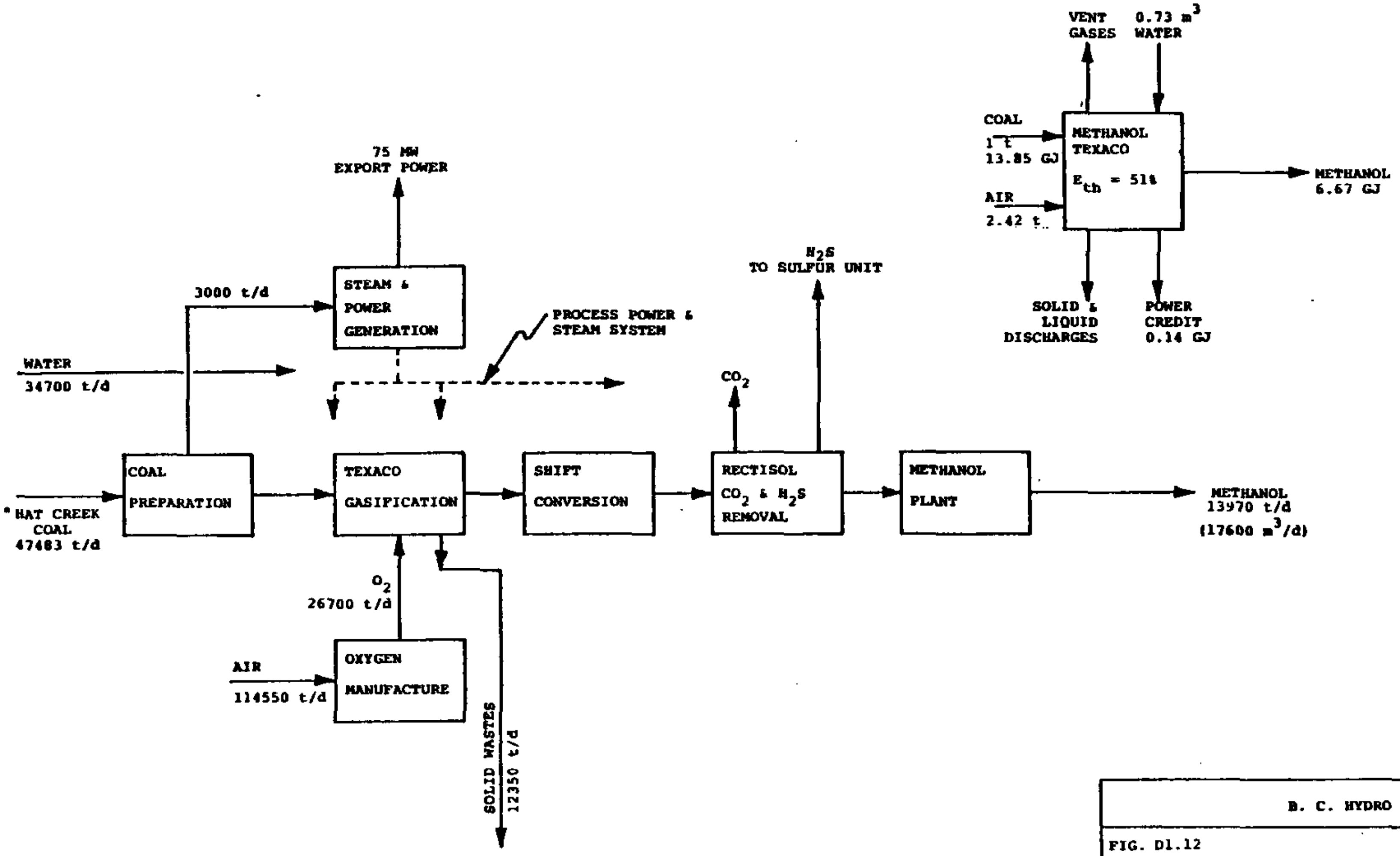

*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.12

METHANOL TECHNOLOGY
TEXACO GASIFICATION PROCESS
LIQUID FUELS MANUFACTURE

NJ 1024

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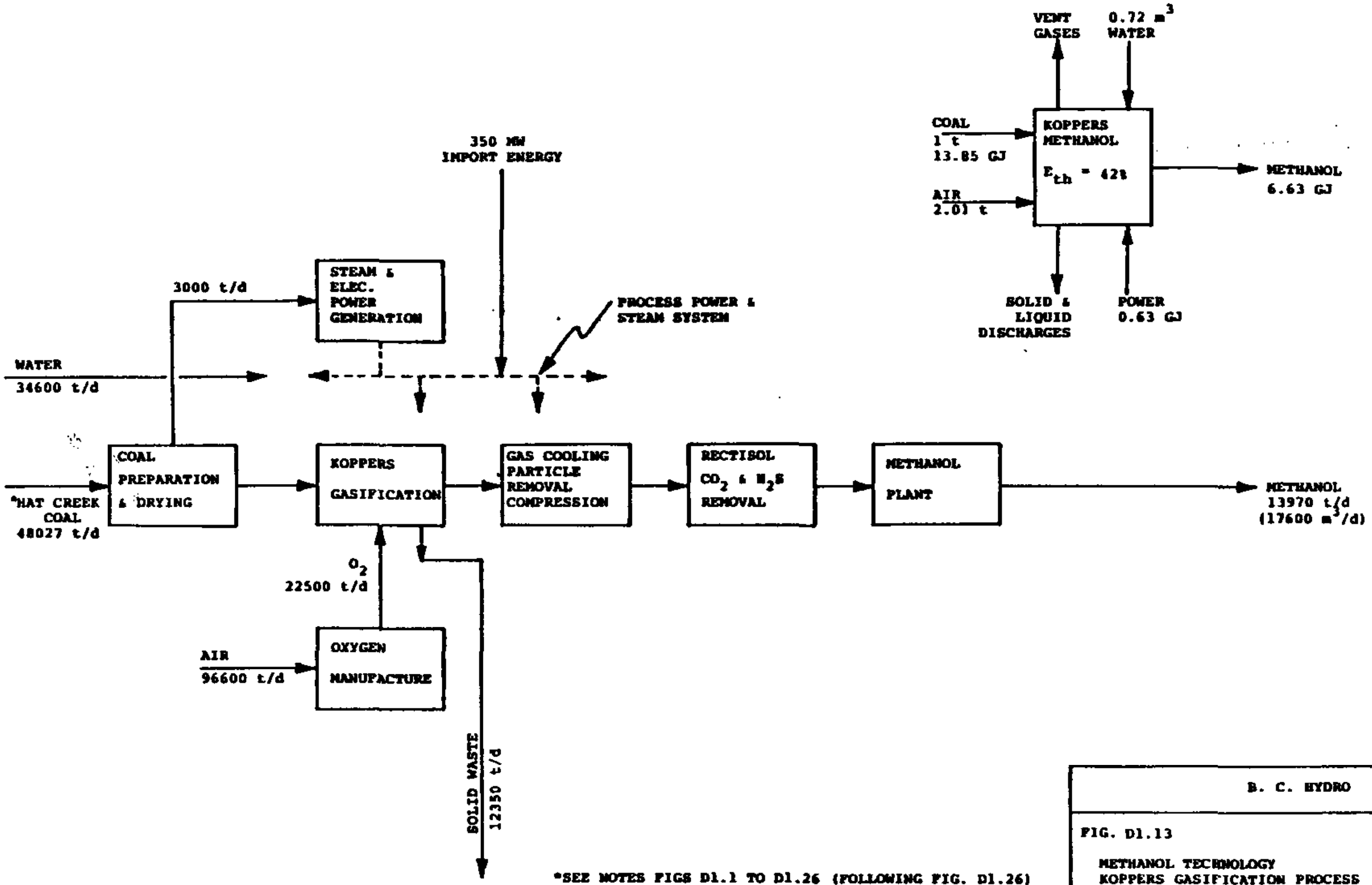

