ALTERNATIVE FUELS FOR AUTOMOTIVE TRANSPORTATION - A FEASIBILITY STUDY VOLUME III - APPENDICES

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VOLUME III - APPENDICES

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PREFACE

This report is the result of a research team effort at the Institute of Gas Technology. In addition to the authors, the major contributors to the study were J. Fore, P. Ketels, W. Kephart, and K. Vyas.

This report consists of three volumes:

Volume I — Executive Summary

Volume II — Technical Section

Volume III — Appendices.

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SUMMARY

This volume contains two appendices:

A.. Properties of Potential Alternative Fuels for Automotive Transportation.

B. Detailed Process Descriptions and Economics for Candidate Fuels From Coal and Oil Shale

As a summary of the complete data sheets given in Appendix A, the pertinent properties of 18 fuels are summarized in Table 1. In the case of coal, a relatively clean, solvent-refined coal would be required for automotive use, but for the sake of characterization, an Illinois coal (raw) is described here. [The technical section (Volume II) of this report considers solvent-refined coal.] Hydrazine is included because it is a fuel for fuel cells; direct or flame combustion in a heat engine is not implied. Many vegetable oils are (theoretically) useful as engine fuels (in external combustion, heat-engine cycles), so we have tabulated the properties of cottonseed oil as an example because there were sufficient data for characterization.

Appendix B presents detailed process descriptions for gasoline and distillate oils from coal, gasoline and distillate oils from oil shale, methyl alcohol from coal, and substitute natural gas (SNG) from coal. Either these processes are at or near commercialization, or sufficient data on the components of these processes have been published to allow characterization and reasonable estimates of economics. The economics have been calculated by using discounted cash flow (DCF) financing in accordance with the method contained in The Supply-Technical Advisory Task Force — Synthetic Gas-Coal.[1] The described processes are "pattern" processes for fuel synthesis, and other processes would be equally or less favored. Certain portions of the process descriptions and calculations in Appendix B have been derived from IGT in-house source material that has been made available to this as well as to other projects. This information includes personal communications that cannot be referenced. The synthesized fuels are candidates for use as alternative fuels for automotive transportation, but they are not necessarily the selected (chosen or recommended)

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[1] Synthetic Gas-Coal Task Force, The Supply-Technical Advisory Task Force—Synthetic Gas-Coal. Prepared for the Supply-Technical Advisory Committee, National Gas Survey, Federal Power Commission, April 1973.

Table 1. SELECTED PROPERTIES OF 18 FUELS

Fuel	Chemical Formula	Molecular Weight	Melting Point, °F	Boiling Point, °F	Density, lbm/cu ft (77°F, 1 atm)	Net Heating Value Vapor, Btu/cu ft	Net Heating Value Liquid, Btu/lbm	Flammability Limits, % in air	Ignition Temp, °F
Acetylene	C_2H_2	26.04	−114	−119	0.070	1448	20,776	2.5-80	581
Ammonia	NH_3	17.03	−108	−28	0.045	365	8,001	15.5-26.6	1204
Carbon Monoxide	CO	28.01	−341	−313	0.074	322	4,347	12.5-74.2	1128
Coal	Mix	--	--	--	84	--	--	--	1100
Diesel Oil	Mix	165-240	--	325-650	53.4	--	18,480	0.7-5.0	490
Ethanol	C_2H_5OH	46.07	−179	600-1000	49.0	1451	11,929	3.3-18.9	738
Fuel Oil (Heavy)	Mix	300-1000	--	100-400	60.6	--	17,160	--	765
Gasoline	Mix	55-145	--	--	45.5	--	19,291	1.4-7.6	495
Hydrogen	H_2	2.02	−431	−423	0.0053	275	51,623	4-74	1065
Hydrazine	N_2H_4	32.05	34.5	236	63.1	--	6,500	4.7-100	74-518
Kerosene	Mix	150-230	--	300-480	50.6	--	19,092	0.7-5.0	491
Methane	CH_4	16.04	−296	−258	0.052	913	21,520	5.0-15.0	1170
Methanol	CH_3OH	32.04	−144	148	49.7	768 ·	9,078	6.0-36.5	878
Methylamine	CH_3NH_2	31.03	−134	20	0.087	1089	12,855	4.9-20.7	806
Naphtha	Mix	84-170	--	--	48	8461	18,864	0.9-6.0	450-530
Propane	C_3H_8	44.09	−306	−44	0.110	2385	19,944	2.1-10.1	808
LPG	Mix	30-60	--	−50	0.117	2399	20,514	2.4-9.6	920-1020
Vegetable (Cottonseed) Oil	Mix	--	20-30	338+	56.9	--	16,113	--	650

Fuel	Heat of Vaporization at Boiling Point, Btu/lbm	Flash Point, °F	Least Detectable Odor, ppm	Least Amount Causing Eye Irritation, ppm	Least Amount Causing Throat Irritation, ppm	Maximum Allowable for 8-hr Exposure, ppm	Octane Number Research Method	Octane Number Motor Method	Cetane Number
Acetylene	264	--	No Odor	No Irr	No Irr	d	40	--	--
Ammonia	584	--	20	40	400	100	111	--	--
Carbon Monoxide	--	--	No Odor	No Irr	No Irr	100	130+	--	--
Coal	--	--	--	--	--	-- c	--	--	--
Diesel Oil	155	100	25-50	--	--	500	Low	Low	40-70
Ethanol	370	55	10	5000	5000	1000	106	89	--
Fuel Oil (Heavy)	85	150	--	?	?	--	<0	<0	--
Gasoline	130	−36 to −50	10-50 a	?	?	500	92-100	84-92	18
Hydrogen	--	--	No Odor	No Irr	No Irr	d	130+	--	--
Hydrazine	561	100	--	b	--	l	--	--	--
Kerosene	140	100	25-100 a	500-1000	500-1000	500	<0	<0	40-65
Methane	219	--	No Odor	No Irr	No Irr	d	130	--	--
Methanol	473	52	100	?	?	200	106	92	--
Methylamine	340	0	0.02	10-50	10-50	10	--	--	--
Naphtha	145	20-50	10-50 a	?	?	500	60-70	50-60	--
Propane	150	−156	? a	No Irr	No Irr	30,000-50,000	100	97	--
LPG	180	−155	1	--	--	10,000	108-111	95-98	--
Vegetable (Cottonseed) Oil	--	486	--	--	--	--	--	--	--

a Dependent on impurities, primarily sulfur or aromatics.

b Contact with more than a 5% solution causes severe eye injury.

c Dust: 50 X 10⁶ particles/cu ft.

d Simple asphyxiant above about 100,000 ppm.

D-94-1708

alternatives. The selected fuels depend on the needs for supplemental fuel, as shown by an energy demand and supply projection, and on the application of a fuel selection procedure, as described in the technical section. Table 2 summarizes the pattern processes and the corresponding fuel synthesis costs (as of about December 1973).

Table 2. PATTERN SYNTHESIS PROCESSES AND FUEL COSTS

Raw Material	Synthesized Fuel	Pattern Process	Production Cost (DCF)	
			Volume Units	Energy Units*
Coal	Gasoline	Consol synthetic fuel (CSF) plus refining with hydrocracking	$0.33/gal	$2.81/$10^6$ Btu
Coal	Gasoline and distillate oils	Consol synthetic fuel (CSF) plus refining with catalytic cracking	$0.31/gal	$2.51/$10^6$ Btu
Oil Shale	Gasoline and distillate oils	Gas Combustion Process (Bureau of Mines) plus hydrotreating and refining	$0.25/gal	$2.05/$10^6$ Btu
Coal	Methanol	Koppers-Totzek gasifier and ICI synthesis	$0.23/gal	$3.88/$10^6$ Btu
Coal	SNG (CH_4)	Lurgi gasifier with methanation	$1.84/$10^3$ SCF*	$2.14/$10^6$ Btu

* Based on the low heating value of the fuel.   A-94-1709

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APPENDIX A. Properties of Potential Alternative Fuels for Alternative Transportation

Acetylene

Ammonia

Carbon Monoxide

Coal (and Solvent-Refined Coal)

Diesel Oil

Ethanol

Fuel (Heavy)

Gasoline

Hydrazine

Hydrogen

Kerosene

Methane

Methanol

Methylamine

Naphtha

Propane

LPG

Vegetable (Cottonseed) Oil

ACETYLENE

Chemical Formula	C_2H_2	
Molecular Weight	26.036	

	English Units	Metric (SI) Units
Melting Point	$-114^\circ F$	$-81^\circ C$
Boiling Point	$-119^\circ F$	$-83^\circ C*$
Density		
Vapor	0.06971 lb/cu ft	1.117 kg/cu m
Liquid	8.276 lb/cu ft	132.582 kg/cu m
Specific Gravity		
Vapor	0.9107	0.9107
Liquid		
Heating Value, Vapor		
Volumetric Gross	1,499 Btu/cu ft	5.5842×10^4 kJ/cu m
Volumetric Net	1,448 Btu/cu ft	5.394×10^4 kJ/cu m
Weight Gross	21,500 Btu/lb	5.00×10^4 kJ/kg
Weight Net	20,776 Btu/lb	4.8321×10^4 kJ/kg
Heating Value, Liquid		
Volumetric Gross	177,934 Btu/cu ft	662.854×10^4 kJ/cu m
Volumetric Net	171,922 Btu/cu ft	640.458×10^4 kJ/cu m
Weight Gross	21,500 Btu/lb	5.000×10^4 kJ/kg
Weight Net	20,776 Btu/lb	4.832×10^4 kJ/kg
Air for Combustion, Vapor	———air component per unit of fuel———	
O_2 Volumetric	2.5 cu ft/cu ft	2.5 cu m/cu m
N_2 Volumetric	9.411 cu ft/cu ft	9.411 cu m/cu m
Air Volumetric	11.911 cu ft/cu ft	11.911 cu m/cu m
O_2 Weight	3.0731 lb/lb	3.0731 kg/kg
N_2 Weight		
Air Weight	10.224 lb/lb	10.224 kg/kg
Air for Combustion, Liquid Fuel	13.297 lb/lb	13.297 kg/kg
Products of Combustion, Vapor	———unit of product per unit of fuel———	
CO_2 Volumetric	2 cu ft/cu ft	2 cu m/cu m
H_2O Volumetric	1 cu ft/cu ft	1 cu m/cu m
N_2 Volumetric	9.411 cu ft/cu ft	9.411 cu m/cu m
CO_2 Weight	3.381 lb/lb	3.381 kg/kg
H_2O Weight	0.692 lb/lb	0.692 kg/kg
N_2 Weight	10.224 lb/lb	10.224 kg/kg
SO_2 Weight		

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ACETYLENE, Cont.

	English Units	Metric (SI) Units
Flammability Limits	2.50-80.0%	
Flash Point		
Ignition Temperature	581°F	305°C
Heat of Vaporization at Boiling Point	264 Btu/lb	
Octane Number Research Method Motor Method	40	
Cetane Number		

Toxicity
    Least Detectable Odor   No odor
    Least Amount Causing Eye Irritation   No irritation
    Least Amount Causing Throat Irritation   No irritation
    Least Amount Causing Coughing   No coughing
    Maximum Allowable for Prolonged Exposure   Simple asphyxiant in high concn
    Maximum Allowable for Short Exposure (0.5 hr)   100 mg/liter
    Dangerous for Short Exposure (0.5 hr)   100 mg/liter

Comments

Normal transportation is by furnished air cylinders packed with asbestos
fibers and dissolved in acetone at about 250 psi pressure. Also generated
onsite from reaction between calcium carbide and water in self-regulating
generators similar in principle to "Kipp" generators. As a gas under
pressure, acetylene may decompose violently. The free gas should never
be used, transported, or stored outside of its special cylinders at pressures
in excess of 2 atmospheres.

*Sublimes at 1 atmosphere.

