

CH AP-5873



DKRW Advanced Fuels
Two Riverway, Suite 1780
Houston, Texas 77056 USA
713.425.6520 phone
713.355.3201 fax

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Received
JAN 08 2009
Casper DEQ

September 30, 2008

Chad Schlichtemeier
Wyoming Department of Environmental Quality
Air Quality Division / NSR Program Manager
Herschler Building
122 West 25th Street
Cheyenne, WY 82002

**Subject: Medicine Bow Fuel & Power LLC
Proposed Integrated Gasification and Liquefaction Plant
(PSD Air Quality Permit Application AP-5873)
Response to Public Comment/WDEQ Information Request**

Dear Mr. Schlichtemeier:

This letter is provided in response to a letter from Mr. Andrew Keyfauver, dated August 15, 2008, requesting clarification and response to specific items brought up during the public comment period for the Medicine Bow Fuel & Power, LLC (MBFP) proposed coal-to-liquids plant. Our responses to these questions are directly below each of the five items in that August 15, 2008 letter. The five specific questions from the WDEQ are shown in italics.

Responses to WDEQ Questions

1. Public comments suggest that the applicability of Section 112(j) and 112(g) need to be addressed for the boilers and process heaters as this facility is shown to be a major source of hazardous air pollutants. Therefore, the Division requests that Medicine Bow Fuel & Power, LLC address Section 112 applicability for the facility.

Response: HAP emissions in the Medicine Bow Fuel & Power LLC (MBFP) Air Permit application (as revised May 12, 2008) were based on early engineering information. Subsequent to MBFP's submission of the Air Permit application, MBFP received the Process Design Package (PDP) in August 2008 from Davy Process Technology for their syngas-to-methanol technology which resulted in MBFP reviewing the original HAP calculations. This review found that traditional sample lines in methanol service were the most significant source of methanol emissions within the equipment leak category. Equipment leaks from traditional sample lines result from purging the lines to atmosphere prior to collecting a sample as part of the sampling protocol. The August 2008 Davy PDP includes 6 closed-loop sampling lines which initial engineering had shown to be traditional sample lines. So we have eliminated 6 traditional methanol sample lines from our prior HAP emission calculations based on the most recent engineering information. The 6 closed-loop sample lines can be eliminated since they

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provide 100% emission control because the sample piping is returned to the process piping at some downstream point without any purging to the atmosphere.

As a result, calculated equipment leak emissions (see attached revised emission calculations for methanol) are reduced to 9.1 tpy methanol which is below the 10 tpy threshold established in Section 112 of the Act for major source determination. Therefore, due to this new engineering information Sections 112(j) and 112(g) of the Act will not be triggered for the proposed facility.

Revised equipment leak and total facility emission calculation pages are provided with this letter.

- 2. A public comment suggested that leak detection and repair (LDAR) levels need to be lowered, based on levels set for petroleum refineries in California. The Division requests that Medicine Bow Fuel & Power, LLC address the feasibility of lowering LDAR levels for the plant.*

Response: Although MBFP is not subject to the NSPS for petroleum refineries, the leak definitions in the MBFP Air Permit Application are equivalent to those in the recently promulgated New Source Performance Standard (NSPS) for petroleum refineries (thus Best Available Control Technology (BACT) for refinery leaks) and the Synthetic Organic Chemical Manufacturing Industry (SOCMI) at 40 CFR 60, Subpart VVa and GGGa, respectively, with a 500 ppm leak definition for valves/connectors and 2,000 ppm leak definition for pumps. We agree with the Wyoming DEQ that MBFP leak definitions are BACT for MBFP.

The EPA considered the more stringent California-leak standards (lower than 500 ppm for valves) when promulgating the November 2007 New Source Performance Standards (BACT) for chemical plants and refineries (40 CFR 60, Subparts VVa and GGGa.), but noted that "data gathered from facilities making a first attempt at repair on valves with leaks above 100 or 200 ppm suggests that these attempts do not always reduce emissions." (Summary of Public Comments and Responses, Docket ID NO. EPA-HQ-OAR-2006-0699-0094) EPA assessed a cost effectiveness of \$5,700/ton for the SOCMI and \$16,000/ton for refineries if leak definitions were lowered to less than 500 ppm for valves, and thus concluded that a leak standard below 500 ppm for valves was not cost effective (72FR64864, November 16, 2007). EPA also dismissed lower leak standards for pumps (less than 2,000 ppm) by stating they had no evidence that lowering pump leak standards would achieve significant emission reductions at a reasonable cost and noting uncertainties regarding pump repair effectiveness at low leak concentrations (72FR64864). The EPA impact analysis is available in the docket for the regulation, at Docket ID No. EPA-HQ-OAR-2006-0699.

MBFP agrees with EPA's decision to dismiss leak standards that would be lower than we have proposed in our application.

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3. *A public comment suggested that the Medicine Bow IGL Plant is subject to the refinery NSPS and NESHAP regulations based on an applicability determination by EPA in 1980. The Division requests that Medicine Bow Fuel & Power, LLC address the applicability of the refinery NSPS/NESHAP standards for the Medicine Bow IGL Plant.*

Response: The public comment making this suggestion refers to two separate documents:

- A 1980 EPA letter titled "Applicability Determination for Solvent Refilled Coal Plants" obtained from the EPA's Applicability Determination Index (ADI);
- Chapter 3 of a 1981 comparative technical and economic assessment of selected synfuel technologies, titled "Selected Technical and Economic Comparisons of Synfuel Options, Final Report," written for the United States' Office of Technology Assessment as a background document to assist in preparation of a larger study report titled "Increased Automobile Fuel Efficiency and Synthetic Fuels: Alternatives for Reducing Oil Imports." The Chapter 3 title is 'Overview of Selected Synthetic Fuel Conversion Processes.' A copy of the full chapter is provided with this letter, for your reference.

As stated in the public comment, the attached 1980 ADI letter notes that NSPS Subpart J requirements for petroleum refineries applies to affected facilities at solvent refined coal (SRC) plants. The letter also notes that "determinations of applicability of solvent refined coal plants to the NSPS for petroleum refineries should be handled on a case-by-case basis, thus, it may not be applicable to all SRC plants."

The public comment on the MBFP permit goes on to state the "SRC II" process, which is one of the two types of SRC technologies, is similar to the MBFP methanol-to-gasoline (MTG) process, with "no distinction that would render the 1980 determination from EPA invalid." For the reasons discussed below, MBFP disagrees with this comment, based on a review of the SRC II and the MTG process technologies, petroleum refineries, the EPA's 1980 determination, and the definition of 'petroleum refinery.'

MBFP disagrees with the public comment based on the following from Chapter 4 (attached) from that same 1981 report (i.e. Selected Technical and Economic Comparisons of Synfuel Options, Final Report) where it states on page 4.-28:

"4.6 REFINING SYNTHETIC LIQUIDS

The direct liquefaction and oil shale syfuels have to be further upgraded to end-use product quality in order to be comparable with indirect liquid products such as methanol from coal or gasoline from methanol (from coal). In a wider sense, this is also desirable in order to achieve comparability with synthetic natural gas (SNG) which can be used for a wide range of end use applications in its 'raw' manufactured state.

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Houston, Texas 77056 USA
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The indirect processes produce refinery output (or intermediate) grade products, without the need for the "refining" of crude liquids."

The attached Chapter 4 also includes on pages 4-38 through 4-40 EXHIBITS 4-21, 4-22 and 4-23 which are flow diagrams for the "refining" of the SRC-II produced oil.

As discussed in that 1981 report, the SRC II process is a direct catalytic liquefaction process to convert coal to a crude oil-like liquid. It involves mixing hydrogen with a coal-slurry, reacting the mixture with steam and oxygen, and allowing reactions to take place in a dissolver vessel operating at high pressure and temperature (2,000 psi, 820-870° F). The coal is dissolved with the resulting solution resembling a crude oil which is then fractionated (sent through distillation columns like those used in a refinery) to recover primary products such as naphtha, fuel oil and a vacuum residue. The naphtha and fuel oil products can be further treated in downstream units. All fuel gases in these downstream units, such as catalytic crackers and naphtha reformers, will emit sour gases that will require processing to remove the sulfur. This is the same for refineries which route fuel gases to a gas processing unit to reduce sulfur content.

In contrast, the proposed MBFP facility will employ an indirect liquefaction process to produce methanol and then gasoline from methanol (not the crude oil like product of SRC II). The MBFP syngas is sent to an Acid Gas Recovery Unit where 99.8% of the sulfur is removed. Any residual sulfur in the syngas is removed in the sulfur beds, reducing the sulfur levels to the part per billion concentration necessary to protect the MBFP methanol catalyst. This is the same as in chemical processes, where sulfur has to be removed to prevent catalyst poisoning in downstream units.

The cleaned syngas produced at the proposed MBFP facility will be directed through methanol converter reactors, where the syngas will pass over a highly selective copper-based catalyst on the reactor's shell-side. Any residual sulfur in the cleaned syngas is mostly captured as a poison on the methanol catalyst, so the methanol and methanol offgases will have a sulfur content of less than 10 ppb. Carbon dioxide (CO₂) and carbon monoxide (CO) in the syngas will combine with hydrogen (H₂) to create methanol (CH₃OH). Tubes in the reactor will carry steam, which will provide temperature control for the reaction. The methanol will then be directed to the gasoline synthesis (MTG) unit, where it undergoes multiple complex reactions in reactor vessels to convert the methanol to olefins, paraffins, and aromatics, without molecular hydrogen production (without producing the SRC II crude like product that must then be distilled into liquid hydrocarbon products as happens in a refinery). The reactor effluent will be separated into a gas/vapor phase to be recycled to the reactor inlet, a liquid water phase containing a small percentage of alcohols, ketones, and acids that will be treated, and a liquid hydrocarbon phase referred to as "raw gasoline." Since the methanol is extremely low in sulfur, the gasoline produced from methanol and all associated gas streams will also be extremely low sulfur. The gas streams will not require processing as in a refinery.

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The raw gasoline stream will be directed to a stabilizer to remove LPG product. The liquid product from the stabilizer will be directed to a gasoline splitter (distillation column) so that heavy gasoline containing durenene can be separated from the light gasoline and treated for durenene removal in a hydrotreating process. Once durenene is removed, the heavy gasoline will be re-combined with the light gasoline and directed to gasoline storage tanks. This entire process as well as the emissions profile of MBFP is much different than the SRC II process and cannot be considered as a similar process except to note that solid coal is the feed to both processes.

Thus, the SRC II process has similarities to a typical petroleum refinery due to the fact that it produces a crude oil with significant sulfur content which can then be sent through the crude distillation, cracking, and reforming processes found in most petroleum refineries (but not MBFP). These similarities between the SRC II process and a petroleum refinery can be seen in a comparison of their respective process flow diagrams. A basic process flow diagram for the SRC II process is included with this submittal (see attached Chapter 3 page 3-12 Figure 3.4 and Chapter 4 pages 4-38 through 4-40 for Exhibits 4.-21, 4-22 and 4-23) which can then be compared to a typical petroleum refinery flow diagram such as the one provided as Figure 1.1 in Gary and Handwerk's text "Petroleum Refining - Technology and Economics, 4th Ed." Or alternatively available on Wikipedia if you type in the word "refinery". The first processing step for the liquid hydrocarbon in both the SRC II flow diagram and a general petroleum refining flow diagram is crude oil distillation. In the SRC II process, this crude oil distillation takes place in the dissolver, and also in a fractionation vessel and a "letdown/flash" system, whereas in a typical petroleum refinery, the distillation takes place in a stabilizer, atmospheric distillation tower, and a vacuum distillation tower. We have drawn a box around the letdown/flash system and fractionation system on the SRC II diagram to illustrate the distillation portion of the process. Note that products from the letdown/flash and fractionation systems in the SRC II process are similar to the products from the crude distillation unit in a petroleum refinery. Although not clearly shown on the diagram, the SRC II products will likely require additional treating in order to crack and reform hydrocarbons into gasoline and fuel oil products and to remove sulfur and aromatic compounds prior to sale, just as with a typical petroleum refinery. The proposed MBFP facility will not utilize the same distillation, cracking, and reforming processes found in petroleum refineries or the SRC II process. Also MBFP fuel gases will have sulfur in the single digit part per billion range, two orders of magnitude less than the refinery specification for fuel gases, and do not require further processing as in a refinery.

Furthermore, MBFP does not consider the proposed facility to fall under the regulatory definition of a petroleum refinery. A "petroleum refinery" is defined at 40 CFR 60.101(a) as "any facility engaged in producing gasoline, kerosene, distillate fuel oils, residual fuel oils, lubricants, asphalt (bitumen) or other products through distillation of petroleum or through redistillation, cracking, or reforming of unfinished petroleum derivatives." The term

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“cracking” is used in the context of petroleum refineries to mean the breaking down of higher molecular weight hydrocarbons to lighter components. (Gary and Handwerk, 2001, Appendix A). Cracking can be accomplished through application of heat or catalytic means, and it can be simplistically visualized as ‘breaking’ long-chain hydrocarbons into smaller-chain hydrocarbons. The term “reforming” is used to describe a process where hydrocarbon molecular structures are re-arranged to form higher-octane aromatics with only a minor amount of cracking. (Gary and Handwerk, 2001, pg. 189) Typically, cyclization and isomerization reactions occur catalytically in a reformer. Neither ‘cracking’ nor ‘reforming’ is clarified in the regulations, and thus is taken to have these meanings. Although the proposed facility will produce gasoline, it will not be produced through distillation, redistillation, cracking, or reforming processes. Rather, as described earlier, syngas is converted into methanol, which is then processed via dehydration, oligomerization (polymeration), and cyclization into a gasoline product. The gasoline product will require some treatment to remove the ‘light-end’ smaller hydrocarbons (LPG), and to remove durene from the heavier constituents prior to storage, but otherwise will be a finished product.

Therefore, MBFP disagrees with the commenters who assert that affected facilities at the proposed facility are subject to petroleum refinery NSPS regulations, on the basis that the proposed facility will not meet the definition of ‘petroleum refinery.’

4. *The Division requests that Medicine Bow Fuel & Power, LLC provide clarification on whether power generated at the facility will be exported to the electrical grid. The application states that it is not expected to be exported (page 1-1). If power is to be exported to the electrical grid Medicine Bow Fuel & Power, LLC will need to address the applicability of standards for electric generating units (EGUs).*

Response: MBFP confirms the statement made on page 1-1 of the application that no power generated at the facility will be exported to the electrical grid.

5. *A public comments suggested BACT needed to be applied to the sour water stripper at the facility during startup. The Division requests that Medicine Bow Fuel & Power, LLC clarify operation of the sour water stripper during startup and normal operations. If this source is vented during startup or normal operations an evaluation of control measures and/or work practices must be conducted to minimize emissions from this source during operations.*

Response: MBFP has confirmed through review of the Project Feasibility study that no emissions will be vented to atmosphere from the sour water stripper. During both normal operations and above 20% design flow during startup operations, the sour gas from the sour water stripper will be directed to the SRU and consumed in the SRU furnace. Effluent from the SRU is compressed and recycled to the Selexol system, so no emissions result from the sour gas stream in these situations.

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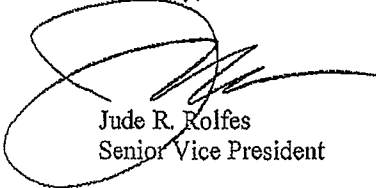
The Startup/Shutdown Emissions Minimization Plan erroneously states on page 2 that the sour water stripper will be vented during startup. As noted above the vent will be directed to flare or other combustion device for ammonia destruction during low flow conditions during startup. A corrected Startup/Shutdown Emission Minimization Plan is attached.

Conclusion:

A CD containing an electronic version of this letter and all enclosures will be sent to you under separate cover.

MBFP appreciates this opportunity to provide additional comment/clarifications to the WDEQ on issues raised during the public comment period. We hope this information is useful for you, and encourage you to contact us if you have any more questions or if you need clarification on any of the points raised in this letter.

Sincerely,



Jude R. Rolfes
Senior Vice President

cc: Andrew Keyfauber (WDEQ)
Robert Moss (DKRW)
Susan Bassett (URS)

Enclosures Revised Emission Calculation Pages for Methanol Equipment Leaks
Copy of 1980 ADI Letter (US EPA to J.Snydor)
Cover Page of 'Selected Technical and Economic Comparisons of Synfuel
Options, Final Report, April 1981 prepared for the Office of Technology
Assessment (OTA) of the United States Congress.
Chapter 3 of 'Selected Technical and Economic Comparisons of Synfuel
Options, Final Report, April 1981
Chapter 4 of 'Selected Technical and Economic Comparisons of Synfuel
Options, Final Report, April 1981
Revised Startup/Shutdown Emission Minimization Plan

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DEQ 002924

Medicine Bow Fuel & Power Industrial Gasification & Liquefaction Plant
Emission Summary Sheet

Normal Operations (8760 hr/yr)

ID No.	Description	Usage	Operation (hr/yr)	Potential Emissions (t/yr)					HAPs Emissions (t/yr)																	
				NO _x	CO	VOC	SO ₂	PM ₁₀	1,3-Butadiene	Acetaldehyde	Acrolein	2,2,4-Trimethylpentane	Benzene	Carbonyl Sulfide	Dichlorobenzene	Ethyl Benzene	Formaldehyde	Hexane	Mercury	Methane	Naphthalene	Propylene Oxide	Toluene	Xylene	TOTALS	
CT-1	Turbine and HRSG Train 1	General Electric, 66 MW	8,760	50.6	46.2	6.6	10.8	43.8	1.37E-03	1.27E-01	2.03E-02	3.81E-02		1.02E-01	2.25E-01		4.33E-05		4.13E-03	6.98E-03	9.21E-02	4.13E-01	2.03E-01	1.23E+00		
CT-2	Turbine and HRSG Train 2	General Electric, 66 MW	8,760	50.6	46.2	6.6	10.8	43.8	1.37E-03	1.27E-01	2.03E-02	3.81E-02		1.02E-01	2.25E-01		4.33E-05		4.13E-03	6.98E-03	9.21E-02	4.13E-01	2.03E-01	1.23E+00		
CT-3	Turbine and HRSG Train 3	General Electric, 66 MW	8,760	50.6	46.2	6.6	10.8	43.8	1.37E-03	1.27E-01	2.03E-02	3.81E-02		1.02E-01	2.25E-01		4.33E-05		4.13E-03	6.98E-03	9.21E-02	4.13E-01	2.03E-01	1.23E+00		
AB	Auxiliary Boiler	Heater, 66 MMBtu/hr ¹	8,760	14.2	23.8	1.8	0.2	2.2				5.95E-04	3.40E-04		2.13E-02	5.10E-01				1.73E-04			9.64E-04	5.33E-01		
B-1	Catalyst Regenerator Heater	Heater, 21.53 MMBtu/hr ¹	8,760	4.6	7.8	0.5	0.1	0.7				1.94E-04	1.11E-04		6.93E-03	1.86E-01				5.64E-05			3.14E-04	1.74E-01		
B-2	Reactivation Heater	Heater, 12 MMBtu/hr ¹	8,760	2.7	4.5	0.3	0.0	0.4				1.12E-04	6.42E-05		4.01E-03	9.62E-02				3.26E-05			1.82E-04	1.01E-01		
B-3	HGT Reactor Charge Heater	Heater, 2 MMBtu/hr ¹	8,760	0.5	0.8	0.1	0.0	0.1				2.00E-05	1.14E-05		7.15E-04	1.72E-02				5.82E-06			3.24E-05	1.79E-02		
Tanks	Storage Tanks	Product Storage	8,760				102.6				5.27E-01			3.79E-02	4.95E-01		2.39E+00					5.67E-01	1.60E-01	4.17E+00		
EL	Equipment Leaks	Fugitives	8,760				58.5				7.90E+00	2.34E-01					5.76E+00						1.40E+01	9.00E+00		
CS	Coal Storage & Processing	Conveyance (point) & Fugitives	8,760					61.1																1.40E+01	9.00E+00	
FW-Pump	Firewater Pump ²	Engine, 575 HP	500	1.5	0.1	0.3	0.0	0.0	3.77E-05	7.39E-04	8.91E-05	8.99E-04				1.14E-03				6.17E-05		2.49E-03	3.94E-04	2.75E-04	6.14E-03	
FL-1	HP / Emergency Flare ³	Flare, 0.816 MMBtu/hr	8,760	0.5	1.0	3.0	0.0																		0.00E+00	
FL-2	LP Flare ³	Flare, 0.204 MMBtu/hr	8,760	0.1	0.2	0.7	0.0																		0.00E+00	
Total Emissions				175.8	176.8	187.4	32.7	195.8	0.00	0.38	0.08	0.00	8.54	0.23	0.00	0.34	0.71	1.28	0.00	9.16	0.81	0.02	0.28	1.81	0.77	23.59

Notes:
¹ Emissions from auxiliary boiler and process heaters assume operation at full design capacity, firing natural gas; however, the equipment may not always fire at full load, and in many cases, will be firing a lower-BTU fuel gas mixture instead of gas.
² SO₂ emissions from the Firewater Pump are based on burning ultra-low sulfur diesel (15 ppm).
³ Flare emissions include pilot emissions for 8760 hr/yr.

Malfunctions and Other Events

ID No.	Description	Usage	Operation (hours) ¹	Potential Emissions (tons)					HAPs Emissions (t/yr)																
				NO _x	CO	VOC	SO ₂	PM ₁₀	1,3-Butadiene	Acetaldehyde	Acrolein	2,2,4-Trimethylpentane	Benzene	Carbonyl Sulfide	Dichlorobenzene	Ethyl Benzene	Formaldehyde	Hexane	Mercury	Methane	Naphthalene	Propylene Oxide	Toluene	Xylene	TOTALS
CO2 VS	CO2 Vent Stack	CO2 Vent Stack	50		63.97	0.23																			2.35E-01
FL-1	HP / Emergency Flare	Flare, 0.816 MMBtu/hr	40	7.83	84.99	0.12	150.16																		0.00E+00
FL-2	LP Flare	Flare, 0.204 MMBtu/hr	8	1.15E-02	2.25E-04	6.79E-04	14.40																		0.00E+00
GP-1	Gasification Preheater	Heater, 21.00 MMBtu/hr	500	0.26	0.43	0.03	3.09E-03	0.04				1.08E-05	6.18E-06		3.86E-04	9.26E-03								1.75E-05	9.89E-03

Notes:
¹ The hours shown are annual estimates, except for the Gasification Preheater which is based on 500 hours per preheating event for one grafter.

Medicine Bow Fuel & Power Industrial Gasification & Liquefaction Plant
Equipment Leaks Emission Summary

Process Stream	Service Type	Controlled Emissions		Uncontrolled Emissions	
		SOCMI Factors		SOCMI Factors	
		VOC Emissions (ton/yr)	HAP Emissions (ton/yr)	VOC Emissions (ton/yr)	HAP Emissions (ton/yr)
Acid Gas	Gas	0.09	0.09	0.12	0.12
Flare KO Drum Drainage	Gas	4.50	1.45	6.70	2.16
Gasifier Vent	Gas	0.14	0.14	0.22	0.22
Gasoline (Gas)	Gas	9.30	3.00	12.38	3.99
Gasoline (Light Liquid)	Light Liquid	10.42	3.36	36.22	11.67
Gasoline (Heavy Liquid)	Heavy Liquid	0.17	0.05	0.26	0.09
LPG	Light Liquid	0.77	0.00	2.21	0.00
Methanol Gas	Gas	0.99	0.99	1.28	1.28
Methanol Pure Liquid	Light Liquid	0.47	0.47	1.44	1.44
Methanol Product (MeOH 1)	Light Liquid	4.83	4.82	13.78	13.75
Methanol Product (MeOH 2)	Light Liquid	0.06	0.06	0.54	0.54
Methanol Product (MeOH 3)	Light Liquid	0.06	0.06	0.54	0.54
Methanol Product (MeOH 5)	Gas	0.35	0.35	0.50	0.50
Mixed Fuel Gas	Gas	0.40	0.01	1.77	0.06
MTG Fuel Gas	Gas	3.88	0.04	5.44	0.06
Propylene	Gas	22.11	0.00	24.36	0.00
Total		58.51	14.89	107.74	36.41
		Controlled Emissions		Uncontrolled Emissions	
		SOCMI Factors		SOCMI Factors	
Individual HAPs		HAP Emissions (lb/hr)	HAP Emissions (ton/yr)	HAP Emissions (lb/hr)	HAP Emissions (ton/yr)
Carbonyl Sulfide (COS)		0.05	0.23	0.08	0.35
Methanol (MeOH)		1.54	6.76	4.13	18.11
C6 - C10 Aromatics (Assumed to be Benzene)		1.80	7.90	4.10	17.96
Total		3.40	14.89	8.31	36.41

Medicine Bow Fuel & Power Industrial Gasification & Liquefaction Plant
Methanol Product (MeOH 1) Process Stream

Stream Name: Methanol Product (MeOH 1)
Service Type: Light Liquid
Hours of Operation: 8760
This piping is included in the LDAR program.

Chemical Name	CAS Number	VOC	HAP	Molecular Weight (lb/lb-mol)	Weight %	Mole Fraction	Mole Percent
CO	630-08-0	N	N	28.01	0.02%	8.44E-06	0.02%
H2	1333-74-0	N	N	2.02	0.00%	3.19E-06	0.01%
CO2	124-38-9	N	N	44.01	0.30%	6.92E-05	0.22%
H2O	7732-18-5	N	N	18.02	3.18%	1.75E-03	5.49%
CH4	74-82-8	N	N	16.04	0.03%	1.59E-05	0.05%
Ar	7440-37-1	N	N	39.95	0.08%	1.61E-05	0.05%
N2	7727-37-9	N	N	28.01	0.03%	1.14E-05	0.04%
H2S	7783-06-4	N	N	34.08	0.00%	0.00E+00	0.00%
COS	463-58-1	Y	Y	60.07	0.00%	0.00E+00	0.00%
NH3	7664-41-7	N	N	17.03	0.00%	0.00E+00	0.00%
O2	7782-44-7	N	N	32.00	0.00%	0.00E+00	0.00%
SO2	7446-09-5	N	N	64.06	0.00%	0.00E+00	0.00%
Cl2	7782-50-5	N	Y	70.91	0.00%	0.00E+00	0.00%
HCl	7647-01-0	N	Y	36.46	0.00%	0.00E+00	0.00%
MeOH	67-58-1	Y	Y	32.04	96.19%	3.00E-02	94.01%
Ethanol	64-17-5	Y	N	46.07	0.05%	1.04E-05	0.03%
Dimethyl Ether	115-10-6	Y	N	46.07	0.03%	7.31E-06	0.02%
Methyl Acetate	78-20-9	Y	N	74.08	0.08%	1.10E-05	0.03%
Propanol	71-23-8	Y	N	60.10	0.02%	4.00E-06	0.01%
Butanol	71-36-3	Y	N	74.12	0.02%	2.60E-06	0.01%
Acetone	67-64-1	Y	N	58.08	0.00%	3.31E-07	0.00%
MEK	78-93-3	Y	N	72.11	0.00%	1.33E-07	0.00%
Ethane	74-84-0	N	N	30.07	0.00%	0.00E+00	0.00%
Ethylene	74-85-1	Y	N	28.05	0.00%	0.00E+00	0.00%
Propane	74-98-6	Y	N	44.10	0.00%	0.00E+00	0.00%
Propylene	115-07-1	Y	N	42.08	0.00%	0.00E+00	0.00%
Isobutane	75-28-5	Y	N	58.12	0.00%	0.00E+00	0.00%
N-Butane	106-97-8	Y	N	58.12	0.00%	0.00E+00	0.00%
Bulylene	25167-87-3	Y	N	56.11	0.00%	0.00E+00	0.00%
Isopentane	78-78-4	Y	N	72.16	0.00%	0.00E+00	0.00%
C4 - C12 Paraffins	N/A	Y	N	114.23	0.00%	0.00E+00	0.00%
C4 - C12 Olefins	N/A	Y	N	112.21	0.00%	0.00E+00	0.00%
C6 - C10 Naphthenes	N/A	Y	N	112.21	0.00%	0.00E+00	0.00%
C6 - C10 Aromatics	N/A	Y	Y	78.11	0.00%	0.00E+00	0.00%
TOTALS					100.00%	3.19E-02	100.00%

Assumed Octane
Assumed Octene
Assumed Cyclooctane
Assumed Benzene

Weight % TOC	96.42%
Weight % VOC	95.40%
Weight % HAP	96.19%

Fugitive Emissions - SOCM Factors				Controlled Emissions				Uncontrolled Emissions	
Equipment Type	SOCM Emission Factor ¹ (kg/hr-source)	% Control With LDAR ^{2,3}	Source Count	TOC Emission Rate (kg/hr)	VOC Emission Rate (kg/hr)	Hours of Operation	VOC Emissions (tpy)	VOC Emissions (tpy)	
Valves-Gas	0.00597	92.00%	0	0.0000	0.0000	8760	0.00E+00	0.00E+00	
Valves-Light Liquids	0.00403	86.00%	134	0.0625	0.0625	8760	6.03E-01	5.03E+00	
Valves-Heavy Liquids	0.00023		0	0.0000	0.0000	8760	0.00E+00	0.00E+00	
Pump Seals-Light Liquids	0.01990	73.90%	22	0.1102	0.1101	8760	1.06E+00	4.07E+00	
Pump Seals-Heavy Liquids	0.00862		0	0.0000	0.0000	8760	0.00E+00	0.00E+00	
Compressor Seals-Gas	0.22800		0	0.0000	0.0000	8760	0.00E+00	0.00E+00	
Relief Valves-Gas/Vapor	0.10400		0	0.0000	0.0000	8760	0.00E+00	0.00E+00	
Connectors	0.00183	93.00%	96	0.0119	0.0119	8760	1.14E-01	1.83E+00	
Open-ended Lines	0.00170		16	0.0262	0.0262	8760	2.53E-01	2.53E-01	
Sampling Connections	0.01500		20	0.2893	0.2892	8760	2.79E+00	2.79E+00	
Totals				0.50	0.50		4.83	13.78	

¹ EPA-453/R-95-017 Protocol for Equipment Leak Emission Estimates (Table 2-1).
² EPA-453/R-95-017 Protocol for Equipment Leak Emission Estimates (Table 5-2). Assumes monthly monitoring with leak definition of 10,000 ppmv.
³ Assumes monthly monitoring with leak definition of 2,000 ppmv for pumps in light liquid service. See Pump LDAR Control Effectiveness Calculation page.