*SEE NOTES FIGS D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.13

METHANOL TECHNOLOGY
KOPPERS GASIFICATION PROCESS
LIQUID FUELS MANUFACTURE

MJ 1024

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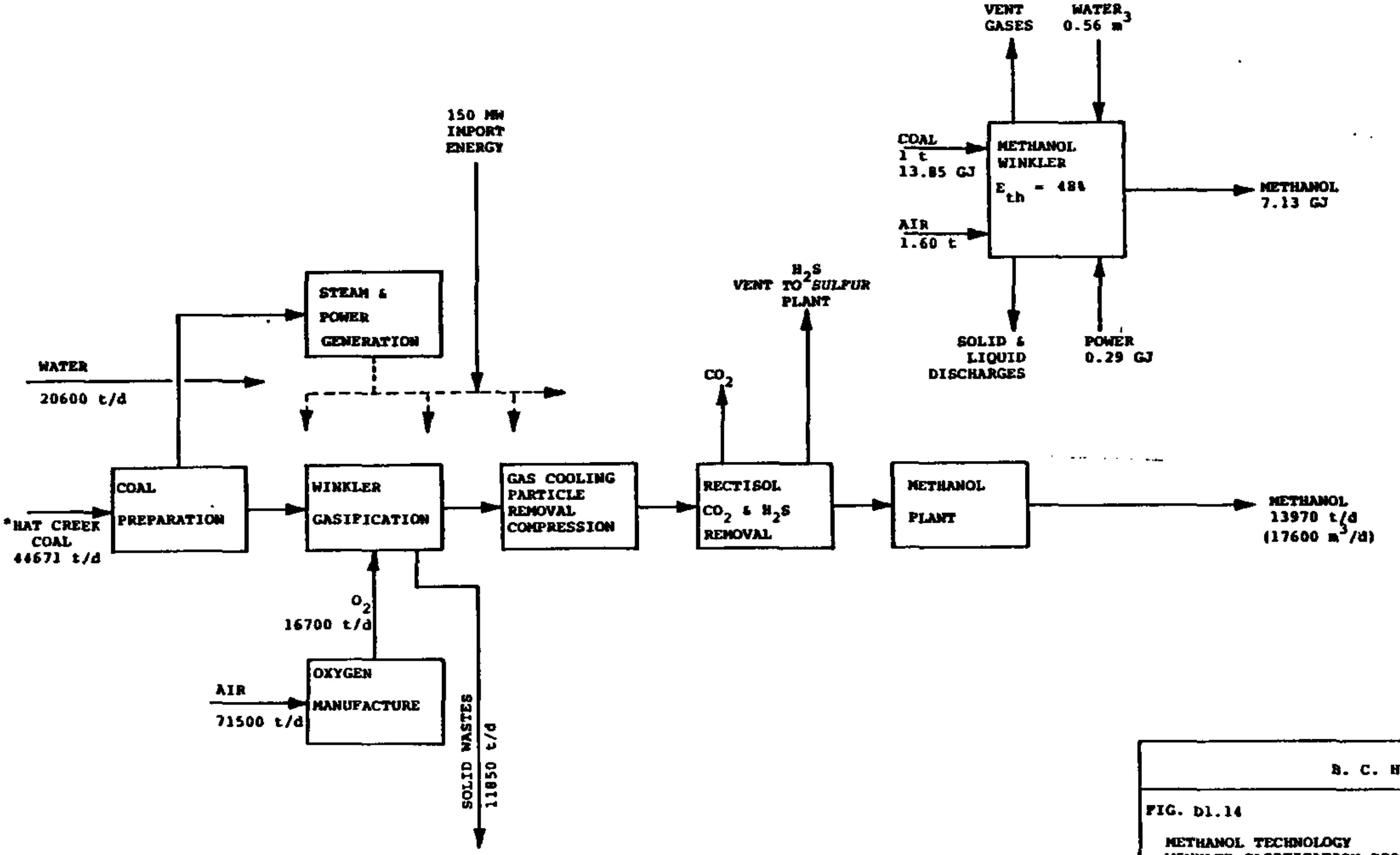
CONSULTANTS, INC.



*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.14

METHANOL TECHNOLOGY
WINKLER GASIFICATION PROCESS
LIQUID FUELS MANUFACTURE

NJ 1024

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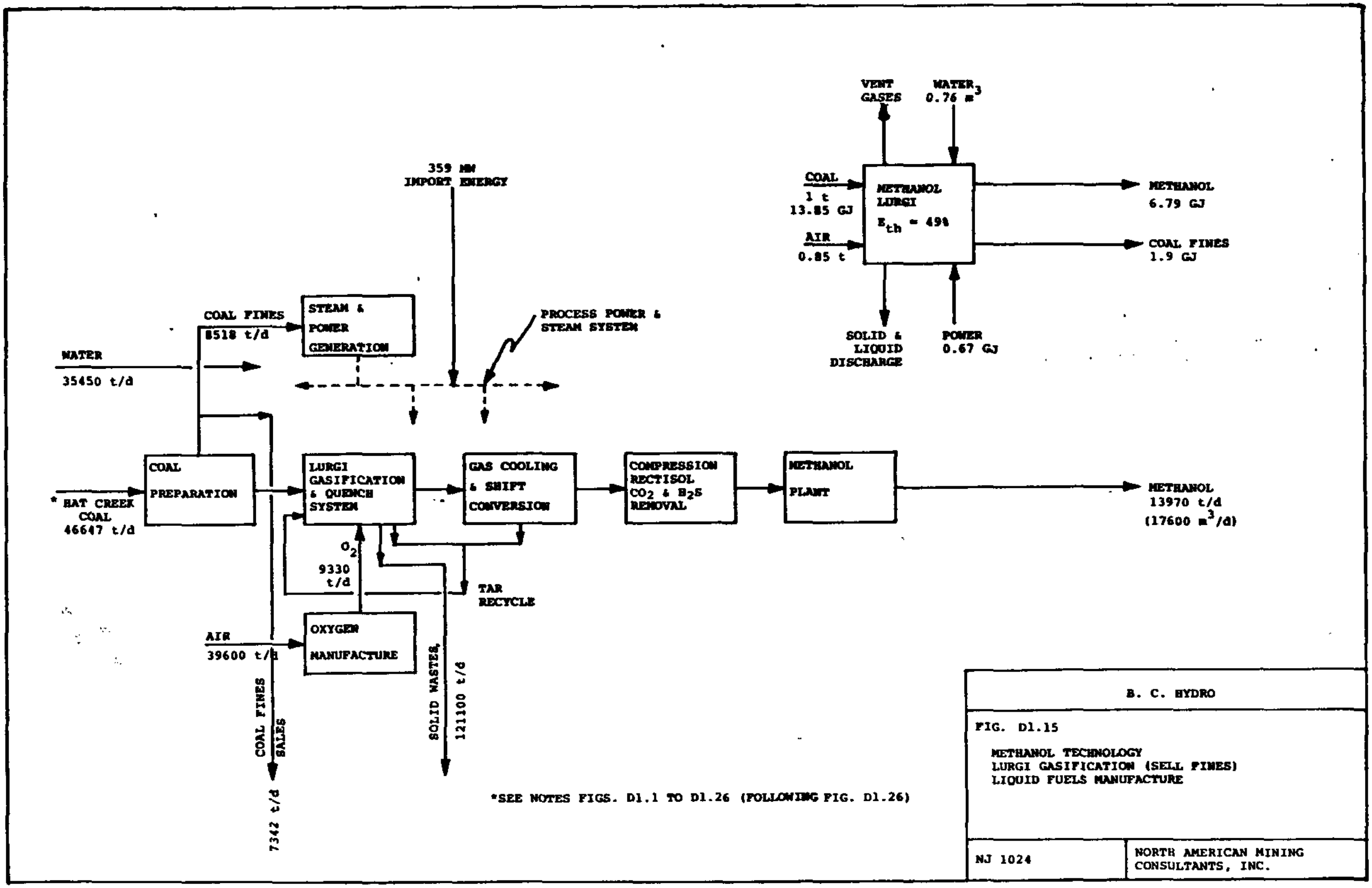