AMMONIA

Chemical Formula	NH₃	
Molecular Weight	17.031	

	English Units	Metric (SI) Units
Melting Point	$-108^\circ F$	$-77.7^\circ C$
Boiling Point	$-28^\circ F$	$-33.35^\circ C$
Density		
Vapor	0.0456 lb/cu ft	0.721 kg/cu m
Liquid	48.1 lb/cu ft	77.056 kg/cu m
Specific Gravity		
Vapor	0.5961	0.5961
Liquid	0.674	0.674
Heating Value, Vapor		
Volumetric Gross	441.1 Btu/cu ft	1.6432×10^4 kJ/cu m
Volumetric Net	365.1 Btu/cu ft	1.3068×10^4 kJ/cu m
Weight Gross	9668 Btu/lb	2.2249×10^4 kJ/kg
Weight Net	8001 Btu/lb	1.8609×10^4 kJ/kg
Heating Value, Liquid		
Volumetric Gross	465.031 Btu/cu ft	1721×10^4 kJ/cu m
Volumetric Net	384.848 Btu/cu ft	1434×10^4 kJ/cu m
Weight Gross	9668 Btu/lb	2.2249×10^4 kJ/kg
Weight Net	8001 Btu/lb	1.8609×10^4 kJ/kg
Air for Combustion, Vapor	———air component per unit of fuel———	
O_2 Volumetric	0.75 cu ft/cu ft	0.75 cu m/cu m
N_2 Volumetric	2.823 cu ft/cu ft	2.823 cu m/cu m
Air Volumetric	3.579 cu ft/cu ft	3.579 cu m/cu m
O_2 Weight	1.409 lb/lb	1.409 kg/kg
N_2 Weight	4.688 lb/lb	4.688 kg/kg
Air Weight	6.097 lb/lb	6.097 kg/kg
Air for Combustion, Liquid	———air component per unit of fuel———	
O_2 Volumetric	801.7 cu ft/cu ft	801.7 cu m/cu m
N_2 Volumetric	3017.5 cu ft/cu ft	3017.5 cu m/cu m
Air Volumetric	3819.2 cu ft/cu ft	3819.2 cu m/cu m
O_2 Weight	1.409 lb/lb	1.409 kg/kg
N_2 Weight	4.688 lb/lb	4.688 kg/kg
Air Weight	6.097 lb/lb	6.097 kg/kg
Products of Combustion, Vapor	———unit of product per unit of fuel———	
CO_2 Volumetric	--	--
H_2O Volumetric	1.5 cu ft/cu ft	1.5 cu m/cu m
N_2 Volumetric	3.323 cu ft/cu ft	3.323 cu m/cu m
CO_2 Weight	--	--
H_2O Weight	1.587 lb/lb	1.587 kg/kg
N_2 Weight	5.511 lb/lb	5.511 kg/kg
SO_2 Weight		

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AMMONIA, Cont.

	English Units	Metric (SI) Units
Products of Combustion, Liquid	———unit of product per unit of fuel———	
CO_2 Volumetric		
H_2O Volumetric	1603.3 cu ft/cu ft	1603.3 cu m/cu m
SO_2 Volumetric		
N_2 Volumetric	3551.9 cu ft/cu ft	3551.9 cu m/cu m
CO_2 Weight	--	--
H_2O Weight	1.587 lb/lb	1.587 kg/kg
N_2 Weight	5.111 lb/lb	5.111 kg/kg
SO_2 Weight		
Ash Weight		

Flammability Limits	15.50-26.60%	
Flash Point		
Ignition Temperature	$1204^{\circ}F$	$651^{\circ}C$
Heat of Vaporization at Boiling Point	584.4 Btu/lb	1356 kJ/kg

Octane Number
  Research Method   >111
  Motor Method

Cetane Number

Toxicity
  Least Detectable Odor   20 ppm
  Least Amount Causing Eye Irritation   40 ppm
  Least Amount Causing Throat Irritation   400 ppm
  Least Amount Causing Coughing   400 ppm
  Maximum Allowable for Prolonged Exposure   0.076 mg/liter 100 ppm
  Maximum Allowable for Short Exposure (0.5 hr)   700 ppm
  Dangerous for Short Exposure (0.5 hr)   1700 ppm

Comments

Normal transportation is in cylinders as a liquid, in tank cars, trucks, and pipelines. At concentrations of 100 ppm in air, there is noticeable irritation to the eyes and nasal passages.

CARBON MONOXIDE

Chemical Formula	CO	
Molecular Weight	28.01	

	English Units	Metric (SI) Units
Melting Point	-340.6°F	-207°C
Boiling Point	-312.7°F	-191.5°C
Density		
Vapor	0.0740 lb/cu ft	1.186 kg/cu m
Liquid	6.268 lb/cu ft	100.41 kg/cu m
Specific Gravity		
Vapor	0.9672	0.9672
Liquid		
Heating Value, Vapor		
Volumetric Gross	321.8 Btu/cu ft	1.1988×10^4 kJ/cu m
Volumetric Net	321.8 Btu/cu ft	1.1988×10^4 kJ/cu m
Weight Gross	4347 Btu/lb	1.0110×10^4 kJ/kg
Weight Net	4347 Btu/lb	1.0110×10^4 kJ/kg
Heating Value, Liquid		
Volumetric Gross	27,264 Btu/cu ft	6.341 kJ/kg
Volumetric Net	27,264 Btu/cu ft	6.341 kJ/kg
Weight Gross	4347 Btu/lb	1.011×10^4 kJ/kg
Weight Net	4347 Btu/lb	1.011×10^4 kJ/kg
Air for Combustion, Vapor	———air component per unit of fuel———	
O_2 Volumetric	0.5 cu ft/cu ft	0.5 cu m/cu m
N_2 Volumetric	1.882 cu ft/cu ft	1.882 cu m/cu m
Air Volumetric	2.382 cu ft/cu ft	2.382 cu m/cu m
O_2 Weight	0.571 lb/lb	0.571 kg/kg
N_2 Weight	1.900 lb/lb	1.900 kg/kg
Air Weight	2.471 lb/lb	2.471 kg/kg
Air for Combustion, Liquid	———air component per unit of fuel———	
O_2 Volumetric	42.30 cu ft/cu ft	42.30 cu m/cu m
N_2 Volumetric	160.1 cu ft/cu ft	160.1 cu m/cu m
Air Volumetric	202.4 cu ft/cu ft	202.4 cu m/cu m
O_2 Weight	0.571 lb/lb	0.571 kg/kg
N_2 Weight	1.900 lb/lb	1.900 kg/kg
Air Weight	2.471 lb/lb	2.471 kg/kg
Products of Combustion, Vapor	———unit of product per unit of fuel———	
CO_2 Volumetric	1.0 cu ft/cu ft	1.0 cu m/cu m
H_2O Volumetric	--	--
N_2 Volumetric	1.882 cu ft/cu ft	1.882 cu m/cu m
CO_2 Weight	1.571 lb/lb	1.571 kg/kg
H_2O Weight	--	--
N_2 Weight	1.900 lb/lb	1.900 kg/kg
SO_2 Weight		

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CARBON MONOXIDE, Cont.

	English Units	Metric (SI) Units
Products of Combustion, Liquid	——unit of product per unit of fuel——	
CO_2 Volumetric	53.45 cu ft/cu ft	53.45 cu m/cu m
H_2O Volumetric	--	
SO_2 Volumetric		
N_2 Volumetric	160.1 cu ft/cu ft	160.1 cu m/cu m
CO_2 Weight	1.571 lb/lb	1.571 kg/kg
H_2O Weight	--	
N_2 Weight	1.900 lb/lb	1.900 kg/kg
SO_2 Weight		
Ash Weight		
Flammability Limits	12.5-74.2%	
Flash Point		
Ignition Temperature	$1128^\circ F$	$608.9^\circ C$
Heat of Vaporization		
Octane Number		
Research Method	130+	
Motor Method		
Cetane Number		
Toxicity		
Least Detectable Odor		No odor
Least Amount Causing Eye Irritation		No irritation
Least Amount Causing Throat Irritation		No irritation
Least Amount Causing Coughing		No coughing
Maximum Allowable for Prolonged Exposure		100 ppm
Maximum Allowable for Short Exposure (0.5 hr)		400 ppm
Dangerous for Short Exposure (0.5 hr)		1500 ppm

Comments

Normal transportation is in cylinders under pressure. Continued exposure to concentrations of carbon monoxide greater than 100 ppm in air will cause headache, palpitation of the heart, confusion of mind, and nausea. Doses above 0.3% (in air) for 1 hr or more are often fatal.

COAL (Northern Illinois, ASTM Rank-Class II, Group 5)

Chemical Analysis (ultimate) 61.8% C, 4.3% H_2, 12% H_2O, 9% Ash, 8% O_2, 8.8% S, 1.2% N_2

	English Units	Metric (SI) Units
Density		
Solid	~84 lb/cu ft	1.345 kg/cu m
Specific Gravity		
Solid	~1.346	1.346
Heating Value, Solid		
Volumetric Gross	940,800 Btu/cu ft	3505 X 10^4 kJ/cu m
Volumetric Net	932,400 Btu/cu ft	3473 X 10^4 kJ/cu m
Weight Gross	11,200 Btu/lb	2.605 X 10^4 kJ/cu m
Weight Net	11,100 Btu/lb	2.583 X 10^4 kJ/kg
Air for Combustion, Solid Fuel		
O_2 Volumetric	1927.986 cu ft/cu ft	1927.986 cu m/cu m
N_2 Volumetric	7295.838 cu ft/cu ft	7295.838 cu m/cu m
Air Volumetric	9219.447 cu ft/cu ft	9219.447 cu m/cu m
O_2 Weight	1.942 lb/lb	1.942 kg/kg
N_2 Weight	6.461 lb/lb	6.461 kg/kg
Air Weight	8.402 lb/lb	8.402 kg/kg
Products of Combustion		
Solid Ash	7.56 lb/cu ft coal	121.111 kg/cu m
CO_2 Volumetric	1625.624 cu ft/cu ft	1625.624 cu m/cu m
H_2O Volumetric	549.743 cu ft/cu ft	549.743 cu m/cu m
SO_2 Volumetric	36.836 cu ft/cu ft	36.836 cu m/cu m
CO_2 Weight	2.264 lb/lb	2.264 kg/kg
H_2O Weight	0.501 lb/lb	0.501 kg/kg
SO_2 Weight	0.176 lb/lb (Min.)	0.176 kg/kg (Min.)
Flammability Limits (Coal Dust)	50 oz coal/1000 cu ft air	0.044 g coal/cu m
Ignition Temperature	$1100^{\circ}F$	$593^{\circ}C$

Toxicity
Least Detectable Odor: Solid Coal Chunks Nontoxic [*]
Maximum Allowable for Prolonged Exposure: (dust) 50 X 10^6 particles/cu ft[*]

Comments:

Bulk transportation in cars, trucks, barges, steamers, etc. Has been transported by pipeline as a slurry. Locally in power plants as airborne dust.

[*] Toxicity: Chronic inhalation of coal dust can cause lung disease. The maximum allowable concentration or threshold limit of coal dust has not been researched adequately.

SOLVENT-REFINED COAL
(Pittsburgh and Midway Coal Mining Co.)

Chemical Composition: Composition of SRC varies with input composition
to solvent-refining process. For this sample, the
input is shown below:

	Raw Coal	SRC
	Wt %	
Carbon	70.7	88.2
Hydrogen	4.7	5.2
Nitrogen	1.1	1.5
Sulfur	3.4	1.2
Oxygen	10.3	3.4
Ash	7.1	0.5
Moisture	2.7	--

Molecular Weight (approx)   12

	English Units	Metric (SI) Units
Heating Value, Solid		
Weight Gross	15,768 Btu/lbm	3.667 X 10^4 kJ/kg
Weight Net	15,120 Btu/lbm	3.517 X 10^4 kJ/kg
Air for Combustion, Solid Fuel		
N_2 Weight	9.10 lb/lb	9.10 kg/kg
O_2 Weight	2.74 lb/lb	2.74 kg/kg
Air Weight	11.83 lb/lb	11.83 lb/lb
Products of Combustion, Solid Fuel		
CO_2 Weight	3.23 lb/lb	3.23 kg/kg
H_2O Weight	0.464 lb/lb	0.464 kg/kg
N_2 Weight	9.10 lb/lb	9.10 kg/kg
SO_2 Weight	0.024 lb/lb	0.024 kg/kg

Toxicity: Solid SRC particles are nontoxic.

Comments:

Bulk transportation of solid SRC is the same as that for coal.

DIESEL OIL

Chemical Formula	Hydrocarbon Mixture 86-87% C, 13-14% H, 0.5% S	
Molecular Weight	165 to 240 (12 to 17 carbon atoms/molecule)	

	English Units	Metric (SI) Units
Pour Point	20 to 32°F (max)	−7 to 0°C (max)
Boiling Point	325-650°F	160-343°C
Density Liquid (7.12 lb/gal)	53.41 lb/cu ft	855.63 kg/cu m
Specific Gravity Liquid	0.856	0.856
Heating Value, Liquid Volumetric Gross Volumetric Net Weight Gross Weight Net	1,046,301 Btu/cu ft 987,017 Btu/cu ft 19,590 Btu/lb 18,480 Btu/lb	3897.764×10^4 kJ/cu m 3676.915×10^4 kJ/cu m 4.556×10^4 kJ/kg 4.298×10^4 kJ/kg
Heating Value, Vapor Weight Gross Weight Net	19,590 Btu/lb 18,480 Btu/lb	4.556×10^4 kJ/kg 4.298×10^4 kJ/kg
Heat of Vaporization	155 Btu/lb	360 kJ/kg
Flash Point	100°F	38°C
Ignition Temperature	490°F	254°C
Flammability Limits	0.7-5.0%	
Cetane Number	40-68	
Air for Combustion, Vapor Fuel O_2 Weight N_2 Weight Air Weight	3.281 lb/lb 10.919 lb/lb 14.201 lb/lb	3.281 kg/kg 10.919 kg/kg 14.201 kg/kg
Air for Combustion, Liquid Fuel O_2 Volumetric N_2 Volumetric Air Volumetric O_2 Weight N_2 Weight Air Weight	2071 cu ft/cu ft 7839 cu ft/cu ft 9910 cu ft/cu ft 3.281 lb/lb 10.919 lb/lb 14.201 lb/lb	2071 cu m/cu m 7839 cu m/cu m 9910 cu m/cu m 3.281 kg/kg 10.919 kg/kg 14.201 kg/kg

DIESEL OIL, Cont.