HAP Emissions - SOCM Factors				Controlled Emissions		Uncontrolled Emissions	
HAP	Individual HAP Weight %	VOC Weight %	Hours of Operation	HAP Emissions (lb/hr)	HAP Emissions (ton/yr)	HAP Emissions (lb/hr)	HAP Emissions (ton/yr)
COS	0.00%	96.40%	8760	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Cl2	0.00%	96.40%	8760	0.00E+00	0.00E+00	0.00E+00	0.00E+00
HCl	0.00%	96.40%	8760	0.00E+00	0.00E+00	0.00E+00	0.00E+00
MeOH	96.19%	96.40%	8760	1.10E+00	4.82E+00	3.14E+00	1.37E+01
C6 - C10 Aromatics	0.00%	96.40%	8760	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Total				1.10	4.82	3.14	13.75

Control Number: J020

Category: NSPS
Region: DSSE
Date: 05/03/1982
Title: Oil Shale Facilities
Recipient: Dattilo, A. A.
Author: Reich, Edward E.
Comments:

Subparts: Part 60 D Foss. Fuel Fired Steam Gen. (post 8/17/71), Part 60 J Petroleum Refineries

References: 60.100, 60.101, 60.101(g), 60.40

Abstract:

Are the facilities used for the processing of oil shale subject to the requirements of Subpart J?

The oil shale facility is a petroleum refinery. The process heaters and steam boiler are subject to the Subpart J since the gas being burned is fuel gas.

Letter:

Control Number: J020

May 03, 1982

Mr. A. A. Dattilo
Davy McKee Corporation
6200 Oak Tree Blvd.
Cleveland, Ohio 44131

Dear Mr. Dattilo:

This letter is in response to your request dated March 29, 1982 for a determination of applicability. Specifically, you have asked whether the facilities used for the processing of oil shale are subject to the requirements of the New Source Performance Standards for steam generators (Subpart D) or for petroleum refineries (Subpart J). We have determined that the oil shale facility is a petroleum refinery as defined in 40 CFR 60.101 and the process units are subject to the requirements of Subpart J.

In determining the applicability of Subpart J, it is necessary to examine the definitions of petroleum refinery, petroleum, fuel gas, and fuel gas combustion device in 60.101. As stated in 60.101, "petroleum refinery" means any facility engaged in producing gasoline, kerosene, distillate fuel oils, residual fuel oils, lubricants or other products through distillation of petroleum.... The definition of refinery accurately describes the processing being performed at the oil shale facility. Furthermore, the definition of "petroleum" is defined as "crude oil removed from the earth and the oils derived from tar sands, shale, and coal." The oil shale facility, therefore, is within the scope of these definitions.

Since it has been determined that the oil shale facility is a petroleum refinery as that term is defined in Subpart J, it is necessary to determine whether your described process units are designated affected facilities. Listed among the affected facilities are fuel gas combustion devices; these devices are defined in 60.101(g) as "any equipment, such as process heaters, boilers, and flares used to combust fuel gas...." We have determined that the process heaters and steam boiler which you have described are subject to the requirements in Subpart J. This determination is based upon your description of the gases being burned in the combustion devices meeting the definition of "fuel gas. Again looking at 60.101, "fuel gas" means any gas which is generated at a petroleum refinery and which is combusted. Fuel has also includes natural gas when the natural gas is combined and combusted in any proportion with a gas generated at a refinery.

After examining your information on the oil shale facility and its process units along with the definitions in 60.101, we have concluded the process units are subject to the requirements in Subpart J.

We have consulted with the Offices of Air Quality Planning and Standards and General Counsel. They concur with this determination. If you have any further questions concerning this matter, please contact Ann Eastham of my staff at (202) 382-2876.

Edward E. Reich
(signed)

cc: Larry Jones - OAQPS
Gail Lacy - OAQPS
Rich Ossias - OGC

*Selected Technical and Economic
Comparisons of Synfuel Options*

October 1982

NTIS order #PB83-147363

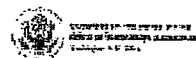
**INCREASED AUTOMOBILE
FUEL EFFICIENCY AND
SYNTHETIC FUELS**

Alternatives For Reducing Oil Imports

Background Paper #2

Selected Technical and Economic Comparisons
of Synfuel Options

October 1982



DEQ 002931

Preface

This volume contains papers written for OTA to assist in preparation of the report *Increased Automobile Fuel Efficiency and Synthetic Fuels: Alternatives for Reducing Oil Imports*. OTA does not endorse these papers. In several instances, the OTA report reaches somewhat different conclusions because of additional information which was obtained later. These papers, however, may prove valuable for readers needing more detailed or specific information than could be accommodated in the final assessment report, and are being made available for such purposes.

SELECTED TECHNICAL AND ECONOMIC COMPARISONS

OF SYN-FUEL OPTIONS

FINAL REPORT

April 1981

Prepared For The
Office of Technology Assessment
United States Congress

E. J. Bentz & Associates Inc.

7 9 1 5 R i c h f i e l d R o a d
S p r i n g f i e l d V i e t n a m 2 2 1 5 3

DEQ 002933

NOTICE

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DEQ 002934

CHAPTER 3 : OVERVIEW OF SELECTED SYNTHETIC FUEL CONVERSION PROCESSES

<u>Section No.</u>		<u>Page</u>
3.1	General Synfuel Processes.	3-1
3.2	Coal Gasification.	3-2
3.3	Coal Liquefaction.	3-8
	3.3.1 General	3-8
	3.3.2 Liquid Solvent Refined Coal (SRII)	3-11
	3.3.3 Exxon Donor Solvent	3-14
	3.3.4 H-Coal.	3-17
	3.3.5 Fischer-Tropsch Process	3-21
	3.3.6 Methanol Process.	3-26
3.4	Oil Shale Retorting.	3-31
	3.4.1 General	3-31
	3.4.2 Surface Retorting	3-31
	3.4.3 Modified In Situ Retorting.	3-36
3.5	Comparison of the various Synfuel Systems With Respect to Resource Requirements	3-40

Chapter 3: OVERVIEW OF SELECTED SYNTHETIC FUEL
CONVERSION PROCESSES

3.1 General Synfuel Processes

The General term "synfuel processes" applies to the following:

1. Upgrading of coal. to gaseous, liquid or solid products with improved characteristics.
2. conversion of the kerogen in oil shale to gaseous or liquid fuels or products.
3. Recovery of petroleum crudes from non-conventional oil resources such as heavy oils and tar sands.

Upgrading of coal by subjecting it to a reaction with steam at high temperatures and pressures in the presence of air or oxygen, or to hydrogen, with or without a catalyst, is called conversion. The coal can be converted to gaseous (gasification) or liquid (liquefaction) hydrocarbons. The products have a much lower content of sulfur than the original coal. Oil shale can be retorted by subjecting it to high temperature and pressure, also producing gaseous or liquid hydrocarbons. Catalysts are used in synfuel processes when there is need to accelerate the reaction rates and affect the product state.

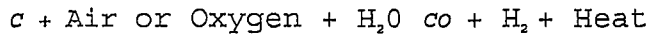
In this report, the following processes are included:

1. Coal gasification
 - to medium Btu gas: generic
 - to high Btu gas: generic
2. Coal Liquefaction
 - by pyrolysis (none included)
 - by solvent extraction: liquid solvent refined coal (SRC II)
Exxon donor solvent (EDS)
 - by catalytic liquefaction: H-coal
 - by indirect liquefaction: Fischer-Tropsch (FT) Methanol
3. Oil shale retorting using:
 - true in situ retorting (none included)
 - modified in situ: generic
 - surface retorting: generic

3.2 coal Gasification

The process by which coal is gasified involves reactions of devolatilization of coal with steam at elevated pressures and temperatures to produce CO and H₂O. Gasification of coal involves basically

the following reaction between steam and carbon:



There are many processes by which coal can be gasified producing low-, medium- or high-Btu gas. The definitions of the heat content of each of them are not rigorous. Low-Btu gas is a mixture of carbon monoxide, hydrogen and nitrogen. It has a heating value of less than 300 Btu per standard cubic foot (Reference No. 25). This gas is of interest to industry either as a combustible fuel or as a raw material from which ammonia, methanol, and other compounds may be synthesized. Due to the low heating value, it cannot command high enough prices to justify long distance transport. Medium-Btu gas is a mixture of methane carbon monoxide, hydrogen, and other gases. It has a heating value between 300 and 700 Btu per standard cubic foot (Reference No. 25). It is suitable as a fuel for industrial consumers, but because of its low heating value, is not economic to transport over great distances. High-Btu gas consists essentially of methane. It has a heating value of approximately 1000 Btu per standard cubic foot, and is compatible with natural gas in that it can be substituted for natural gas in existing pipeline systems.

Coal gasification processes can be divided into three major process types according mainly to the way in which the feedstock coal, steam, and the product **gases** are **contacted**. They are:

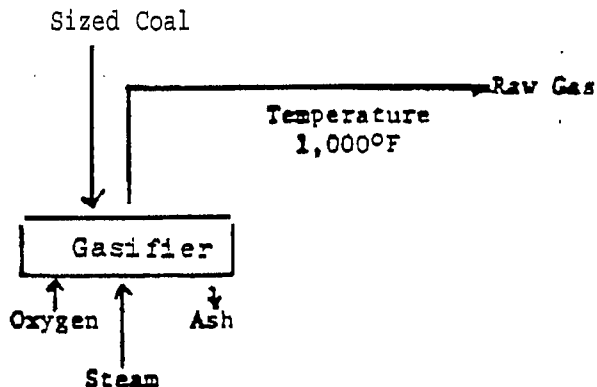
1. Fixed bed gasification in which the crushed, sized coal is fed from the top of the reactor vessel. Steam, air or oxygen are blown upwardly.
2. Fluidized bed gasification in which the finely sized coal particles are "fluidized" by the steam, air or oxygen, which are piped through them.
3. Entrained bed gasification: in which the even finer coal particles are blown into the reacting gas stream prior to entry into the reactor. The coal particles are suspended in the gas phase, and are filtered and recycled until a product gas with a suitable heating value is produced.

Figure 3.1 (Reference No. 31) describes the main features of these three **processes**.

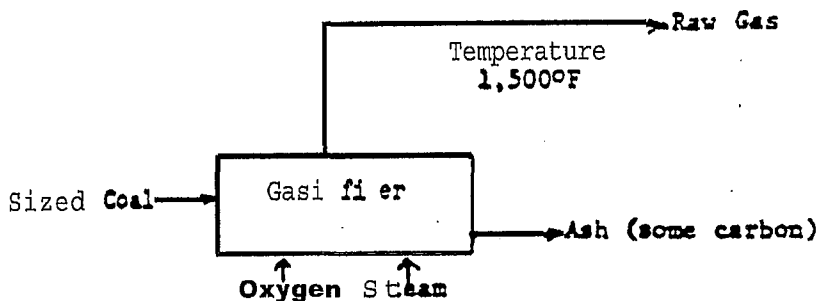
1

Usually, low-Btu gas has a heating value below 200 Btu per SCF; and medium-Btu gas ranges in heating value between 300 - 350 Btu per SCF.

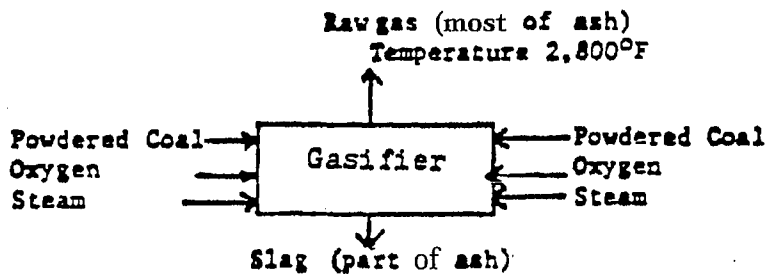
Figure 3.1: Basic Coal Gasification Processes



Fixed Bed Gasification e.g. Lurgi



Fluid Bed Gasification e.g. Winkler



Entrained Gasification e.g. Koppers-Totzek

SOURCE : Reference 31

Figure 3.2 (Reference No. 31) is a schematic diagram of coal gasification. It represents the whole coal gasification fuel cycle, including the production of low-, medium- or high-Btu gas. All of these gasification processes share a number of process steps. If high-Btu, pipeline-quality gas is desired, essentially all of the following process steps are required. In some cases, some of them may be omitted, depending on the type of coal being processed and the type of gas product desired. The process steps are as follows (Reference No. 25).

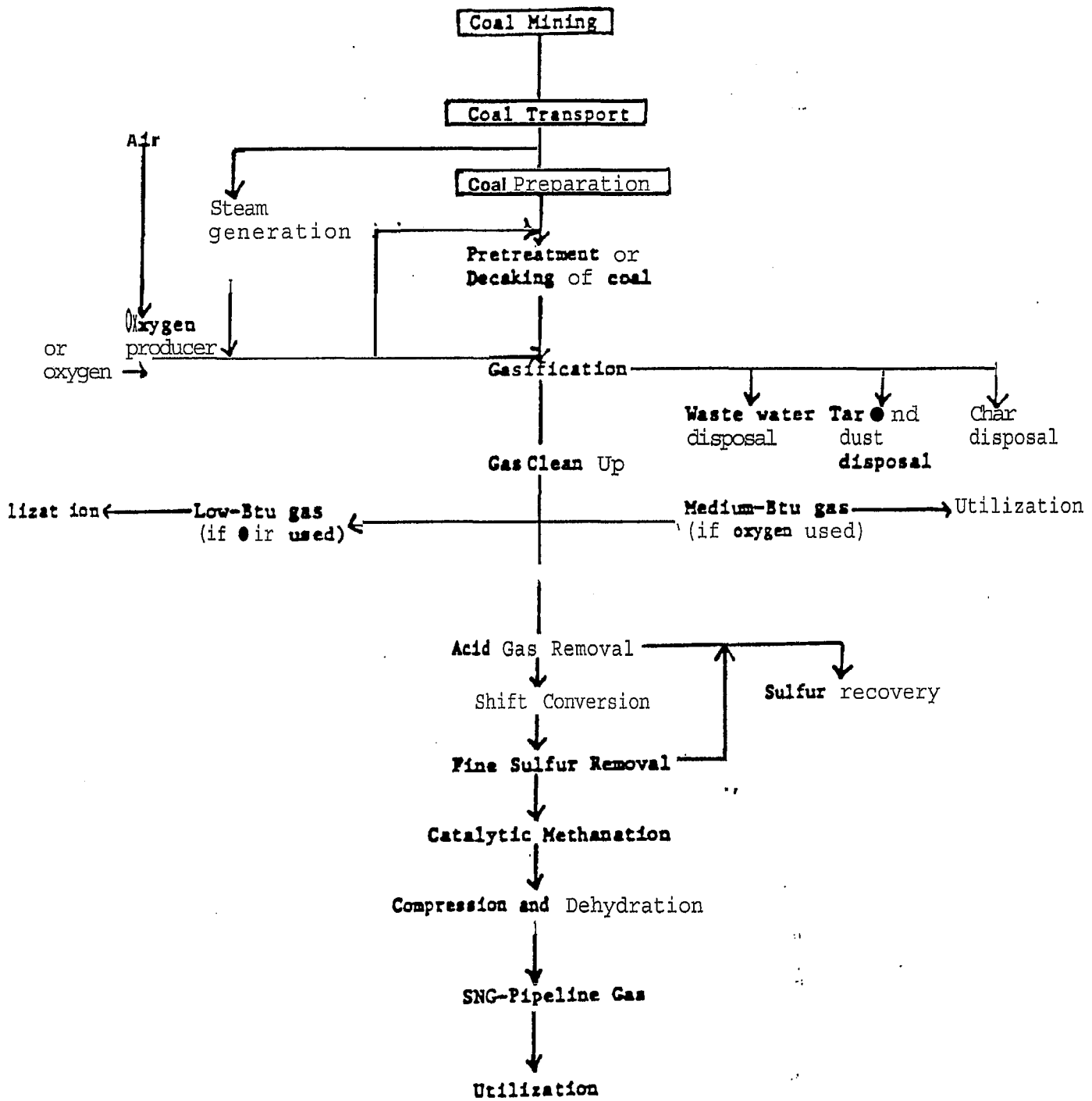
1. Pretreatment of coal² (if sizing or caking are problems) .
2. Primary gasification of coal.
3. Secondary gasification of carbonaceous residue from primary gasifier.
4. Removal of CO₂, H₂S, and other acid gases.
5. shift conversion for adjustment of the carbon monoxide/hydrogen mole ratio to the desired 1:3.
6. Catalytic methanation of the carbon monoxide/hydrogen **mixture** to form methane.

Pretreatment

The coal received at the plant must be further cleaned and crushed or ground before it can enter the gasifier. Extraneous materials such as shale, rocks, metal, etc. are removed by conventional cleaning methods. For fluidized or entrained gasification processes, the coal needs to be finely ground. Crushing and sizing may also be required for other processes. In the case of certain bituminous coals called caking coals, agglomeration of the material is observed when they are heated. Treatment is needed if they are to be gasified by fluidized or Moving bed processes, or even in fixed bed reaction. The caking characteristics are destroyed when the coal is heated to low temperatures in the presence of air or oxygen.

2 Pretreatment of coal by partial oxidation with air or oxygen is not in general a cost-effective approach to destroying the **caking** characteristics of certain coals, such as Eastern bituminous coals, because of the loss of Btu values of the coal in producing CO₂ & H₂O. The **caking** problem is a serious problem in the processing of such coals and limits the applicability of current commercial gasifiers such as the dry-bottom Lurgi to Western Subbituminous coals and lignite.

Figure 3.2: Schematic Diagram of Coal Gasification



SOURCE: Reference 31

Primary Gasification

This is the heart of the process, and is basically a pyrolysis process of the raw coal. The coal feed is contacted with synthesis gas (carbon monoxide and hydrogen). The coal is devolatilized according to the following general reaction (Reference No. 25).

COAL + HEAT (Pyrolysis) + Methane, water, tars, phenols,
hydrogen sulfide, hydrogen,
carbon dioxide, char, etc.

The pressures used for gasification range from atmospheric pressure to 1000 psi. The heat required to maintain the endothermic gasification reaction is supplied from burning coal. Air or oxygen are also needed to support the combustion reaction. If air is used, the product is low Btu gas ranging from essentially a carbon monoxide/hydrogen mixture (Koppers-Totzek process) to mixtures containing various proportions of carbon monoxide, carbon dioxide, hydrogen, water, methane, hydrogen sulfide, nitrogen, and typical products of pyrolysis such as tar, oils, phenols, etc. If oxygen is used, medium Btu gas results.

The bulk of the original coal is transformed into a solid char. Certain coals are more "reactive" to gasification than others. Thus the type of coal being processed determines to a large extent the amount of char produced, and the analysis of the gaseous products. The char is usually gasified by additional processing steps, or is marketed.

Secondary Gasification

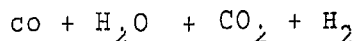
Secondary gasification involves the gasification of char from the primary gasifier. This is usually done by reacting the hot char with water vapor to produce carbon monoxide and hydrogen.

If the desired final product is either low- or medium-Btu gas, secondary gasification is usually followed by scrubbing and cleaning. Carbon dioxide and sulfur compounds are partially removed, and the resulting gas is used directly. If high-Btu gas is desired, shift conversion and methanation are further required.

Shift Conversion

In most gasification processes, a shift reaction is employed prior to methanation. Its purpose is to react

a portion of the carbon monoxide with steam to form more hydrogen.

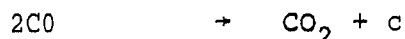
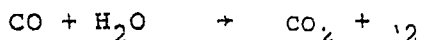
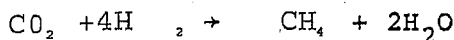
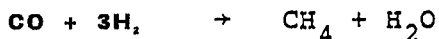


By this exothermic reaction the ratio of carbon monoxide to hydrogen may be increased to 1:3 mole ratio needed to produce methane. Otherwise, deactivation of the catalyst used in the methanation takes place.

The catalytic shift conversion reaction is a well-known process, but it has not been applied on the large scale required for commercial coal gasification. For coal gas shifting, conventional iron-chromium catalysts may be used; however, the coal gas stream must be purified prior to shifting (Reference No. 25).

Methanation

If carbon monoxide and hydrogen are present in the mole ratio of 1:3, the coal gas can be reacted in the presence of a catalyst to produce methane. Group VII transition elements such as iron, cobalt, nickel, ruthenium, rhodium, palladium, osmium, iridium, and platinum have been found to be effective catalysts. The following exothermic reactions occur simultaneously within the methanation unit (Reference No. 25).



Special care must be taken to prevent deactivation of the catalyst by temperatures above 750°F. It can also be poisoned by carbon deposition. These can be circumvented by ensuring that the mixture of carbon monoxide and hydrogen shall be fed to the methanator in the ratio of 1:3. Scrubbing of sulfur from the synthesis gas feed is employed to alleviate sulfur poisoning of the catalyst.

The final step to prepare high Btu gas for marketing is to remove water to specified levels. The product gas usually undergoes compression prior to storage or marketing.

3.3 Coal Liquefaction

303.1 General

Coal liquefaction processes are conversion processes in which liquids are the primary products. Some gases and solid char may also be produced.

There are two basic routes to coal liquefaction, namely direct and indirect liquefaction. In direct processes, slurried crushed coal is reacted directly with hydrogen at high temperature and pressure conditions to produce liquid hydrocarbons. In indirect liquefaction, coal is first gasified to produce a hydrogen-and carbon monoxide mixture. Further recombination with the aid of a catalyst produces liquid products.

Direct liquefaction is further broken down into three generic processes, namely: pyrolysis, solvent extraction, and catalytic liquefaction. The yields and physical properties of the produced liquid products depend directly on the reactor conditions and degree of hydrogenation.

Pyrolysis

In pyrolysis processes, coal is heated to temperatures above 750°F. It is converted into gases, liquids, and char. The latter accounts for more than 50 percent of the weight of the feed coal and requires hydrogenation. Some amount of solids remain in the raw gas and liquid products. They consist of unreacted coal and ash, and can be relatively easily removed from the gas stream. But the liquid requires filtration, distillation, or some other treatment to remove the solids.

Solvent Extraction

This process makes use of coal derived liquids known as "donor" solvents to increase the fraction of the coal that goes into solution. The "donor" solvents act as source of hydrogen to the coal products, and are reacted together at temperatures up to 950°F. Hydrogen may be supplied under pressure in the extraction step, or it may be used to hydrogenate the solvent prior to recycle. In some processes the unreacted coal is used to generate the necessary hydrogen. In other processes, the hydrogen is generated from by-product gases or from additional raw coal.

Catalytic Liquefaction

In this process, pulverized coal is mixed with 1-1.5 parts of recycle solvent. A suitable catalyst is used to add hydrogen. Most processes of this type operate in the liquid phase with catalyst dispersed throughout or in a fixed bed. Some processes now in the development stage involve the injection of catalyst-impregnated coal into a stream of hot hydrogen at about 950° F for a very short time (Reference No. 25).

Indirect Liquefaction

Two stage conversion of coal typifies indirect liquefaction processes. Coal is first reacted with steam and oxygen to produce a gas composed primarily of carbon monoxide and hydrogen. This gas stream is subsequently purified to *remove sulfur*, nitrogen, and ash. The product gas is then catalytically reacted to yield liquid hydrocarbon products.

Figure 3.3 (Reference No. 31) presents a schematic diagram of the basic liquefaction processes. Each of them produces several types of products and some gas, which may be used within the plant.

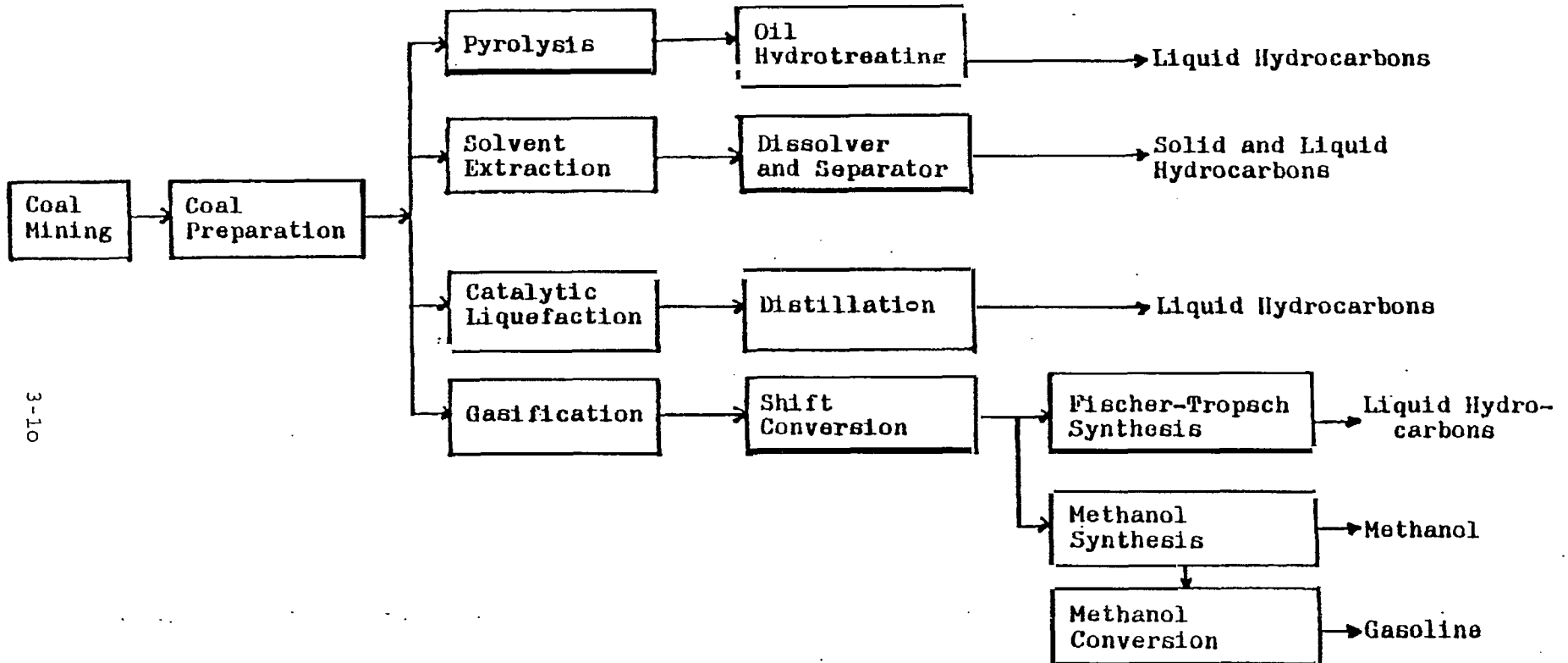
Removal of solids from coal liquids is a critical step in most of these liquefaction processes. Although there is currently a trend toward elimination of the solid-liquid separation step by the recovery of a solids-laden vacuum bottoms stream for gasification, most existing plant designs call for some type of physical/chemical solids removal system.³ The three processes receiving the most current interest are critical solvent deashing, antisolvent deashing, and pressure filtration (Reference No. 25).

Separation of ash and unreacted coal particulate from coal liquids is difficult because of the small size and large quantity of the solid particles, the small density difference between solids and the liquid, and the high viscosity and melting point of the liquids. The Kerr McGee Corporation has been developing a separation technique which utilizes solvents such as benzene, toluene, xylene, pyridene, and cresols near their critical temperature and pressure, hence the term solvent deashing (Reference No. 25).

3

Solid/liquid separation is a critical step only in direct liquefaction process. Most modern coal hydroliquefaction processes in the pilot plant stage of development, such as SRCII, EDS, H-Coal (syncrude mode) do not require a solid/liquid separation stage.

Figure 3.3; Schematic Diagram of the Basic Liquefaction Processes*



3-10

ejbka

* Only major products are shown. The last process represents indirect liquefaction.

SOURCE: Reference 31

3.3.2 Liquid Solvent Refined Coal (SRCII)

The SRCI process was developed to convert high-sulfur, high-ash coals to low-sulfur and ash solid fuels. The SRCII is the same kind of process, except the product is a liquid rather than a solid. This is achieved by adding more hydrogen through the following steps:

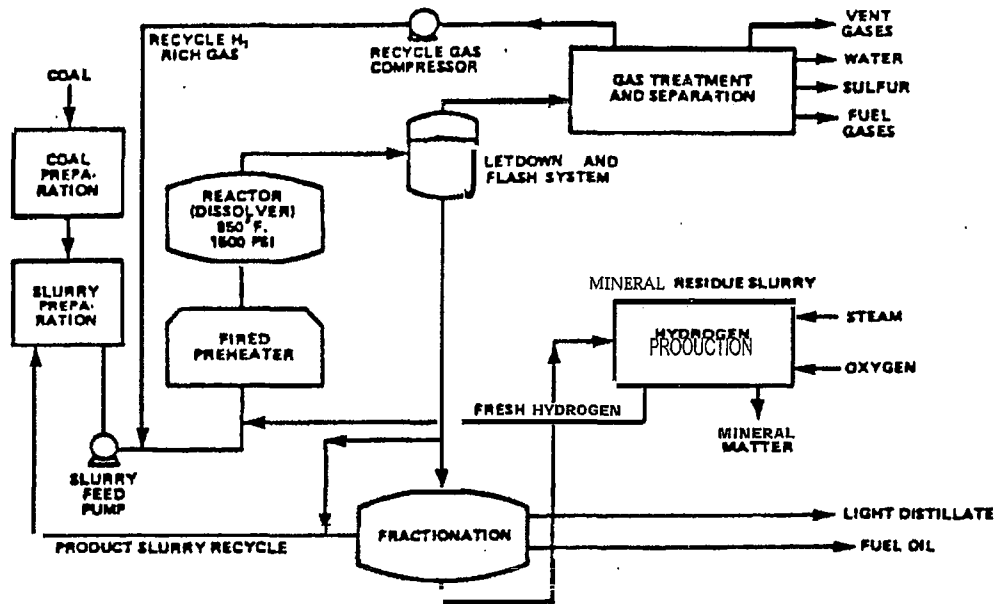
1. Recycling of a portion of the product slurry as solvent for the feed coal.
2. Higher residence time in dissolver.
3. Higher pressure.
4. Use, of vacuum distillation to separate solids from liquid, rather than the troublesome filtration step employed in SRCI .