*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.15

METHANOL TECHNOLOGY
LURGI GASIFICATION (SELL FINES)
LIQUID FUELS MANUFACTURE

NJ 1024

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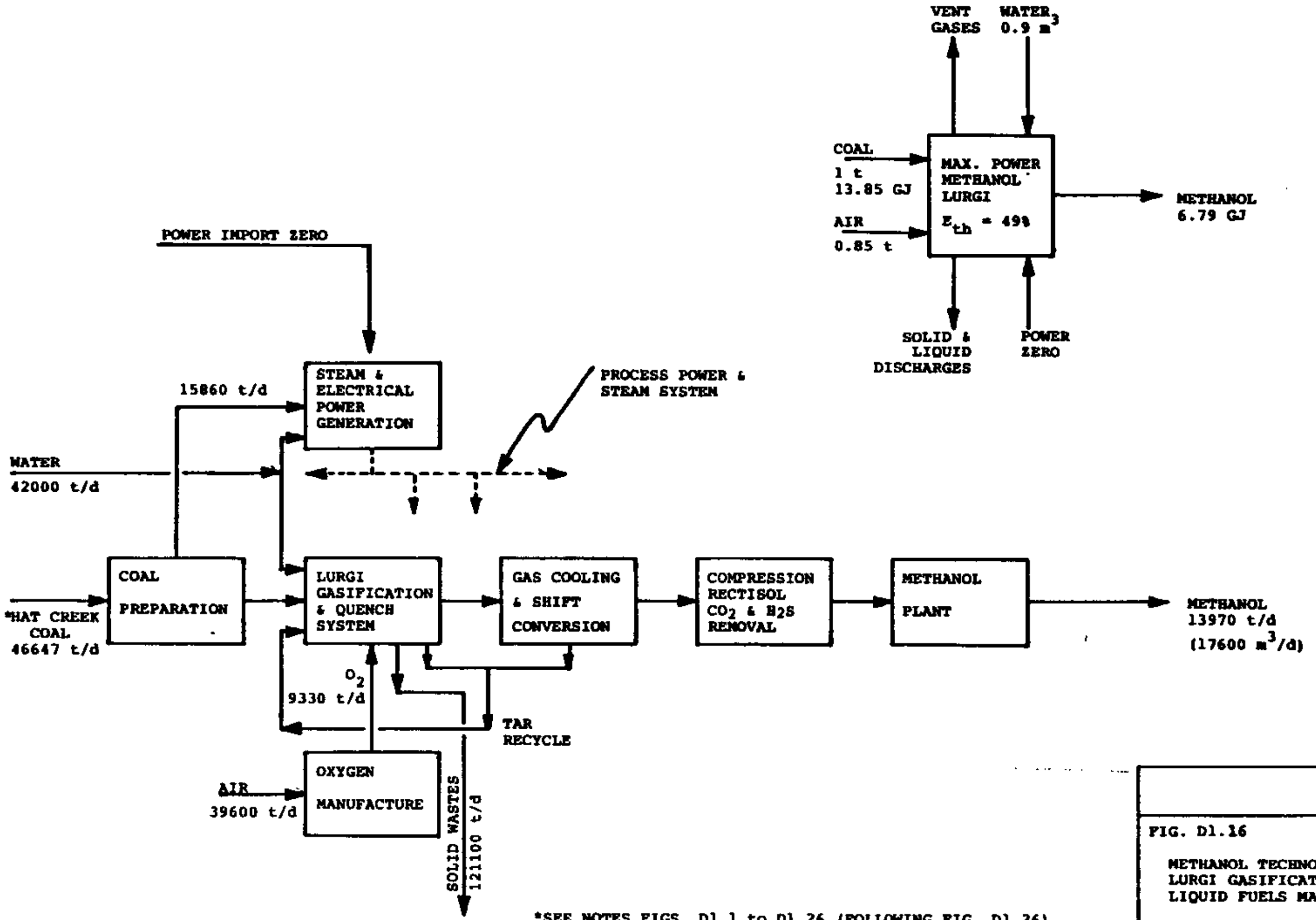

*SEE NOTES FIGS. D1.1 to D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.16

METHANOL TECHNOLOGY
LURGI GASIFICATION (MAX. POWER GENERATION)
LIQUID FUELS MANUFACTURE

NJ 1024

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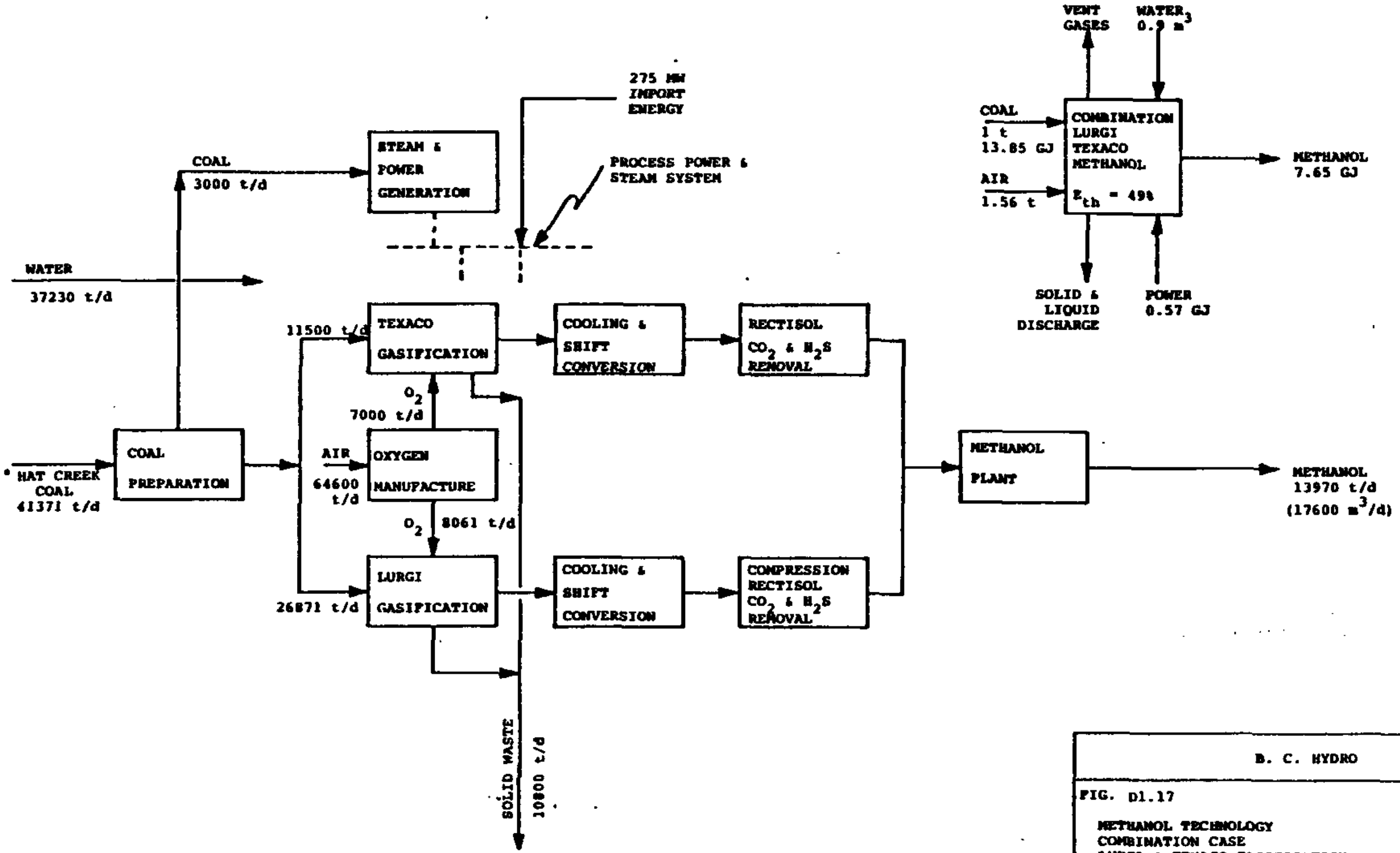
CONSULTANTS, INC.