	English Units	Metric (SI) Units
Products of Combustion, Vapor Fuel		
CO_2 Weight	3.218 lb/lb	3.218 kg/kg
H_2O Weight	1.057 lb/lb	1.057 kg/kg
N_2 Weight	11.019 lb/lb	11.019 kg/kg
SO_2 Weight	0.007 lb/lb	0.007 kg/kg
Products of Combustion, Liquid Fuel		
CO_2 Volumetric	1469 cu ft/cu ft	1469 cu m/cu m
H_2O Volumetric	1186 cu ft/cu ft	1186 cu m/cu m
N_2 Volumetric	7911 cu ft/cu ft	7911 cu m/cu m
SO_2 Volumetric	2.3 cu ft/cu ft	2.3 cu m/cu m
CO_2 Weight	3.218 lb/lb	3.218 kg/kg
H_2O Weight	1.057 lb/lb	1.057 kg/kg
N_2 Weight	11.019 lb/lb	11.019 kg/kg

Toxicity
    Least Detectable Odor   25-100 ppm*
    Maximum Allowable for Prolonged Exposure   500 ppm

Comments

Normal transportation is by railway cars, trucks, barges, and pipelines.

*Amounts detectable by odor depend on impurities, aromatics, and sulfur compounds. Diesel oil is fuel oil No. 1 or No. 2. Fuel oil No. 1 is very similar to kerosene in chemical and physical properties. It is slightly toxic, and inhalation of high concentration of vapor can cause headache, stupor, and nausea.

ETHANOL

Chemical Formula	C_2H_4OH	
Molecular Weight	46.067	

	English Units	Metric (SI) Units
Melting Point	$-179^\circ F$	$-117^\circ C$
Boiling Point	$172^\circ F$	$77.7^\circ C$
Density		
Vapor	0.1216 lb/cu ft	1.948 kg/cu m
Liquid	48.98 lb/cu ft	784.65 kg/cu m
Specific Gravity		
Vapor	1.5890	1.5890
Liquid	0.785	0.785
Heating Value, Vapor		
Volumetric Gross	1600.3 Btu/cu ft	5.9616×10^4 kJ/cu m
Volumetric Net	1450.5 Btu/cu ft	5.4035×10^4 kJ/cu m
Weight Gross	13,161 Btu/lb	3.0610×10^4 kJ/kg
Weight Net	11,929 Btu/lb	2.7745×10^4 kJ/kg
Heating Value, Liquid		
Volumetric Gross	644,678 Btu/cu ft	2401.61×10^4 kJ/cu m
Volumetric Net	584,282 Btu/cu ft	2176.62×10^4 kJ/cu m
Weight Gross	13,161 Btu/lb	3.0610×10^4 kJ/kg
Weight Net	11,929 Btu/lb	2.7745×10^4 kJ/kg
Air for Combustion, Vapor	———air component per unit of fuel———	
O_2 Volumetric	3 cu ft/cu ft	3 cu m/cu m
N_2 Volumetric	11.293 cu ft/cu ft	11.293 cu m/cu m
Air Volumetric	14.293 cu ft/cu ft	14.293 cu m/cu m
O_2 Weight	2.084 lb/lb	2.084 kg/kg
N_2 Weight	6.934 lb/lb	6.934 kg/kg
Air Weight	9,018 lb/lb	9.018 kg/kg
Air for Combustion, Liquid	———air component per unit of fuel———	
O_2 Volumetric	1208.4 cu ft/cu ft	1208.4 cu m/cu m
N_2 Volumetric	4548.8 cu ft/cu ft	4548.8 cu m/cu m
Air Volumetric	5757.2 cu ft/cu ft	5757.2 cu m/cu m
O_2 Weight	2.084 lb/lb	2.084 kg/kg
N_2 Weight	6.934 lb/lb	6.934 kg/kg
Air Weight	9.018 lb/lb	9.018 kg/kg
Products of Combustion, Vapor	———unit of product per unit of fuel———	
CO_2 Volumetric	2.0 cu ft/cu ft	2.0 cu m/cu m
H_2O Volumetric	3.0 cu ft/cu ft	3.0 cu m/cu m
N_2 Volumetric	11.293 cu ft/cu ft	11.293 cu m/cu m
CO_2 Weight	1.922 lb/lb	1.922 kg/kg
H_2O Weight	1.170 lb/lb	1.170 kg/kg
N_2 Weight	6.934 lb/lb	6.934 kg/kg
SO_2 Weight		

ETHANOL, Cont.

	English Units	Metric (SI) Units
Products of Combustion, Liquid	—unit of product per unit of fuel—	
CO_2 Volumetric	805.6 cu ft/cu ft	805.6 cu m/cu m
H_2O Volumetric	1208.4 cu ft/cu ft	1208.4 cu m/cu m
SO_2 Volumetric		
N_2 Volumetric	4548.7 cu ft/cu ft	4548.7 cu m/cu m
CO_2 Weight	1.922 lb/lb	1.922 kg/kg
H_2O Weight	1.170 lb/lb	1.170 kg/kg
N_2 Weight	6.934 lb/lb	6.934 kg/kg
SO_2 Weight		
Ash Weight		
Flammability Limits	3.28-18.95%	
Flash Point	$55^{\circ}F$	$13^{\circ}C$
Ignition Temperature	$738^{\circ}F$	$392^{\circ}C$
Heat of Vaporization at Boiling Point	370 Btu/lb	860 kJ/kg
Octane Number		
Research Method	106	
Motor Method	89	
Cetane Number		

Toxicity
Least Detectable Odor	10 ppm	
Least Amount Causing Eye Irritation	5000 ppm	
Least Amount Causing Throat Irritation	5000 ppm	
Least Amount Causing Coughing		
Maximum Allowable for Prolonged Exposure	1000 ppm	
Maximum Allowable for Short Exposure (0.5 hr)		
Dangerous for Short Exposure (0.5 hr)		

Comments

Normal transportation is by railway cars, tank trucks, individual drums, or other size containers; ethanol could be transported in liquid pipelines. Because it dissolves readily in water, ethanol is easily adulterated. By law it must be denatured to prevent consumption as beverage.

FUEL OIL (No. 6)

Chemical Formula		Hydrocarbon mixture 86% C, 10.6% H, 2% S, 0.3-0.4% Ash
Molecular Weight		350-975 (25-70 carbon atoms/molecule)

	English Units	Metric (SI) Units
Melting Point		
Boiling Range	$600°F-1000°F$	$315°-540°C$
Density		
Vapor		
Liquid 8.094 lb/gal	60.65 lb/cu ft	971 kg/cu m
Specific Gravity		
Vapor		
Liquid	0.972	0.972
Heating Value, Liquid		
Volumetric Gross	1,101,000 Btu/cu ft	4101.9×10^4 kJ/cu m
Volumetric Net	1,030,754 Btu/cu ft	3877.1×10^4 kJ/cu m
Weight Gross	18,155 Btu/lb	4.222×10^4 kJ/kg
Weight Net	17,160 Btu/lb	3.991×10^4 kJ/kg
Air for Combustion, Vapor	———air component per unit of fuel———	
O_2 Volumetric		
N_2 Volumetric		
Air Volumetric		
O_2 Weight	3.181 lb/lb	3.181 kg/kg
N_2 Weight	10.586 lb/lb	10.586 kg/kg
Air Weight	13.767 lb/lb	13.767 kg/kg
Air for Combustion, Liquid	———air component per unit of fuel———	
O_2 Volumetric	2.277 cu ft/cu ft	2.277 cu m/cu m
N_2 Volumetric	8.626 cu ft/cu ft	8.626 cu m/cu m
Air Volumetric	10.903 cu ft/cu ft	10.903 cu m/cu m
O_2 Weight	3.181 lb/lb	3.181 kg/kg
N_2 Weight	10.586 lb/lb	10.586 kg/kg
Air Weight	13.767 lb/lb	13.767 kg/kg

FUEL OIL (No. 6) Cont.

	English Units	Metric (SI) Units
Products of Combustion, Liquid	————unit of product per unit of fuel————	
CO_2 Volumetric	1560 cu ft/cu ft	1560 cu m/cu m
H_2O Volumetric	2315 cu ft/cu ft	2315 cu m/cu m
SO_2 Volumetric	variable	variable
N_2 Volumetric		
CO_2 Weight	191 lb/cu ft	30.6×10^2 kg/cu m
H_2O Weight	116 lb/cu ft	18.6×10^2 kg/cu m
N_2 Weight		
SO_2 Weight		
Ash Weight	1.8 lb/cu ft	28.8 kg/cu m
Flammability Limits		
Flash Point	150°F	66°C
Ignition Temperature	765°F	407°C
Heat of Vaporization at 1 atm	$\sim$85 Btu/lb	40 kJ/kg
Octane Number		
Research Method		
Motor Method		
Cetane Number	15-30	

Toxicity
Least Detectable Odor   Unknown
Least Amount Causing Eye Irritation
Least Amount Causing Throat Irritation
Least Amount Causing Coughing
Maximum Allowable for Prolonged Exposure   Unknown
Maximum Allowable for Short Exposure (0.5 hr)
Dangerous for Short Exposure (0.5 hr)

Comments

Normal transportation is by rail tank cars, tank trucks, and barges in drums. No. 6 fuel oil is a residual oil that is very viscous. It often requires heating to allow flow or pumping. There is no legal limit on the sulfur content, which typically varies from 1 to 2% by weight.

Chemical Formula	Hydrocarbon mixture 84.9% C, 15% H, 0.1% (or less) S	
Molecular Weight	85-145 (6-10 carbon atoms/molecule)	

	English Units	Metric (SI) Units
Melting Point		
Boiling Range	100°-400°F	38°-204°C
Density		
Vapor		
Liquid	45.5 lb/cu ft	728.1 kg/cu m
Specific Gravity		
Vapor		
Liquid	0.7275	0.7275
Heating Value, Vapor		
Volumetric Gross		
Volumetric Net		
Weight Gross	20,700 Btu/lb	4.814×10^4 kJ/kg
Weight Net	19,291 Btu/lb	4.487×10^4 kJ/kg
Heating Value, Liquid		
Volumetric Gross	941,850 Btu/cu ft	3508.655×10^4 kJ/cu m
Volumetric Net	877,740 Btu/cu ft	3269.827×10^4 kJ/cu m
Weight Gross	20,700 Btu/lb	4.814×10^4 kJ/kg
Weight Net	19,291 Btu/lb	4.487×10^4 kJ/kg
Air for Combustion, Vapor	——air component per unit of fuel——	
O_2 Volumetric		
N_2 Volumetric		
Air Volumetric		
O_2 Weight	3.455 lb/lb	3.455 kg/kg
N_2 Weight	11.494 lb/lb	11.494 kg/kg
Air Weight	14.949 lb/lb	14.949 kg/kg
Air for Combustion, Liquid	——air component per unit of fuel——	
O_2 Volumetric	1858 cu ft/cu ft	1858 cu m/cu m
N_2 Volumetric	7030 cu ft/cu ft	7030 cu m/cu m
Air Volumetric	8888 cu ft/cu ft	8888 cu m/cu m
O_2 Weight	3.455 lb/lb	3.455 kg/kg
N_2 Weight	11.494 lb/lb	11.494 kg/kg
Air Weight	14.946 lb/lb	14.946 kg/kg
Products of Combustion, Vapor	——unit of product per unit of fuel——	
CO_2 Volumetric		
H_2O Volumetric		
N_2 Volumetric		
CO_2 Weight	3.004 lb/lb	3.004 kg/kg
H_2O Weight	1.342 lb/lb	1.342 kg/kg
N_2 Weight	11.494 lb/lb	11.494 kg/kg
SO_2 Weight		

GASOLINE, Cont.