Figure 3.4 is a schematic diagram of the SRCII process (Reference No. 35) . Table 3.1 summarizes the components, resource requirements, and potential impacts of this process (Reference No. 17) . The feed coal is first pulverized to less than 1/8" size, dried and mixed with process derived solvent in a slurry mix tank (Reference No. 35) . Feed coal is limited to those containing certain trace mineral elements which may be required to act as catalysts for the breaking of solids to liquids in the liquefaction reaction⁴ (Reference No. 291. However, in cases where the problem is concentration rather than the presence of specific trace elements, a recycle of residue may broaden the allowable coal feeds (Reference No. 29) . The coal slurry is then mixed with hydrogen generated by gasification of the vacuum bottoms from the liquefaction step and reacting with steam and oxygen in a gasifier-converter. The slurry is pumped through a preheater (700 to 750°F) and passed through a dissolver (2000 psi, 820 to 870°F) to dissolve about 90 percent of the coal (Reference No. 35) . The following additional reactions take place in the dissolver (Reference No. 35) .

1. The coal is depolymerized and hydrogenated.
2. The solvent is hydrocracked to form lower molecular weight hydrocarbons, ranging from light oil to methane.
3. Much of the organic sulfur is removed in the form of hydrogen sulfide.

The *slurry stream* from the dissolver is split into two. One is recycled to provide solvent for coal slurry mixing. The other is fractionated to recover the primary

⁴ Opinions differ about the role of the trace minerals as catalysts. The primary "catalyst" in the SRCII process may well be the pyritic mineral matter contained in the coal and not "trace mineral elements. "



SOURCE : Reference 35

Figure 3.4

A SCHEMATIC DIAGRAM OF THE
SOLVENT REFINED COAL (SRC-II) PROCESS

ENERGY SYSTEM:

- SIZE:**
- 36,000 bbl/day (7)
 - 0.335×10^{12} Btu/day (2)
 - 90% plant availability (2)
 - operation 329 days/year (2)
 - 110.22×10^{12} Btu/year (2)
 - 31,552 tons per day ROM coal (a)
 - efficiency 84% (d)
 - plant size 30 years (a)
 - 6.7×10^6 Btu/bbl (3)

DESCRIPTION:

Coal is crushed to about 200 mesh and dried to three percent moisture and slurried with a solvent. A 70 to 85 percent hydrogen mixture is added. This mixture is first preheated, then hydrogenated in the reactor. The temperature is 425 to 495°C; pressure 58 to 136 atmospheres. Liquefaction produces a mixture of gases, vapor, liquids, and solids. The gases and vapors are separated from the liquids and solids and the vapors condensed. The liquid is separated from the solids by filtration (SRC 11).

EQUIPMENT:

- crushers
- SRC reactor
- drying unit
- pre-heaters
- gas recovery plant
- filtering unit

ENVIRONMENTAL CONCERNS:

- air emissions
- water pollution from runoff and leachates
- solid waste
- occupational hazards and health effects
- noise pollution
- odor

*The data presented are based on a conceptual design of a commercial facility. The data will be updated when more current data become available. The data should not be used directly for comparison with other coal liquefaction processes.

- (1) After coal preparation includes solid waste from mine burial and tailings pond.
- (2) Representative values collected for analysis purposes.

SOURCE:

- (a) Hittman Associates, Inc., Standards of Practice Manual for Solvent Refined Coal Liquefaction Process, 1978.
- (b) Hittman Associates, Inc., Environmental Assessment of Coal Liquefaction, 1978.
- (c) TMS, Characteristics and Data in the Area of Coal from Environmental Data Book, Volume IV, 1978.
- (d) Flack and Kypar, Boiler Fuel and Building Blocks, 1978.

SOURCE: Reference 17

RESOURCES USED:
(For 10^{12} Btu Produced)

FUEL (DAY) (c)
 coal-bituminous, plant spec 40,919 tons
 energy content 12,821 Btu/lb

COAL ANALYSIS (DAY) (a)

	% by weight
moisture	2.70
volatile matter	35.09
fixed carbon	51.70
sulfur	1.18
ash	7.13

LAND (c)
 Acres
 plant facilities 3.5
 solid waste 0.25
 (per year)

WATER (c)
 Acq.-Ft.
 cooling towers 64.5
 plant use 11.0
 steam 13.3
 miscellaneous 0.9
 total 89.5

COSTS
 construction NA
 operation NA

PERSONNEL (c)
 construction NA
 operation & maintenance NA

RESIDUALS AND POLLUTANTS:
(Per 10^{12} Btu Produced)

AIR POLLUTANTS (c)

	Tons
particulates	1.8
SO ₂	2.0
NO _x	10.0
hydrocarbons	1.0
CO	1.8
arsenic	1.7
radium	2.9
mercury	1.3
zinc	2.3
chromium	3.9
nickel	6.2
lead	1.5
polyuclear organic material	

WATER POLLUTANTS
 no direct discharge to any watercourse

SOLID WASTE (c)

	Tons
dry ton equivalent	36,141 (1)

ENERGY EFFICIENCY (2)
 equivalent (SRC 11)
 149,254
 (energy content = 6.7×10^6 Btu/bbl)

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products which consist of naphtha, low sulfur fuel oil, and a vacuum residue which is separated from the solution in a filtration unit. The residue consists of heavy oil, ash and undissolved organic material from the coal (Reference No. 25).

The gases from the dissolver are treated to remove hydrogen sulfide and carbon dioxide. Liquid petroleum gases and pipeline gas are separated in a cryogenic separation unit. Unreacted hydrogen is recovered and recycled.

Recent developments have resulted in increased efficiency of the SRCII process. A combination of solid and liquid products are produced. A wide range of products can be obtained depending on the severity of recycling. Table 3.2 (Reference No. 25) shows the properties of a typical mix of products.

3.3.3. Exxon Donor Solvent (EDS)

The process is similar to SRCII, except that the major portion of the hydrogen supplied as part of the solvent is chemically combined rather than in the form of a free dissolved gas (Reference No. 29). A schematic diagram of the process is illustrated in Figure 35 (Reference No. 35). Crushed coal is liquefied in a reactor at 800-880°F and 1500 - 2000 psig (Reference No. 25). The reaction is non-catalytic, in the presence of molecular hydrogen and the hydrogen-donor solvent, which transfers hydrogen to the coal. The product from the liquefaction reactor is separated into two portions. One part is sent to the solvent hydrogenation unit to produce donor solvent. It is a catalytically hydrogenated recycle stream which is fractionated from the middle boiling range of the liquid product, and has a boiling range of 400 - 850°F (Reference No. 25). After hydrogenation, the solvent is mixed with fresh coal feed, heated in a furnace, and pumped into the liquefaction reactor.

The other portion from the product liquefaction reactor is a slurry. It is separated by distillation into gas, naphtha, middle distillate, and a bottom product that contains heavy liquid, untreated coal and mineral matter. The vacuum bottoms slurry is cooked to produce additional liquids.

The major advantages of the EDS process are:

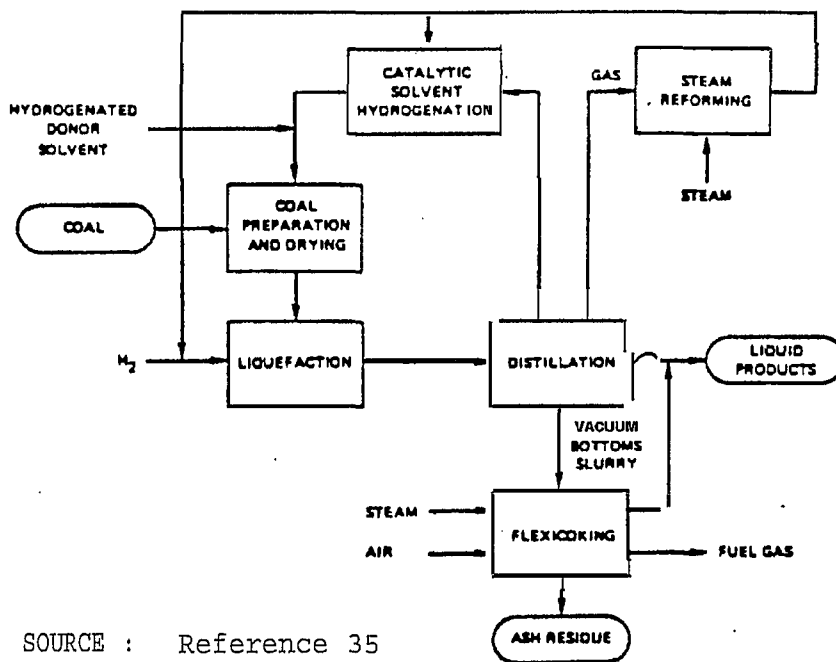
1. High yields of low sulfur liquids are obtained from bituminous and sub-bituminous coals or lignites (Reference No. 25). A yield

TABLE 3.2
TYPICAL PROPERTIES OF SRC FUELS
USING RECYCLE SRC II PROCESS

	<u>Solid Fuel</u>	<u>Distillate Fuel</u>
Gravity: 'API	-18.3	5.0
Approximate Boiling Range: 'F	800+	400-800
Fusion Point: 'F	350	
Flash Point: 'F		168
Viscosity: SUS at 100°F		50
Sulfur*: Percent	0.8	0.3
Nitrogen*: Percent	2.0	0.9
Heating Value: Btu/lb.	16,000	17,300

* Assuming Western Kentucky coal feed with 4% Sulfur and 2% Nitrogen.

SOURCE: Reference 15



SOURCE : Reference 35

Figure 3.5
A Schematic Diagram of the
Donor Solvent Liquefaction Process

of 2.6 barrels of liquids per ton of dry coal is typical for an Illinois bituminous coal (Reference No. 25).

2. The only by-products of significance are ammonia and elemental sulfur (Reference No. 25).
3. There is wide flexibility in product distribution by varying liquefaction conditions or adjusting solvent properties (Reference No. 25).

The typical properties of the products from the EDS process are shown in Table 3.3 (Reference No. 25). An estimated heat balance is given in Table 3.4 (Reference No. 35).

3.3.4 H-Coal

The H-coal process converts coal to hydrocarbon liquids by hydrogenation with a cobalt-molybdenum catalyst. An ebullated bed reactor is employed. The liquid products may range from a heavy boiler fuel to a synthetic crude product (Reference No. 25).

Figure 3.6 (Reference No. 35) is a schematic diagram of the H-coal process. Coal is first crushed to minus 60 mesh, dried, and then slurried with recycled oils at pressures of approximately 200 atmospheres (Reference No. 25). " Mixing of the slurry with compressed hydrogen follows, and the mixture is preheated. The material is pumped to the bottom of the ebullated bed reactor, with the upward flow of slurry through the reactor maintaining the catalyst in a fluidized state (i.e. random motion). The catalyst needs periodic additions of fresh catalyst and withdrawals of spent portions. Typical temperatures of the slurry entering the reactor are 650 - 700°F (Reference No. 25). The finely divided coal and ash particles flowing through the ebullating bed are removed with liquid and vapor products.

The reactor effluent is separated into recycle and net product streams. Conventional processing equipment is used. The liquid stream is distilled to produce a mixture of light distillate and a heavy distillate product. Gaseous products composed of hydrocarbon gas, hydrogen sulfide and ammonia are separated. A portion of the heavy distillate is recycled as the slurring medium.

The operating conditions of the H-Coal process can be altered to produce various types of primary products. For

TABLE 33

DONOR SOLVENT PRODUCT ANALYSES

	Heavy Naphtha ¹		200°C+ Fuel Oil	
	Raw Liquid	Hydrotreated Liquid	Raw Liquid	Hydrotreated Liquid
Nominal Boiling Range, °C	70/200	70/200	200/540	200/540
Distillation, 15/5°C				
10 wt. %	106	92	247	239
50 wt. %	180	157	368	347
90 wt. %	199	182	433	412
Density (g/cm ³)	0.87	0.80	1.08	1.01
Elemental Analysis, Wt. %				
C	85.60	86.80	89.40	90.80
H	10.90	12.90	7.70	8.60
O	2.82	0.23	1.83	0.32
N	0.21	0.06	0.66	0.24
S	0.47	0.005	0.41	0.04
Higher Heating Value MJ/kg	42.6	44.9	39.8	42.1

¹Excludes C₆/70°C naphtha cut

SOURCE: Reference 25

Table 3.4

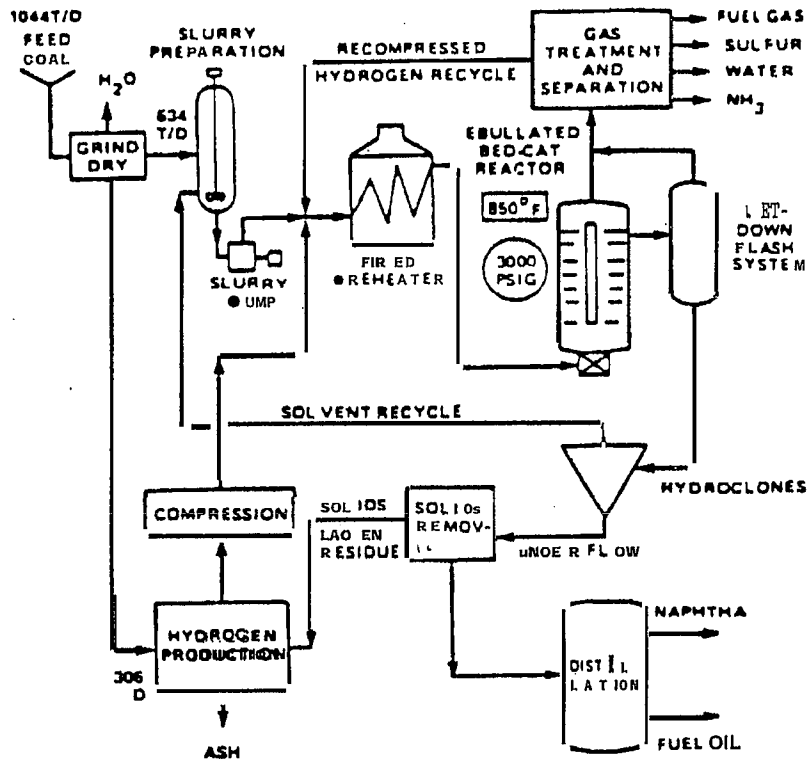
Estimated Heat Balance for a Commercial Scale EDS Plant

	Btu/day (10 Btu's)	Percent of Total Energy Input
<u>System Products</u>		
Liquids	323,071	61.72
Sulfur, ammonia	8,309	1.59
<u>System Losses</u>		
Ash, combustibles and sensible heat	26,082	5.13
Stack losses	20,039	3.83
Energy losses via water and air	136,853	26.14
Liquefaction and solvent hydrogenation (9.80%)		
Flexicoking (6.44%)		
Hydrogenation and recovery (6.72%)		
" By-product recovery, offsites, and miscellaneous (3.18%)		
Other miscellaneous	8,309	1.59
<u>Energy Input</u>		
Coal (cleaned)*	488,761	93.37
Electrical power**	34,702	6.63

* Coal - Illinois No. 6; 10,574 Btu/lb as received prior to cleaning

** Power based on 8,500 Btu/kwh to generate

SOURCE: Reference 35



SOURCE : Reference 35

Figure 3.6

SCHMATIC DIAGRAM OF THE
H-COAL PROCESS

example, relatively high temperatures and high hydrogen partial pressures are used to produce a synthetic crude products. Vacuum distillation is used to separate the solids from the liquid phase. If gas and oil are desired, lower temperatures and pressures are used (Reference No. 25). Conversion and yield structure are determined by reactor conditions, catalyst replacement rate, and recycle slurry oil composition (Reference No. 29).

Table 3.5 (Reference No. 25) summarizes the properties of both the fuel oil syncrude products from H-coal.

Table 3.6 (Reference No. 17) summarizes the components, resources and potential impacts from H-coal process. It requires between 14,000 and 20,000 standard cubic feet of hydrogen for each ton of coal produced. Hydrogen consumption depends on the type of product produced, with less hydrogen required during the production of residual oil (Reference No. 25).

3.3.5 Fischer-Tropsch Process

A commercial plant using a modification of this process is currently operating in South Africa (Reference No. 36). This is the **only commercial sized** plant producing synfuels. Table 3.7 (Reference No. 35) is an overview of this plant.

In the Fischer-Tropsch process the coal is initially gasified (for description of gasification see section 3.2 of this report). The synthesis gas is then converted to largely aliphatic hydrocarbons using an iron or cobalt catalyst.

Figure 3.7 (Reference No. 35) is a schematic diagram of the SASOL I plant, which utilizes the Fischer-Tropsch process. Thirteen high pressure gasifiers convert coal in the presence of steam and oxygen to medium Btu gas containing mainly carbon monoxide, tars and oils. The product gas is then cleaned of carbon dioxide, hydrogen sulfide, organic sulfur, ammonia, and phenols. The cleaned gas is then subjected to the catalytic Fischer-Tropsch reaction which produces a mixture of gases, liquid hydrocarbons, and an aqueous chemical mixture that must be further processed to set the desired plant output.

The cleaned gas from the Lurgi gasifiers is partitioned into two streams. One stream is reacted in a fixed bed catalytic reactor to produce straight chain and medium boiling oils, diesel oil, LPG, and some alcohols. Operating conditions are 450°F and 360 psig (Reference No. 35). The other stream is combined with reformed product gas to increase the hydrogen to carbon ratio. It is reacted in a fluidized bed reactor

TABLE 3.5

PROPERTIES OF H-COAL DISTILLATES
FROM ILLINOIS NO. 6 COAL LIQUIDS

<u>Property</u>	<u>Fuel Oil</u>		<u>Syncrude</u>	
	<u><203°C distillate</u>	<u>>203°C distillate</u>	<u><197°C distillate</u>	<u>>197°C distillate</u>
Specific gravity, 60°/60°F	0.864	0.979	0.838	1.025
Gravity, 'API	32.3	13.0	37.4	6.6
Pour point, ASTM D-97, F	<5	<5	<5	<5
Color, ASIM D-1500 or (BuMines description)	NPA6	Brownish black	NPA4-1/2	Brownish black
Kinematic viscosity @ 100°F, ASTM D-455, C _s	1.08	3.87	0.96	14.90
Saybolt viscosity, SUS, 100°F		39		77
Sulfur (Bomb) ASTM D-129, wt-pet	0.13	0.29	0.06	0.35
Nitrogen, Kjeldahl, Wt-pet	0.420	0.446	0.212	0.871
Carbon residue (Conradson) ASTM-524, Wt-pet	0	2.33	0	5.44

SOURCE: Reference 25

TABLE 3.6

H-Coal

EXACT SYSTEM:	RESOURCE USED: (Per 10 ⁶ Btu Produced)	RESIDUAL AND PRODUCTS: (Per 10 ⁶ Btu Produced)
<p>SIZE • 30,000 bbl/day • 32 x 1011 Btu/day • 103.11 x 10¹¹ Btu/year • 90% plant availability • operate 128.5 days/year • efficiency 70% • plant life 30 years</p> <p>DESCRIPTION • Coal is crushed to about 3/4 inch and then ground to minus 60 mesh and mixed with recycled oil. The slurry is then preheated and fed to the reactor bed hydrogenated as it comes about 350°C. Gases and vapors are then withdrawn from the top of the reactor and passed through condensers for product separation. The heavier product is withdrawn as a side-stream from the reactor and pumped to an atmosphere distillation unit.</p> <p>COMPONENTS • crushers • drying unit • hydrogen plant (gasifier) • B-coal reactor • gas recovery plant • distillation unit</p> <p>ENVIRONMENTAL CONCERNS • H₂ emissions • sulfur oxides • particulate emissions from transfer and handling • occupational hazards and health effects • noise • odor</p>	<p>FUEL coal, bituminous energy content 37,736 cases 11,900 Btu/lb</p> <p>COAL ANALYSIS moisture 11.4 volatile matter 31.7 fixed carbon 31.0 sulfur 2.1 ash 17.4</p> <p>LAD (1) plant facility solid waste disposal</p> <p>ACIDIC WATER USE process water 101.4 cooling water 7.8 waste water 3.3 potable 2.2 other (utility, quench, etc) 171.8 Total 297.9</p> <p>COSTS construction (2) operation 2,105,000 2,700,000</p> <p>EMERGENCY construction operation & maintenance 4.1</p>	<p>AIR POLLUTANTS particulates SO₂ NO_x hydrocarbons • formal • benzene • toluene • xylene • ethyl benzene • styrene • naphthalene • anthracene • phenanthrene • benzo(a)pyrene • benzo(b)fluoranthene • benzo(k)fluoranthene • dibenz(a,h)anthracene • indeno(1,2,3-cd)perylene • benzofluoranthene • fluorene • acenaphthylene • acenaphthene • fluorene • perylene • benzofluoranthene • fluorene • perylene</p> <p>WATER POLLUTANTS • direct discharge into sea water source • effluent • effluent • effluent</p> <p>SOLID WASTE (1) 5,705 1.1 x 10⁻³</p> <p>ENERGY PRODUCTS liquid fuel (energy content Btu/bbl) 136,736 6.4 x 10⁶</p>

The data presented are based on a conceptual design of a commercial facility. The data will be updated when more content data become available. The data should not be used directly for comparison with other coal liquefaction processes.

(1) This represents load committed to use over the lifetime of the plant, divided by the annual output of the plant, expressed in trillion Btu. (403 acres x 103.11 = 3.9 acres)

(2) This represents total cost of construction the plant, divided by the annual output of the plant, expressed in trillion Btu. (1650 million x 103.11 = 16.3 million).

(3) Does not include solid waste from sludge burial and tailings pond.

SOURCE: Nitman Associates, Inc., Environmental Assessment of Coal Liquefaction, 1978.
 U.S. Environmental Data Book, Volume IV, 1978.
 U.S. Energy Research and Development Administration, Synthetic Liquid Fuel Development: Assessment of Critical Factors, ERDA 76-17973.
 Fluor Engineers and Constructors, Inc., H-Coal, Commercial Evaluation Report, March 1976.

SOURCE: Reference I7

TABLE 3.7

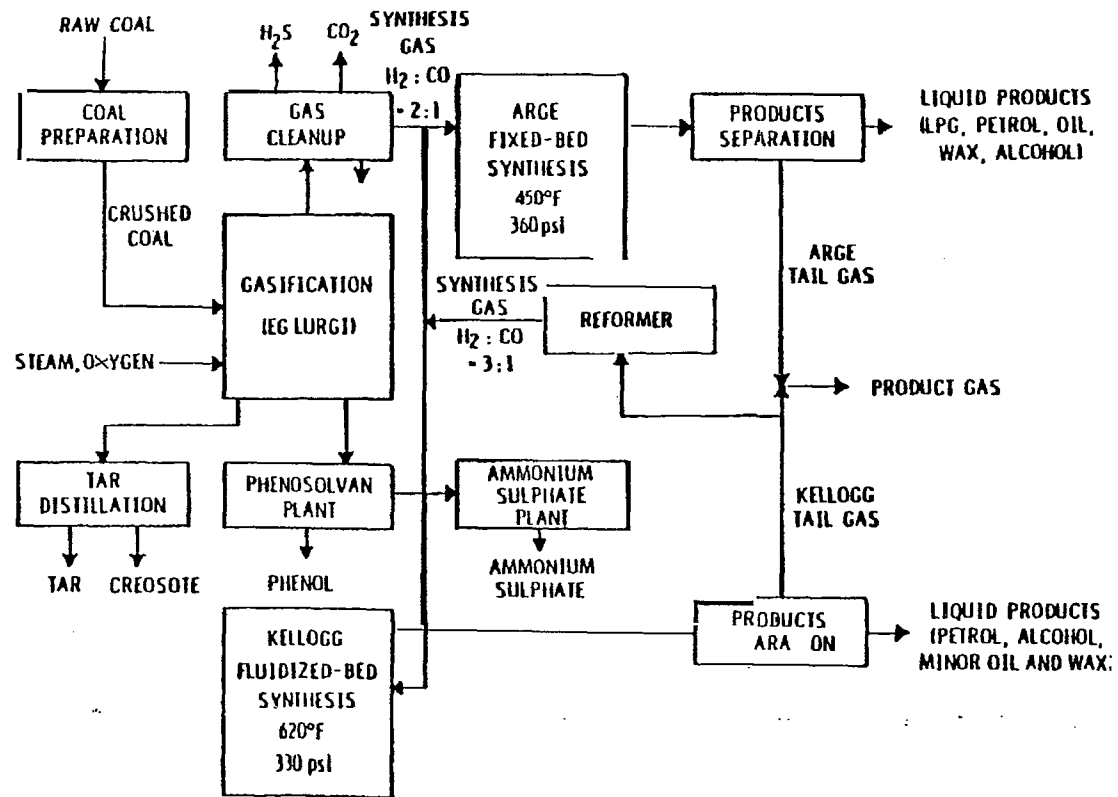
Overviews on SASOL I and SASOL II, based on reference 8, follow:

SASOL 1

LOCATION: Sasolburg, South Africa
DESCRIPTION: Gasification in Lurgi gasifiers
Two Fischer-Tropsch synthesis units;
1) ARGE fixed-bed unit, temp. 230°C;
press. 23 atm.; catalyst, pelleted
precipitated iron.
2) Kellogg SYNTHOL process, high-
velocity entrained-flow reaction
using a doubly promoted iron
catalyst.
SIZE: 10,000 bpd
STATUS: in commercial production since 1956
YEARS OPERATION: 24
COAL TYPE: Subbituminous
MAJOR PRODUCTS: Liquid fuels, chemicals, and fuel gas.

SASOL II

LOCATION: Secunda, South Africa
DESCRIPTION: Gasification in Lurgi gasifiers,
Fischer-Tropsch synthesis unit using the
Kellogg SYNTHOL process
SIZE: Nominal 40,000 bpd
STATUS: Anticipate ready for commissioning in 1980
COAL TYPE: Subbituminous
MAJOR PRODUCTS: Liquid fuels (gasoline is the major product).
SOURCE: Reference 35



SOURCE: Reference 35

Figure 3.7

Fischer-Tropsch Synthesis

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at 620% and 330 psig, (Reference No. 35). The main products are gasoline, fuel oil fractions, and various chemical products. The gasoline has a lower octane rating than the one derived from petroleum crude. The products produced do not fit well into existing markets. However, Mobil Oil Corporation has developed catalysts that improve the quantity and quality of gasoline (Reference No. 29).

3.3.6 Methanol Process

The production of methanol from synthesis gas is a specialized application of the Fischer-Tropsch reaction. Whereas the F-T process produces liquid fuels and chemical products, the Mobil methanol process produces gasolines. The schematic outline of this process is given in Figures 3.8 and 3.9 (Reference No. 35). Table 3.8 (Reference No. 35) presents a comparison of the thermal efficiencies of the Fischer-Tropsch and the Mobil methanol-to-gasoline process.

In the Mobil methanol liquefaction process, synthesis gas is produced from coal by any of the medium-Btu coal-gasification processes. The synthesis gas is converted to methanol by a number of catalytic processes. The reaction is exothermic. The yield of methanol is optimized by using high pressures and low temperatures, optimum type and shape of catalysts, and of recycling of the unreacted gases.

The conversion of methanol to gasoline is a separate catalytic conversion process. The Mobil conversion process dehydrates methanol, then rearranges the carbon and hydrogen atoms. The zeolite catalysts employed in the process (called ZSM-5 class catalysts) have a unique ~~channel~~ ^{channel}. The pore openings are of the right size to limit the size of the product molecules that can pass through them. The conversion proceeds to conventional high quality gasoline (Reference No. 25).

Table 3.9 (Reference No. 25) summarizes the overall material and energy balances of the methanol-to-gasoline conversion process.

Table 3.9 (Reference No. 25) shows typical product yields produced from methanol by this conversion process.

⁵ Even though no commercial demonstration plants of the "indirect" coal-methanol-gasoline process has been built as of this date, this route is considered by many authorities to be a very promising way to get gasoline from coal. There are several proposed studies and plants under instruction in the U.S. using this process (see Appendix chart). Also, New Zealand Liquid Fuels Trust Board (Report No. LF 5502, 10/31/79) has a large Mobil-M gasoline plant under construction (expected to become operational by 1983-5).

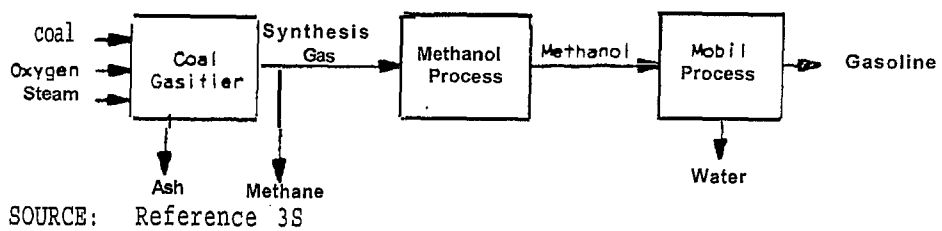
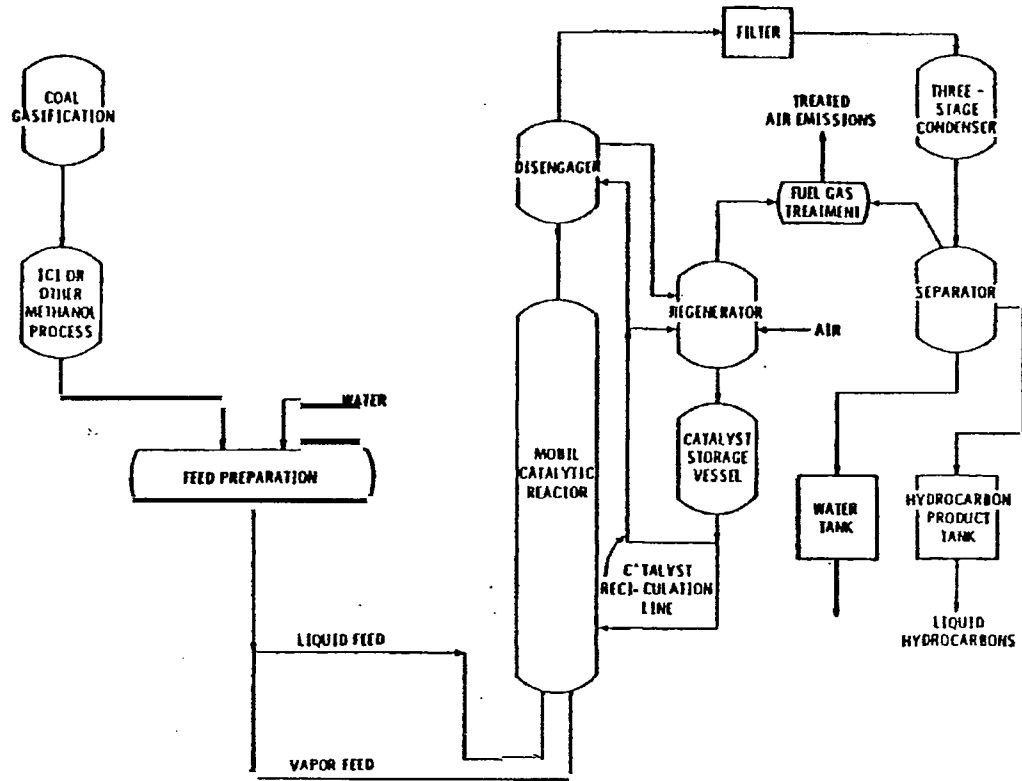


Figure 3.8

Synthesized Gasoline From Coal



SOURCE: Reference 35

Figure 3.9

Mobil Catalytic Process

Table 3.8
Thermal Efficiencies

	<u>Methanol-to-Gasoline</u> ⁷		<u>Fischer-Tropsch</u> ⁷	
	Btu/hour (1C ⁶ Btu)	Percent of Input	Btu/hour (106 Btu)	Percent of Input
<u>Input</u>				
coal	19,383		19,708	
Coal Fines (excess)	(872)		—	
Methanol			3	
Total Input	18,511		19,711	
<u>output</u>				
SNG	6,067	32.8	7,243	36.8
C3 LPG	247	1.3	176	0.9
C ₄ LPG	385	2.1	26	0.1
10 RVP Gasoline	4,689	25.3	2,842	14.4
Diesel Fuel			514	2.6
Heavy Fuel Oil			147	0.7
subtotal	11,388	61.5	10,948	55.5
Alcohols			290	1.5
sulfur	19	0.1	19	0.1
Ammonia	83	0.5	83	0.4
Power	18	0.1	11	0.1
Total Output	11,508	62.2	11,351	57.6

⁶ Thermal efficiencies are highly dependent on product mix.