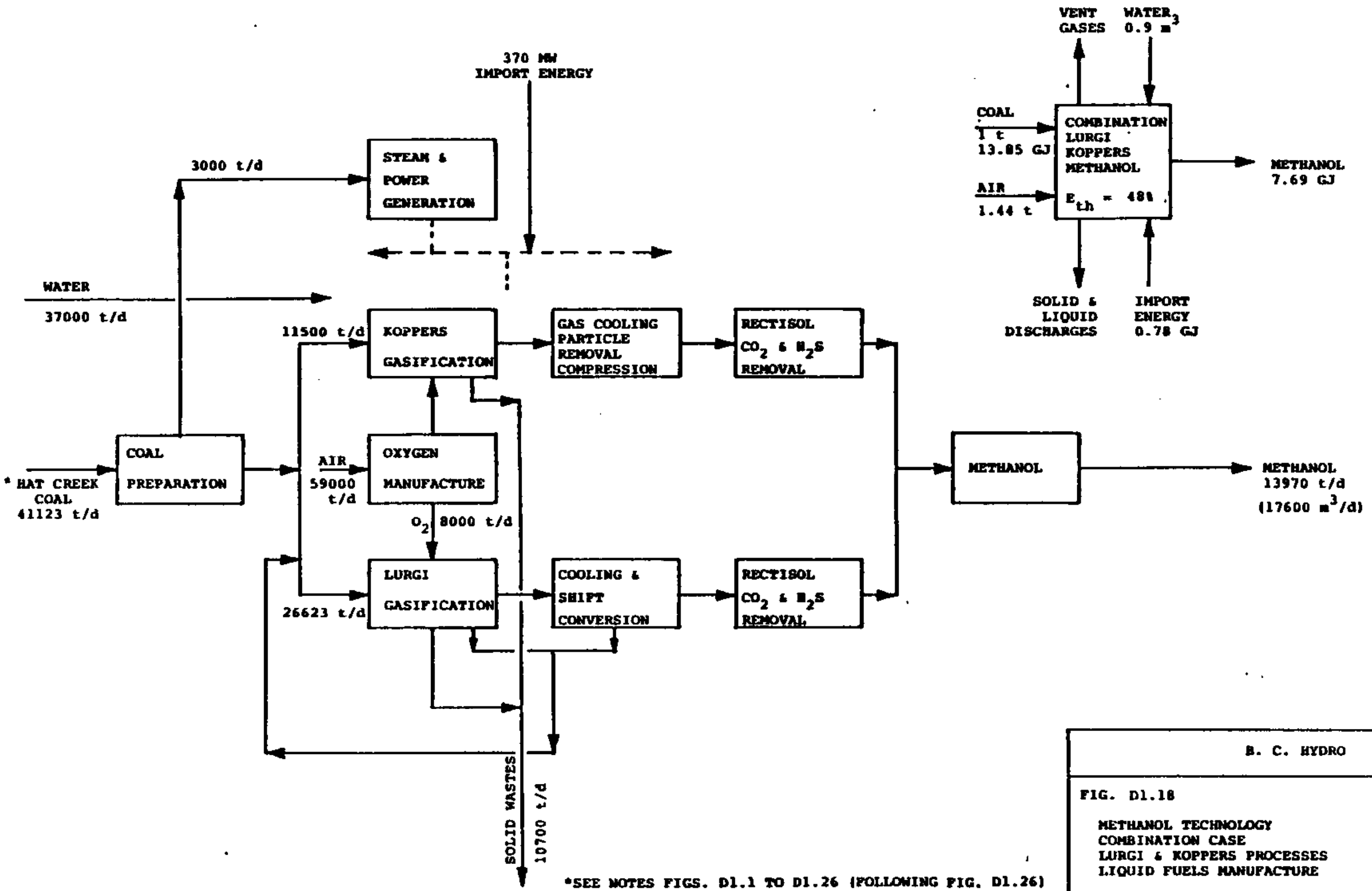
*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.17

METHANOL TECHNOLOGY
COMBINATION CASE
LURGI & TEXACO GASIFICATION
LIQUID FUELS MANUFACTURE

NJ 1024	NORTH AMERICAN MINING CONSULTANTS, INC.



*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

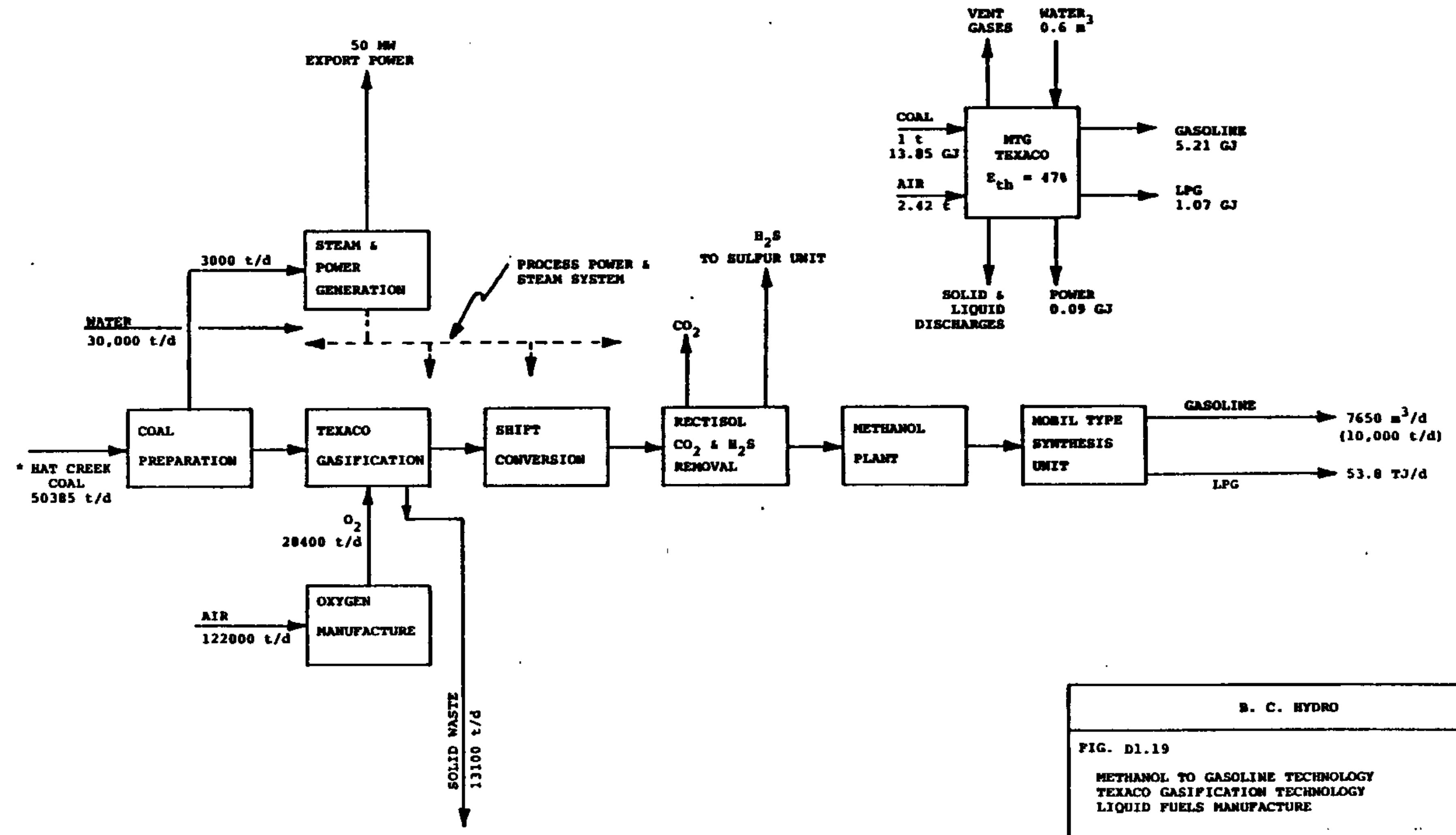
B. C. HYDRO

FIG. D1.18

METHANOL TECHNOLOGY
COMBINATION CASE
LURGI & KOPPERS PROCESSES
LIQUID FUELS MANUFACTURE

NJ 1024

NORTH AMERICAN MINING
CONSULTANTS, INC.