	English Units	Metric (SI) Units
Products of Combustion, Liquid	——unit of product per unit of fuel——	
CO_2 Volumetric	1156 cu ft/cu ft	1156 cu m/cu m
H_2O Volumetric	1280 cu ft/cu ft	1280 cu m/cu m
SO_2 Volumetric	5.1 cu ft/cu ft	5.1 cu m/cu m
N_2 Volumetric		
CO_2 Weight	3.11 lb/lb	3.11 kg/kg
H_2O Weight	2.70 lb/lb	2.70 kg/kg
N_2 Weight		
SO_2 Weight	0.002 lb/lb	0.002 kg/kg
Ash Weight		
Flammability Limits	1.4-7.6%	
Flash Point (range)	-36° to -50°F	-38° to -45°C
Ignition Temperature	495°F	257°C
Heat of Vaporization at 1 atm	130 Btu/lb	300 kJ/kg
Octane Number		
Research Method	92-100	
Motor Method	84-92	
Cetane Number	18.0	

Toxicity

Least Detectable Odor	10-50 ppm*
Least Amount Causing Eye Irritation	Unknown
Least Amount Causing Throat Irritation	Unknown
Least Amount Causing Coughing	
Maximum Allowable for Prolonged Exposure	500 ppm
Maximum Allowable for Short Exposure (0.5 hr)	
Dangerous for Short Exposure (0.5 hr)	

Comments

Normal transportation is by pipeline, railway cars, trucks, and drums.
Toxicity: detectable amount depends on sulfur content, additives, and
aromatic hydrocarbon content. Further, the maximum allowable level
for prolonged exposure is dependent on the aromatic hydrocarbon content.

HYDRAZINE

	English Units	Metric (SI) Units
Chemical Formula	N_2H_4	
Molecular Weight	32.05	

	English Units	Metric (SI) Units
Melting Point	34.52°F	1.4°C
Boiling Point	236.3°F	113.5°C
Density		
Vapor	0.0848 lb cu/ft	1.358 kg/cu m
Liquid	63.0864 lb/cu ft	1010.638 kg/cu m
Specific Gravity		
Vapor	1.108	1.108
Liquid	1.011	1.011
Heating Value, Vapor*		
Volumetric Gross	232.33 Btu/cu ft	86.549 kJ/cu m
Volumetric Net	179.55 Btu/cu ft	66.887 kJ/cu m
Weight Gross	2597 Btu/lb	6040 kJ/kg
Weight Net	2007 Btu/lb	4667 kJ/kg
Heating Value, Liquid†		
Volumetric Gross	296,454 Btu/cu ft	11,043,500 kJ/cu m
Volumetric Net	271,784 Btu/cu ft	10,124,500 kJ/cu m
Weight Gross	7090 Btu/lb	16,490 kJ/kg
Weight Net	6500 Btu/lb	15,118 kJ/kg
Air for Combustion, Vapor	——air component per unit of fuel——	
O_2 Volumetric	1.0 cu ft/cu ft	1.0 cu m/cu m
N_2 Volumetric	3.764 cu ft/cu ft	3.764 cu m/cu m
Air Volumetric	4.764 cu ft/cu ft	4.764 cu m/cu m
O_2 Weight	0.998 lb/lb	0.998 kg/kg
N_2 Weight	3.286 lb/lb	3.286 kg/kg
Air Weight	4.301 lb/lb	4.301 kg/kg
Air for Combustion, Liquid	——air component per unit of fuel——	
O_2 Volumetric	744 cu ft/cu ft	744 cu m/cu m
N_2 Volumetric	2786 cu ft/cu ft	2786 cu m/cu m
Air Volumetric	3530 cu ft/cu ft	3530 cu m/cu m
O_2 Weight	0.998 lb/lb	0.998 kg/kg
N_2 Weight	3.286 lb/lb	3.286 kg/kg
Air Weight	4.284 lb/lb	4.284 kg/kf
Products of Combustion, Vapor	——unit of product per unit of fuel——	
CO_2 Volumetric		
H_2O Volumetric	2 cu ft/cu ft	2 cu m/cu m
N_2 Volumetric	4.764 cu ft/cu ft	4.764 cu m/cu m
CO_2 Weight		
H_2O Weight	1.122 lb/lb	1.122 kg/kg
N_2 Weight	4.179 lb/lb	4.179 kg/kg
SO_2 Weight		

	English Units	Metric (SI) Units
Products of Combustion, Liquid	——unit of product per unit of fuel——	
CO_2 Volumetric		
H_2O Volumetric	2651.773 cu ft/cu ft	2651.773 cu m/cu m
SO_2 Volumetric		
N_2 Volumetric	4040.208 cu ft/cu ft	4040.208 cu m/cu m
CO_2 Weight		
H_2O Weight	1.122 lb/lb	1.122 kg/kg
N_2 Weight	4.179 lb/lb	4.179 kg/kg
SO_2 Weight		
Ash Weight		
Flammability Limits	4.7-100.0%	
Flash Point	$100^{\circ}F$	$36^{\circ}C$
Ignition Temperature	See Comment 2	See Comment 2
Heat of Vaporization at Boiling Point	561 Btu/lb	1302 kJ/kg

Octane Number
    Research Method
    Motor Method

Cetane Number

Toxicity
    Least Detectable Odor
    Least Amount Causing Eye Irritation
                                                       Contact with an excess
                                                        of 5% aqueous solution
                                                        causes severe injury.

    Maximum Allowable for Prolonged Exposure   1 ppm
    Maximum Allowable for Short Exposure (0.5 hr)
    Dangerous for Short Exposure (0.5 hr)

Comments

Dangerous to transport in undiluted state, can dissociate without access to air. Normally transported as a hydrate, $N_2H_4 \cdot H_2O$, or fuming liquid that can be dissolved in water for additional safety in handling. In hydrated form, it can be used as fuel-cell fuel. Ignition temperatures vary probably due to catalytic action: in contact with iron rust, $74^{\circ}F$ ($23^{\circ}C$); black iron, $270^{\circ}F$ ($132^{\circ}C$); stainless steel, $313^{\circ}F$ ($156^{\circ}C$); glass, $518^{\circ}F$ ($270^{\circ}C$).

*Liquid heating value is for $N_2H_4 \cdot H_2O$.

†Vapor heating value is for N_2H_4.

HYDROGEN

Chemical Formula H_2

Molecular Weight 2.016

	English Units	Metric (SI) Units
Melting Point	$-430.85^{\circ}F$	$-257.14^{\circ}C$
Boiling Point	$-423.0^{\circ}F$	$-252.78^{\circ}C$

Density
Vapor Gas at 2000 psi 0.667	0.005327 lb/cu ft	0.08533 kg/cu m
Liquid at N.B.P.	4.43 lb/cu ft	70.968 kg/cu m

Specific Gravity
Vapor	0.06959	0.06959
Liquid	0.07099	0.07099

Heating Value, Vapor
Volumetric Gross	325.0 Btu/cu ft	1.2107×10^4 kJ/cu m
Volumetric Net	275.0 Btu/cu ft	1.0245×10^4 kJ/cu m
Weight Gross	61,100 Btu/lb	14.2108×10^4 kJ/kg
Weight Net	51,623 Btu/lb	12.0067×10^4 kJ/kg

Heating Value, Liquid
Volumetric Gross	270,274 Btu/cu ft	1006.85×10^4 kJ/cu m
Volumetric Net	228,693 Btu/cu ft	851.95×10^4 kJ/cu m
Weight Gross	61,100 Btu/lb	14.2108×10^4 kJ/kg
Weight Net	51,623 Btu/lb	12.0067×10^4 kJ/kg

Air for Combustion, Vapor
————air component per unit of fuel————
O_2 Volumetric	0.5 cu ft/cu ft	0.5 cu m/cu m
N_2 Volumetric	1.882 cu ft/cu ft	1.882 cu m/cu m
Air Volumetric	2.382 cu ft/cu ft	2.382 cu m/cu m
O_2 Weight	7.937 lb/lb	7.937 kg/kg
N_2 Weight	26.406 lb/lb	26.406 kg/kg
Air Weight	34.344 lb/lb	34.344 kg/kg

Air for Combustion, Liquid
————air component per unit of fuel————
O_2 Volumetric	415.8 cu ft/cu ft	415.8 cu m/cu m
N_2 Volumetric	1565.1 cu ft/cu ft	1565.1 cu m/cu m
Air Volumetric	1980.9 cu ft/cu ft	1980.9 cu m/cu m
O_2 Weight	7.937 lb/lb	7.937 kg/kg
N_2 Weight	26.406 lb/lb	26.406 kg/kg
Air Weight	34.344 lb/lb	34.344 kg/kg

Products of Combustion, Vapor
————unit of product per unit of fuel————
CO_2 Volumetric	--	--
H_2O Volumetric	1.0 cu ft/cu ft	1.0 cu m/cu m
N_2 Volumetric	1.882 cu ft/cu ft	1.882 cu m/cu m
CO_2 Weight	--	--
H_2O Weight	8.937 lb/lb	8.937 kg/kg
N_2 Weight	26.407 lb/lb	26.407 kg/kg
SO_2 Weight		

HYDROGEN, Cont.

	English Units	Metric (SI) Units
Products of Combustion, Liquid	—unit of product per unit of fuel—	
CO_2 Volumetric	--	
H_2O Volumetric	831.6 cu ft/cu ft	831.6 cu m/cu m
SO_2 Volumetric		
N_2 Volumetric	1565.1 cu ft/cu ft	1565.1 cu m/cu m
CO_2 Weight	--	
H_2O Weight	8.937 lb/lb	8.937 kg/kg
N_2 Weight	26.407 lb/lb	26.407 kg/kg
SO_2 Weight		
Ash Weight		
Flammability Limits	4.00-74.2%	
Flash Point		
Ignition Temperature	1065°F	574°C
Heat of Vaporization		
Octane Number		
Research Method	130+	
Motor Method		
Cetane Number		
Toxicity		
Least Detectable Odor	No odor	
Least Amount Causing Eye Irritation	No irritation	
Least Amount Causing Throat Irritation	No irritation	
Least Amount Causing Coughing	No coughing	
Maximum Allowable for Prolonged Exposure		
Maximum Allowable for Short Exposure (0.5 hr)		
Dangerous for Short Exposure (0.5 hr)		

Comments

Normally distributed as a compressed gas in high-pressure container
(cylinder). Can be shipped by pipeline. Can also be transported as a
cryogenic liquid in insulated or vacuum-jacketed tanks. Trucks and
railroads are commonly used for long-distance bulk transport of hydrogen
as a cryogenic liquid. Hydrogen can be combined with many metals and
alloys to form metal hydrides. Titanium, iron, and magnesium are examples.

Hydrogen is an odorless, colorless, and nontoxic gas. It burns with a non-
luminous flame, and it can be easily combusted catalytically (without flame)
because of its low ignition energy. For safe use as a fuel, an odorant and
possibly an illuminant would be required.

KEROSENE

	English Units	Metric (SI) Units
Chemical Formula	Hydrocarbon mixture, 85-86% C, 14-15% H, 0.5% S (max)	
Molecular Weight	150-230 (11-16 carbon atoms/molecule)	

	English Units	Metric (SI) Units
Melting Point		
Boiling Range	300°-480°F	150°-250°C
Density		
Vapor		
Liquid	50.61 lb/cu ft	810.87 kg/cu m
Specific Gravity		
Vapor		
Liquid	0.811	0.811
Heating Value, Vapor		
Volumetric Gross		
Volumetric Net		
Weight Gross	20,500 Btu/lb	4.768 X 10^4 kJ/kg
Weight Net	19,092 Btu/lb	4.440 X 10^4 kJ/kg
Heating Value, Liquid		
Volumetric Gross	1,037,505 Btu/cu ft	3864.997 X 10^4 kJ/cu m
Volumetric Net	966.246 Btu/cu ft	3599.537 X 10^4 kJ/cu m
Weight Gross	20,500 Btu/lb	4.768 X 10^4 kJ/kg
Weight Net	19,092 Btu/lb	4.440 X 10^4 kJ/kg
Air for Combustion, Vapor	——air component per unit of fuel——	
O_2 Volumetric		
N_2 Volumetric		
Air Volumetric		
O_2 Weight	3.455 lb/lb	3.455 kg/kg
N_2 Weight	11.495 lb/lb	11.495 kg/kg
Air Weight	14.950 lb/lb	14.950 kg/kg
Air for Combustion, Liquid	——air component per unit of fuel——	
O_2 Volumetric	2066.23 cu ft/cu ft	2066.23 cu m/cu m
N_2 Volumetric	7820.62 cu ft/cu ft	7820.62 cu m/cu m
Air Volumetric	9886.85 cu ft/cu ft	9886.85 cu m/cu m
O_2 Weight	3.455 lb/lb	3.455 kg/kg
N_2 Weight	11.495 lb/lb	11.495 kg/kg
Air Weight	14.950 lb/lb	14.950 kg/kg
Products of Combustion, Vapor	——unit of product per unit of fuel——	
CO_2 Volumetric		
H_2O Volumetric		
N_2 Volumetric		
CO_2 Weight	3.114 lb/lb	3.114 kg/kg
H_2O Weight	1.341 lb/lb	1.341 kg/kg
N_2 Weight	11.495 lb/lb	11.495 kg/kg
SO_2 Weight		

KEROSENE, Cont.