⁷ The indirect liquefaction processes shown here may be considered as gasification processes for SNG, with the major coproduct being gasoline, e.g., for the "Fischer-Tropsch process" shown, the yield of SNG is 1.45 BOE/ton of coal, with a gasoline yield of 0.58 BOE/ton of coal. It is thus not representative of the SASOL-II process which emphasizes the production of liquid fuels.

⁸ Direct thermal equivalent value (thermal efficiencies are highly dependent on product mix (see Section 7.5)).

SOURCE : Reference 35

TABLE 3.9

METHANOL-TO-GASOLINE BALANCES

	<u>Methanol</u> →	<u>Hydrocarbons</u> +	<u>Water</u>
Material Balance	100 tons	44 tons	45 tons
Energy Balance:	100 Btu	95 Btu	0 Btu

YIELDS FROM METHANOL

Average Bed Temperature, °F	775°F
Pressure, psig	25
Space Velocity (WHSV)	1.0
Yields, wt % of charge	
Methanol + Ether	0.2
Hydrocarbons	43.5
Water	56.0
co, CO ₂	0.1
Coke, Other	0.2
	<u>100.0</u>
Hydrocarbon products, wt %	
Light gas	5.6
Propane	5.9
Propylene	5.0
i-Butane	14.5
n-Butane	1.7
Butenes	7.3
C ₅ + Gasoline	60.0
	<u>100.0</u>
Gasoline (including alkylates), wt, % (96 RON, 9 RVP)	88.0
LP Gas, wt %	6.4
Fuel Gas, wt %	5.6
	<u>100.0</u>

SOURCE: Reference 25

3.4 Oil Shale Retorting

3.4.1. General

Oil shale resources vary widely in their oil yields. High grade shale is normally defined as a deposit that averages 30 or more gallons of oil per ton of shale. Low grade shale averages 10 to 30 gallons per ton⁹ (Reference No. 7) . Several factors determine whether or not an oil shale deposit is recoverable. These include oil yield (usually equal or above 20 gallons per ton) , zone thickness, overburden thickness, the presence of other materials in the shale, availability of needed resources such as water and services, and location relative to markets.

There are two major routes for converting oil shale to liquid or gaseous fuels. They are:

1. Conventional mining followed by surface retorting (heating) , and
2. In situ (in place) retorting

In addition, there is modified in situ. In this process, the permeability (i.e., void volume) of oil shale deposits is increased in order to enhance the in situ retorting by removing some of the shale. The methods of rein@ or increasing the permeability of the oil shale deposits are explained in reference 8.

3.4.2. Surface Retorting

In surface retorting of oil shale, the heating takes place above ground. The shale is crushed to the right size, and fed into a retorting vessel. Heating the shale to between 800°F and 1000°F removes about 75 percent of the kerogen from the shale (Reference No. 8) . Different retorting processes apply heat to the shale in different ways. Gas or non combustible solids such as sand or ceramic balls can be used as heat carriers. The vapor produced during the heat@ is condensed to form crude shale oil. It can be further upgraded and refined to produce more marketable products.

As a generic surface retorting process, TOSCO II is described. Its schematic diagram is given in Figure 3.10 (Reference No. 8).

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Shale deposits yielding less than 10 gallons of oil per ton are normally omitted from USGS resource estimates.

Raw oil shale is crushed to 1/2 inch and preheated to 500° F. It is mixed with hot ceramic balls 3/4 inch in diameter and at 1200°F in a retorting Pyrolysis drum (Reference No. 25) . About two tons of balls mix with every ton of shale. The oil shale is heated to 900°F, releasing hydrocarbon vapors from the kerogen. The spent shale and the balls pass to the sealed accumulator vessel, in which the balls are separated from the shale by a heavy duty rotating cylinder with numerous holes. The balls are lifted by a bucket elevator to the gas fired ball heater, which heats the balls to 1270°F by direct contact heat exchanger. The spent shale goes through

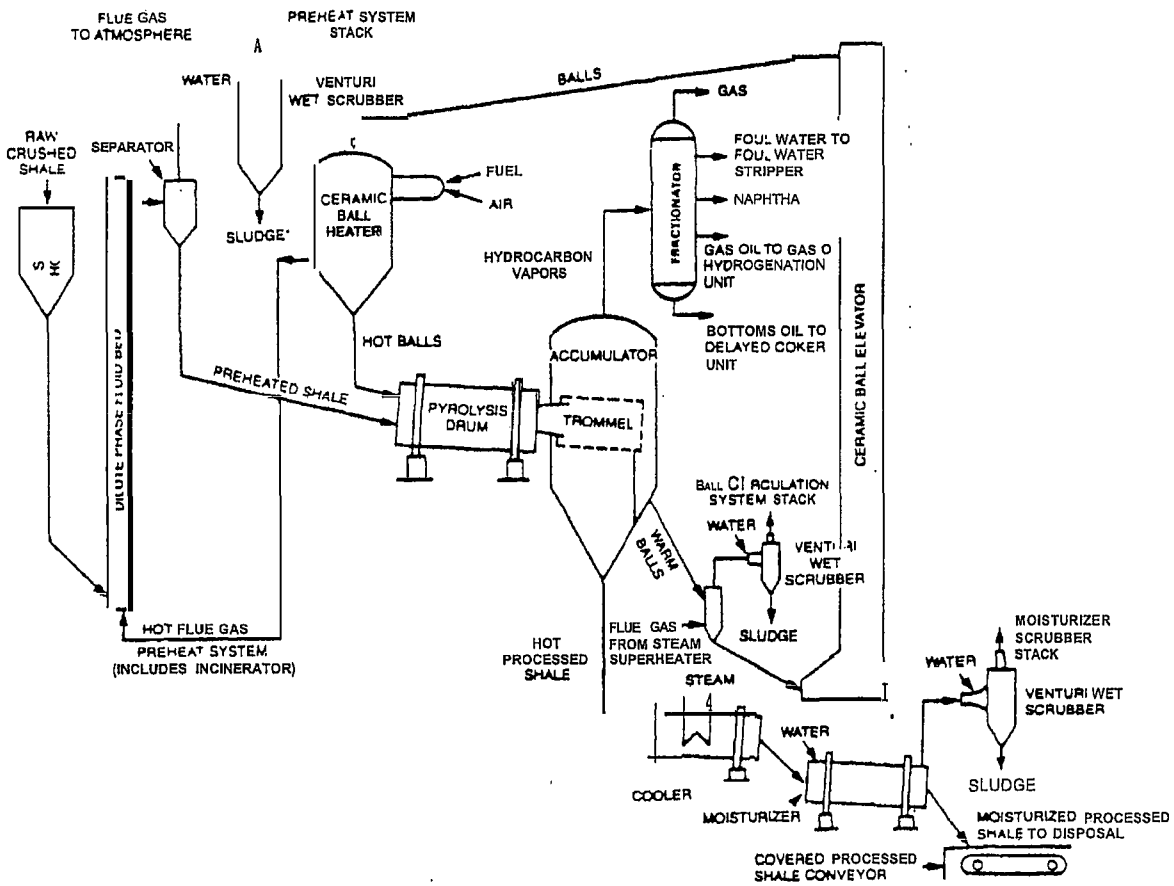
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FIGURE 3.10

The TOSCO II Oil Shale Retorting System



SOURCE Oil Shale Retorting Technology prepared for OTA by Cameron Engineers, Inc. 1978

a special heat exchanger which cools the shale for disposal and produces steam for plant use. Then the spent shale is quenched with water and moisturized to 14 percent, a level proper for disposal.

Hot flue gas from the ball heater is used to lift raw shale to a point at which it can subsequently flow by gravity into the pyrolysis drum. The flue gas also heats the raw shale to approximately 500°F.

Table 3.10 (Reference No. 25) summarizes the basic material balance for a TOSCO II retort module.

TABLE 3.10

BASIC MATERIAL BALANCE FOR
A TOSCO II RETORT MODULE

Oil Shale

Feed rate, TPSD	10,700
Fischer Assay, GPT	20

Pipelineable Shale Oil Product

production rate, BPSD	4,500
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Properties

Gravity, *API	28.6
Viscosity (SSU @ 30°F)	800
Pour Point, °F	30

Table 3.11 (Reference No. 35) summarizes the energy balance for a plant producing 47,000 barrels per day. Table 3.12 (Reference No. 17) summarizes the components, resource requirements and potential impacts of surface oil shale retorting.

Table 3.11

Estimated Energy Balance For a TOSCO II Plant
 producing 47,000 BPSD* Upgraded Shale Oil
 From 35 Gallons Per Ton Oil Shale

	Btu/hour (10 Btu's)	Percent of Total Energy Input
<u>Product Output</u>		
Product oil	10.30	58.00
LPG	0.70	3.94
Diesel fuel	0.11	0.62
<u>System Losses</u>		
Spent shale and moisture	1.78	10.02
Residual carbon (coke)	0.93	5.24
Ammonia	0.11	0.62
Sulfur	0.06	0.34
Cooling water	1.07	6.02
Water evaporat on on shale	0.25	1.41
Losses (includ ing flue gas heat)	2.45	13.79
<u>Energy Input</u>		
Raw shale	17.76	100.0
Steam	17.00	95.72
Electrical energy	0.53	2.98
	0.23	1.30

* BPSD = barrels per stream day

SOURCE: Reference 35

PROCESS SYSTEM:

- 100,000 gal/day of crude shale oil
- 20 x 10³ gal/day
- 20 x 10³ gal/day
- 5.6 x 10³ gal/barrel
- operates 328.5 days/year
- total annual output 93.2 x 10³ Btu
- plant life 30 years
- plant efficiency (thermal) 6%

DESCRIPTION:

• heated ceramic balls are fed into a horizontal cylindrical reactor and react with hydrogen gas (17% H₂, 83% N₂).
 • Diatomic hydrogen gas (H₂) is fed into the reactor and reacts with the ceramic balls.
 • The end of the reactor and are collected and fed into a fractionator for product recovery. The ceramic balls are recycled in a vertical ball heater where reheating for further use occurs.

COMMENTS:

- horizontal cylindrical reactor
- fractionator and heater
- nepheline hydroflour
- gas oil hydroflour
- hydrogen plant
- by-product recovery

ENVIRONMENTAL CONCERNS:

- air quality deterioration
- effects due to hydrocarbons
- modification of biological environment
- deterioration of water quality due to hydrocarbons and runoff
- solid waste disposal
- socio-economic problems due to high influx of personnel in previously sparsely populated areas

(1) Land use value depression had committed to use for the facility, divided by annual production, measured in trillion Btu.
 (2) Costs are total costs for plant construction, divided by annual output, measured in trillion Btu.

SOURCES: (a) Environmental Protection Agency, Monitoring Environmental Impacts of the Coal and Oil Shale Industries, 600/7-1-015, February 1972.
 (b) University of Denver, Denver Research Institute, An Environmental Analysis Report on the TONGUE HILL Shale Process, March 1972.
 (c) University of Oklahoma, Energy Alternatives: A Comparative Analysis, 1971.
 (d) Bechtel Corporation, Bechtel Supply Planning Model - 1971.
 (e) Cameron Engineering Incorporated, FATHSHEE Plant Handbook, 1973.

SOURCE: Reference 17

RESOURCES USED:
(for 10³ Btu Produced)

- raw shale
- oil content
- OPERATION*
oil
- water
- agent shale
- gas

LAND (1)
 rearing, upgrading and
 effluent facilities
 operations, water conser-
 vation and ground
 water treatment
 water treatment plant
 dust control on shale ash
 molasses evaporator
 dust scrubbers
 re-vegetation
 fire and drinking
 total

MANPOWER
 construction
 materials
 equipment
 fuel cost
 total
 operation & maintenance
 construction
 operation & maintenance

COSTS (2)

1,840,000
 811,000
 1,833,000
 4,324,000
 84
 2.1
 11.5

RESIDUAL AND PRODUCTS:
(for 10³ Produced)

- AIR POLLUTANTS (b):
particulates
SO₂
hydrocarbons
CO
- WATER POLLUTANTS
charge to city water-treat-
ment plant
- LAND VALUE (b)
agent shale
- ENERGY PRODUCT (b)
refined shale oil
blast content

210,000 tons
 35 gallons/ton
 (by weight)
 11.8
 1.4
 81.8
 2.6
 1.42
 14.5
 16.5
 31.3
 3.9
 7.9
 10.4
 1.1
 3.4
 37.7

196,000
 172,000 barrels
 2.6 x 10³ Btu/barrel

3.4.3 Modified In Situ Retorting

Occidental modified in situ oil shale retorting process is selected as representative. It involves the mining out of about 10 to 25 percent of the shale deposit. This mined portion would presumably be retorted by one of the surface retorting processes, or if its oil content is too low, will be treated as waste (Reference No. 37).

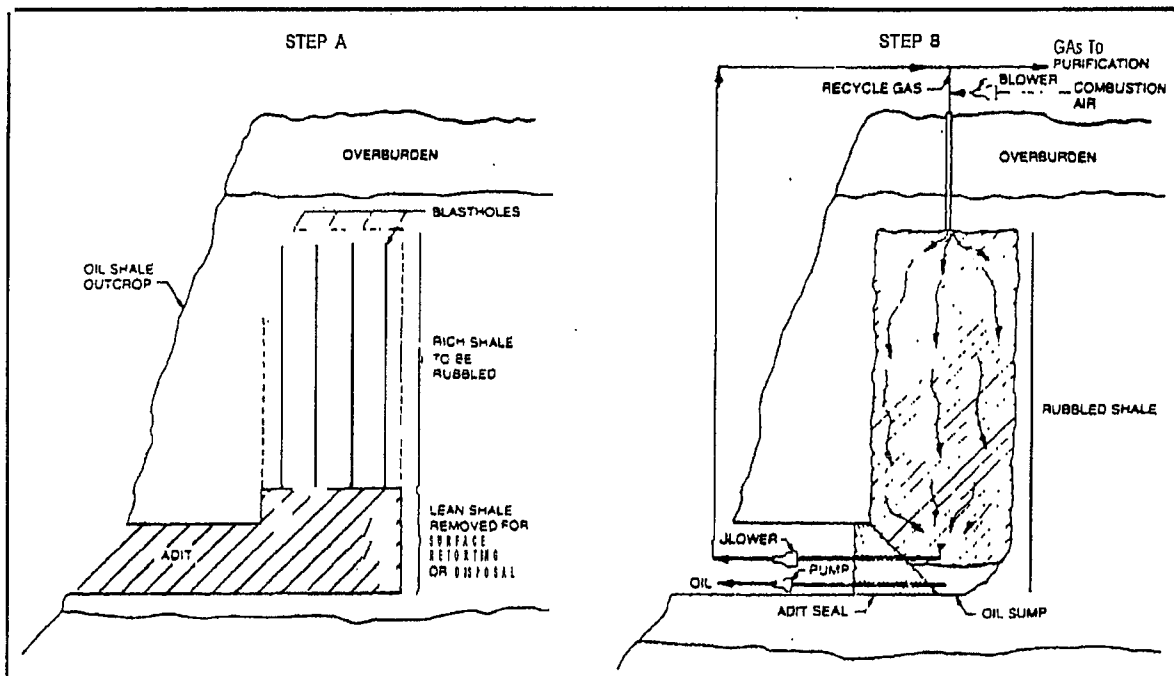
Figure 3.11 (Reference No. 8) represents in schematic form a generic modified in situ oil shale retorting process. Figure 3.12 (Reference No. 37) is a more detailed description of the Occidental modified in situ retorting process. As observed in Figure 3.12, in steps A or the pre-detonation phase, drifts (chambers) are excavated at the top and bottom of the shale deposit, which is about 300 feet-thick. An interconnecting shaft is dug to connect the drifts. Rooms with a volume of about 15 to 20 percent of the eventual volume of the planned chamber are then mined. Shot holes are drilled to allow blasting of the shale oil to produce the desired fragmentation.

In the burn phase, the explosives in the shot holes are detonated. A rubble-filled chamber is created which can function as a batch retort. The percentage of void space and the particle size distribution of the rubble are a function of the explosive loading. Connections are made to air/gas recycle and air supply compressors. An outside heat source (e.g., off gas or oil from other retorts) is used for heating the rubble at the top of the retort. Oil shale and hydrocarbon gases are produced which move downward. Residual carbon is left on the spent shale.

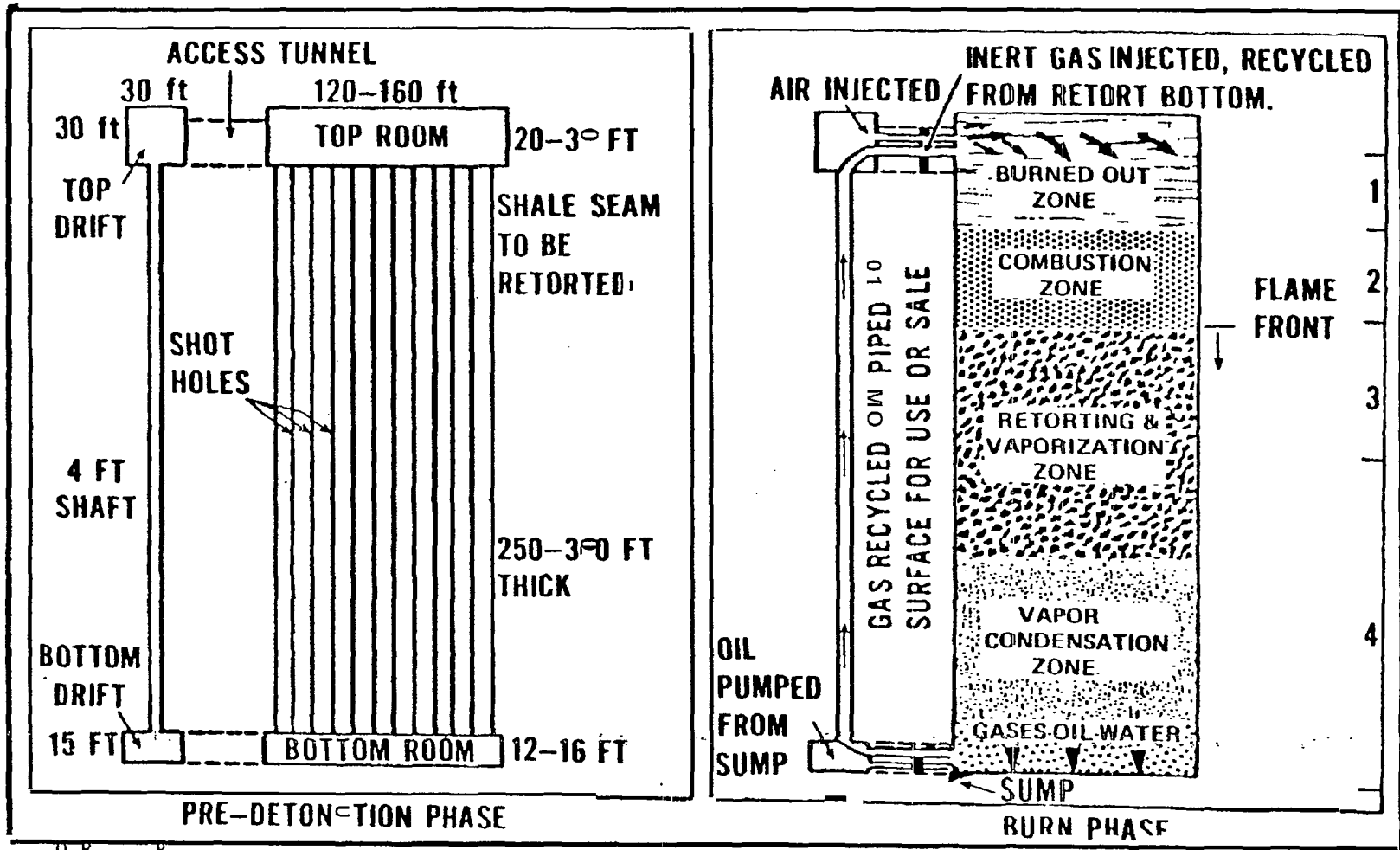
The retorting reaction is terminated after a predetermined amount of the rubble has been retorted by halting the external heating supply. The residual carbon is utilized to continue the combustion process, which now does not need external heating. The flame front moves downwards, preceded by the liquid and gaseous products retorted from the shale by the hot, oxygen-deficient combustion gases. The liquid hydrocarbons collect in a sump, from which they are pumped to the surface. The gaseous by-products are used partially, with steam, as a recycle stream to control the oxygen content of the inlet gas. The four distinct zones that develop during the retorting are shown in Figure 3.11.

Table 3.13 (Reference No. 17) summarizes the components, resource requirements, and potential impacts of modified in situ retorting.

Figure 3.11: Modified in Situ Retorting.



SOURCE: T. A. Sader, "Recent Trends in Oil Shale—Part I: Mining and Shale Oil Extraction Processes," *Mineral Industries Bulletin*, vol. 18, No. 1, January 1975, p. 18.



U.S. PATENT OFFICE, 1964
 Figure 3.12: Occidental Modified In Situ Retorting

3-38

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Modified In-Situ Shale Retorting (Occidental)

PART 3 13

ENERGY SYSTEMS

- 1111 = 34,700 tons of raw shale mined/day (d)
- 102,000 tons of raw shale retorted in-situ/day (c)
- 25 million gallons of steam produced per day
- 1012 = 1012 kw/day
- 1013 = 1013 Btu/year
- operate 378.5 days/year
- plant life 30 years
- thermal efficiency 88% overall (d)

DESCRIPTION

• In modified in-situ approximately 15-20% of the deposit is mined using conventional mining techniques in the ground and first used for either chemical, hydraulic, or electric energy. Prior to fracturing of the deposit, parallel wells (productions and injection) are drilled on two opposite sides of the deposit. A retorting fluid (hot steam or gas) is injected within the formation. After ignition, retorting takes place and the oil mist, gas and steam produced are forced to the surface through the production wells. The gas is burned at the base of the combustion chamber. The retorting products are refined using techniques similar to surface refining.

COMMENTS

- underground retort created from blasting procedure
- fractionator and reher
- asphaltic hydrofiner
- gas oil hydroliner
- hydrogen plant
- by-product

ENVIRONMENTAL CONSEQUENCES

- air quality deterioration
- health effects due to hydrocarbons
- deterioration of biological environment
- deterioration of water quality due to leachate and runoff
- solid waste disposal
- socio-economic problems due to high influx of personnel in previously sparsely populated areas

*Approximately one barrel of water/barrel of oil is produced during retorting by the release of interstitial water and the combustion of hydrocarbons.

- SOURCES: (a) General Engineers, Incorporated, Synthetic Fuels Handbook, 1975.
 (b) BESS Base Group, Environmental Characteristics for Energy Technologies and End Uses, Revision 8, 1978.
 (c) Department of Energy, Draft Environmental Impact Statement for the (Oxydized) Kentucky Oil Shale Leasing Program, 1978.
 (d) Environmental Protection Agency, A Preliminary Assessment of the Environmental Impacts from Oil Shale Development, 600/7-77-040, July 1977.
 (e) Amfield Oil, Inc., Lease & Occidental Oil Shale, Inc., Modifications to Detailed Developmental Plan, Oil Shale Tract C-4, February, 1977.

SOURCE: Reference 17

RESIDUALS AND PRODUCTS:
(Per 1012 Btu Produced)

AIR POLLUTANTS (e.o.)	Tons
particulates	19.8
SO ₂	4.21
hydrocarbons	31.7
CO	0.31
	10.6

WATER POLLUTANTS
 These are assumed to be direct discharge of effluent into any water course.

SOLID WASTE (d)	Tons
spent shale	99,000
	137,400

ENERGY PRODUCTS
 Refined shale oil

RESOURCES USED:
(Per 1012 Btu Produced)

FUEL (c,d)	1012 Btu Produced
mined shale	117,000 tons
mined shale	125,000 tons
oil content	25 gallons/inch
COMPOSITION (e)	(by weight)
organic material	MA
inorganic material	MA
spent shale	MA

LAND (c)

Area	1012-310/year
Permanent disposal	345
surface facilities	882/year
active well: area	

WATER (c)

Acres-ft	(1012-310)
retorting and upgrading	34.4
power generation	0.9
vegetation	28.9
steam injection	1.4
miscellaneous	34.4
total	100.0

COSE (b)

Barrels (1012)	1012-310
construction	11,000,000
operation & maintenance	MA

PERSONNEL (b)

Months	MA
construction	30.3
operation & maintenance	MA

3.5 comparison of the Various Synfuel Systems With Respect to Resource Requirements

In order to estimate the resource requirements of the coal and oil shale fuel cycles we need first to assess their energy utilization efficiencies. These are summarized in Table 3.14.

The *resource* requirements of coal and oil shale energy systems per 10⁶ Btu of product delivered to end user are given in Tables 3.15 and 3.16. Tables 3.17 and 3.18 convert these requirements to energy systems producing 50,000 barrels of oil equivalent per day.

Manpower requirements for operating and maintenance labor of coal conversion plants are given in Reference 29.

They are:

Plant operators	
Operating supervisors	
Maintenance labor	
Maintenance labor supervisors	30
Administration	30
Total	355

These manpower requirements are for a basic (ESCOE) coal conversion plant that consumes 25,000 tons of coal per day with 22.4 million Btu/ton and produces 50,000 bbl/day liquids output.

Very considerable variations exist in the literature in respect to manpower requirements for the other phases of the fuel cycle. They depend on such variables as methods of mining, location of mine, kind of transportation system and extent of beneficiation. A table indicating the ranges of variables is given in the footnote in respect to the conversion plants.

¹⁰ Limitations of Data Sources: Evaluations carried out in this report are often subject to great uncertainties because:

- (1) The information available is only of preliminary nature. There are no full scale operating synfuel plants in the U.S. (subject to U.S. siting considerations), so that data needs to be extrapolated from pilot plants with many uncertainties of scale and dissimilarities associated with the extrapolation, as well as specific siting and feedstock characteristics discussed below.

10 (cont'd)

- (2) **There are variations among sources which are often due to different assumptions or local influences. Changes in design account for some differences as the technology changes and the environmental regulations change. Many of the assumptions are not stated - or even referenced. Budget and time limitations, however, necessitate the need to use existing data bases, rather than the development of new data.**

Even estimating the range of uncertainties is often a value judgement process, unless more extensive on-site interviewing with site and process specific sources of information are developed.

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Table 3. 4 Resource Utilization Efficiencies of Generic Synthetic Fuel Energy Systems

(In Percent)

	1		2		3		4		5		6	
	<u>Coal Gasification</u>		<u>Coal Liquefaction</u>		<u>Coal Liquefaction</u>		<u>Coal Liquefaction</u>		<u>Oil Shale Retorting</u>		<u>Oil Shale Retorting</u>	
	<u>Medium-Btu</u>	<u>High-Btu</u>	<u>Direct</u>	<u>Indirect</u>	<u>Direct</u>	<u>Indirect</u>	<u>Surface</u>	<u>Modified in Situ</u>	<u>Surface</u>	<u>Modified in Situ</u>	<u>Surface</u>	<u>Modified in Situ</u>
Beneficiation ^a	96.4-97.3	96.4-97.3	96.4-97.3	96.4-97.3	96.4-97.3	96.4-97.3	96.4-97.3	96.4-97.3	96.4-97.3	96.4-97.3	96.4-97.3	100
Transportation to Conversion Plant ^b	98.5	98.5	98.5	98.5	98.5	98.5	98.5	98.5	99.5	99.5	99.5	100
Conversion to Fuel ^c	83	59	64-70	48-57	64-70	48-57	67	61	67	61	67	61
Upgrading and Refining ^d	N.A. ^e	N.A. ^e	75-95 ^f	95-100 ^f	75-95 ^f	95-100 ^f	77 ^g	77 ^g	77 ^g	77 ^g	77 ^g	77 ^g
Distribution to End User	96.9 ^h	97.1 ⁱ	98.8 ^j	98.8 ^j	98.8 ^j	98.8 ^j	98.8 ^j	98.8 ^j	98.8 ^j	98.8 ^j	98.8 ^j	98.8 ^j
Overall Energy Efficiencies	76.4-79.2	54.4-54.9	45.0-63.0	42.8-54.0	45.0-63.0	42.8-54.0	48.9-49.3	46.4	48.9-49.3	46.4	48.9-49.3	46.4

SOURCE: E. J. Bentz & Associates

3-41

Notes for Table 3.14

- a. Estimates of losses of coal and oil shale from beneficiation (in terms of Btu's) vary broadly among authors, depending on the assumed degree of upgrading and the kind of coal or oil shale used. Estimates vary from 0% (Reference 37a) ; 2.7-3.6% (Reference 7) ; and 12.5% for intensive beneficiation (Reference No. 17) .
- b. Average value of losses are 1.5% (time from Reference No. 7) . In the case of oil shale, where distances are shorter, 0.5% is assumed.
- c. The net efficiencies (rather than the process efficiencies) were used. The efficiencies for coal conversion processes are derived from Roger and Hill. (Reference 29) . In the case of H-Coal, the syncrude efficiency was used. In the case of oil shale retorting processes, the efficiencies are derived from DOE (Reference No. 17) .
- d. Data on efficiencies of upgrading and refining syncrudes is very limited and unreliable (see Section 1.7) .
- e. N.A. means not applicable.
- f. Overall yields for SRC II of finished fuels range between 83 and 98 liquid volume percent of SRC II syncrude, depending on the product slate and how refinery fuel and hydrogen plant feed are supplied. An average of the net product yields ranging between 88 and 91 was assumed (Reference No. 22) . However, these values apparently do not include coal use for the production of hydrogen needs for the upgrading process. If coal-derived hydrogen is to be used (as against hydrogen from nuclear fission or from biosynthesis) , then the upgrading and refining efficiencies for coal conversion products become 75 percent. However, in some cases it may be expected that all of the hydrogen and energy required for the Upgrading/refining process would be obtained from residuals, higher boiler fractions, and methane produced in the process or plant refinery (which may include the use of Petroleum derived vacuum . In the case of indirect liquefaction processes, all the needed hydrogen is accounted for in the gasifier, and higher upgrading efficiencies can be achieved, depending on product slate .
- g. Derived from Reference 26a. However, MIS oil is easier to upgrade, so that higher efficiency may be in order.
- h. Derived from Reference 17.
- i. Derived from Reference 7.
- j. Derived from Reference 7 and 10.