*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.19

METHANOL TO GASOLINE TECHNOLOGY
TEXACO GASIFICATION TECHNOLOGY
LIQUID FUELS MANUFACTURE

NJ 1024

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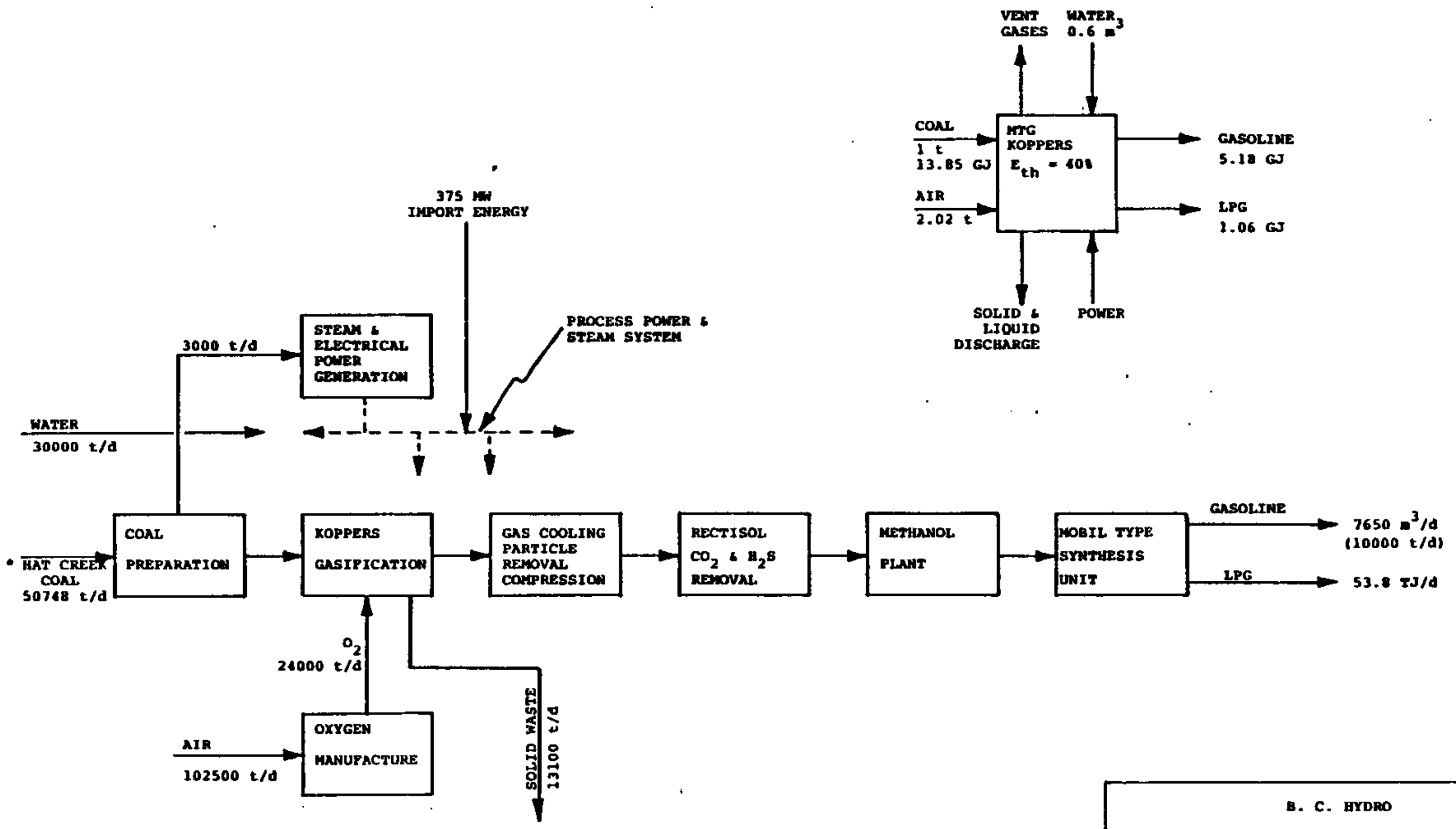

*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.20

METHANOL TO GASOLINE TECHNOLOGY
KOPPERS GASIFICATION PROCESS
LIQUID FUELS MANUFACTURE

NJ 1024

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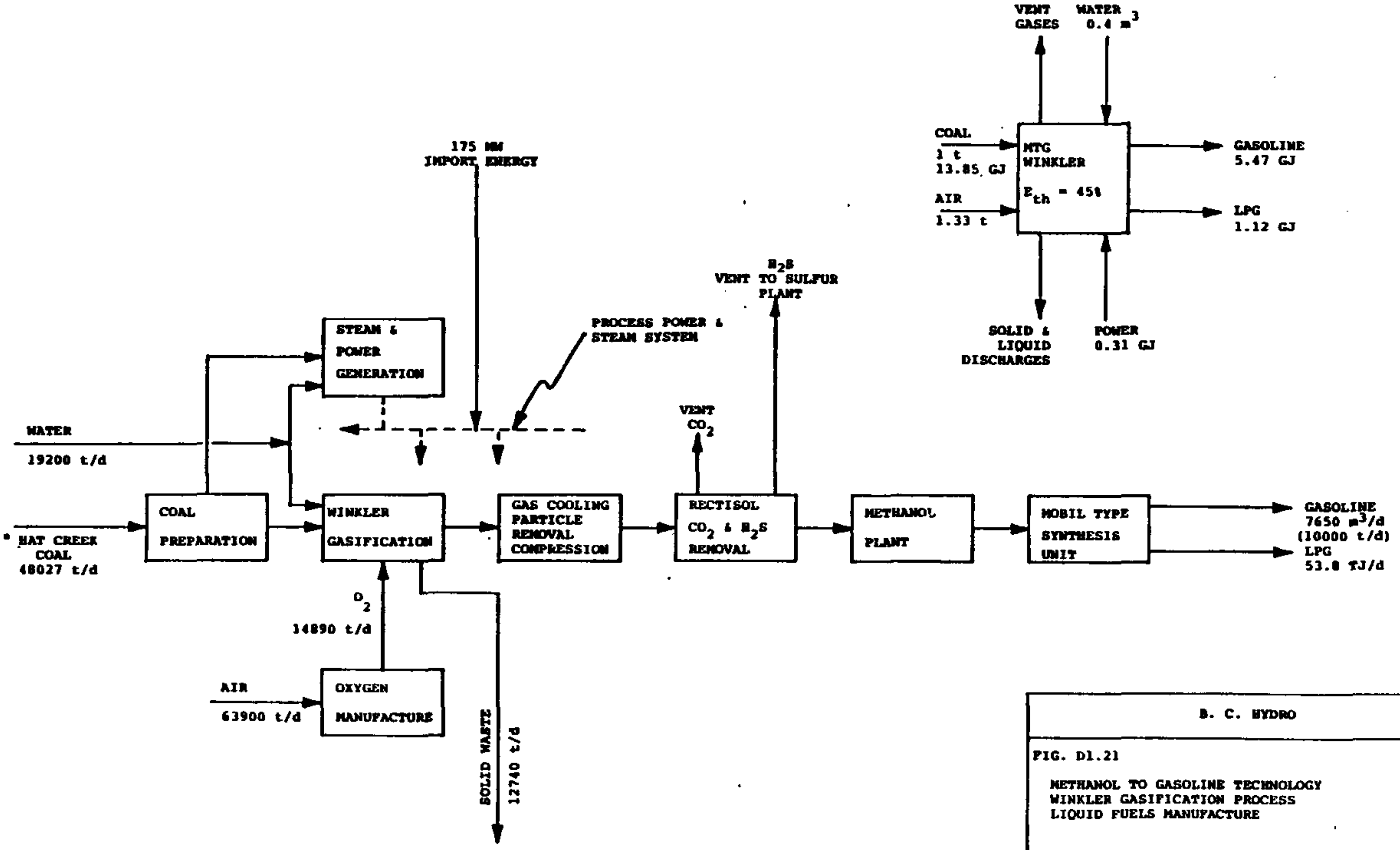