	English Units	Metric (SI) Units
Products of Combustion, Liquid	unit of product per unit of fuel	
CO_2 Volumetric	1347.16 cu ft/cu ft	1347.16 cu m/cu m
H_2O Volumetric	1426.38 cu ft/cu ft	1426.38 cu m/cu m
SO_2 Volumetric		
N_2 Volumetric	7820.62 cu ft/cu ft	7820.62 cu m/cu m
CO_2 Weight	3.114 lb/lb	3.114 kg/kg
H_2O Weight	1.341 lb/lb	1.341 kg/kg
N_2 Weight	11.495 lb/lb	11.495 kg/kg
SO_2 Weight	0.01 lb/lb	0.01 kg/kg
Ash Weight		

Flammability Limits	0.7-5.0%	
Flash Point	$100^{o}F$	$38^{o}C$
Ignition Temperature	$491^{o}F$	$255^{o}C$
Heat of Vaporization		
Octane Number		
Research Method		
Motor Method		
Cetane Number	40-65	

Toxicity
Least Detectable Odor	25-100 ppm*
Least Amount Causing Eye Irritation	500-1000 ppm
Least Amount Causing Throat Irritation	500-1000 ppm
Least Amount Causing Coughing	
Maximum Allowable for Prolonged Exposure	500 ppm
Maximum Allowable for Short Exposure (0.5 hr)	
Dangerous for Short Exposure (0.5 hr)	

Comments

Normally carried in tanks or drums; can be transported by tank truck.
Could be pipelined in liquid (hydrocarbon) fuel pipelines.

*Toxicity: Leat detectable odor depends on impurities, aromatics, and
sulfur contents. Kerosene and No. 2 distillate oil are essentially the
same. The difference (if any) is that kerosene has a slightly higher gravity.
Inhalation of high concentrations of vapor can cause headache, stupor,
nausea, and vomiting.

METHANE

Chemical Formula	CH_4
Molecular Weight	16.041

	English Units	Metric (SI) Units
Melting Point	$-296.5^{\circ}F$	$-182.5^{\circ}C$
Boiling Point	$-258.5^{\circ}F$	$-161.4^{\circ}C$
Density		
Vapor at 2000 psi	7.08 lb/ft	113.42 kg/cu m
Liquid	26.5 lb/cu ft	424.5 kg/cu m
Specific Gravity		
Vapor	0.5543	0.5543
Liquid	0.1135	0.1135
Heating Value, Vapor (1 atm)		
Volumetric Gross	1013 Btu/cu ft	3.7745×10^4 kJ/cu m
Volumetric Net	913.1 Btu/cu ft	3.4016×10^4 kJ/cu m
Weight Gross	23,879 Btu/lb	5.5539×10^4 kJ/kg
Weight Net	21,520 Btu/lb	5.0052×10^4 kJ/kg
Heating Value, Liquid		
Volumetric Gross	16,9065 Btu/cu ft	629.815×10^4 kJ/cu m
Volumetric Net	152,362 Btu/cu ft	567.591×10^4 kJ/cu m
Weight Gross	23,879 Btu/lb	5.5539×10^4 kJ/kg
Weight Net	21,520 Btu/lb	5.0052×10^4 kJ/kg

Air for Combustion, Vapor	———air component per unit of fuel———	
O_2 Volumetric	2.0 cu ft/cu ft	2.0 cu m/cu m
N_2 Volumetric	7.528 cu ft/cu ft	7.528 cu m/cu m
Air Volumetric	9.528 cu ft/cu ft	9.528 cu m/cu m
O_2 Weight	3.990 lb/lb	3.990 kg/kg
N_2 Weight	13.275 lb/lb	13.275 kg/kg
Air Weight	17.265 lb/lb	17.265 kg/kg

Air for Combustion, Liquid	———air component per unit of fuel———	
O_2 Volumetric	333.7 cu ft/cu ft	333.7 cu m/cu m
N_2 Volumetric	1256.1 cu ft/cu ft	1256.1 cu m/cu m
Air Volumetric	1589.9 cu ft/cu ft	1589.9 cu m/cu m
O_2 Weight	3.990 lb/lb	3.990 lb/lb
N_2 Weight	13.275 lb/lb	13.275 kg/kg
Air Weight	17.265 lb/lb	17.265 kg/kg

Products of Combustion, Vapor	———unit of product per unit of fuel———	
CO_2 Volumetric	1.0 cu ft/cu ft	1.0 cu m/cu m
H_2O Volumetric	2.0 cu ft/cu ft	2.0 cu m/cu m
N_2 Volumetric	7.528 cu ft/cu ft	7.528 cu m/cu m
CO_2 Weight	2.744 lb/lb	2.744 kg/kg
H_2O Weight	2.246 lb/lb	2.246 kg/kg
N_2 Weight	13.275 lb/lb	13.275 kg/kg
SO_2 Weight		

	English Units	Metric (SI) Units
Products of Combustion, Liquid	——unit of product per unit of fuel——	
CO_2 Volumetric	166.9 cu ft/cu ft	166.9 cu m/cu m
H_2O Volumetric	333.7 cu ft/cu ft	333.7 cu m/cu m
SO_2 Volumetric		
N_2 Volumetric	1256.1 cu ft/cu ft	1256.1 cu m/cu m
CO_2 Weight		
H_2O Weight		
N_2 Weight		
SO_2 Weight		
Ash Weight		
Flammability Limits	5.00-15.00%	
Flash Point		
Ignition Temperature	$1170^\circ F$	$632^\circ C$
Heat of Vaporization at Boiling Point	219 Btu/lb	510 kJ/kg
Octane Number		
Research Method	130	
Motor Method		
Cetane Number		
Toxicity		
Least Detectable Odor	No odor	
Least Amount Causing Eye Irritation	No irritation	
Least Amount Causing Throat Irritation	No irritation	
Least Amount Causing Coughing		
Maximum Allowable for Prolonged Exposure	90,000 ppm	
Maximum Allowable for Short Exposure (0.5 hr)		
Dangerous for Short Exposure (0.5 hr)		

Comments

Normal transportation is by pipeline under pressure or as a cryogenic liquid in bulk.

Methane is considered a nontoxic chemical. In concentrations above 9% in air, it acts as a simple asphyxiant. When used as a gaseous fuel, it is odorized by mercaptan compounds for recognition.

METHANOL

	English Units	Metric (SI) Units
Chemical Formula	CH_3OH	
Molecular Weight	32.041	

	English Units	Metric (SI) Units
Melting Point	−143.82	−97.68°C
Boiling Point	148.1°F	64.1°C
Density		
Vapor	0.0846 lb/cu ft	1.355 kg/cu m
Liquid	49.72 lb/cu ft	796.51 kg/cu m
Specific Gravity		
Vapor	1.1052	1.1052
Liquid	0.796	0.796
Heating Value, Vapor		
Volumetric Gross	867.9 Btu/cu ft	3.233×10^4 kJ/cu m
Volumetric Net	768.0 Btu/cu ft	2.8610×10^4 kJ/cu m
Weight Gross	10,259 Btu/lb	2.3861×10^4 kg/kg
Weight Net	9,078 Btu/lb	2.114×10^4 kJ/kg
Heating Value, Liquid		
Volumetric Gross	510,077 Btu/cu ft	1900.18×10^4 kJ/cu m
Volumetric Net	451,358 Btu/cu ft	1681.44×10^4 kJ/cu m
Weight Gross	10,259 Btu/lb	2.3861×10^4 kJ/kg
Weight Net	9,078 Btu/lb	2.1114×10^4 kJ/kg

Air for Combustion, Vapor — air component per unit of fuel —

	English Units	Metric (SI) Units
O_2 Volumetric	1.5 cu ft/cu ft	1.5 cu m/cu m
N_2 Volumetric	5.646 cu ft/cu ft	5.646 cu m/cu m
Air Volumetric	7.146 cu ft/cu ft	7.146 cu m/cu m
O_2 Weight	1.498 lb/lb	1.498 kg/kg
N_2 Weight	4.984 lb/lb	4.984 kg/kg
Air Weight	6.482 lb/lb	6.482 kg/kg

Air for Combustion, Liquid — air component per unit of fuel —

	English Units	Metric (SI) Units
O_2 Volumetric	881.56 cu ft/cu ft	881.56 cu m/cu m
N_2 Volumetric	3318.1 cu ft/cu ft	3318.1 cu m/cu m
Air Volumetric	4199.8 cu ft/cu ft	4199.8 cu m/cu m
O_2 Weight	1.498 lb/lb	1.498 kg/kg
N_2 Weight	4.984 lb/lb	4.984 kg/kg
Air Weight	6.482 lb/lb	6.482 kg/kg

Products of Combustion, Vapor — unit of product per unit of fuel —

	English Units	Metric (SI) Units
CO_2 Volumetric	1.0 cu ft/cu ft	1.0 cu m/cu m
H_2O Volumetric	2.0 cu ft/cu ft	2.0 cu m/cu m
N_2 Volumetric	5.646 cu ft/cu ft	5.646 cu m/cu m
CO_2 Weight	1.374 lb/lb	1.374 kg/kg
H_2O Weight	1.125 lb/lb	1.125 kg/kg
N_2 Weight	4.984 lb/lb	4.984 kg/kg
SO_2 Weight		

METHANOL, Cont.

	English Units	Metric (SI) Units
Products of Combustion, Liquid	——unit of product per unit of fuel——	
CO_2 Volumetric	587.71 cu ft/cu ft	587.71 cu m/cu m
H_2O Volumetric	1175.4 cu ft/cu ft	1175.4 cu m/cu m
SO_2 Volumetric		
N_2 Volumetric	3318.1 cu ft/cu ft	3318.1 cu m/cu m
CO_2 Weight	1.374 lb/lb	1.374 kg/kg
H_2O Weight	1.125 lb/lb	1.125 kg/kg
N_2 Weight	4.984 lb/lb	4.984 kg/kg
SO_2 Weight		
Ash Weight		
Flammability Limits	6.72-36.50%	
Flash Point	52°F	11°C
Ignition Temperature	878°F	470°C
Heat of Vaporization at Boiling Point	473 Btu/lb	1100 kJ/kg
Octane Number		
Research Method	106	
Motor Method	92	
Cetane Number		
Toxicity		
Least Detectable Odor	100 ppm	
Least Amount Causing Eye Irritation	Unknown	
Least Amount Causing Throat Irritation	Unknown	
Least Amount Causing Coughing		
Maximum Allowable for Prolonged Exposure	200 ppm	
Maximum Allowable for Short Exposure (0.5 hr)		
Dangerous for Short Exposure (0.5 hr)		

Comments

Normal transportation is in bulk or by container. Methanol dissolves readily in water, so it is easily contaminated or adulterated. The main toxic effect of methanol is on the nervous system, particularly the optic nerves.

METHYLAMINE

Chemical Formula	CH_3NH_2

Molecular Weight   31.08

	English Units	Metric (SI) Units
Melting Point	$-134^\circ F$	$-92.5^\circ C$
Boiling Point	$20.3^\circ F$	$-6.5^\circ C$
Density		
Vapor	0.0872 lb/cu ft	1.396 kg/cu m
Liquid 5.833 lb/gal	43.638 lb/cu ft	699.0 kg/cu m
Specific Gravity		
Vapor	1.0797	1.0797
Liquid	0.699	0.699
Heating Value, Vapor		
Volumetric Gross	1292 Btu/cu ft	4.813×10^4 kJ/cu m
Volumetric Net	1088.7 Btu/cu ft	4.056×10^4 kJ/cu m
Weight Gross	14,819 Btu/lb	3.447×10^4 kJ/kg
Weight Net	12,384 Btu/lb	2.904×10^4 kJ/kg
Heating Value, Liquid		
Volumetric Gross	646,671 Btu/cu ft	2409.0×10^4 kJ/cu m
Volumetric Net	560,966 Btu/cu ft	2069.8×10^4 kJ/cu m
Weight Gross	14,819 Btu/lb	3.447×10^4 kJ/kg
Weight Net	12,855 Btu/lb	2.904×10^4 kJ/kg
Air for Combustion, Vapor	————air component per unit of fuel————	
O_2 Volumetric	3.667 cu ft/cu ft	3.667 cu m/cu m
N_2 Volumetric	13.804 cu ft/cu ft	13.804 cu m/cu m
Air Volumetric	17.471 cu ft/cu ft	17.471 cu m/cu m
O_2 Weight	3.558 lb/lb	3.558 kg/kg
N_2 Weight	11.776 lb/lb	11.776 kg/kg
Air Weight	15.338 lb/lb	15.338 kg/kg
Air for Combustion, Liquid	————air component per unit of fuel————	
O_2 Volumetric	1835 cu ft/cu ft	1835 cu m/cu m
N_2 Volumetric	6908 cu ft/cu ft	6908 cu m/cu m
Air Volumetric	8743 cu ft/cu ft	8743 cu m/cu m
O_2 Weight	3.558 lb/lb	3.558 kg/kg
N_2 Weight	11.776 lb/lb	11.776 kg/kg
Air Weight	15.338 lb/lb	15.338 kg/kg
Products of Combustion, Vapor	————unit of product per unit of fuel————	
CO_2 Volumetric	1.630 cu ft/cu ft	1.630 cu m/cu m
H_2O Volumetric	4.074 cu ft/cu ft	4.074 cu m/cu m
N_2 Volumetric	14.617 cu ft/cu ft	14.617 cu m/cu m
CO_2 Weight	2.187 lb/lb	2.187 kg/kg
H_2O Weight	2.223 lb/lb	2.223 kg/kg
N_2 Weight	12.370 lb/lb	12.370 kg/kg
SO_2 Weight		

METHYLAMINE, Cont.