Table 3.15 Fossil Carbon Consumption of Generic Synthetic Fuel Energy Systems
(In 10^{-3} ton of fossil carbon/ 10^6 Btu fuel delivered to end user)

	1	2	3	4	5	6
	<u>Coal</u>	<u>Gasification</u>	<u>Coal Liquefaction</u>		<u>Oil Shale Retorting</u>	
	<u>Medium-Btu</u>	<u>High-Btu</u>	<u>Direct</u>	<u>Indirect</u>	<u>Surface</u>	<u>Modified in Situ</u>
Beneficiation	1.2-1.6	1.7-2.2	2.0-2.7	2.1-2.8	0.9-1.2	0
Transportation to Conversion Plant	0.7	0.9	1.1	1.2	0.2	0
Conversion to Fuel	0.7	25.1	22.2-26.7	33.5-40.5	11.4	18.7
Upgrading and Refining	0	-	3.7-18.5	0-3.9	7.9	11.0
Distribution to End User	1.4	1.8	0.9	0.9	0.4	0.6
Overall Consumption	9.1-10.3	27.6-27.9	27.4-40.7	35.8-44.8	17.5-17.6	25.7

SOURCE: E. J. Bentz & Associates

3-43

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Notes to Table 3.15

- a This table summarizes the consumption of fossil carbon contained in the feedstocks or products during the various phases of the various synfuel cycles.
- b The numbers in the table are based on the following assumptions:
- (i) The resource utilization efficiencies are those developed in Table 3.14.
 - (ii) The carbon content of bituminous coal averages 87.8%, lignites - 72.5% and sub-bituminous - 73.5%. The carbon content of the kerogen (i. e., crude shale oil) averages 80.5%. (Ref. 26b). For convenience, an average figure of 80% for the carbon content of coals and kerogen is used.
 - (iii) The loss in fossil carbon is directly proportional to the loss in coal or kerogen.
 - (iv) The Btu content of a ton of coal is 24×10^6 Btu and of ton crude shale oil is 36×10^6 Btu.
- c A sample calculation for medium Btu coal gasification is as follows:
- A ton of feedstock bituminous coal has 24×10^6 Btu, of which 18.34×10^6 to 19.01×10^6 Btu is delivered to the end users (74.4 to 79.2% overall energy efficiency - see Table 3.14). Since a ton of feedstock coal has 80% fossil carbon content, and 20.8% to 23.6% of it is consumed during the medium Btu coal gasification fuel cycle, (see Table 3.14), the total fossil carbon consumption of the cycle is between 0.1664-0.1888 tons per 18.34×10^6 to 19.01×10^6 Btu delivered to end users. This translated to 0.009 to 0.010 tons of fossil carbon per 10^6 Btu.

Table 3.16 Water Consumption of Synthetic Fuel Energy Systems (Generic)
(In gallons per 10⁶ Btu product delivered to end user)

	1		2		3		4		5		6	
	<u>Coal Gasification</u>		<u>Coal Liquefaction</u>		<u>Coal Liquefaction</u>		<u>Coal Liquefaction</u>		<u>Oil Shale Retorting</u>		<u>Oil Shale Retorting</u>	
	<u>Medium-Btu</u>	<u>High-Btu</u>	<u>Direct</u>	<u>Indirect</u>	<u>Direct</u>	<u>Indirect</u>	<u>Direct</u>	<u>Indirect</u>	<u>Surface</u>	<u>Modified in Situ</u>	<u>Surface</u>	<u>Modified in Situ</u>
Mining ^{a,b}	0.6-0.9	0.6-0.9	0.6-0.9	0.6-0.9	0.6-0.9	0.6-0.9	0.6-0.9	0.6-0.9	0.7-1.1	0.7-1.1	0.7-1.1	0.7-1.1
Beneficiation ^c	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	0	0	0	0
Transportation to Conversion Plant	0	0	0	0	0	0	0	0	0	0	0	0
Conversion ^o Fuel ^d	13-24	13-24	7-26	13-26	7-26	13-26	7-26	13-26	9-32	9-32	9-13	9-13
Upgrading and Refining ^e	0	0	-	-	-	-	-	-	24	24	24	24
Distribution to End User	0	0	0	0	0	0	0	0	0	0	0	0

SOURCE: E. J. Bentz & Associates

3-45

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Notes to Table 3.16

- a The water required for mining and preparation of the coal or shale and for the disposal of ash or spent shale is a function of location, mainly through the amount of material that must be mined or disposed; and the degree of attested surface reclamation. Assuming 2/3 of coal is surface-mined and 1/3 is undergroundd mined, water consumption for surface mining ranges between 0.55 and 0.98 gallons per 10⁶ Btu of product, and for underground mining - 0.75 gallons per 10⁶ Btu of Product (Reference No. 17) .
- b Assume 2/3 of oil shale is surface mined and 1/3 is underground mined. Water consumption or both kinds of operations range between 0.7 and 1.1 gallons per 10⁶ Btu of product (Reference No. 17) .
- c Consumption of 1.2 gallons of water 10⁶ Btu Of product is assumed for beneficiation of coal (Reference No. 17) and none for shale oil.
- d Consumption of water for the conversion of feedstock to fuels depends principally on the overall plant conversion efficiency, degree of water recycling, and the water content of the coal or shale. Consumption figures range from 13-24 gallons per 10⁶ Btu of product for coal gasification; 7-26 for direct coal liquefaction; 13-26 for indirect coal liquefaction; 9-32 for surface shale retorting; and 9-13 for modified in situ shale retorting. (Derived from References 17, 37b,c) .
- e Water consumption for upgrading and refining is not available in the literature. The estimates presented for shale oil upgrading are based on private conversation with Mr. Bobby Hall and Ray Young of the American Petroleum Institute 3/81. For shale oil - 100 gallons per barrel are needed to make the raw shale oil suitable for pumping, and 40 more gallons per barrel to convert it to transportation fuels. Polling of a large number of oil companies and API experts did not result in water consumption estimates for upgrading of coal liquids (namely: Robert Howell, Bonner and Moore, Fred Wilson Texaco, Patton, Nanny, Hall and Young of API - 3/81) .

Table 3.17* Annual Feedstock Requirements for Generic Synthetic Fuel Energy Systems Producing 50,000 bbl Oil Equivalent per Day to End User
(In millions of tons or barrels of oil)

	<u>Coal Gasification</u>		<u>Coal Liquefaction</u>		<u>Oil Shale Retorting</u>	
	<u>Medium-Btu</u>	<u>High-Btu</u>	<u>Direct</u>	<u>Indirect</u>	<u>Surface</u>	<u>Modified in Situ</u>
Mining ⁵	5.6-5.8	8.0-8.1	7.0-9.8	8.2-10.3	62.2-62.7	N.A. ⁷
Beneficiation ⁵	5.4-5.6	7.8	6.8-9.4	7.9-9.9	60.5-60.6	N.A. ⁷
Transportation to Conversion Plant ⁵	5.3-5.5	7.7	6.7-9.3	7.8-9.8	60.2-60.3	N.A. ⁷
Conversion to Fuel ⁶	18.9	18.8	19.4-24.6	18.5-19.4	24.0	24.0
Upgrading and Refining ⁶	18.8	18.8	18.5	18.5	18.5	18.5
Distribution to End User ⁶	18.3	18.3	18.3	18.3	18.3	18.3

* These are the quantities of coal, shale or equivalent oil leaving the indicated phase of the fuel cycle.

SOURCE: E. J. Bentz & Associates

3-47

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Notes to Table 3.17

1. Same assumptions and references as those in Table 3.14.
2. Oil has energy content of 5.8×10^6 Btu/barrel.
3. Coal has energy content of 24×10^6 Btu/ton.
4. Oil shale has energy content of 3.45×10^6 Btu/ton (based on 25 gallons of oil per ton) .
5. Tons of coal or shale.
6. Barrels of oil equivalent.
7. N.A. is not applicable.

Table 3.18* Annual Water Consumption of Generic Synthetic Fuel Energy Systems Producing 50,000 bbl Oil Equivalent per Day to End User (In million gallons per year)

	<u>Coal Gasification</u>		<u>Coal Liquefaction</u>		<u>Oil Shale Retorting</u>	
	<u>Medium-Btu</u>	<u>High-Btu</u>	<u>Direct</u>	<u>Indirect</u>	<u>Surface</u>	<u>Modified in Situ</u>
Mining	64-95	64-95	64-95	64-95	74-120	74-120
Beneficiation	130	130	130 "	130	0	0
Transportation to Conversion Plant	0	0	0	0	0	0
Conversion to Fuel	1400-2500	1400-2500	740-2800	1400-2800	950-3400	950-1400
Upgrading and refining	0	0			2500	2500
Distribution to End User	0	0	0	0	0	0

* Same assumptions and references as in Table 3.16.

SOURCE: E. J. Bentz & Associates

3-49

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Table 1 Footnote to Chapter 3: Manpower Requirements of Generic Synfuel Plants Producing 50,000 Barrels of Oil Equivalent per Day

	<u>1</u> <u>Coal Gasification</u>	<u>2</u> <u>Coal Liquefaction</u>	<u>3</u> <u>Oil Shale Retorting</u>	<u>4</u> <u>Oil Shale Retorting</u>
	<u>Medium-Btu-High-Btu</u>	<u>Direct & Indirect</u>	<u>Surface</u>	<u>Modified in Situ</u>
Peak Construction (men)	1,500-4,800 ^a	2,200-8,000 ^b	330 ^d	4,900 ^d
Construction (man-years)	3,400 - 10,800 ^a	7,500-25,000 ^b	1100 ^d	16,000 ^d
Operation and Maintenance (men)	320-500 ^a	355-3800 ^c	1200 ^d	-

^a DOE, 1980, Comparative Assessment of Health and Safety Impacts of Coal Use. DOE/EV 0069.

^b The lower value is derived from DOE/EV 0069; ^c the upper value - from Reference 34.

^c The lower value is derived from Reference 29 ; ^c the upper value - from Reference 34.

^d Derived from Reference 17 and assuming 5 year construction of plant peaking at 30% of $\omega \omega$ 1 man-years labor requirements (Reference 34).

SOURCE: E. J. Bentz & Associates

CHAPTER 4 : SYNTHETIC FUEL PROCESSES CONVERSION COST
AND PRODUCT ECONOMICS

<u>Section No.</u>		<u>Page</u>
4.1	Conversion Costs and Product Economics	4-1
4.2	Scale of Production	4-7
4.3	Product Quality	4-8
4.4	Estimating Methods.	4-11
4.5	Product Upgrading	4-22
4.6	Refining synthetic Liquids.	4-28
4-7	Transportation and Other Infra-structure Costs	4-36
4-8	Addendum to Chapter 4: Basis for Cost Assumptions	
	1. Basic Conversion Plant. . . * . . .	4-43
	2. Assumptions of Product Upgrading * . . .	4-44
	3. Refining Cost Assumptions .O . . .	4-46

The following evaluation of a wide range of alternate fuels produced from coal attempts to build upon prior work in the field that has, successively, estimated the plant construction and operating costs for each process, standardized the bases of estimation (time of construction, size of plant, location, financing methods, etc.) and evaluated the quality of product produced.

Such work has been sponsored by the Department of Energy since the early 1970's. The most recent work was performed by the Engineering Societies Commission on Energy, Inc. (ESCOE).¹ That work collected prior analyses performed for DOE and others, made adjustments in each to account for differing assumptions regarding input prices, plant scale, financing methods and costs, and thus reevaluated them on a more common basis. The differences in product quality were factored for value based on current price relationship among natural petroleum products.

Our approach will differ in several regards:

First of all we shall use the baseline ESCOE plant models, capital costs and operating cost relationships, updated to a uniform 1980 dollar basis.

Second we shall scale all plants to a common output plant size in order to retain comparability at other, downstream stages of processing and use.

Third we shall deal with differences in product quality directly, and on a cost of product basis, by considering the additional costs required to upgrade lower quality products and make them comparable with the higher grade synfuels.

Fourth we shall then examine the methods and costs of further processing and transporting the generic synfuel products to make them available to end use markets.³

The ESCOE capital estimates were all adjusted to a 1980 dollar basis by the use of the Wholesale Price Index - Industrial Commodities Index. Others have frequently used the Chemical Engineering Plant Index, however we feel that no significant historical difference exists and the WPI Index basis is a more suitable bench mark for further forecasting since it is a component

¹ Coal Conversion Comparison, ESCOE Report FE-2468-51, July, 1979.

² ESCOE scaled all plants to a common input size in order to simplify the costs - auxiliaries and off-sites are normalized.

³ We did not examine differences in end use efficiency that exist or are possible. This should be subsequently examined.

of Us. macro-economic forecasting models and the Chemical Construction Index is not.

Exhibit 4-1 displays the original capital cost estimates of ESCOE. Exhibit 4-2 updates these estimates to a uniform 1980 cost basis.

Operating costs are more complex. The major cost categories are:

- . Coal
- . Utilities
 - Water
 - Power
- . Catalysts and Chemicals
- . Labor
- . Overhead
- . Maintenance

Coal prices are uniform to all processes - as are assumed costs of water, power and labor. The costs of overhead are a uniform fraction of operating and maintenance labor - they include administrative personnel costs as well as G&A expenses. The maintenance rule is made uniform among systems-although differences should exist on the basis of system approach.

The original ESCOE operating cost variables are shown on Exhibit 4-3. These unit prices provide the bases for updating the ESCOE costs to the values shown on Exhibit 4-4.

The cost of producing hydrogen for product upgrading is partially imbedded in other estimates. The uniform condition is that hydrogen is demanded at a greater level than could be supplied from excess char, residue, or filtrate from the process plant. Therefore a hydrogen plant must be built at the upgrading plant site. This plant is designed to reform synthesis gas.⁵ The cost of hydrogen can then be based on the hydrogen plant's costs - including syngas feed at the estimated syngas product costs of our companion syngas plant. Alternately we could capitalize a coal gasification plant in this area, however that seems to be an even more unrealistic mode of system optimization.

In the long run, as product slate demand for synthetic coal liquids becomes clarified, the optimization of an integrated coal-to-product plant can be designed in a much more sophisticated manner.

⁴The input costs were in certain instances drawn from original sources cited by ESCOE.

⁵Or reform synthetic fuel product - the cost is comparable \$6.25 - 6.75/MM BTU.

EXHIBIT 4-1

PLANT CAPITAL REQUIREMENTS
MAJOR ON-SITE PLANT COST IN MILLIONS OF MID 1978 \$

<u>Category</u> <u>Process</u>	<u>SRC-II</u>	<u>EDS</u>	<u>H-FO</u>	<u>H-Syn</u>	<u>FT</u>	<u>M</u>	<u>Lur.</u>	<u>West.</u> <u>Syn.</u>
Coal Preparation	63	63	84	84	63	63	90	63
H ₂ or Gasification	253	190	138	158	228	228	143	22
O ₂ Plant	129	-	67	87	117	175	114	80
Gas Shift	H	-	30	35	-	40	30	H
Acid Gas and Sulfur Plants	60	60	57	57	57	57	136	57
Reactor Section	195	180	140	210	55	106	90	-
Conversion	-	-	-	-	100	75	20	42
Gas Plant	30	-	30	25	25	10	12	-
Flexicoker	-	160	-	-	-	-	-	-
Pollution Systems	44	44	40	40	40	40	55	24
Solvent Hydro. or Catalyst Prep.	-	82	-	-	3	-	-	-
Compression	-	H	H	H	H	-	28	-
Total less Int. Including Indirects	1262	1270	955	1134	1121	1212	1151	684

- Notes: 1. M includes HF Alkylation.
 2. Some EDS cost included in Flexicoker.
 3. All costs shown above are considered bare cost and have not been confirmed with process developer.

SOURCE: E. J. Bentz & Associates

4-3

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EXHIBIT 4-2

TOTAL CONVERSION PLANT INVESTMENT - 50,000 BBL LIQUIDS/DAY PLANT BASIS

(Million 1980 Dollars)

	<u>ESCOE Basis</u> <u>1978 \$</u>	<u>ESCOE Basis</u> ¹ <u>1980 \$</u>	<u>50,000 BBL/ Day Output</u> <u>Basis</u>	<u>Capital</u> <u>Cost/ Daily BBL</u>	<u>Total</u> <u>BTU</u> <u>El4/Yr.</u>	<u>Capital</u> <u>Cost/ MM BTU/ Yr.</u>	<u>Reference:</u> <u>Tons of</u> <u>Coal/Day</u>
<u>Coal Liquids</u>							
<u>Direct Liquefaction</u>							
SRC-II	\$1,262.	\$1,565.	1310.8	\$26,210	1,081	\$12.13	20,938
EDS	1,279.	1,574.	1422	28,440	1.072	13.26	22,584
H-Coal Syn. Oil	1,134.	1,407.	1252	25,040	1.115	11.23	22,242
H-Coal F.O.	955.	1,185.	980.9	19,620	1.048	9.36	20,695
<u>Indirect Liquefaction</u>							
Fischer/Tropsch	1,121.	1,391.	1730	34,600	1.112	15.56	31,095
Mobil 'M'	1,212.	1,676. ²	1396.4	27,930	8016	17.42	20,833
Methanol	1,195.	1,482.	608.4	12,170	428	14.20	10,263
Methanol/SNG	1,587.	2,225. ²	2132.9	42,650	905	23.57	26,174
<u>Coal Gasses</u>							
<u>High BTU</u>							
Lurgi	1,151.	1,427.	1313.1	(26,260) ³	1.067 ³	12.30	23,000
<u>Low BTU</u>							
Westinghouse	684.	851.	889.5	(11,790) ³	1.067 ³	5.58	17,313
<u>Shale Oil</u>							
Surf. Retort.	700. ⁴	798.	798.	15,960	.953 ⁵	11.93	N.A.

¹ESCOE - 25,000 tons coal/day input basis revised to reflect 20% contingency vs. 10% and 1980 dollars.

²Mobil Research Center Basis - \$ 1977 - 27,300 ton coal input revised \$ 1980 and 1.73% markup of plant.

³Evaluated at average heating value of coal liquids 6.5 MM/BTU/bbl. x 50,000 bbl./day = 325 billion BTU/day (1.067 El4 BTU /yr.)

⁴OTA Basis - 3rd quarter 1979, 50,000 bbl. basis.

⁵Evaluated at average daily value of 5.8 MM/BTU/bbl.

SOURCE: E. J. Bentz & Associates

EXHIBIT 4-3

COST = ATA
ESCOE)

Process	Capital C	OPERATING & MAINTENANCE COST					Local Tax & Ins. 2d	Total M 2
		Fuel 1	Catalyst & Chem. 2a	Labor 2b	Maintenance 2c			
SRC-I	1092	246.	3.0	13.8	33.	55.	104.8	
SRC-II	1262	246.	6.0	12.2	38.2	43.	119.0	
EDS	1270	246.	6.0	12.2	38.5	64.	121.0	
H Coal: Fuel Oil	955	246.	6.0	12.2	29.	48.	95.0	
Syncrude	1134	246.	7.0	12.2	34.3	57.	111.0	
FT	1121	246.	7.0	12.2	34.	56.	109.0	
Methanol	1195	246.	7.0	12.2	34.	60.	113.2	
M-Gasoline	1212	246.	8.5	12.2	35.5	65.	121.0	
CO ₂ Acceptor SNG	1084	246.	5.9	12.7	34.6	54.	107.0	
Syngas	942	246.	3.2	12.7	22.5	47.	73.0	
HYGAS	980	246.	4.8	12.2	23.4	49.	69.0	
BIGAS	998	246.	5.8	12.2	23.9	50.	91.9	
Synthane	870	246.	4.5	12.9	24.3	44.	82.0	
Lurgi	1151	246.	4.5	12.7	36.7	58.	112.0	
CE Power	1268	246.	3.0	12.0	38.	63.	116.0	
West Power	1066	246.	3.0	12.0	32.	53.	100.0	
Westinghouse Syngas	684	246.	4.5	12.2	20.5	34.3	71.4	
Shale Oil	798	H	H	H	-	H	-	

SOURCE: E. J. Bentz & Associates

4-5

efb&a

E* HIB T 4-4

ANNUAL OPERATING COST - 50,000 BBL LIQUIDS/DAY PLANT

(Million 1980 Dollars)

Synthetic Coal Liquids	Capital Charges @ 30% of Capital	Feedstock Coal (Shale) @ \$30/Ton (00)	Other Operating Costs						Total Operating Cost	Average Cost/ BBL Liq.	Cost/ MM/BTU
			Catalysts & Chems. (escalated @ 20%)	Labor	Maintenance 3% of Total Capital	Utilities & Supplies (50% of Chem.)	Taxes & Ins. 3%	Total Other			
<u>Direct Liquefaction</u>											
SRC-II	393.2	206.3	6	11.9	39.3	3	39.3	99.5	699.	42.557	6.47
EDS	426.6	222.6	6	13.	42.7	3	42.6	107.3	756.5	46.058	7.06
H-Coal S	375.6	219.2	7	12.7	37.6	3.5	37.5	212.6	807.4	49.157	7.24
H-Coal F.O.	294.3	203.9	6	11.8	29.4	3	29.4	79.6	577.8	35.178	5.51
<u>Indirect Liquefaction</u>											
Fischer- Tropsch	519.	306.4	7	17.7	51.9	3.5	51.9	132.	957.4	58.29	9.61
Mobil 'M'	418.9	205.3	7	11.9	41.9	3.5	41.9	106.2	730.4	44.469	9.11
Methanol	182.5	101.1	3.5	5.9	18.3		18.3	48.	331.6	20.189	7.75
Methanol/SNG	639.9	256.9	6	13.7	64.0		64.	150.7	1047.5	(63.775)	11.57
<u>Synthetic Gas</u>											
Hi BTU Lurgi	393.9	226.7	4	13.7	39.4	2	39.4	98.5	719.1	(43.78)	6.74
Med. BTU Westinghouse	266.9	170.6	3	10.3	26.7	1.5	26.7	68.2	505.7	(30,788)	4.74
<u>Shale Oil</u>											
Surf. Retort.	239.4	230.0	-	-	-	-	-	322.3	791.7	48.20	8.31

SOURCE: E. J. Bentz & Associates

4.2 SCALE OF PRODUCTION

Chemical process plant economics are highly sensitive to scale. Typical scaling factors or rules, are of the order of 60% - 70%. This means that as plant size doubles the cost only increases by 60% - 70%. In the case of decreased scale - the factor works in the opposite direction, a decrease in scale to $1/2$ plant scale leads to only about $1/3$ decrease in cost, which in turn leads to almost 30% more capital being required per unit of output. In very capital intensive processes, the importance of this to product cost is great. Coal conversion processes typically have $1/2$ of their costs derived from capital charges, therefore a doubling of scale could reduce total unit costs by as much as 15% - 20%.

For this reason the question of plant scale must be very carefully examined. ESCOE, in ordering the various estimates to the values shown in Exhibit 4-1 applied "typical chemical engineering scaling factors". It is beyond the scope of the present effort to audit that undertaking. However, it is incumbent upon us to avoid the distortion of fairly presented uniform cost data by another exponential adjustment of capital costs. We must rescale the liquids' plants since they have been standardized on an 'input' basis, whereas we must examine costs on a plant 'output' basis, since we are also examining downstream processes and costs, which in turn require uniform scale assumptions.

Several difficulties are present:

1. The optimal size of plant and vessels for various systems is not known, due to the fact that most processes are now being explored at 5 - 10% pilot plant scale.
2. In a shift from uniform input scale to a uniform output scale, the most efficient processes will suffer the greatest penalty for their relative downsizing. This is not realistic.
3. We are not aware of the relative changes that took place in the initial (ESCOE) standardization, hence are blind to the compound effect of a second scaling adjustment.

For these reasons, with the emphasis upon the above factors, in order of their ranking, we have chosen to restate costs on an output basis through a linear method of cost adjustment.

The principal justification for this apparently unsound procedure is found in the first factor above - there is no evidence of commercial scale economy available in the case of any processes, with the exception of gasification plants (or gasifier reactors). In that case, multiple train plants appear at sub-commercial plant scale. In general, the bulk of the solid feed stock is so great, that initial reactor vessel sizes become

limited by available fabricating (rolling, bending, heat-treating) facilities, as well as transportation constraints. Subsequent plant stage economics do not determine. The gasifier-reactor vessel size limitations are such that returns to scale may be limited at a relatively low level of output.

For this predominant reason, we have used a unitary cost scaling factor to shift from uniform input sized plants (25,000 tons of coal per day) to a uniform output basis - 50,000 bbl. per day. A normal procedure would otherwise unfairly penalize the most efficient processes. In the final analysis, efficiency will determine economic advantage.

4.3 PRODUCT QUALITY (Reference No. 38)

The issue of product quality was resolved in a somewhat indirect manner by ESCOE. Their 'rating scale' value system (a measure of ordinal utility or value) which was based on present product price relationships is not a suitable method for long range economic analyses. During the long-run, values change, end use patterns and conversion technology developments can create a surplus of a once premier product, or contrariwise, create a shortage of a previously unwanted by-product. Distillates and gasoline have traded places once and are perhaps posed to trade places again in their relative values.

The setting of widespread synthetic fuels production and use creates an entirely new framework for evaluating the 'normal' refinery slate of petroleum derived products. We have created a slate of products that to some degree reflects the range of compounds present in crude oil and in some degree reflects the technology (now) available to separately produce these compounds. In some instances the products were specifically sought, in other cases markets were sought for by-products that were available.

When coal is introduced in lieu of crude oil to a substantial degree, the available range of products and by-products may be the same, but the proportions of availability will be quite different, as will be the cost of producing different fractions. "

The proportion of each fraction that can be derived from crude oils is highly variable depending upon the nature of the feedstock and the nature of the refining processes used. In general, increasing the lighter fraction (-350°F) involves more severe reforming, and higher cost. The use of a heavy, sour feedstock crude oil worsens this condition. The use of coal as the feedstock significantly exaggerates this condition in certain synthetic processes - such as direct liquefaction. Indirect liquefaction processes are specific for alcohols, gasolines and the light ends.

It is reasonable to visualize a population of crude oil and coal "refineries" with individually more specialized or limited

product slates than are found in the universe of conventional refineries.

Broad slate coal synthetic liquids plants are unlikely to be widely deployed. This can be expected for several reasons:

1. Product upgrading is difficult and expensive once outside of the basic process.
2. A fair range of limited slate coal-conversion processes are becoming available, that more selectively produce various fractions.

The costs of achieving a given level of product quality increases in a slightly non-linear fashion as the percent hydrogen is increased or the boiling range is lowered. Exhibit 4-5 shows this relationship graphically. Benchmark products and costs are shown for several direct and indirect liquefaction processes. The indirect processes - which catalytically synthesize liquids from synthesis gas are specific for gasolines, alcohols and LPG. The direct catalytic hydrogenation processes tend to produce naphthenes and crude oil equivalent range compounds. The hydrogen solvent systems tend to produce a more limited range of product with a substantial (20 - 35%) naphtha fraction, the majority product in the distillate boiling range (350°F - 750°F).

Increased yield of the higher quality products can be achieved by:

- . Increased coking of bottoms
- . Adding more hydrogen
 - . To process stream
 - . By hydrotreatment of products

The cost of the former is seen in the difference between SRC II and EDS on Exhibit 4-4. The Exxon donor solvent system cokes the bottoms (or heavy distillates) to yield more naphtha and LPG as follows:

<u>SRC II</u>		<u>EDS</u>	
(18%)	13,000 bbl Naphtha	(36%)	27,500 Naphtha
(8%)	6,400 bbl #2 Fuel Oil	(15%)	10,000 LPG
(73%)	<u>52,900</u> bbl Distillate	(49%)	<u>37,200</u> bbl Distillate
	72,300		75,400

Similarly changing the H Coal process from a fuel oil to a synthoil mode increases cost as it lowers the average boiling range.

The distribution of product quality that is typical of each process is shown on the following page. (Exhibit 4-5).

EXHIBIT 4-5

SYNTHETIC PLANT PRODUCT YIELDS

QUANTITY - BBLs/DAY OUTPUT

	Approx. API ^o	Direct Liquids				Indirect Liquids				Synthetic Gas ³		Shale Oil
		SRC-II	EDS	H Coal Coal (Syn.)	H Coal Fuel Oil	Fischer/ Tropsch	Mobil 'M'	Meth- anol	Meth- anol SNG	Lurgi H BTU	West. Lo BTU	Shale Oil Bit. (Surf.)
SNG (Low) MM BTU/Day											880	
Methanol (High) MM BTU/Day								140	300			
LPG	125 ^o	4,610			23,380	6,080						
Propane (C ₃)	148 ^o		2,950									
Butane (C ₄)	110 ^o		3,160									
Methanol	NA							50,000	48,740			
Gasoline (C ₅)	62 ^o				82,640	43,920						
Naphtha (C ₅₊)	40 ^o	10,625	17,970	28,380	15,070	1,490 ¹ 2,490 ²		1,260	2,025			
Fuel Oil (Heavy Distillate)	18 ^{o+}	35,000		21,620	34,930							50,000 ⁴
Fuel Oil (Resid.)	5 ^o		25,920									

¹ Light (Diesel) Fuel Oil API-56^o, ² Heavy Fuel Oil API-41^o.