*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.21

METHANOL TO GASOLINE TECHNOLOGY
WINKLER GASIFICATION PROCESS
LIQUID FUELS MANUFACTURE

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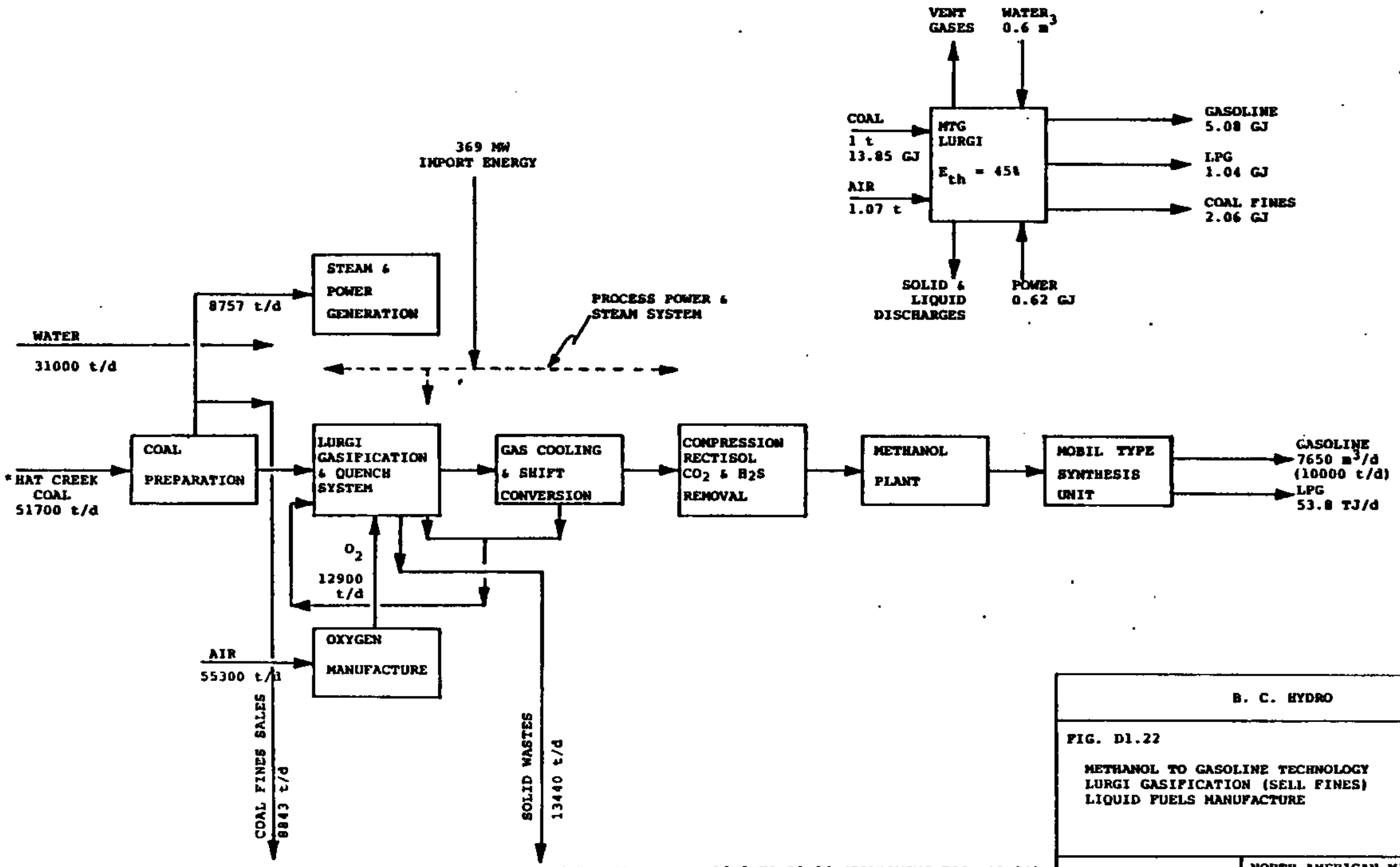

*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.22

METHANOL TO GASOLINE TECHNOLOGY
LURGI GASIFICATION (SELL FINES)
LIQUID FUELS MANUFACTURE

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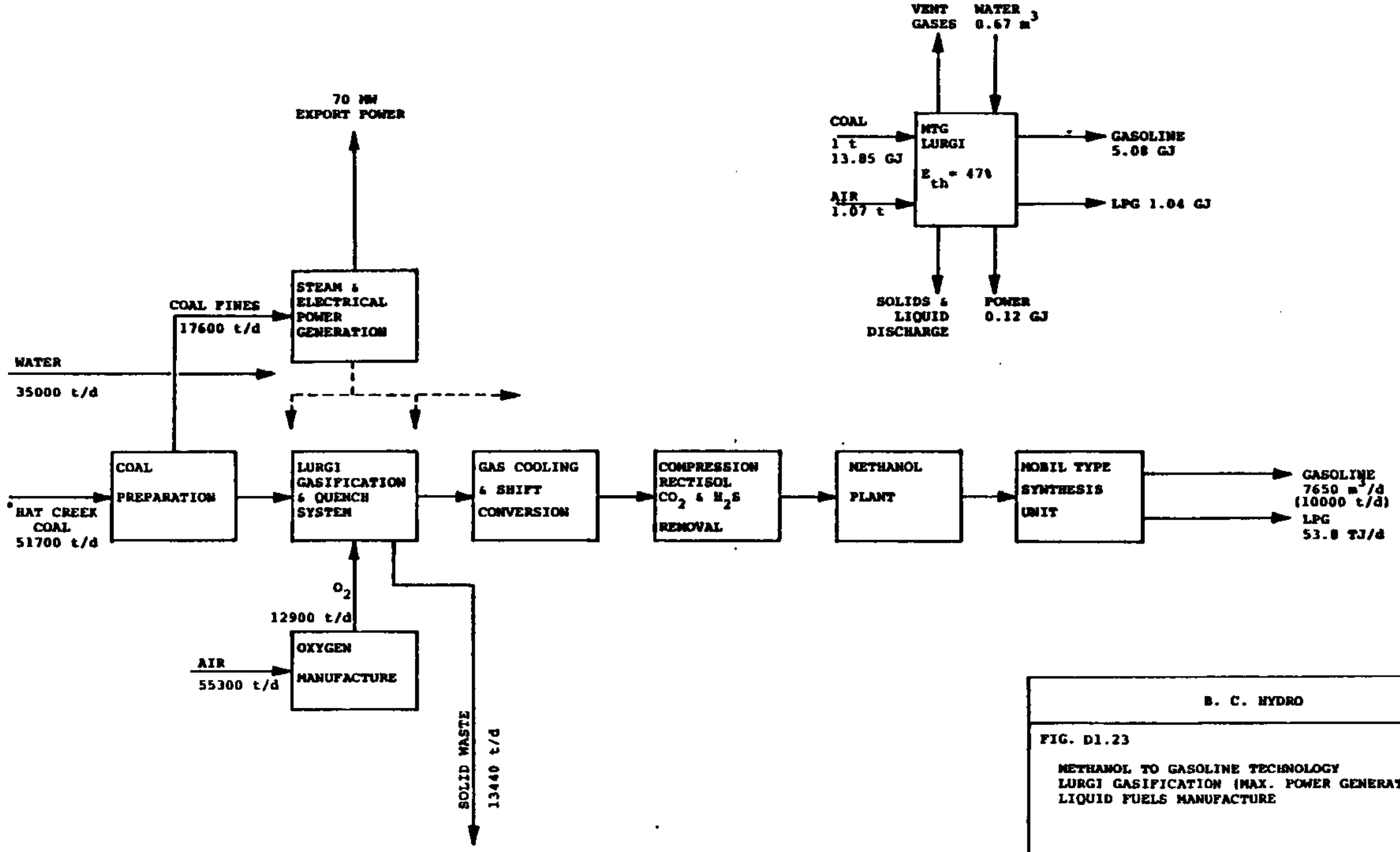