	English Units	Metric (SI) Units
Products of Combustion, Liquid	——————unit of product per unit of fuel——————	
CO_2 Volumetric	815.8 cu ft/cu ft	815.8 cu m/cu m
H_2O Volumetric	2038.8 cu ft/cu ft	2038.8 cu m/cu m
SO_2 Volumetric		
N_2 Volumetric	7315.2 cu ft/cu ft	7315.2 cu m/cu m
CO_2 Weight	2.187 lb/lb	2.187 kg/kg
H_2O Weight	2.223 lb/lb	2.223 kg/kg
N_2 Weight	12.470 lb/lb	12.370 kg/kg
SO_2 Weight		
Ash Weight		
Flammability Limits	4.9-20.7%	
Flash Point	$0^\circ F$	$-18^\circ C$
Ignition Temperature	$806^\circ F$	$430^\circ C$
Heat of Vaporization at Boiling Point	340 Btu/lb*	790 kJ/kg*

Octane Number
    Research Method
    Motor Method

Cetane Number

Toxicity

Least Detectable Odor	0.02 ppm
Least Amount Causing Eye Irritation	10-50 ppm
Least Amount Causing Throat Irritation	10-50 ppm
Least Amount Causing Coughing	
Maximum Allowable for Prolonged Exposure	10 ppm
Maximum Allowable for Short Exposure (0.5 hr)	
Dangerous for Short Exposure (0.5 hr)	

Comments

Normal transportation is by bulk or container at moderate pressures. Methyl amine readily dissolves in water so it can easily be contaminated or adulterated. Synonyms: Monomethylamine, aminomethane. Is a strong irritant to the respiratory tract. CH_2NH_2 is a colorless gas or liquid with a strong ammoniacal odor.

*Estimated.

NAPHTHA

Chemical Formula	Hydrocarbon mixture 86-87% C, 13-14% H, 0-1% S
Molecular Weight	84-170 (6-12 carbon atoms/molecule)

	English Units	Metric (SI) Units
Melting Point		
Boiling Range	150-300°F	65-150°C
Density		
Vapor	~0.45 lb/cu ft	7.21 kg/cu m
Liquid	48 lb/cu ft	769 kg/cu m
Specific Gravity		
Vapor	~5.8	
Liquid	0.77	
Heating Value, Vapor		
Volumetric Gross	9104.5 Btu/cu ft	33.917 X 10^4 kJ/cu m
Volumetric Net	8460.5 Btu/cu ft	31.518 X 10^4 kJ/cu m
Weight Gross	20,300 Btu/lb	4.721 X 10^4 kJ/kg
Weight Net	18,864 Btu/lb	4.387 X 10^4 kJ/kg
Heating Value, Liquid		
Volumetric Gross	974,400 Btu/cu ft	3629.9 X 10^4 kJ/cu m
Volumetric Net	905,472 Btu/cu ft	3373.1 X 10^4 kJ/cu m
Weight Gross	20,300 Btu/lb	4.721 X 10^4 kJ/kg
Weight Net	18,864 Btu/lb	4.387 X 10^4 kJ/kg
Air for Combustion, Vapor	——air component per unit of fuel——	
O_2 Volumetric	18.394 cu ft/cu ft	18.394 cu m/cu m
N_2 Volumetric	69.619 cu ft/cu ft	69.619 cu m/cu m
Air Volumetric	88.013 cu ft/cu ft	88.013 cu m/cu m
O_2 Weight	3.470 lb/lb	3.470 kg/kg
N_2 Weight	11.547 lb/lb	11.547 kg/kg
Air Weight	15.018 lb/lb	15.018 kg/kg
Air for Combustion, Liquid	——air component per unit of fuel——	
O_2 Volumetric	1968.6 cu ft/cu ft	1968.6 cu m/cu m
N_2 Volumetric	7450.9 cu ft/cu ft	7450.9 cu m/cu m
Air Volumetric	9419.5 cu ft/cu ft	9419.5 cu m/cu m
O_2 Weight	3.470 lb/lb	3.470 kg/kg
N_2 Weight	11.547 lb/lb	11.547 kg/kg
Air Weight	15.018 lb/lb	15.018 kg/kg
Products of Combustion, Vapor	——unit of product per unit of fuel——	
CO_2 Volumetric	11.896 cu ft/cu ft	11.896 cu m/cu m
H_2O Volumetric	12.886 cu ft/cu ft	12.886 cu m/cu m
N_2 Volumetric	69.619 cu ft/cu ft	69.619 cu m/cu m
CO_2 Weight	3.103 lb/lb	3.103 kg/kg
H_2O Weight	1.367 lb/lb	1.367 kg/kg
N_2 Weight	11.547 lb/lb	11.547 kg/kg
SO_2 Weight		

NAPHTHA, Cont.

	English Units	Metric (SI) Units
Products of Combustion, Liquid	——unit of product per unit of fuel——	
CO_2 Volumetric	1273.2 cu ft/cu ft	1273.2 cu m/cu m
H_2O Volumetric	1379.0 cu ft/cu ft	1379.0 cu m/cu m
SO_2 Volumetric		
N_2 Volumetric	7450.9 cu ft/cu ft	7450.9 cu m/cu m
CO_2 Weight	3.103 lb/lb	3.103 kg/kg
H_2O Weight	1.367 lb/lb	1.367 kg/kg
N_2 Weight	11.547 lb/lb	11.547 kg/kg
SO_2 Weight		
Ash Weight		

Flammability Limits	0.90-6.00%	
Flash Point	20°-50°F	-7° to $+10^{\circ}$C
Ignition Temperature	450°-531°F	232°-277°C
Heat of Vaporization at 1 atm	145 Btu/lb	336 kJ/kg

Octane Number	
Research Method	60-70
Motor Method	50-60
Cetane Number	

Toxicity
Least Detectable Odor	10-50 ppm*
Least Amount Causing Eye Irritation	Unknown
Least Amount Causing Throat Irritation	Unknown
Least Amount Causing Coughing	
Maximum Allowable for Prolonged Exposure	500 ppm
Maximum Allowable for Short Exposure (0.5 hr)	
Dangerous for Short Exposure (0.5 hr)	

Comments

Normal transportation is by pipeline, tank truck, drums, or other closed containers.

*Toxicity: Detectable amounts depend on sulfur and aromatic hydrocarbon content.

PROPANE (Pure)

Chemical Formula C_3H_8

Molecular Weight 44.092

	English Units	Metric (SI) Units
Melting Point	$-305.88\,^{\circ}F$	$-187.71\,^{\circ}C$
Boiling Point	$-43.7\,^{\circ}F$	$-42.07\,^{\circ}C$

Density
Vapor	0.1196 lb/cu ft	1.91599 kg/cu m
Liquid	31.8 lb/cu ft	509.44 kg/cu m

Specific Gravity
Vapor	1.5617	1.5617
Liquid		

Heating Value, Vapor
Volumetric Gross	2590 Btu/cu ft	9.6485×10^4 kJ/cu m
Volumetric Net	2385 Btu/cu ft	8.8848×10^4 kJ/cu m
Weight Gross	21,661 Btu/lb	5.0380×10^4 kJ/kg
Weight Net	19,944 Btu/lb	4.6387×10^4 kJ/kg

Heating Value, Liquid
Volumetric Gross	688,645 Btu/cu ft	2565.40×10^4 kJ/cu m
Volumetric Net	634,138 Btu/cu ft	2362.34×10^4 kJ/cu m
Weight Gross	21,661 Btu/lb	5.0380×10^4 kJ/kg
Weight Net	19,944 Btu/lb	4.6387×10^4 kJ/kg

Air for Combustion, Vapor
	———air component per unit of fuel———	
O_2 Volumetric	5.0 cu ft/cu ft	5.0 cu m/cu m
N_2 Volumetric	18.821 cu ft/cu ft	18.821 cu m/cu m
Air Volumetric	23.821 cu ft/cu ft	23.821 cu m/cu m
O_2 Weight	3.629 lb/lb	3.629 kg/kg
N_2 Weight	12.074 lb/lb	12.074 kg/kg
Air Weight	15.703 lb/lb	15.703 kg/kg

Air for Combustion, Liquid
	———air component per unit of fuel———	
O_2 Volumetric	1329.4 cu ft/cu ft	1329.4 cu m/cu m
N_2 Volumetric	5004.2 cu ft/cu ft	5004.2 cu m/cu m
Air Volumetric	6333.6 cu ft/cu ft	6333.6 cu m/cu m
O_2 Weight	3.629 lb/lb	3.629 kg/kg
N_2 Weight	12.074 lb/lb	12.074 kg/kg
Air Weight	15.703 lb/lb	15.703 kg/kg

Products of Combustion, Vapor
	———unit of product per unit of fuel———	
CO_2 Volumetric	3.0 cu ft/cu ft	3.0 cu m/cu m
H_2O Volumetric	4.0 cu ft/cu ft	4.0 cu m/cu m
N_2 Volumetric	18.821 cu ft/cu ft	18.821 cu m/cu m
CO_2 Weight	2.994 lb/lb	2.994 kg/kg
H_2O Weight	1.634 lb/lb	1.634 kg/kg
N_2 Weight	12.074 lb/lb	12.074 kg/kg
SO_2 Weight		

PROPANE (Pure) Cont.

	English Units	Metric (SI) Units
Products of Combustion, Liquid	——unit of product	per unit of fuel——
CO_2 Volumetric	797.6 cu ft/cu ft	797.6 cu m/cu m
H_2O Volumetric	1063.5 cu ft/cu ft	1063.5 cu m/cu m
SO_2 Volumetric		
N_2 Volumetric	5003.9 cu ft/cu ft	5003.9 cu m/cu m
CO_2 Weight	2.994 lb/lb	2.994 kg/kg
H_2O Weight	1.634 lb/lb	1.634 kg/kg
N_2 Weight	12.074 lb/lb	12.074 kg/kg
SO_2 Weight		
Ash Weight		
Flammability Limits	2.1-10.1%	
Flash Point	-156°F	-104°C
Ignition Temperature	808°F	431°C
Heat of Vaporization at Boiling Point	150 Btu/lb	340 kJ/kg
Octane Number		
Research Method	111	
Motor Method	97	
Cetane Number		
Toxicity		
Least Detectable Odor	Unknown*	
Least Amount Causing Eye Irritation	No irritation	
Least Amount Causing Throat Irritation	No irritation	
Least Amount Causing Coughing		
Maximum Allowable for Prolonged Exposure	3-5%	
Maximum Allowable for Short Exposure (0.5 hr)		
Dangerous for Short Exposure (0.5 hr)		

Comments

Normal transportation is pipeline, bulk, or containers at moderate pressure. Liquid propane is contained at 110 psig at 70°F.

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*Propane has a faint odor that varies with (trace) impurities.

COMMERCIAL PROPANE (LPG)

Chemical Formula	Hydrocarbon mixture C_3H_8, C_4H_{10}, C_2H_4, C_3H_6	
Molecular Weight	30-60	

	English Units	Metric (SI) Units
Melting Point		
Boiling Point	$-50^\circ F$	$-45^\circ C$
Density		
Vapor	0.1169 lb/cu ft	1.8735 kg/cu m
Liquid 4.24 lb/gal	31.8 lb/cu ft	509.435 kg/cu m
Specific Gravity		
Vapor	1.52	1.52
Liquid	0.509	0.509
Heating Value, Vapor		
Volumetric Gross	2522 Btu/cu ft	9.395×10^4 kJ/cu m
Volumetric Net	2399 Btu/cu ft	8.937×10^4 kJ/cu m
Weight Gross	21.56 Btu/lb	5.014×10^4 kJ/kg
Weight Net	20,51 Btu/lb	4.771×10^4 kJ/kg
Heating Value, Liquid		
Volumetric Gross	685,068 Btu/cu ft	2554.09×10^4 kJ/cu m
Volumetric Net	652,345 Btu/cu ft	2430.17×10^4 kJ/cu m
Weight Gross	21,560 Btu/lb	5.013×10^4 kJ/kg
Weight Net	20,514 Btu/lb	4.771×10^4 kJ/kg
Air for Combustion, Vapor	———air component per unit of fuel———	
O_2 Volumetric	4.9 cu ft/cu ft	4.9 cu m/cu m
N_2 Volumetric	18.49 cu ft/cu ft	18.49 cu m/cu m
Air Volumetric	23.4 cu ft/cu ft	23.4 cu m/cu m
O_2 Weight	3.60 lb/lb	3.60 kg/kg
N_2 Weight	11.98 lb/lb	11.98 kg/kg
Air Weight	15.58 lb/lb	15.58 kg/kg
Air for Combustion, Liquid	———air component per unit of fuel———	
O_2 Volumetric	1353.04 cu ft/cu ft	1353.04 cu m/cu m
N_2 Volumetric	5115.87 cu ft/cu ft	5115.87 cu m/cu m
Air Volumetric	6468.91 cu ft/cu ft	6468.91 cu m/cu m
O_2 Weight	3.60 lb/lb	3.60 kg/kg
N_2 Weight	11.98 lb/lb	11.98 kg/kg
Air Weight	15.58 lb/lb	15.58 kg/kg
Products of Combustion, Vapor	———unit of product per unit of fuel———	
CO_2 Volumetric	3 cu ft/cu ft	3 cu m/cu m
H_2O Volumetric	3.8 cu ft/cu ft	3.8 cu m/cu m
N_2 Volumetric	18.5 cu ft/cu ft	18.5 cu m/cu m
CO_2 Weight	3 lb/lb	3 kg/kg
H_2O Weight	1.6 lb/lb	1.6 kg/kg
N_2 Weight	12 lb/lb	12 kg/kg
SO_2 Weight		

LPG, Cont.