³ 1.067 EJ4 BTU/yr. is equivalent to 50,000 BBL/day of typical Synthetic Liquids ≈ 6.5 Million BTU/bbl.

⁴ Synthetic light crude oil equivalent - approx. 20^o API.

SOURCE: E. J. Bentz & Associates

4-10

ejbka

The range of quality is not entirely a function of the API gravity, the boiling range or hydrogen content, however, these related indices are sufficient for our purposes. We can relate the cost of producing a synthetic fuel to this scale. Exhibit 4-6 shows a graph of the production cost of the whole liquid product from various synthetic processes versus the average (50% distillation) boiling range of the synthetic product.

This chart shows the increase in average cost per million btu's as the average distillation range of the liquid is lowered. Thus gasoline costs more to produce via indirect processes such as Mobil 'M' or Fischer Tropesch, than naphthas, distillates and fuel-oils.

This scale illustrates the relative costs of the ESCOE liquid fuel processes. It also contrasts the (1978) earlier ESCOE cost estimates with later estimates of shale oil costs developed by the Office of Technology Assessment (1980). The oil shale liquids, which reside in a higher boiling range than the coal liquids, appear significantly more expensive on this scale. In order to reconcile this discontinuity it is necessary to digress briefly.

4.4 ESTIMATING METHODS

The accuracy of complex systems cost estimating has been the subject of several studies. These studies have been primarily behavioral rather than conceptual. As larger, more complex systems projects have been conceived, -the amount of unknown and untried system components have necessarily increased due to the great cost of large system prototypes. Pilot or process demonstration units and models are developed at extremely small scale for the same economic reasons; the subsequent scale-up is of a high order. Estimates drawn from bench or small scale pilot plants are subject to much greater estimating error.

Two overriding conclusions have been reached in this matter:

1. Cost estimates tend to decrease in variation from actual costs as the elapsed time between estimate and construction is shortened.
2. The accuracy of the estimate is related to the degree of detail of the design engineering.

Chemical process plants,⁸ public works,⁹ and weapons systems¹⁰ development and estimating histories have been analyzed, with

⁶ Syngas (fuels) are not suitable related to boiling point measurement.

⁷ "An Assessment of Oil Shale Technologies", OTA - June 1980.

⁸ A Review of Cost Estimates in New Technologies: Implications for Energy process Plants, Rand Corp. for the Dept. of Energy July 1979.

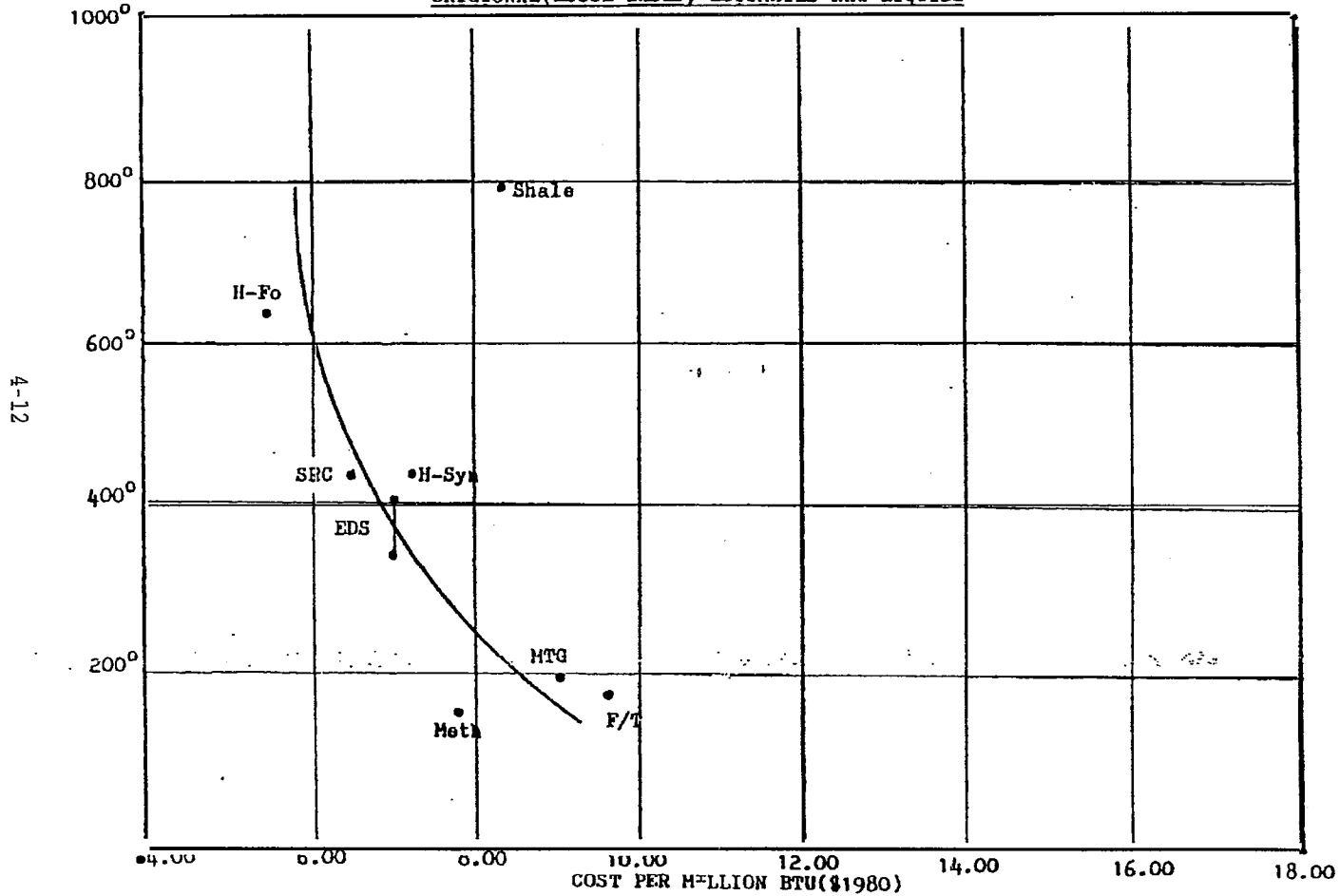
⁹ "Systematic Errors in Cost Estimates for Public Investment Projects", Hufschmidt & Gerin, in The Analysis of Public Output, Columbia Univ. Press 1970.

¹⁰ The Weapons Acquisition Process: An Economic Analyses, Peck & Scherer, Harvard Univ. 1962.

EXHIBIT 4-6

50% Boiling Point
Degrees F

SYNTHETIC FUELS PRODUCTION COSTS
ORIGINAL(ESCOE BASED) ESTIMATES-RAW LIQUIDS



SOURCE: E. J. Bentz & Associates, Springfield, Virginia

essentially the same range of variances found between early estimates and actual results - growth in costs have occurred of the order of 2-3 times the original estimate. The average of actual to estimated costs ($\frac{Ca}{Ce}$), were found to be as follows:

<u>System Type</u>	<u>Actual Cost/ Estimated Cost (Ratio)</u>
Weapons System	1.40 - 1.89
Public Works	1.26 - 2.14
Major Construction	2.18
Energy Process Plants	2.53

The weapons system cost overruns were higher in the 1950's (1.89) than in the 1960's (1.40) most likely, because of the greater degree of pioneering efforts and the greater lack of experience with large weapons systems at that time.

Exhibit 4-7 below shows the cost growth experience in pioneering energy systems as a function of the type of estimate employed (or available at that time). It can be seen that the preliminary estimates were nearly double that of the initial estimates - (84% above the first estimate) and the definitive estimates increased almost as much again from the preliminary estimates (134% above the first, or 50% above the preliminary estimate).

The ESCOE data were largely taken from preliminary estimates, based on Process Demonstration Unit (PDU) development experience, in one or two cases from pilot plant experience (at less than 1% scale) or from foreign commercial experience under different site and environmental conditions. The OTA shale oil values were derived from a very highly definitized engineering analysis. The degree of evolution which that estimate had undergone can be seen on Exhibit 4-E.

If the other ESCOE liquid synfuel plants were to increase by as much as have typically occurred between preliminary and definitive estimates, the costs would increase by about another 50%.¹² That would result in a shift of the cost line on Exhibit 4-6 as shown on Exhibit 4-9.

Such an interrelation of the quality of the ESCOE estimates would resolve the discrepancy between the ESCOE estimates and the OTA estimates (for oil, shale liquids) and produce a more continuous scale of synfuel cost relationships.

An alternative method of calibrating the various estimates for consistency with respect to the status of process estimates as well as the methods employed in the estimating process? would be to select

¹¹ Average increase from preliminary to definitive cost estimates for energy process plants.

¹² - Reference 3.

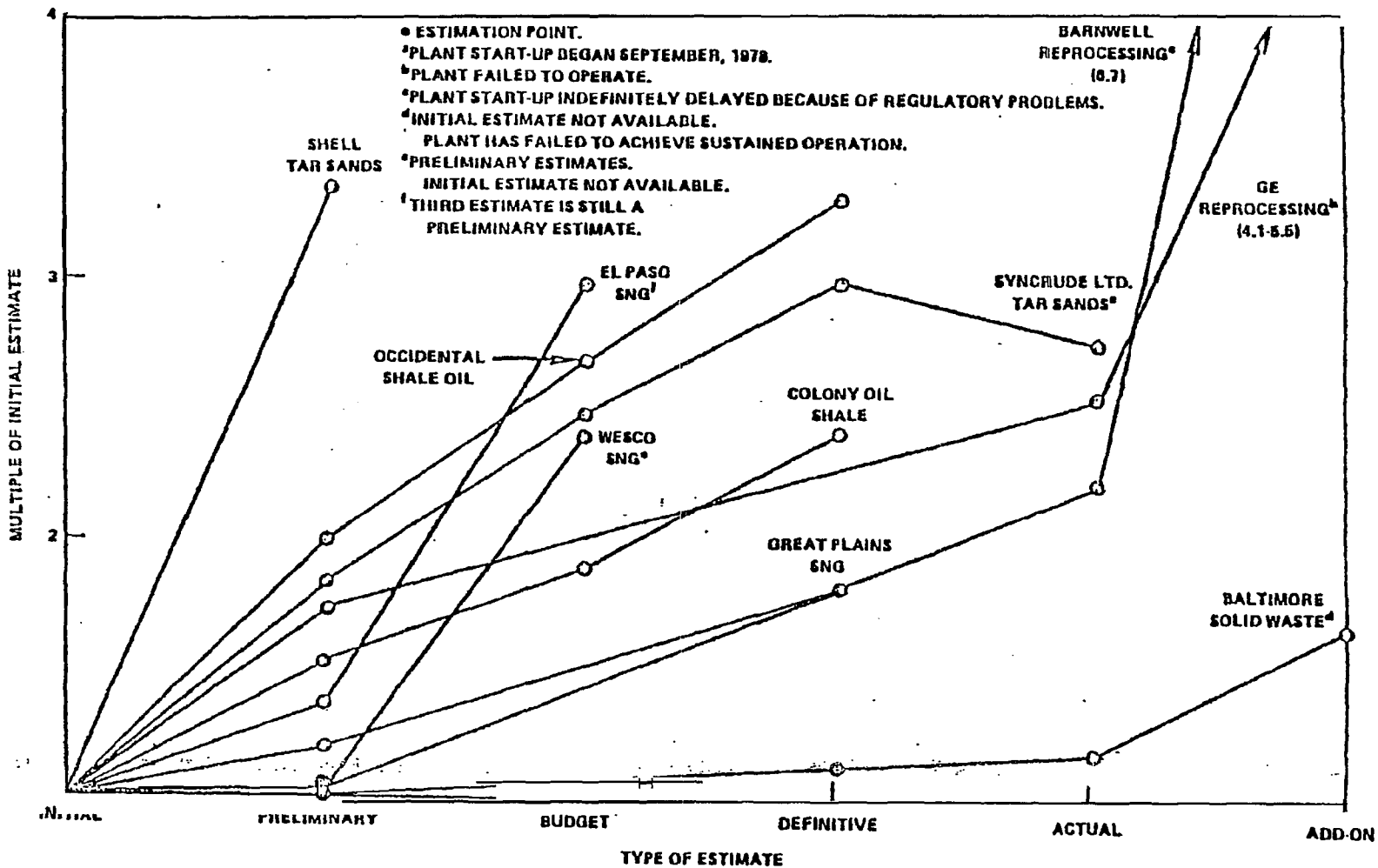
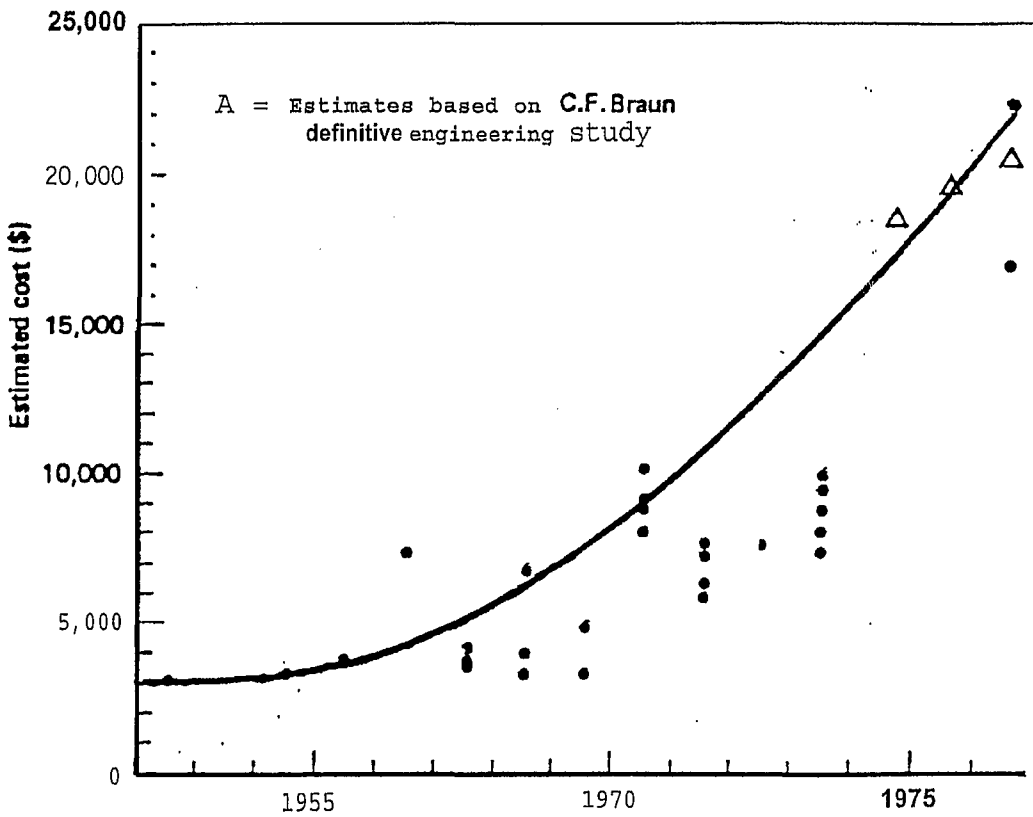


EXHIBIT 4-7
 COST GROWTH IN PIONEER ENERGY PROCESS PLANTS (CONSTANT DOLLARS)

Source: "A Review of Cost Estimation in New Technologies: Implications for Energy Process Plants," DOE, E-24-81, July 1979.

EXHIBIT 4-8

HISTORY OF SHALE OIL CAPITAL COST ESTIMATES



Estimated surface shale oil facility construction costs
(capital costs/barrel/calendar day; constant \$ 1977)

a sub-set of processes that were developed on the basis of the same level of engineering definition or maturity - preferably the most advanced projects in this sense.

There have been more recent, updated design and estimating efforts undertaken in the case of:

- 1) Indirect Liquefaction - Mobil MTG. ¹³
- 2) Methanol ¹³
- 3) High BTU Gasification ¹⁴
- 4) Direct Liquefaction - H-Coal ¹⁵

These estimating efforts are essentially comparable with the (OTA) Oil Shale estimates in terms of the relative engineering and development maturity of the process plants involved.

Exhibit 4-9 also reflects the liquid fuel costs of 'generic' synfuel processes based on the selected "best estimates" noted above. These are not meant to be truly generalized processes (or generic processes) , they are nonetheless representative, advanced members of each synthetic liquid product class.

The costs of these processes are shown in detail on Exhibit 4-10.

The effect of using the latest, or best estimates is approximately the same as was achieved by the use of the Rand Corp. (and others) cost estimating error factors. The original ESCOE values are increased by about 50% on average.

The satisfactory conjunction of factored cost estimates arrived at by the use of statistical variances derived from past estimating histories with the "generic" estimates taken from the most advanced projects, gives us an improved measure of confidence in the adjustment of ESCOE synfuel production costs to the higher levels displayed on Exhibits 4-9 and 4-10. The revised functional form of the liquid fuels is displaced to the right on Exhibit 4-9 by about \$3.00-\$4.00 per million BTU's. The relative costs are not appreciably affected considering the probable differences in residual (estimating) error contained in these estimates. It seems most reasonable, however, to presume that the majority of the estimating errors have been accounted for, and the values we are employing are normalized to the greatest practical degree possible at the present time: i.e., barring further engineering or demonstration plant design and construction experience.

¹³ Liquefaction Technology Assessment - Phase I ORNL-5664 Feb. 1981.

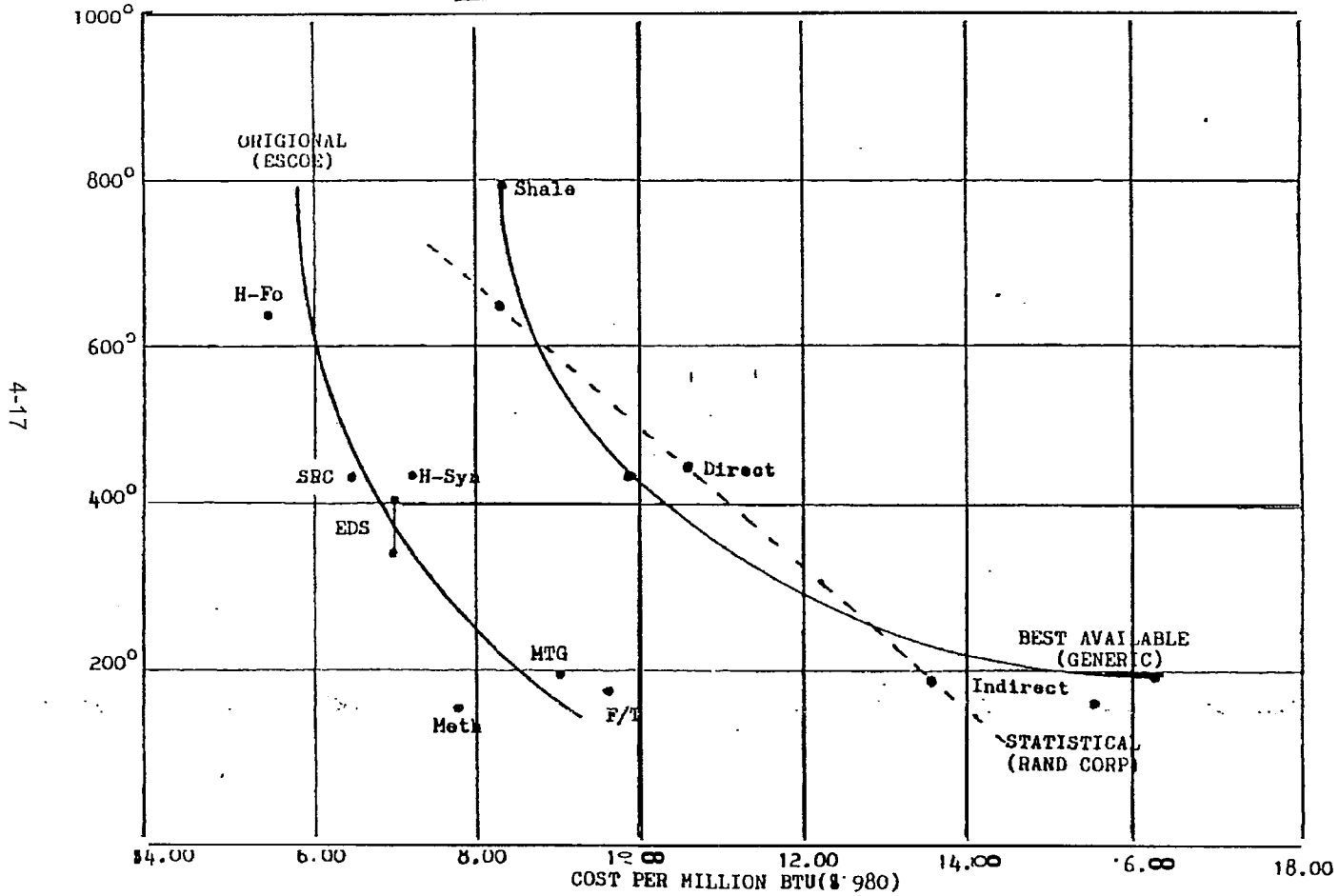
¹⁴ Unpublished Analyses

¹⁵ Rand Corporation - Unpublished Analyses.

EXHIBIT 4-9

50% Boiling Point
Degrees F

SYNTHETIC FUELS PRODUCTION COSTS
FACTORED COST ESTIMATES-GENERIC RAW LIQUIDS



SOURCE: E. J. Bentz & Associates, Springfield, Virginia

EXHIBIT 4-10

BEST AVAILABLE ESTIMATES - TYPICAL SYN FUEL PROCESSES

	Revised Capital Est. \$ 1979	Revised Capital Est. \$ 1980	Capital Recovery @ 30%	Feedstock Cost (From 4-4)	Other* Oper. Costs (From 4-4)		Total Revised Estimate	Cost Per Barrel \$ 1980	Cost Per MM BTU \$ 1980	Cost 1 Gal. \$ 1980
<u>Direct Liquids</u>										
H Coal (Synfuel)	\$2,200	\$2,200	\$ 660	\$219.2	\$212.6		\$1,091.8	\$66.47	\$ 9.79	\$ 1.58
<u>Indirect Liquids</u>										
Mobil MTG (Lurgi)	2,685	3,054	916.2	205.3	159.8		1,281.3	78.01	16.18	1.86
SNG/Methanol (ICI- Lurgi)	1,849	21,035	631.1	256.9	160	1048 SNG	688. 360.	11.88	15.53	1.00
<u>High BTU Gas</u>										
Lurgi (BGC)	1,600	1,820	546.	226.7	113.7		886.4	-	8.30	-

*Adjusted for capital cost changes.

SOURCE: E. J. Bentz & Associates

4-18

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The H-coal process (in the synfuel mode) has been used as a surrogate for direct coal liquids. Updated estimates of an unpublished nature were used that draw from the cumulative pilot plant histories and the most recent demonstration plant estimates. The Mobil Methanol-to-Gas (MTG) and methanol estimates were drawn from a recently published study by Fluor Corporation for Oak Ridge National Laboratory of indirect liquefaction processes. The study provided a (nearly) 100% gasoline option which virtually eliminates the by-product costing problems. The methanol estimates were Methanol/SNG joint production process schemes. The systems could have been adjusted in keeping with the 100% gasoline MTG process scheme by eliminating the direct costs of methanol to gasoline stages. Alternately the by-product value of SNG could be directly priced by using the high BTU gas plant costs from the SNG estimate below. Both synthesis gas processes are Lurgi systems.

The SNG process estimate was taken from unpublished estimates drawn from advanced commercial design and estimating efforts. An advanced Lurgi gasifier - the British Gas Corporation slagging bed version - is used.

The costs of direct and indirect liquids - increase by about 50% - to remain in approximately the same relative cost relationship that the ESCOE based data displayed. The hi-BTU gas estimates only increased about 25% above the earlier ESCOE values. This appears to be reasonable considering the relatively more mature status of (Lurgi) gasification technology. The OTA oil shale liquids estimate of \$48.20/bbl reflects the precommercial stage of development. The level that we are attempting to standardize at, versus the development stage of the foregoing direct and indirect liquid systems.

Continuing Cost Escalation

The earlier analyses of Rand Corp. and others suggested that the potential cost increase from even a definitive estimate to the actual project costs of pioneer plants and major developmental systems is typically another twenty percent increase in cost. We can add that increment to arrive at an upper value for all systems.

There have been and continue to be other relevant post-commercial trends of commercial series production plants that were not considered by the authors of the cost escalation - studies cited above.

Historical data regarding the chemical process industry and petroleum refining industry demonstrates a strong pattern of capital productivity improvement or technology advance, during post-development years. This can be demonstrated for the entire sector as well as in the micro-industrial setting of a single chemical industry segment.

A capital productivity rate of less than 2%/year can return the 20% (actual cost to definitive cost estimate potential increase

during the first 10 years of commercial deployment. In 20 years at least a 35% reduction in the capital outlay per barrel of product can be expected.¹⁶

These two viewpoints provide us with minimum and maximum estimates of the most probable range of expected production costs for synthetic fuels. Exhibit 4-11 illustrates the range of expected values for synfuel liquids based on these estimating limits.

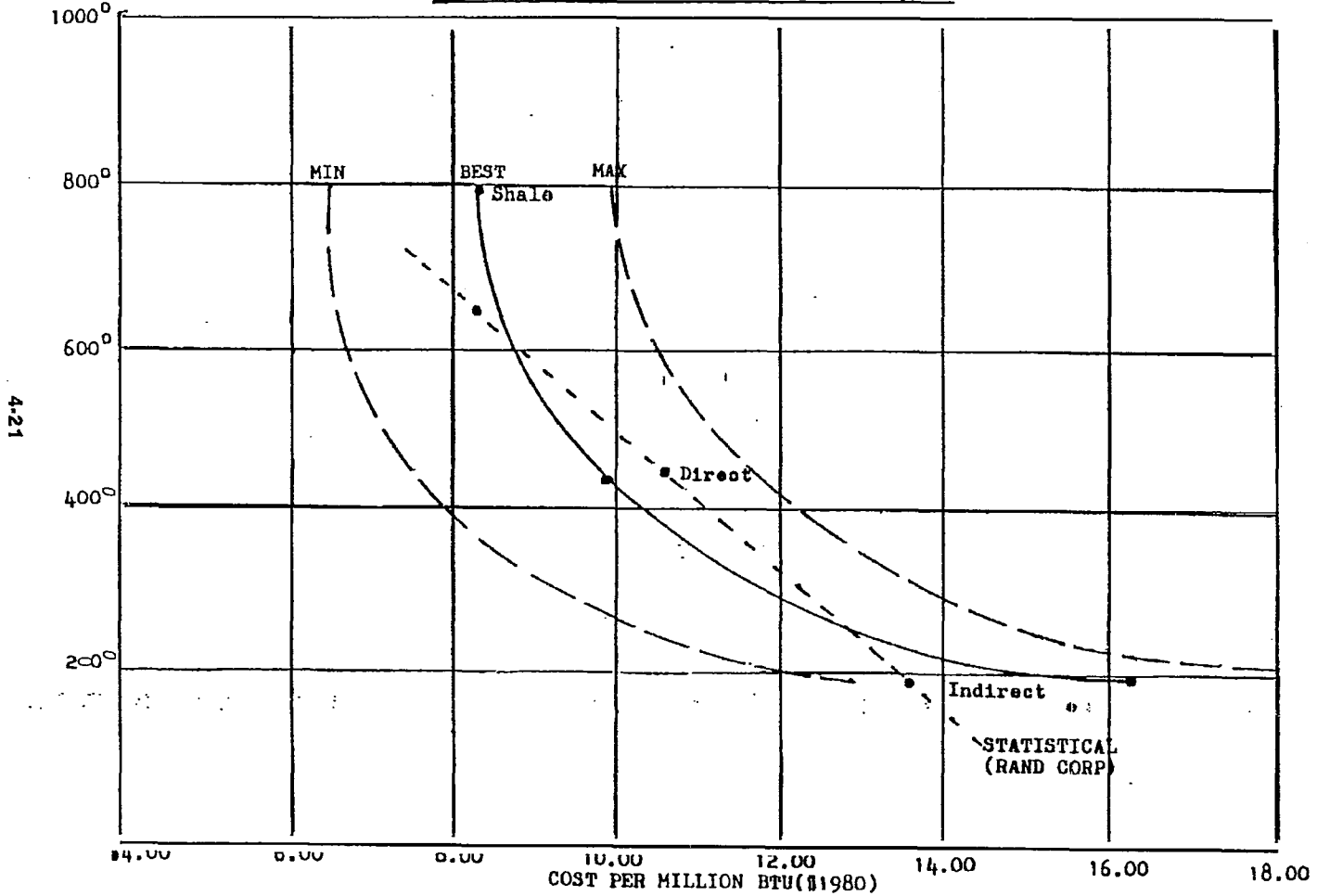
This scale of values will be used to provide individual product (or by-product) costs. The presence of a significant amount of petroleum in the total supply equation, for as far as we can see, creates many cost and pricing complexities. We do not wish to complicate synthetic fuel supply economics with World Oil Price disruptions, or any free-market or administered market conditions. We will close our eyes to all of these dimensions and construct our cost schedule on the basis of coal based liquid, gas and solid fuel options or opportunity costs.

¹⁶This rate (1.4%) has been experienced by the entire chemical industry throughout the entire post war period (1949 to date). Specific industry sectors have experienced much greater rates of productivity improvement; viz, synthetic methanol experienced more than a 4% / year productivity gain for over 20 years.

EXHIBIT 4-11

50% Boiling Point
Degrees F

SYNTHETIC FUELS PRODUCTION COSTS
FACTORED COST ESTIMATES-GENERIC RAW LIQUIDS



SOURCE: E. J. Bentz & Associates, Springfield, Virginia

4.5 PRODUCT UPGRADING (References 39, 40)

The typical (direct liquefaction) coal liquids possess several characteristics that require upgrading in order to:

- . Provide product stability
- . Permit mixture with conventional petroleum liquids . . . or
- . Permit common use of pipelines and other infrastructure.

The principal differences result from:

Lower levels of hydrogen - 9 - 10% versus 11 - 14% for petroleum and 11 - 12% for shale oils.

Higher levels of heteroatoms in both liquids and shale oil (nitrogen and oxygen compounds) than are found in petroleum feedstocks.

The lower hydrogen and higher heteroatom conditions are resolved together by hydrotreatment. Raising the hydrogen levels up above 10% results in the removal of most of the nitrogen and oxygen heteroatoms, and also decreases the aromaticity of the coal liquids and shale oils.