*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.23

METHANOL TO GASOLINE TECHNOLOGY
LURGI GASIFICATION (MAX. POWER GENERATION)
LIQUID FUELS MANUFACTURE

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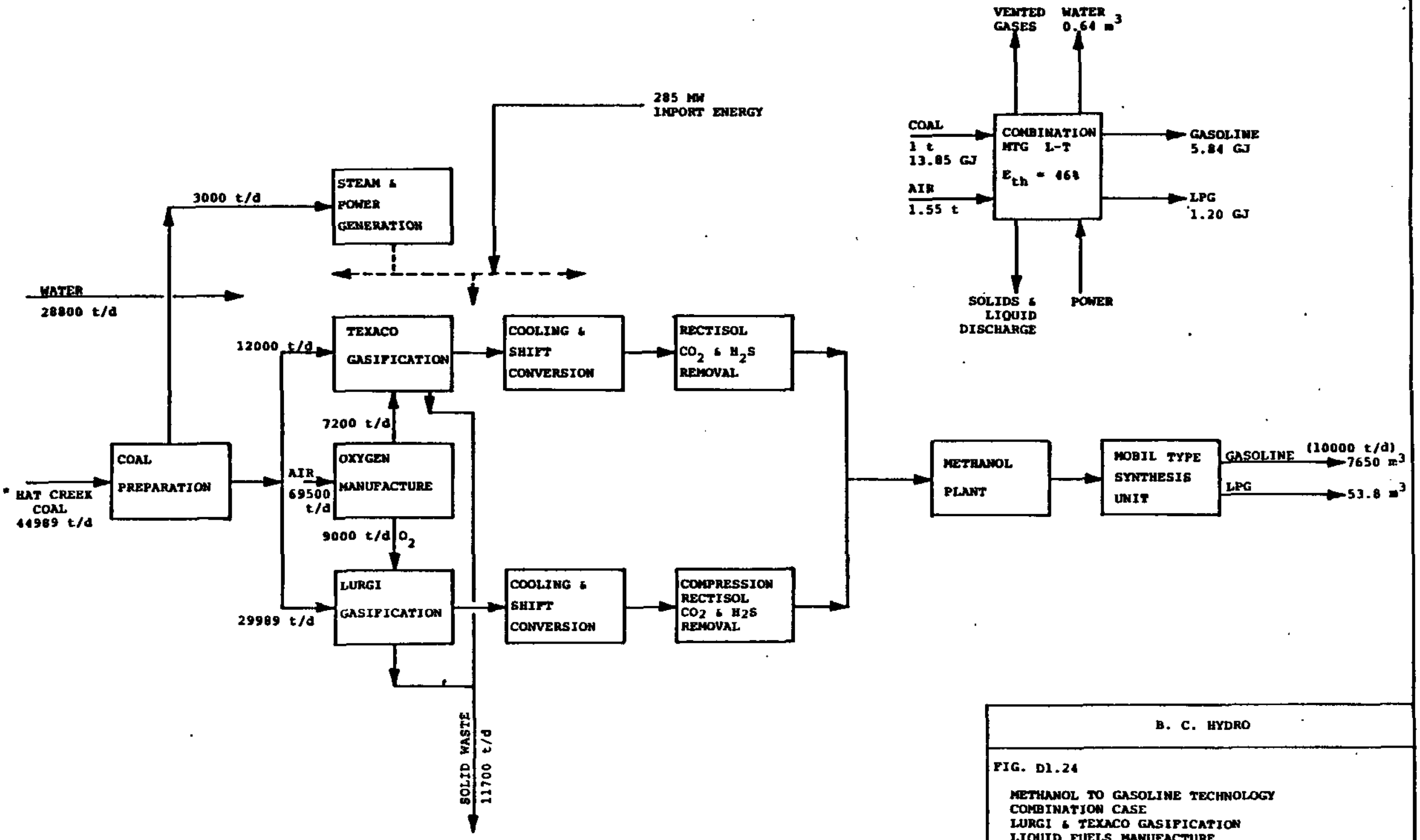

*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.24

METHANOL TO GASOLINE TECHNOLOGY
COMBINATION CASE
LURGI & TEXACO GASIFICATION
LIQUID FUELS MANUFACTURE

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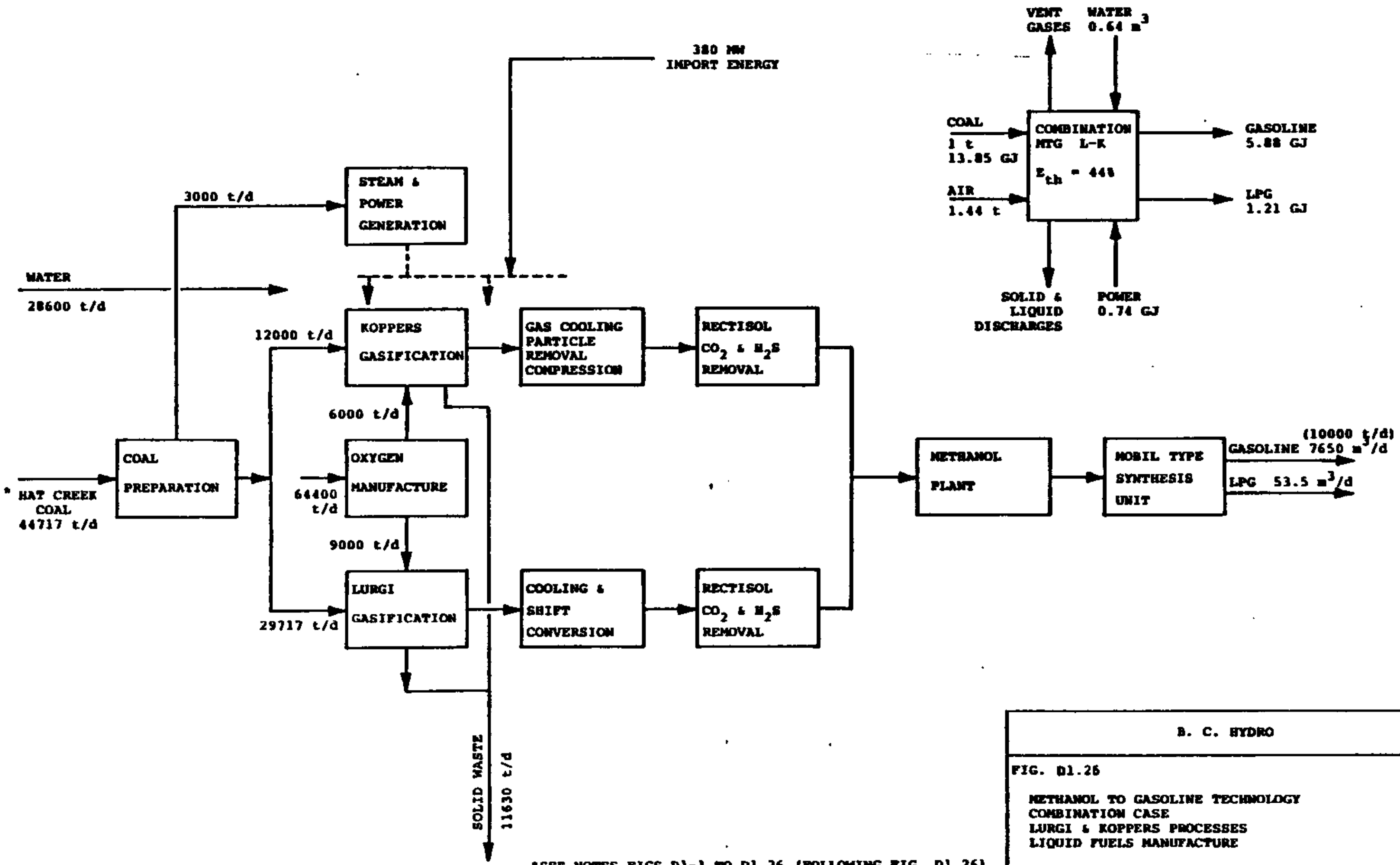