	English Units	Metric (SI) Units
Products of Combustion, Liquid	———unit of product per unit of fuel———	
CO_2 Volumetric	815.47 cu ft/cu ft	815.47 cu m/cu m
H_2O Volumetric	664.65 cu ft/cu ft	664.65 cu m/cu m
SO_2 Volumetric		
N_2 Volumetric	5129.84 cu ft/cu ft	5129.84 cu m/cu m
CO_2 Weight	3 lb/lb	3 kg/kg
H_2O Weight	1.6 lb/lb	1.6 kg/kg
N_2 Weight	12 lb/lb	12 kg/kg
SO_2 Weight		
Ash Weight		
Flammability Limits	2.4-9.6%	
Flash Point	$-160°$ to $-150°F$	$-107°$ to $-100°C$
Ignition Temperature	$920°-1020°F$	$493°-547°C$
Heat of Vaporization at 1 atm	100 Btu/lb	420 kJ/kg

Octane Number
    Research Method   108-111
    Motor Method   95-98

Cetane Number

Toxicity
    Least Detectable Odor   About 1 ppm*
    Least Amount Causing Eye Irritation   Unknown
    Least Amount Causing Throat Irritation   Unknown
    Least Amount Causing Coughing
    Maximum Allowable for Prolonged Exposure   1% or more*
    Maximum Allowable for Short Exposure (0.5 hr)
    Dangerous for Short Exposure (0.5 hr)

Comments

Normal transportation is by bulk, or containers at moderate pressures.

*Commercial LPG is odorized with mercaptans; methyl mercaptans can
  be detected by odor at 0.002-0.005 ppm. LPG in high concentrations can
  act as an asphyxiant. Any toxic effects would result from contaminants
  (gases other than propane, propylene, butane, or ethane).

VEGETABLE (Cottonseed) OIL

Chemical Formula	Carbohydrate mixture 77.2% C, 12% H, 10.8% O	
Molecular Weight	Contains C_{16} (paluritic) and C_{18} (oleic and linoleic) fatty acids	

	English Units	Metric (SI) Units
Melting Point	23°-30°F	-5° to -1°C
Boiling Range	170°C and above make point	300°C
Density		
Vapor		
Liquid	56.94 lb/cu ft	912.18 kg/cu m
Specific Gravity		
Vapor		
Liquid	0.9125	0.9125
Heating Value, Vapor		
Volumetric Gross		
Volumetric Net		
Weight Gross	17,270 Btu/lb	4.017×10^4 kJ/kg
Weight Net	16,113 Btu/lb	3.748×10^4 kJ/kg
Heating Value, Liquid		
Volumetric Gross	983,354 Btu/cu ft	3663×10^4 kJ/cu m
Volumetric Net	917,474 Btu/cu ft	3418×10^4 kJ/cu m
Weight Gross	17,270 Btu/lb	4.017×10^4 kJ/kg
Weight Net	16,113 Btu/lb	3.748×10^4 kJ/kg
Air for Combustion, Vapor	———air component per unit of fuel———	
O_2 Volumetric		
N_2 Volumetric		
Air Volumetric		
O_2 Weight	2.896 lb/lb	2.896 kg/kg
N_2 Weight	9.538 lb/lb	9.538 kg/kg
Air Weight	12.434 lb/lb	12.434 kg/kg
Air for Combustion, Liquid	———air component per unit of fuel———	
O_2 Volumetric	1948.7 cu ft/cu ft	1948.7 cu m/cu m
N_2 Volumetric	7094.4 cu ft/cu ft	7094.4 cu m/cu m
Air Volumetric	9043.1 cu ft/cu ft	9043.1 cu m/cu m
O_2 Weight	2.896 lb/lb	2.896 kg/kg
N_2 Weight	9.538 lb/lb	9.538 kg/kg
Air Weight	12.024 lb/lb	12.024 kg/kg
Products of Combustion, Vapor	———unit of product per unit of fuel———	
CO_2 Volumetric		
H_2O Volumetric		
N_2 Volumetric		
CO_2 Weight	2.830 lb/lb	2.830 kg/kg
H_2O Weight	1.066 lb/lb	1.066 kg/kg
N_2 Weight	9.528 lb/lb	9.528 kg/kg
SO_2 Weight		

VEGETABLE (Cottonseed) OIL, Cont.

	English Units	Metric (SI) Units
Products of Combustion, Liquid	———unit of product per unit of fuel———	
CO_2 Volumetric	1377.426 cu ft/cu ft	1377.426 cu m/cu m
H_2O Volumetric	1275.691 cu ft/cu ft	1275.691 cu m/cu m
SO_2 Volumetric		
N_2 Volumetric	7293.154 cu ft/cu ft	7293.154 cu m/cu m
CO_2 Weight	2.830 lb/lb	2.830 kg/kg
H_2O Weight	1.066 lb/lb	1.066 kg/kg
N_2 Weight	9.528 lb/lb	9.528 kg/kg
SO_2 Weight		
Ash Weight		

Flammability Limits

Flash Point	486°F	252°C
Ignition Temperature	650°F	343°C

Heat of Vaporization

Octane Number
    Research Method
    Motor Method

Cetane Number

Toxicity*
Least Detectable Odor	Odorless
Least Amount Causing Eye Irritation	Unknown
Least Amount Causing Throat Irritation	Unknown
Least Amount Causing Coughing	
Maximum Allowable for Prolonged Exposure	Unknown
Maximum Allowable for Short Exposure (0.5 hr)	
Dangerous for Short Exposure (0.5 hr)	

Comments

Normal transportation is in bulk or by containers. Cottonseed oil contains fatty acids (C_{16}, C_{18} molecules) distributed in a complex glyceride structure. This oil is presented as an example of several vegetable oils that might be used in external-combustion engines. Such oils are corn oil, peanut oil, and soybean oil.

*Normally not considered toxic.

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APPENDIX B. Detailed Process Descriptions and Economics
for Candidate Fuels From Coal and Oil Shale

Gasoline and Distillate Fuels From Coal

Three processing routes are employed to manufacture liquid fuels
from coal:

1. Pyrolysis involves heating the coal to drive out the naturally occurring
   oils and volatile matter. The syncrude produced then is hydrotreated
   for quality improvement and desulfurization. Pyrolysis processes
   produce significant quantities of by-product gas and char that must
   be disposed of economically. Three processes based on this principle
   are under development: COED, TosCoal, and Garrett.

2. Selective hydrogenation of the coal involves destructive dissolution
   (usually in a hydrogen-donor solvent), ash filtration, and hydro-
   cracking to produce a liquid hydrocarbon fuel. In this process, pyritic
   sulfur goes with the ash, which is insoluble in the solvent, and
   organic sulfur goes with the liquid fuels because it is soluble in the
   solvent. The syncrude produced then is treated with hydrogen to
   improve its quality and, at the same time, to remove organic
   sulfur. This route is very energy-efficient compared with other
   methods for producing liquid fuels from coal. Four processes are
   under development: CSF, H-Coal, Synthoil, and SRC. In SRC, a
   solid fuel is produced if the syncrude is allowed to cool before
   hydrotreatment.

3. Gasification of coal is carried out to produce synthesis gas. After
   purification, the clean gas containing appropriate proportions of
   carbon monoxide and hydrogen is converted by the Fischer-Tropsch
   Process to hydrocarbon oil. The two chemical equations that
   generalize the formation of hydrocarbons in the Fischer-Tropsch
   synthesis are as follows:

$$nCO + 2nH_2 \rightarrow (CH_2)_n + nH_2O \qquad (B-1)$$

$$2nCO + nH_2 \rightarrow (CH_2)_n + nCO_2 \qquad (B-2)$$
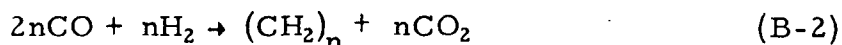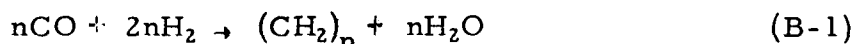
This process was demonstrated in the U.S. about 25 years ago. South
African Coal, Oil, and Gas Corp., Ltd, (SASOL) built a Fischer-
Tropsch synthesis plant at Sasolburg, South Africa, to produce liquid
hydrocarbons from coal in 1955. At present, the plant utilizes about
13,000 tons of coal/day (14.9 million Btu/ton of coal) and produces
71 million SCF/day of 525 Btu/SCF of pipeline gas, in addition to
9000 bbl/day liquid fuels. The process block diagram is given in
Figure B-1, and the following two paragraphs describe the process
in brief.

COAL → COAL PREPARATION → GASIFICATION → GAS CLEANUP → ARGE FIXED-BED SYNTHESIS → PRODUCTS SEPARATION → LIQUID PRODUCTS

OXYGEN, H₂S + CO₂, OIL, SYNTHESIS GAS → REFORMER, ARGE TAIL GAS → PRODUCT GAS, KELLOGG TAIL GAS, KELLOGG FLUID-BED SYNTHESIS → PRODUCTS SEPARATION → LIQUID PRODUCTS
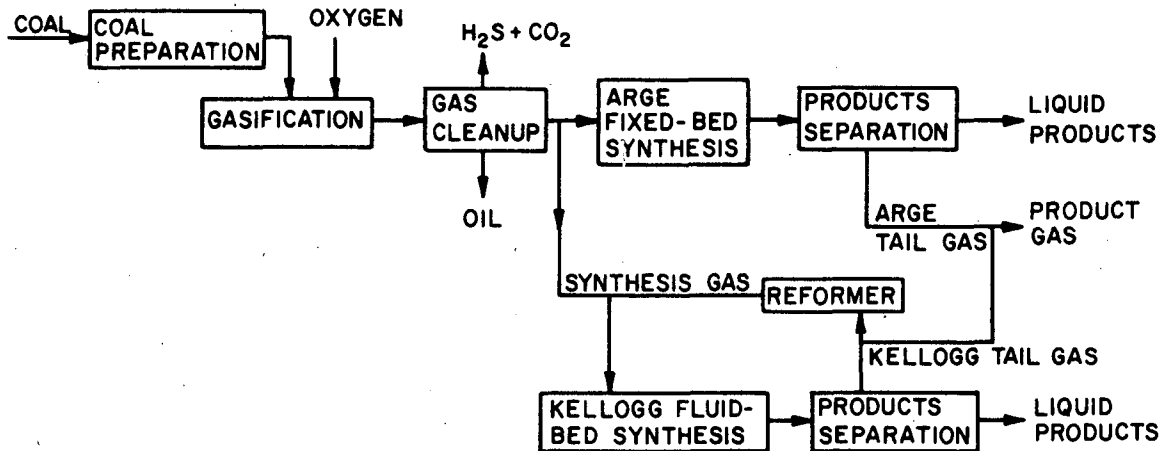
A-94-1628

Figure B-1.   FISCHER-TROPSCH SYNTHESIS AT SASOLBURG

Noncaking coal is crushed to 3/8 to 1-1/2 inches and dried. Dried coal is converted to gas at 350-450 psi in a Lurgi gasifier. The gas is quenched to remove tar and oil, and then carbon dioxide and hydrogen sulfide are removed to produce synthesis gas. A part of the synthesis gas is passed through a fixed catalyst bed contained in vertical tubes (Arge synthesis). Released heat is absorbed by boiling water outside the tubes. The feed gas has an hydrogen-to-carbon monoxide ratio of about 2, and the operating conditions are 430°-490°F and 360 psig. The ratio of recycled gas to fresh feed is about 2.4. The products of fixed-bed synthesis are straight-chain high-boiling-point hydrocarbons, with some intermediate-boiling-point oils, diesel oil, LPG, and oxygenated compounds.