The high aromatic content of coal liquids makes the naphthas excellent high octane blending stock - however the high nitrogen and oxygen percent (2 - 3%) in the heavy naphtha range requires the use of fairly severe hydrotreatment to remove the diolefins and heteroatoms - which are present in the form of phenols and cresols (oxygen).

In the synfuel distillates the nitrogen level is higher and results in unstable compounds with rapid gum formation, making this a very unsatisfactory fuel unless upgraded.

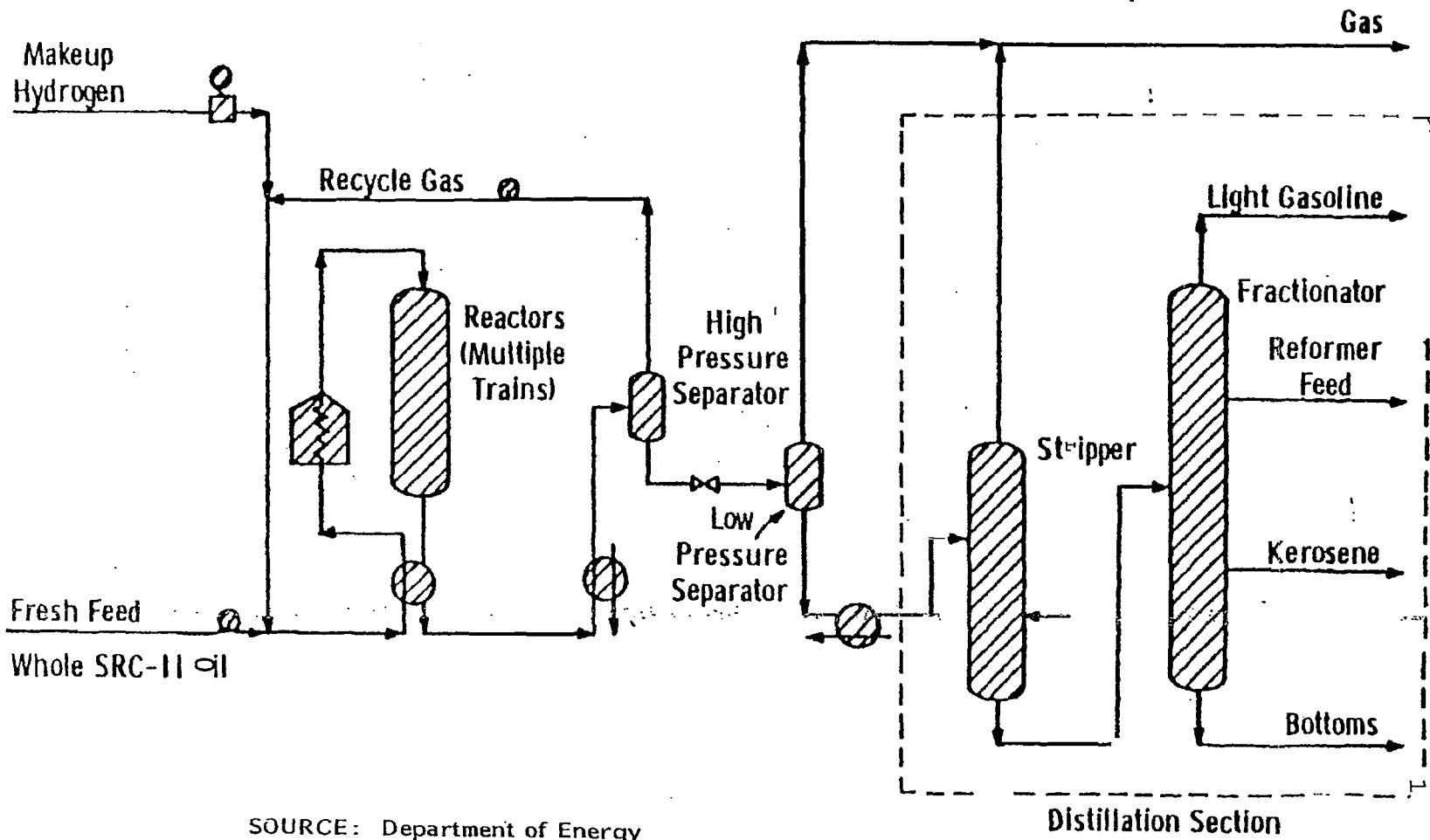
There have been a succession of studies of synthetic liquids upgrading processes sponsored by DOE. They have been conducted on both shale oil and direct coal liquids.

The principal measures examined include:

- . Hydrotreating (Exhibit 4-12)
- . Hydrocracking
- . Fluid Catalytic Cracking

Catalytic reforming as well as hydrocracking are subsequently used to upgrade (naphthas) to finished transportation fuels. (See Exhibit 4-12 below) .

EXHIBIT 4-12: SIMPLIFIED FLOW DIAGRAM OF
 CHEVRON FIRST STAGE HYDROTREATER
 REFINING OF SRC-II OIL



SOURCE: Department of Energy

4-23

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Consideration has also been given to variation in the hydrogen source for hydrotreaters - the partial oxidation of raw coal liquids, reforming of refinery products and overheads, or outside gasses.

An additional issue is the location of upgrading facilities; at the coal liquids (or shale oil - retort) plant, or at a conventional refinery, or both.

The factors which favor the synthetic oil plant location are:

- . available residue for hydrogen manufacturing
- . local upgrading permits common carrier transportation
- . upgraded synthetic product can be blended with petroleum feedstock (in pipelines and at refineries)

The factors that favor a refinery location for upgrading are:

- . Superior prospects for system optimization
- . Availability of hydrogen from naphtha reformers
- . Uses available refinery capacity idled by lack of petroleum feedstock.

An alternative approach could be to perform a minimum amount of upgrading at the synfuels plant to facilitate transportation and storage, with product finishing and blending performed at a larger refinery site. The coal liquids in general do not require further cracking because they lie in the atmospheric gas-oil and naphtha range. The shale oils require cracking to produce more usable product from the higher distillate range such as jet fuel and diesel oils. The heavy distillates from coal liquids; if heavily hydrotreated (to 11% H bywt) can be used as a feedstock for a fluid catalytic cracker (FCC) where the product can be significantly upgraded.

Exhibit 4-13 illustrates the cost of upgrading various direct liquid process cuts.

The raw liquids versus the upgraded liquids are compared below in hydrogen content.

	<u>Raw Liquid</u>	<u>Upgraded</u>
SRC Naphtha	11.33%	11.6%
SRC Distill.	7.71	11.0
H Coal Distillate	10.1	11.4
H Coal Fuel Oil	7.37	10.0

These cases cover the general conditions experienced by the range of most direct coal liquids - the samples being drawn from experimental laboratory investigations performed by Mobil Research and Development Corporation upon SRC light and heavy fractions and

EXHIBIT 4-13

DIRECT LIQUIDS UPGRADING COST

50,000 BBL) - 1980 \$

	<u>SRC-II</u>		<u>H Coal Svn. Crude or Fuel Oil</u>		
	<u>Naphtha</u>	<u>Heavy Distillate</u>	<u>Naphtha</u>	<u>Distillate</u>	<u>Fuel Oil</u>
Operating Labor	.0639	.0855	.0639	.086	.085
Maintenance	1606	5830	1606	409	593
Administration & Support	.0394	.0992	.0394	.077	.100
G&A	<u>.1828</u>	<u>.5757</u>	<u>.1828</u>	<u>.420</u>	<u>.585</u>
	.4467	1.3434	.4467	.992	1.363
Fuel	.3517	.8691	.3517	.525	.154
Utilities	.1089	.9716	.1089	.577	.946
Cat. & Chem.	.0194	.9829	.0194	1.650	1.358
Hydrogen	<u>.5586</u>	<u>6.836</u>	<u>.5586</u>	<u>2.390</u>	<u>7.355</u>
	1.0385	9.4895	1.6386	5.142	9.813
Capital Recovery (30%)	<u>2.577</u>	<u>8.484</u>	<u>2.577</u>	<u>5.449</u>	<u>7.933</u>
Total Upgrading Cost	4.062	19.32	4.062	11.58	19.109
Product					
API	37.5	24.5	Same as	25.7	12.5
H Content (wt %)	11.6	11.0	SRC II	11.4	10.0
BTU/lb.	18,500	18,780	Naphtha	18,970	18,400
Plant Investment					
Total in Million \$/BBL	\$141	\$464.5	\$141	\$298	\$434
	\$8.59	\$28.28	\$8.59	\$18.16	\$26.44

SOURCE: E. J. Bentz & Associates

4-25

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H-coal distillate and fuel oil fractions. This pretty well covers the range of liquids produced by SRC and H-Coal (synfuel and fuel oil mode) and can be extrapolated to the EDS case.

Additional work performed by U.O.P., Chevron and Suntech confirm the general upgrading needs and the best approach - hydro-treatment.

The plant investment required varies from \$140 million dollars for the mild hydrotreatment required of the naphtha cuts (C₅-400°F) to as much as \$465 million for a hydrotreatment plant for the heavy distillate or residual SRC fraction and nearly that for the fuel oil fraction of H Coal fuel oil process plants.

The average upgrading cost is about \$2.00 per million BTU's - varying from \$4.00 to nearly \$20.00 per barrel. The latter figure represents an economic limit which suggests either a lower grade utilization of the heavier products or a different refining approach.

The direct liquids upgrading cost analysis can be compressed to a single representative or "generic" upgraded coal liquid.

The general costs of upgrading are shown on Exhibit 4-14:

Naphtha's	\$ 4.06
L. Distillates	11.58
Heavy Distillates	
Fuel Oil	19.21 (19.11-19.32)

Individual processes such as EDS SRC-II and H-Coal (fuel oil mode) will differ in raw liquid base costs, but since the quality of product tends to vary in a reasonable relationship to their costs¹⁷, the costs of upgrading, which are increasly related to quality, lend to cause a clustering of upgraded direct liquid costs.

If we utilize the costs of H Coal production of raw liquids developed above as a base, the 'generic' costs for upgraded products would be as follows on Exhibit 4-14. The estimated costs of nearly \$75.00 per barrel or over \$12.00 per million btu's is for a product that is equivalent to a high grade refining crude oil feed.

The upgrading of shale oil to a suitable refinery syncrude has been estimated by Chevron to cost \$10.00 per barrel (in 1980 dollars) or \$1.72 per million btu. If this is added to the cost of raw shale-oil liquids at the retort, the total cost of shale oil "syncrude" is:

¹⁷ See Exhibit 4-6 above.

EXHIBIT 4-14

DIRECT LIQUIDS UPGRADED COSTS/BARREL

(\$ 1980)

	<u>Barrels/Day</u>	<u>Cost/BBL</u>	<u>Total Daily Cost</u>
Naphtha	28,380	\$ 4.06	\$115,223
Distillate	<u>21,620</u>	11.58	<u>250,360</u>
	50,000		\$365,583 == \$7.31 Avg.
Raw Liquid Cost (per barrel)			<u>66.47</u>
Total Upgraded Fuel Cost Per Barrel			\$73.78
Total Upgraded Fuel Cost Per MM/BTU			\$12.30

SOURCE: E. J. Bentz & Associates

4-27

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OIL SHALE LIQUIDS COST

(\$1980)

	<u>Per Barrel</u>	<u>Per Million BTU</u>
Retorted Shale Oil	\$48.20	\$ 8.31
Upgrading	<u>10.00</u>	1.72
	\$58.20	\$10.03

These compare favorably with upgraded direct liquefaction production in the 'syncrude' class as shown below:

SYNCRUDE PRODUCTION COSTS

(\$1980)

	<u>Per Barrel</u>	<u>Per Million BTU</u>
Shale Oil	\$58.20	\$10.02
Direct Coal Liquids	21.12	18.5%
Shale Oil Advantage	12%	9%

The shale oil has about a 21%-cost advantage as a refinery feed-stock. This is reduced to less than a 20% cost advantage on a heating value basis. However heating values are not the principal criterion to be applied to refinery feedstocks - quite the opposite - the lighter crude demands a premium. In certain instances the coal liquid with higher aromatic content will be preferred, at other refineries the shale oil, with a higher hydrogen content, and a greater yield of distillate product will be sought.

Exhibit 4-15 illustrates how the process of upgrading shifts the cost of oil shale and coal based synthetic crudes upward by \$1.75 - 2.50 per barrel.

4.6 REFINING SYNTHETIC LIQUIDS

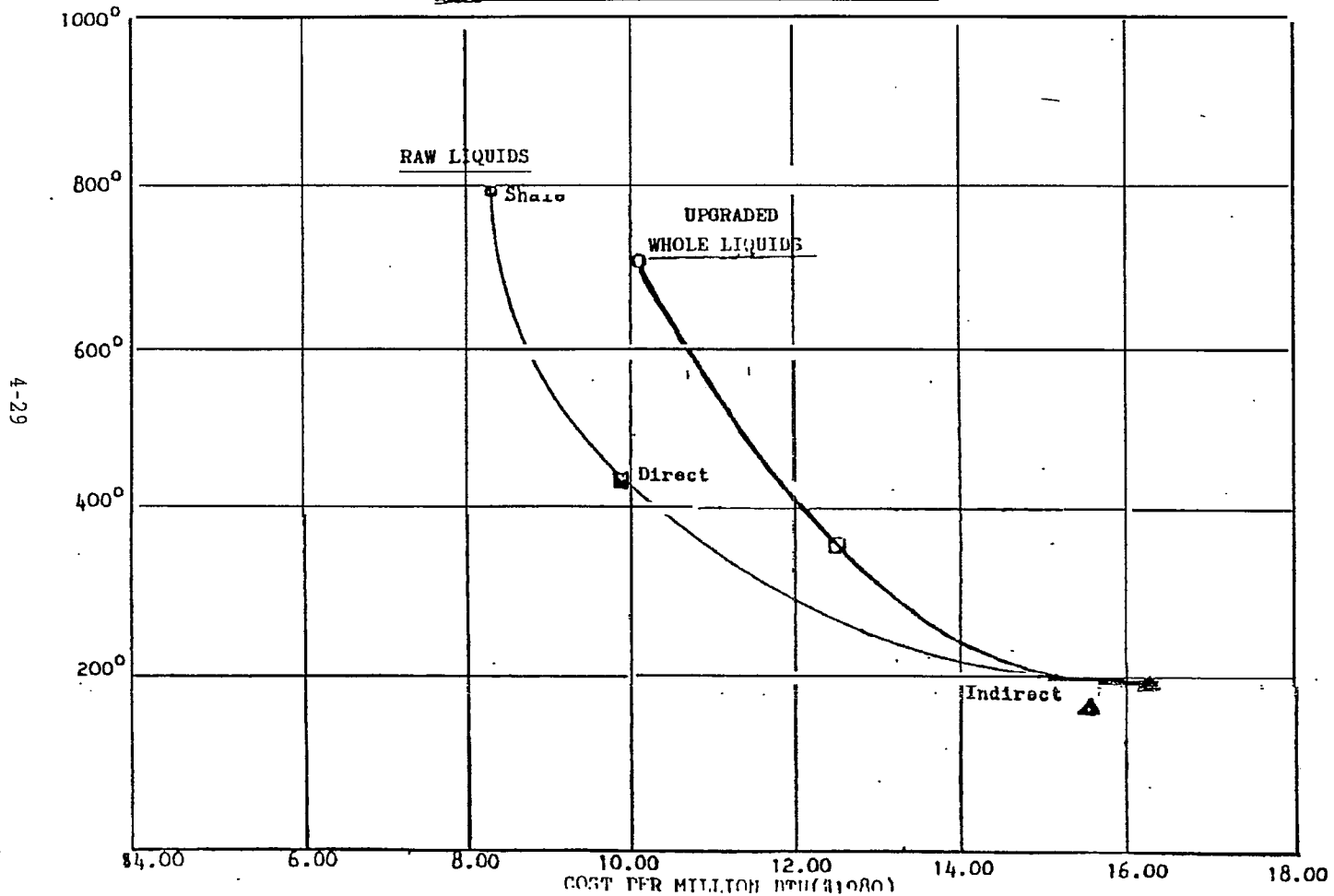
The direct liquefaction and oil shale synfuels have to be further upgraded to end-use product quality in order to be comparable with indirect liquid products such as methanol from coal or gasoline from methanol (from coal). In a wider sense, this is also desirable in order to achieve comparability with synthetic natural gas (SNG) which can be used for a wide range of end use applications in its 'raw' manufactured state.

The indirect processes produce refinery output (or intermediate) grade products, without the need for the "refining" of crude liquids. In order to compare direct liquids and shale liquids with indirect process liquids, we must bring the former

EXHIB T 4-15

50% Boiling Point
Degrees F

SYNTHETIC FUELS PRODUCTION COSTS
BEST AVAILABLE ESTIMATES - UPGRADED LIQUIDS



4-29

SOURCE: E. J. Bentz & Associates, Springfield, Virginia

into a state that is comparable. This requires the refining of the synthetic liquids to finished fuels.

Refining of shale oils and coal liquids will vary in cost depending upon the size, location and degree of integration of the refinery complex. We will assume that this is not done in an existing refinery (perhaps modified to better handle these feedstocks), but is performed at a new refinery integrated at the re-tort or conversion plant site. Such a refinery is under-scale (50,000 bbl/day) and remote from chemical complexes that might make better use of by-products and hence provide higher (by-product) credits or other similar economic benefits.

The costs of upgrading the raw coal and shale liquids to high grade (transportation) fuels is shown below:

REFINERY COSTS FOR SYNTHETIC (RAW) LIQUIDS

(\$1980)

	<u>Cost Per Barrel</u>	<u>Cost Per Million BTU</u>
<u>Shale Oil</u> (Hydrotreat & Hydrocrack)	\$18.50	\$3.19
<u>Coal Liquids</u> (Hydrotreat)	\$18.29	\$4.02

The costs of refining synthetic liquids cannot truly be determined without specifying the product slate produced. The costs of refining a particular feedstock can vary depending upon the product cuts sought. The basis used above is not strictly comparable between the processes. It tends to slant the refinery approach to the type of slate that is favored by the feedstock - Light distillates in the case of shale oil, and gasolines and distillates in the case of coal liquids.

Exhibit 4-16 illustrates the potential variation.

These costs can be seen to vary dramatically if different product slates are sought. If the highest grade transportation fuels are maximized, to provide the highest degree of comparability with indirect liquids. The costs are as follows:

REFINERY SYNTHETIC UNITS TO 100% TRANSPORTATION FUEL

(\$ 1980)

	<u>Shale</u>		<u>Coal</u>	
	<u>\$/BBL</u>	<u>\$/MM BTU</u>	<u>\$/BBL</u>	<u>\$/MM BTU</u>
Raw Liquid	\$48.20	\$ 8.31	\$66.47	\$ 9.79
Upgrading	<u>18.50</u>	<u>N.A.</u>	<u>18.28</u>	<u>N.A.</u>
Total	\$66.70	\$11.50	\$84.75	\$14.61
Average Heat Content\ BBL	5.8 Million BTU		5.8 Million BTU	

EXHIBIT 4-16

PROCESS AND SLATE

(1980 \$)

<u>Feedstock</u>	<u>Hydrotreat & Hydrocrack</u>	<u>Severe Hydrotreat</u>	<u>Moderate Hydrotreat</u>
<u>Coal Liquids*</u>	Motor Gasoline	Motor Gasoline Plus	Motor Gasoline Plus
Product		Jet Fuel	#2 Fuel Oil
Slate	(100%)	(1/3 - 2/3)	(1/3 - 2/3)
Cost	\$20.70	\$18.29	\$12.55
<u>Shale Liquids</u>	<u>Hydrotreat & Hydrocrack</u>	<u>Hydrotreat-FCC</u>	<u>Coking Hydrotreat</u>
	3/4 - 1/4		(4/5 - 1/5)
Product Slate	Motor Gasoline Plus Jet Fuel	Jet Fuel Plus Motor Gasoline	Jet Fuel
Cost/BBL	\$18.50	\$17.00	\$16.00

*SRC-II

SOURCE: E. J. Bentz & Associates

4-31

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By comparison, indirect liquid (methanol to gasoline) costs are about \$78.00 per barrel; approximately in the middle of this range. The cost per million BTU's is lower for shale and coal liquids, refined to a transportation slate consisting of gasoline and distillate fuels (jet fuel and diesel oil). If direct liquids are refined to a 100% gasoline slate the costs would increase to \$87.17 per barrel or above \$19.00 per million BTU's.

Exhibit 4-17 graphically displays the finished fuels in a framework which relates the product quality to the finished fuel cost.

Exhibit 4-18 calculates the total cost of refining coal liquids. A 50,000 barrel per day refinery for coal liquids would cost between \$420 million and \$690 million. The lower case represents a moderate hydrotreatment plant producing #2 fuel oil and gasoline, the upper case represents a hydrotreatment and hydrocracking plant that produces 100% gasoline.

Instead of using other indirect measures of product value,¹⁸ we can use a cost based scale. The lighter fractions cost more to produce from both coal and shale, whether by direct or indirect means. By-product credits do not have to be assigned to determine the cost of a single cut liquid. Upgrading plant has been assigned to individual fractions so that the full cost of the beneficiated product cut is known. The costs of fully refining the product are developed incrementally by determining the cost of creating a 100% gasoline yield, and two subsequently lower grade mixtures.

The alternate product slate refinery costs of Exhibit 4-18 can be used to develop a measurement of the direct costs of products in a multi-product refinery run. The principal cost differences result from the increased capital (per unit of product yielded) and the increased consumption of hydrogen associated with higher grade product slates.

If we take the per barrel cost of producing a 100% gasoline slate, and assign it to the gasoline fraction of a mixed slate as the appropriate cost of that portion of the output, the remainder of the total cost divided by the number of barrels of the other product (jet fuel or #2 fuel oil) will give us the unit cost of the "secondary product".

Exhibit 4-19 shows this costing procedure for the slates presented for direct liquids refining in Exhibit 4-17.

By using this method, we are not artificially lowering the cost of gasoline production by assuming a market equilibrium price

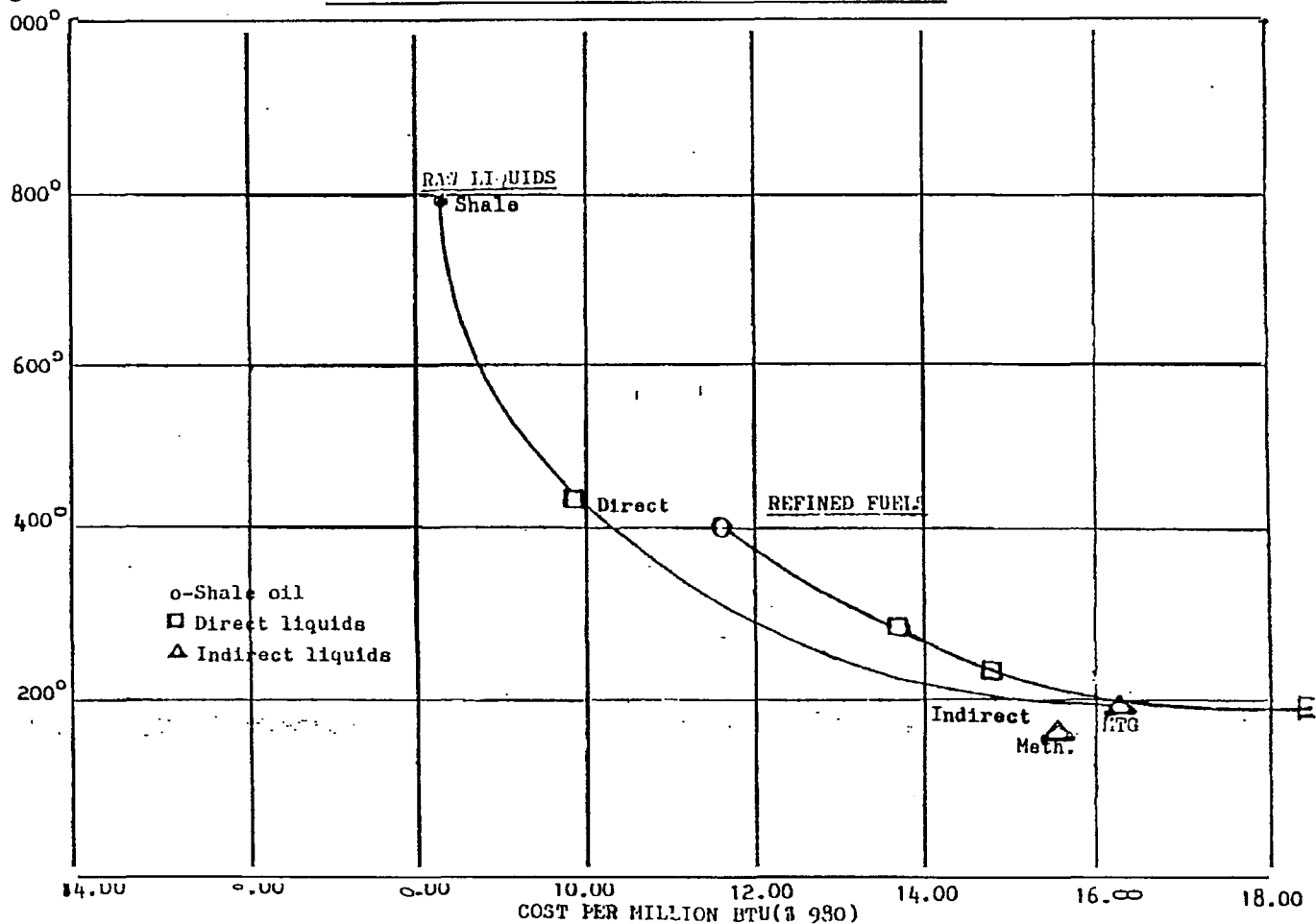
¹⁸Product value ratios are commonly used. They are of absolutely no meaning in a long-term and discontinuous supply context. The use of such ratios is a major violation of the most elementary laws or principles of economics as a measure of utility.

EXHIBIT 4-17

50% Boiling Point
Degrees F

SYNTHETIC FUELS PRODUCTION COSTS
BASED AVAILABLE COST ESTIMATES - REFINED LIQUID FUELS

4-33



SOURCE: E. J. Bentz & Associates, Springfield, Virginia

EXHIBIT 4-18

DIRECT LIQUIDS (SRC-II) REFINING
(50,000 BBL/Day) 1980 \$ per BBL.

<u>\$/BBL</u>	<u>Motor Gasoline (Hydrotreat Plus Hydrocrack)</u>	<u>Motor Gasoline Plus Jet Fuel (severe Hydrotreating)</u>	<u>Motor Gasoline Plus #2 Oil (Moderate Hydrotreat)</u>
Operating Labor	.244	.183	.183
Maintenance	.791	.669	.487
G&A	<u>.852 1.887</u>	<u>.670 1.522</u>	<u>.487 1.157</u>
Fuel	.183	.304	.122
Utilities	.183	.122	.122
Cat. & Chem.	.304	.365	.244
Hydrogen	<u>5.540 6.210</u>	<u>5.750 6.541</u>	<u>3.230 3.718</u>
Capital Recovery @ 30%/Yr.	<u>12.603</u>	<u>10.228</u>	<u>7.67</u>
	\$20.70	\$18.291	\$12.545

SOURCE: E. J. Bentz & Associates

4-34

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EXHIBIT 4-19

SRC II REFINED TO PRODUCT COSTS

		<u>Barrels/Day</u>		<u>Cost/BBL*</u>	<u>Total Daily Cost</u>
CASE I	Motor Gasoline	50,000	@	\$87.17	\$4,358,500
CASE II	Motor Gasoline	15,395	@	87.17	1,341,982
	Jet Fuel	<u>34,605</u>	@	<u>(83.69)**</u>	<u>(2,896,018)**</u>
	Total	50,000	@	\$84.76	\$4,238,000
CASE III	Motor Gasoline	16,995	@	\$87.17	\$1,481,454
	#2 Oil	<u>33,005</u>	@	<u>(74.74)**</u>	<u>(2,466,796)**</u>
	Total	50,000	@	\$78.965	\$3,948,250

Product Costs

Motor Gasoline = \$87.17/bbl (4.95) \$17.61/MM BTU
 Jet Fuel = \$83.69/bbl (5.67) \$14.36/MM BTU
 #2 Oil = \$74.74/bbl (5.825) \$12.83/MM BTU

*Cost from 4-17 plus 4-9.

**Values in parenthesis inferred from weighted average value of motor gasoline and total product.

SOURCE: E. J. Bentz & Associates

4-35

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for a lower grade (by) product. The method used is entirely an assignment of marginal cost to products. It would be more desirable to operate in a reverse manner, i.e., from the lowest product, assigning incremental costs to the higher product on a marginal basis. We, unfortunately, do not have a process estimate for a single slate of the lowest value product. The distillation range of all products is too broad to produce such an artificiality. Therefore we have begun with the marginal gasoline cost and assigned it as a by-product price to the lower value (mixed) slates, permitting us to infer the marginal cost of the lower grade products.

The results of this cost analysis are related to the costs of indirect liquefaction end products and shale products on Exhibit 4-20. The cost series increase as average distillation point is lowered. The average distillation point of most useful transportation fuels lies between 180° - 400 F, with the majority of the compounds contained lying within this range.

There is a persistence of the earlier noted relationship between product quality (as measured by average boiling point) and production costs of finished products. The relationship shows less than unitary cost increases per barrel, all greater than unitary cost increases per million BTU. The latter case is due to the generally lower heating value of the premier fuels that have increased hydrogen content. The increases in cost are about 7 1/2 cents per barrel of liquids for every degree fahrenheit that the boiling range is lowered.

Exhibit 4-21 is a flow sheet of a process (examined by Chevron Research) for hydrotreating and hydrocracking of direct coal liquid (SRC-II) whole oil to produce 100% motor gasoline product. This is the first case on Exhibit 4-16. Exhibits 4-22 and 4-23 illustrate the refining process used to upgrade the whole liquid to gasoline and jet fuel by severe hydrotreating alone, and to a lower quality slate of gasoline and heating oil created by less severe hydrotreating of direct (SRC-II) liquids.

The latter case is more comparable to an upgrading process.

4.7 TRANSPORTATION AND OTHER INFRASTRUCTURE COSTS (Reference 41)

Although we have differentiated between coal liquid's plant site upgrading facilities and finished product refineries, we have really not selected the site for refining. The upgrading must in most cases be done at the site of the coal liquids plant. The degree of upgrading we have embraced (Exhibit 4-15) is sufficient to permit the fuels to be used in as high a use as a combustion turbine, or transported without creating contamination or incompatible sediments.