*SEE NOTES FIGS D1-1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.26

METHANOL TO GASOLINE TECHNOLOGY
COMBINATION CASE
LURGI & KOPPERS PROCESSES
LIQUID FUELS MANUFACTURE

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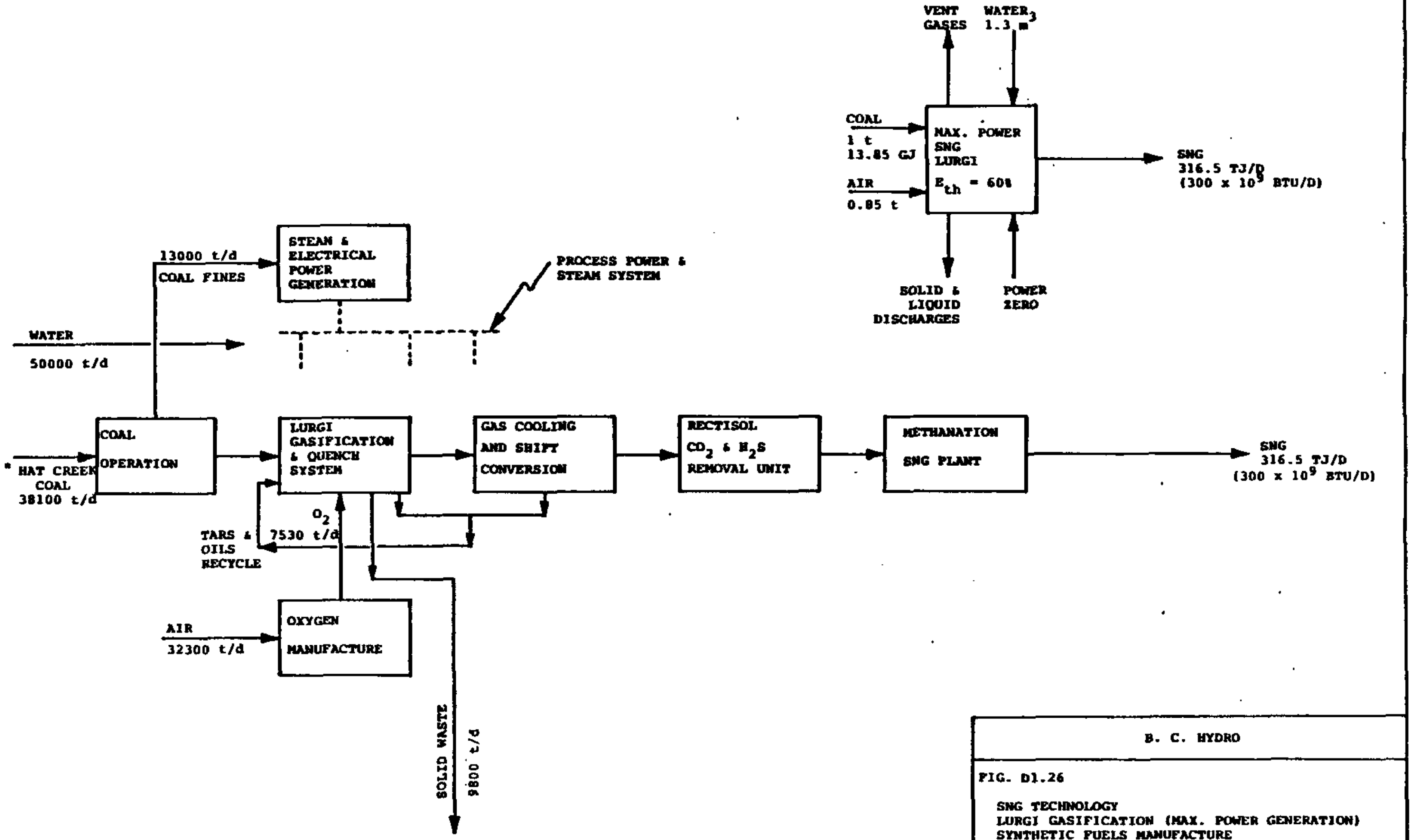

*SEE NOTES FIGS. D1.1 TO D1.26 (FOLLOWING FIG. D1.26)

B. C. HYDRO

FIG. D1.26

SNG TECHNOLOGY
LURGI GASIFICATION (MAX. POWER GENERATION)
SYNTHETIC FUELS MANUFACTURE
COMPARATIVE GAS COST

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General Notes - Figures D1.1 to D1.26

1. Coal requirements are specified on a "corrected-as-received" basis, ad detailed in Table 2.2.*
2. All solid waste streams are expressed on a dry basis.
3. VTBS is the abbreviation for "Vacuum Tower Bottoms Slurry" which is the residue of the coal liquefaction process and which is commonly used as feed material for hydrogen manufacture.
4. Light fuel oil in the SRC-II type process consists of a mixture of naphtha and turbine fuel. Insufficent data exists to estimate the split of light fuel oil into these products for Hat Creek coal.
5. Sulfur and ammonia production for the indirect coal liquefaction cases is summarized in Table 3.11.*

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* See Chapters 2 and 3 of the Report

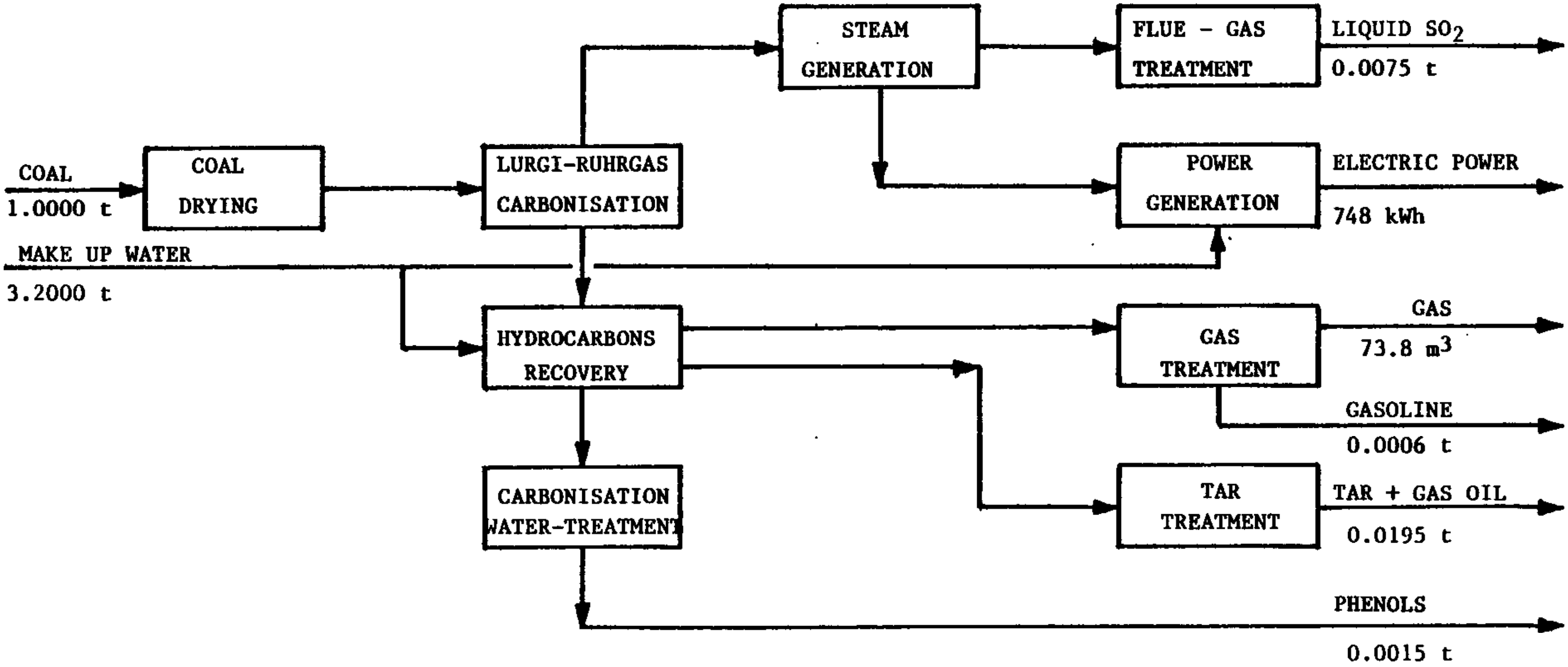

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FIG. D1.27
SCHEME FOR LURGI-RUHRGAS PROCESS

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