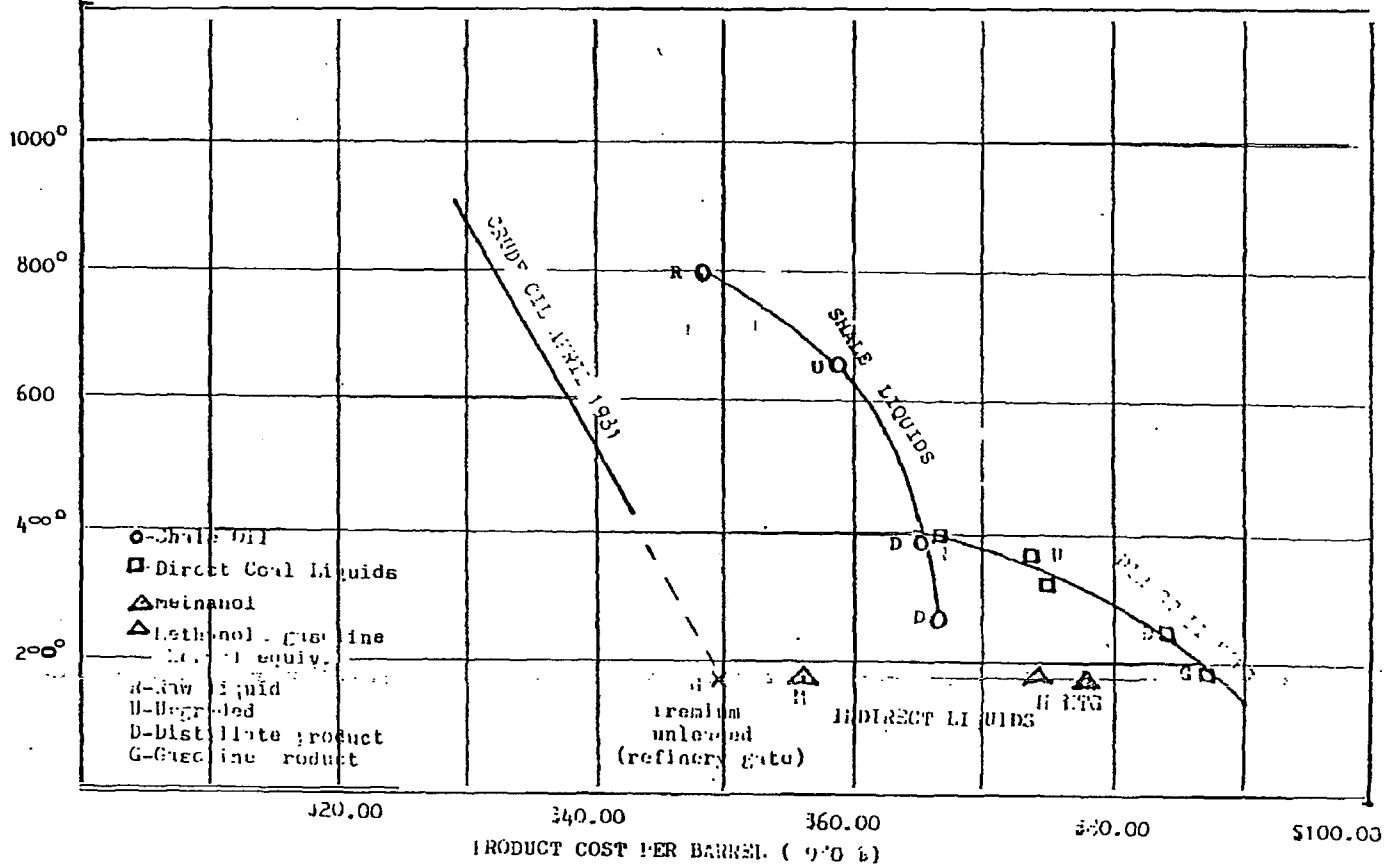
Transportation costs are directly related to the distance involved, and indirectly related to the quantity moved or flow rate.

EXHIBIT 4-20

SYNTHETIC FUEL LIQUIDS

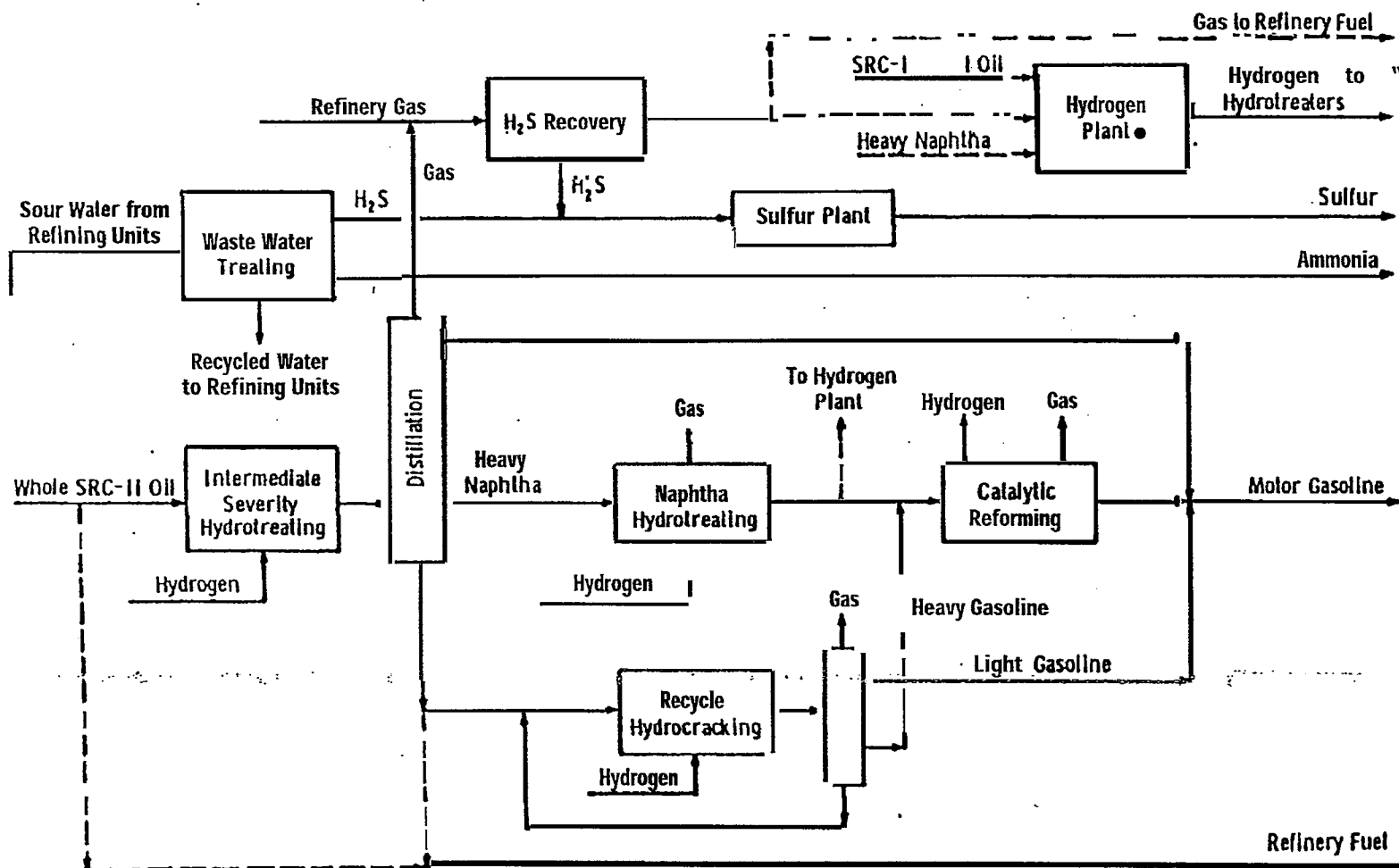
50% BOILING POINT
(DEGREES F)

4-37



SOURCE: E. J. Bentz & Associates, Springfield, Virginia

EXHIBIT 4-21: SCHEMATIC FLOW DIAGRAM
 REFINING OF SRC-11 OIL BY
 HYDROTREATING AND HYDROCRACKING - CASE I



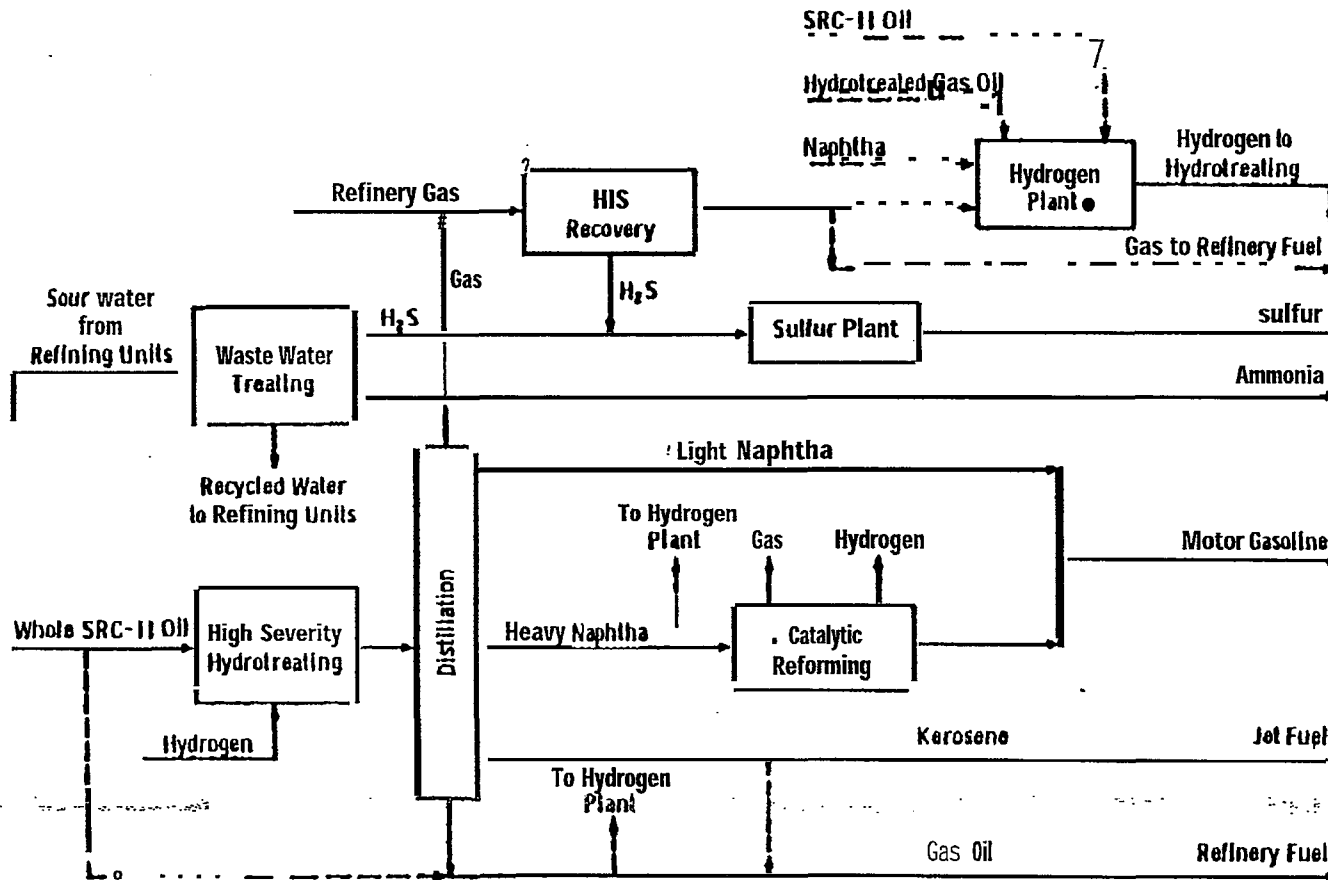
4-58

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* Steam reforming feeding gas and naphtha in Cases 4A, 4B, and 4D.
 Partial oxidation feeding SRC-11 oil in Case 4C.

SOURCE : Department of Energy

EXHIBIT 4-22: SCHEMATIC FLOW DIAGRAM
 REFINING OF SRC-II OIL BY
 HIGH SEVERITY HYDROTREATING - CASE II



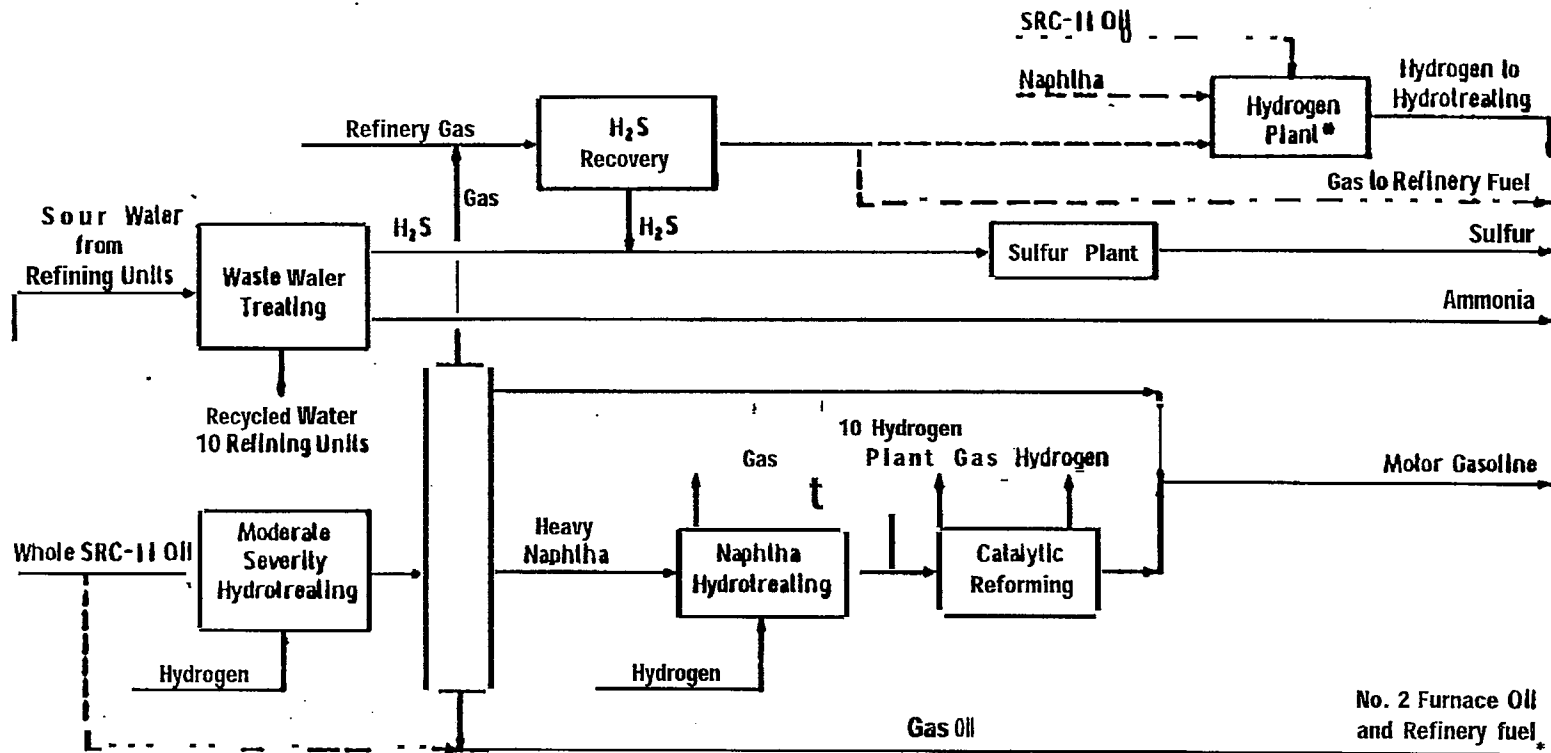
4-39

● Steam reforming feeding gas and naphtha in Cases IA, IB, and ID
 Partial oxidation feeding gas oil and SRC-I oil in Case IC.

SOURCE : Department of Energy

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EXHIBIT 4-23: SIMPLIFIED FLOW DIAGRAM
 REFINING OF SRC-11 OIL BY
 MODERATE SEVERITY HYDROTREATING - CASE I I I



* Steam reforming feeding gas and naphtha in Cases 5A, 5B, and 5D.
 Partial oxidation feeding SRC-11 oil in Case 5C.

SOURCE : Department of Energy

4-40

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We cannot visualize any other form of transportation for these upgraded liquids, or for further refined products except by pipeline. The daily volume required to support a 6" or 8" pipeline is approximately the size of one or two 50,000 bbl./day plants. Considering the geographical concentration of coal and shale deposits it is not difficult to visualize a mining-conversion center adequate to support either:

. An upgraded liquids pipeline to a refining center

or

. A product pipeline to major pipeline junctions or product distribution terminals

The general location of all coal and shale resources is such that deep draft water transportation does not figure prominently in synfuels distribution patterns.

Without siting specific plants and conducting the refinery trade-offs - which would have to be done in context with both the balance of foreign and domestic petroleum supplies and the slate of (regional) demand for all liquids - we cannot develop very meaningful insights into either the operating (product) costs of transportation and distribution, or the capital requirements.

We will have to make some nominal assumptions and then establish unitary relationships. The future energy transportation patterns and infrastructure requirements are impossible to determine without a specific scenario. We shall briefly examine a *cases:

- . Pipelining from Southern Illinois to Houston of syncrudes.
- . Pipelining from Wyoming to St. Louis
- . Pipelining from Western Colorado to L.A. of shale oil.

Southern Illinois to Houston

Raw Liquids
(upgraded) **33c/MM BTU**

Western Colorado to L.A.

Shale Liquids **40\$/MM BTU**

Wyomina to St. Louis

Raw Liquids
m ' * ' **30 \$/MM! BTU**

Methanol **68c/MM BTU**

MTG - Gasoline **37\$/MM BTU**

The additional capital investment required for synthetic fuel transportation is highly speculative to a greater degree. There

is a great deal of existing product and crude liquid pipeline as well as gas pipeline in place, that can equally serve the synthetic fuels industry. In all cases the pipelines are connected to either markets or distribution terminals at the delivery end. In most cases, the input end is originally either at a major refinery (and production) location or at a port location. The refinery connection argues for upgrading of liquids (coal and shale) at mine mouth conversion plant locations, and transportation to the existing refinery districts for product finishing. Such a general pattern would involve the construction of a minimum number of new "crude" synfuel pipelines from coal fields to refining districts.

We assume that the ultimate conditions would lead to the construction of several large diameter pipelines in such a pattern.

Methanol, which does not require refining, obviously will move in different patterns from coal field to the major terminals and markets.

Pipelines of that size (10-12") would cost an average of \$100,000 per mile, considering material, labor, and right of way and other expenses. Terrain would influence the cost, generally increasing construction costs but reducing right of way costs in some cases by an equivalent amount. 20" or greater diameter pipelines would cost \$250,000/mile.

A total construction budget of 50,000 miles of new pipeline of 12" diameter to 20" diameter would cost between \$5 billion and \$12 billion.

4.8 ADDENDUM TO CHAPTER 4: BASIS FOR COST ASSUMPTIONS

1) Basic Conversion Plant (ESCOE)

● Capital Costs

Year: Mid (June-July) 1979 dollars
 Scale: 25,000 tons of coal input
 Base Plant to installed battery limits: 1.63
 Contingency: 10%
 Scaling exponential rule: $C_2 = G$
 C_1

$\lambda = .65$ for vessel size
 $\lambda = .9$ with trains

Outlay of Capital: instantaneous plant

● Revisions to Capital Assumptions in This Report

Year: Mid 1980 (June-July)
 Scale: 50,000 bbl/day liquids output
 Plant to Battery Limits: 1.73
 Contingency: 20%
 Scaling: Linear
 Outlay of Capital: Instantaneous plant

● Operating costs

Coal Feedstock: \$30/ton (delivered)
 Coal: Illinois #6
 Catalysts and Chemicals and Operating Supplies:
 at cost for amounts proscribed by process
 designer's material balance.

● Labor Cost

	#	Rate/Hr
Plant Operators	120	\$ 10.00
Operating Supervisors	25	15.00
Maintenance Labor	150	12.00
Maintenance Labor Supervisors	30	16.00
Administration	30	11.00
Total	355	@ \$11.79/hr avg.

Fringes @ 35% --changed to 40% = total labor rate
 of \$16.50/hr

Maintenance Cost (Materials & Contracts)

3% of total plant capital cost

G & A

Local taxes and insurance, 5% capital cost
changed to total G&A - 5% capital cost

Capital Charge Rate

ESCOE basis not used. 30% of capital used as
recovery rate (as per guidance of OTA staff) .

On-Stream Rate

90%--328.5 days/year

2. Assumptions for Product Upgrading

● Capital

Basis --Instantaneous Plant, mid-1980 dollars
On-stream factor 90% 328.5 stream days.

● Hydrotreater

capitalized for each separate product stream.

● Hydrogen Feedstock Plant Capital

Not included, only cost feedstock "across the
fence" from the plant complex.

● Hydrogen Reformer or manufacturing plant capital
included

● Battery Limits

Includes hydrotreaters, waste water treatment,
sulphur plants (commercial grade)

● Contingency

General -- 25%
Battery Limits--15%
Engineer---4% of investment capitalized
Working Capital--45 days receivables; 30 day
chemicals catalysts; 30 day feedstocks

● Operating costs

Hydrogen Feedstock: Syngas @ \$6.74/mmbtu
raw gas liquids @ \$6=50/mmbtu
includes recovery of production
plant capital.

Hydrogen Pressure: 500 PSIG for SRC light (naptha)
product--2000 PSIG all other
cases.

Plant Size: 20,000 bbl/day upgraded to
50,000 bbl/day for each product
cut

● Royalties

500 PSIG Hydrotreating -o-
1500 PSIG Hydrotreating Fixed Bed \$30/bst feed
Sulphur plant -o-

Waste Water
Initial project \$75,000
First 5,000 units \$14.70/unit
Next 5000-25,000 units \$7.35/unit
Next 25,000 + units \$5.25/unit

● Sales Tax

5% of equipment cost

● Maintenance

4% of depreciated capital/year

● Operating Labor

\$11.00/hr

● Labor Burden

45%

● Administrative and Support Labor

30% of operations and maintenance labor

● G & A

60% of operations and maintenance labor plus
property-tax of 2-1/2% of plant investment

- Utilities

- Fuel \$4/mmbtu
 - Steam \$3.50/1000 lbs
 - Electricity 4c/kwh
 - Water (make-up) 40c/1000 gal

- Hydrogen Bleed was assumed to be:

- 50 SCP/bbl @ 500 PSIG
 - 100 SCP/bbl @ 2000 PSIG

- By-product Credits

- Ammonia (anhydrous) \$100/ton
 - Hydrogen and Hydrocarbon off gasses (C₁-C₄) \$4/mmbtu (\$1.30/MSCF)

3. Refining Cost Assumptions (Chevron Basis)

- 1980 costs: Instantaneous plant (first quarter adjusted to June/July)
- Mid-Continent Location
- Cost correlations based on actual experience of Standard Oil of California, 1960-1970s adjusted for:
 - Lower field productivity
 - Increased safety
 - Improved efficiency and reliability
 - Additional energy conservation
 - Stricter environmental regulations
- 10% Contingency
- Utilities
 - Water 30c/1000 gal
 - Boiler fuel, coal or refinery fuel power 3\$/kwh
- Maintenance
 - 2-1/2%/yr of both on-plant and off-plant facility investment
- G&A
 - Property taxes @ 2 1/2% of both on-plant and off-plant/yr
- Labor
 - Operating-- \$110,000 per shift position/hr (\$18.30/hr including fringes)
 - Support Labor (Administrative, security, technician) 65% of Direct Labor

Rev 0
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Medicine Bow Fuel & Power

40 CFR §60.11(d): At all times, including periods of startup, shutdown, and malfunction, owners and operators shall, to the extent practicable, maintain and operate any affected facility including associated air pollution control equipment in a manner consistent with good air pollution control practice for minimizing emissions. Determination of whether acceptable operating and maintenance procedures are being used will be based on information available to the Administrator which may include, but is not limited to, monitoring results, opacity observations, review of operating and maintenance procedures, and inspection of the source.

The goal of this Plan is to provide guidelines and suggestions for steps that will minimize air emissions during startup and shutdown periods, in accordance with Clean Air Act permits and regulations, including the provisions from 40 CFR 60 as cited above.

Specific startup and shutdown operating procedures for all process units in the Plant shall incorporate the elements of this Plan to the greatest extent possible.

Flaring Associated with Startup – General Comments

- Commission all downstream equipment and prepare them for operation prior to gasifier startup. This will include preparation of the:
 1. Low Temperature Gas Cleanup (LTGC),
 2. Sour Water Stripper,
 3. Acid Gas Removal (AGR),
 4. Sulfur Recovery Unit (SRU) - Claus Plant,
 5. CO₂ compression, and
 6. Methanol synthesis loop.
- Preparation will include completion of commissioning activities and final signoff, establishment of normal operating levels for fluids, preheating of required components, and start of circulating pumps as necessary.

Flaring Associated with Startup – Activities Following Gasifier Startup

Once a gasifier is started up certain conditions must be met prior to introducing syngas to subsequent stages. These conditions include:

- Gasifier
 - One gasifier will be started at a time at 50% design flow rate. Subsequent gasifiers will not be started until the downstream equipment is ready to receive the increase in syngas volume.
 - After light off a leak check of gasifier piping and components is required.
 - A low pressure and normal operating pressure check are required.
 - Raw syngas will be diverted to flare until after checks are complete. At this stage pressure can be bled into downstream piping to equalize pressures and then the control valve can be fully opened and placed in automatic control.

DRAFT (Rev 0)
Startup/Shutdown Emission Minimization Plan
Medicine Bow Fuel & Power

2

- The amount of syngas sent downstream will be determined by the startup and status of downstream units.
- Start-up flaring will be at a reduced rate due to a planned slow ramp up of the plant.
- LTGC
 - Leak checks are required after pressurization, but not to delay input to the AGR system.
 - This stage includes several steam generators needed to ensure the syngas temperature is in spec for downstream components. Failure to cool down the syngas can result in a high temperature scenario requiring flare to avoid damage to downstream equipment and catalysts.
 - The syngas temperature must be monitored as the system heats up to prevent a high temperature trip. Temperature setpoints to be defined by AGR vendor and by catalyst vendors for COS and Sour Shift catalysts.
- Sour Water Stripper
 - The sour water unit will send low pressure sour gas to the Claus plant for conversion of ammonia and H₂S to N₂, H₂O, and SO₂. Base case is to flare this stream during startup until the SRU is started up. The SRU can start operations at approximately 20% design conditions.
- AGR
 - The AGR will be slowly ramped up at an estimated 10% of design syngas flow per hour.
 - Syngas temperature must be maintained below AGR vendor specifications.
 - The clean high pressure syngas must be vented to flare until the total sulfur in the syngas comes into the specification of less than 0.5 ppmv.
 - Start-up flaring will be at a reduced rate due to slow ramp up of plant.
- Claus Plant
 - When the acid gas reaches approximately 40% H₂S content it can be sent from the AGR to the SRU. Prior to this we will assume the acid gas is flared.
 - Start-up flaring will be at a reduced rate due to slow ramp up of plant.
- Methanol Synthesis
 - No syngas can be sent to the Methanol synthesis loop until sulfur is in spec. **Syngas sulfur content must be less than 0.5 ppmv prior to sending to methanol synthesis.**
 - If CO₂ is out of spec (>2% vol) for several hours it will result in high water content in the methanol which is not acceptable.
 - **Syngas flow rate must be at least 50% of design flow rate prior to being sent to methanol synthesis to prevent compressor surge. This rate will be reviewed and verified during compressor design and surge protection design.**
 - After the Methanol step the effluents are primarily low sulfur fuel gases sent to the power block and liquid methanol sent to storage or MTG. No further flaring events as part of startup are expected.

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Medicine Bow Fuel & Power

3

Venting Associated with Startup

- CO₂ Capture
 - CO₂ produced from AGR will need to be vented until sufficient flow is produced to start the compressors. This flow rate is expected to be 25% of design flow rate assuming two compressor trains and a 50% turndown capacity. *This will require confirmation from compressor vendor during FEED engineering.*
 - Start-up venting will be at a reduced rate due to slow ramp up of plant.
 - If during startup export of CO₂ is not feasible then CO₂ will continue to be vented.

- Gasifier heaters
 - Initially all five heaters will be online. Heaters will be started shortly after the refractory is installed to cure the refractory. After refractory cure, the heaters will need to remain in operation to prevent moisture accumulation; otherwise another multiday heater dryout session will be required prior to startup.
 - Medicine Bow will attempt to startup as soon as possible after refractory cure is complete to minimize heater operations. This is the basis of the current plan to commission units from the end of the process to the beginning to ensure that as soon as the gasifiers are commissioned, the plant will be ready to startup and receive syngas. This plan is dependent on the construction and commissioning schedule and a situation may develop where light off is delayed after cure is complete. The time of this delay will determine if the heaters will remain on or be shutdown.
 - As each gasifier is prepared for startup the heaters will be turned off and removed. After full startup is complete, only one heater will be in operation on the spare gasifier.

- MTG heaters
 - These heaters will be brought on line when the unit is prepared to receive methanol and be operated per design.

Startup/Shutdown Emission Minimization Plan Medicine Bow Fuel & Power

- Power block
 - The ASUs, which are the major power load for the plant, will be started several days in advance of the gasifier light-off to establish required temperatures in the cold box to generate purified oxygen. Two turbines with heat recovery steam turbine power will be required to start up both ASUs. If the steam turbine is not available, then all three gas turbines at reduced load will be required to startup the ASUs.
 - During plant startup most process units will begin to draw power in preparation for gasifier light off. The main exceptions are the CO₂ Compressors, Methanol Synthesis compressor, and MTG compressor units. All three gas turbines with heat recovery steam power are required to support the plant as it is prepared for full start-up.

- Fugitive emissions
 - Fugitive emissions will be at a reduced rate until Methanol and gasoline are synthesized
 - Tank emissions will be at a reduced rate initially as storage tanks are filled.

- Aux boiler
 - The boiler will be in operation during startup. At a minimum it will be turned down and floated on the system if the heat recovery steam generators are able to support plant steam requirements. *If more steam is required as defined in the FEED, then the aux boiler may be operated at its maximum rate.* After syngas is routed to methanol and the startup steam loads are reduced and process steam is available, the auxiliary boiler can be reduced to minimum.

- Flare pilots
 - Pilots will be lit as part of preparation for gasifier light off.