The portion of gas that did not go to Arge synthesis goes to a fluid-bed reactor (Kellogg synthesis). A portion of the tail gas from the Kellogg fluid-bed synthesis is reformed with steam to increase the hydrogen-to-carbon monoxide ratio to about 3, and this gas is mixed with the fresh synthesis gas. In the fluid bed, catalyst is circulated along with the synthesis gas. The gas and catalyst leaving the reactor are separated in cyclones, and the catalyst is recycled. The operating conditions are 600°-625°F and 330 psig. The ratio of recycled gas to fresh feed is 2. Products from the fluid-bed synthesis are mainly low-boiling-point hydrocarbons (C_1-C_4) and gasoline, with little intermediate- and high-boiling-point material. Substantial amounts of oxygenated products and aromatics are produced. A portion of the fixed-bed tail gas and a portion of a fluid-bed tail gas are removed and used for utility gas.

The typical produce analysis for a fixed-bed process and a fluid-bed process given in Table B-1. The overall yield per ton of coal fed to the process is given in Table B-2.

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Table B-1. TYPICAL PRODUCTS OF SASOL PROCESS[2]

	Fixed-Bed Process	Fluid-Bed Process
	wt %	

Liquid Product Composition

Liquified Petroleum Gas (C_3-C_4)	5.6	7.7
Petrol (C_5-C_{11})	33.4	72.3
Middle Oils (diesel, furnace, etc.)	16.6	3.4
Waxy Oil or Gatsch	10.3	3.0
Medium Wax, mp 203°-206°F	11.8	
Hard Wax, mp 203°-206°F	18.0	
Alcohols and Ketones	4.3	12.6
Organic Acids	Traces	1.0

	Fixed-Bed Process		Fluid-Bed Process	
	C_5-C_{10}	C_{11}-C_{18}	C_5-C_{10}	C_{11}-C_{14}
	vol %			

Liquid Product Composition

Paraffins	45	55	13	15
Olefins	50	40	70	60
Aromatics	0	0	5	15
Alcohols	5	5	6	5
Carbonyls	Traces	Traces	6	5

Table B-2. PRODUCT YIELD OF SASOL PROCESS

Products	Yield, gal/ton
LPG	0.18
Gasoline	25.72
Kerosene	0.31
Diesel Fuel	2.56
Fuel Oil	0.63
Wax Oil and Wax	2.06
Methanol	0.11
Ethanol	2.17
Methyl Ethyl Ketone	0.15
Acetone	0.12
	34.01
Gas (500 Btu/SCF)	5500 SCF

In this study, the CSF Process (Consol Synthetic Fuel) has been selected for a detailed analysis because —

1. Not enough information is available for the SASOL Process in the following areas:
   a. Material and heat-transfer data for an energy balance to determine process efficiency
   b. Economic data.

2. The SASOL Process is designed to produce not only liquid fuels but also pipeline gas and many other by-products.

3. The SASOL Process route (Fischer-Tropsch) is less efficient than coal dissolution routes.

Description of CSF Process

The CSF Process has been developed by Consolidation Coal Company. A 70 ton/day pilot plant that was operational at Cresap, W. Va., for 40 months with less than 500 hours of operating time was shut down in April 1970 for a detailed study of process and operating problems. The process is designed to produce fuel oil and naphtha from coal. The liquid product can be catalytically reformed in a refinery to produce gasoline and No. 2 fuel oil. The process flow diagram is given in Figure B-2. The process setup and data required for this study have been taken from Reference 7.

Coal Preparation and Extraction

The raw coal is crushed in hammer-mill crushers and partially dried by contact with the flue gas. The partially dried coal is dried further in fluid-bed dryers. Fines smaller than 14 mesh are recovered in multiple-stage cyclones and bag filters. The crushed coal is combined with the recovered fines and heated to 450°F in fluid-bed dryers to remove any remaining moisture. The preheated coal then is slurried with a coal-derived solvent and pumped at 150 psig through a tubular furnace, where it is heated to extraction temperature, 765°F. Extraction mainly occurs in a stirred extraction vessel. The hot vapor from the extractor is sent to the solvent recovery area, and the slurry phase is sent to a residue separation section.

Residue Separation and Solvent Recovery

The untreated coal residue is removed from the slurry in the residue separation section by two-stage hydroclones. Overflow from the first

Figure B-2. FLOW DIAGRAM OF CSF-PROCESS PRODUCTION OF
GASOLINE (50,000 bbl/Day) FROM COAL

stage goes to the solvent recovery section, and the underflow passes to a second stage. The overflow from the second stage is fed back to the first stage; the underflow is sent to the low-temperature carbonization (LTC) system.

Solvent recovery is divided into two sections. The vapor from the extraction section is condensed, the gaseous stream is sent to the gas-cleanup section, and the recovered solvent is returned to the slurry mix tanks. The hydroclone overflow from residue separation is fractionated in a vacuum still. The distillate (including spent solvent and fuel oil) is taken overhead from the fractionator, and a heavier cut from a side stream provides most of the recycle solvent for the extraction section. The bottoms, which contain the extract, residue, and tar, are sent to the extract hydroconversion.

Low-Temperature Carbonization

The hydroclone underflow from residue separation is pumped to the carbonizer, where it is reacted with steam and air. The overhead gaseous product from the low-temperature carbonizer is quenched, and a gas stream is separated from a solvent-tar stream. The solvent-tar stream is delivered to the tar distillation section, and the gas stream is used as a plant fuel after sulfur removal. Char from the LTC section is delivered to the Lurgi gasification system for hydrogen production.

Tar Distillation and Extract Hydroconversion

The heavy liquids from the LTC section are vacuum-distilled in the tar distillation section. The overhead product is distillate fuel, and the bottoms are sent to the residue separation area. The extract from the solvent recovery section is hydrotreated to produce the donor solvent and product oil. Extract hydrogenation is done in four stages operating at 3000 psig and 800°-825°F temperature in the presence of a Co-Mo-Ni catalyst. The overhead vapors are cooled to separate hydrogen from the light oils. The recovered hydrogen is compressed and recycled to the reactors. The hydrotreated liquid product is flashed to 5 psig. The fuel gas is sent to the gas-treatment plant, and gas liquor is sent to the waste-water-treatment plant to recover ammonia and hydrogen sulfide. The hydrotreated liquid product is stabilized to remove C_4 and lighter hydrocarbons and then is fractionated. In the fractionator, hydrogen donor solvent is separated from the light product oil. The hydrogen

donor solvent is sent to the slurry system for makeup solvent, and light product oil is delivered to the refinery to produce gasoline.

Gas Treatment and Sulfur Recovery

The fuel gas is produced at various sections of the plant. This gas is treated in the amine system to remove carbon dioxide and hydrogen sulfide. Some of the fuel gas is used as a fuel for the plant and refinery operation, and the remaining amount is by-product for sale. The hydrogen sulfide stream from the amine system is passed through the sulfur recovery system. The sulfur recovery system consists of a modified Claus plant and a Beavon tail-gas plant.

Hydrogen Production

Hydrogen can be produced from the coal char. To form the char into the proper size pellets, a mixture of tar, char, and dried coal must be coked. These pellets are fed to the gasifier. The raw gas produced is passed through quench systems, shift systems, an acid-gas-removal section, and a methanation section. The gas containing hydrogen and methane is passed through cryogenic separation units to produce a 96%-pure hydrogen stream and a methane-rich stream. A hydrogen stream is used to hydrotreat the extract, and the methane-rich stream is used as a fuel gas. Tables B-3, B-4, and B-5 show the composition of the important streams for a 50,000 bbl/day plant.

Gasoline Production

Gasoline can be produced in the refinery from naphtha and distillate fuel, as shown in Figure B-3. The distillate fuel is first hydrocracked and sent to the distillation section along with naphtha. The 180°-400°F fraction is sent to a reformer to increase the gasoline octane number. Some butane, the C_5-180°F fraction, and reformate (C_5-400°F) are blended, and then some tetraethyl lead is added to meet the specified octane number. The product gas is utilized in the production of hydrogen. Some additional fuel gas is required for plant fuel and hydrogen production.

Table B-3. COMPOSITION OF GASEOUS STREAMS FROM CSF PROCESS*

					Stream Number				
	1	2[†]	3	4	5	6	7	8	9
					mol %				
CO	2.48	0.20	8.42	--	8.49	--	--	--	--
CO_2	15.82	--	12.18	--	12.28	--	--	--	--
H_2	4.95	37.90	--	96.02	--	35.34	--	--	--
H_2O	--	--	5.02	--	5.06	--	--	100.00	--
CH_4	20.79	50.20	4.81[‡]	2.48	--	36.13	--	--	--
C_2H_6	9.34	5.60	--	--	4.85	10.83	--	--	--
C_3H_8	3.09	3.40	--	--	--	8.28	--	--	--
C_4H_{10}	6.80	--	--	--	--	--	--	--	--
H_2S	36.73	--	0.79	--	--	3.31	--	--	--
N_2	--	2.70	68.45	1.50	68.99	6.11	79.00	--	0.54
O_2	--	--	0.33	--	0.33	--	21.00	--	99.46
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Mol/hr	1934.3	13,061	15,173	32,548	15,053	8005	13,087	47,593	11,521
10^6 SCF/hr	0.73	4.95	5.75	12.34	5.71	3.03	4.96	18.04	4.37

*See Figure B-2.

†Dry gas.

‡Assumed C_2H_6.

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Table B-4. COMPOSITION OF LIQUID STREAMS FROM CSF PROCESS[*]

	Stream Number							
	1	2	3	4	5	6	7	8
	wt %							
Butane	2.00	--	--	--	--	--	--	--
Light Oil	--	--	--	100.00	--	--	--	--
Naphtha	--	--	--	--	--	8.19	--	7.75
Hydrodistillate, $C_4/400^0$ F	--	--	--	--	100.00	91.81	--	86.81
Tar Oil, -230^0 C	--	--	--	--	--	--	1.35	--
Tar, $+230^0$ C	--	3.92	4.04	--	--	--	--	--
Extract	--	--	77.66	--	--	--	--	--
Residue	--	--	3.47	--	--	--	--	5.44
Spent Solvent, $400^0/750^0$ F	--	--	--	--	--	--	96.38	--
Solvent, $400^0/750^0$ F	98.00	96.08	14.83	--	--	--	--	--
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
lb/hr	60,783	1,097,742	1,063,100	7867	125,417	136,608	497,017	144,475

[*] See Figure B-2.

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Table B-5. COMPOSITION OF SOLID STREAMS FROM CSF PROCESS[*]

	Stream Number						
	1	2	3	4	5	6	7
	wt %						
C	59.04	68.97	67.59	53.67	51.93	--	84.58
H	4.19	4.90	4.80	2.52	1.80	--	6.36
N	1.10	1.28	1.26	1.25	1.29	--	1.09
O	6.28	7.34	7.19	3.85	3.25	--	0.74
S	3.67	4.29	4.20	5.03	4.80	--	0.80
Moisture	14.40	--	2.00	--	--	--	--
Ash	11.32	13.22	12.96	33.68	36.93	100.00	6.43
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00
lb/hr	2,436,750	1,666,667	427,708	617,300	563,083	263,450	193,467
Extract, lb/hr	--	--	--	66,892	--	--	--
Solvent, lb/hr	--	--	--	995,175	--	--	--
Tar, lb/hr	--	--	--	--	--	--	26,192
Total	2,436,750	1,666,667	427,708	1,679,367	563,083	263,450	219,659

* See Figure B-2.

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Figure B-3. FLOW DIAGRAM OF 50,000 bbl/DAY
GASOLINE REFINERY

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Overall Energy Balance and Efficiencies

The overall energy balance is presented in Table B-6.

Table B-6. ENERGY BALANCE FOR CSF-PROCESS COAL-TO
GASOLINE (50,000 bbl/Day) PLANT

	10^6 Btu/hr
Input	
Coal (as received)	
(1218.4 tons/hr X 2000 X 10,820 Btu/lb)	26,365
Output	
Gasoline (2076.7 bbl/hr X 5.3 million Btu/bbl)	11,007
Butane (187.5 bbl/hr X 4.3 million Btu/bbl)	806
Fuel Tar (4.95 SCF/m X 819 Btu/SCF)	4,054
Sulfur (43.68 tons/hr X 2000 X 398.3 Btu/lb)	348
Ammonia (5.37 tons/hr X 2000 X 9675 Btu/lb)	104
Cooling by Air and Water	3,155
Other (by Difference)*	6,891
Total	26,365

* Includes sensible heat of product streams, heating values of other
unaccounted products, and heat lost to the atmosphere.

The overall efficiency (including by-product credit) of the process is about
61%, and the coal-to-gasoline efficiency is about 42%. The overall efficiency
of the CSF Process producing naphtha and distillate fuel is about 67%, and the
efficiency of the refinery is about 91%. This 67% efficiency of the CSF Process
is achieved by using the Lurgi gasification system for the production of hydro-
gen. However, on the basis of the information given in Ref. 2, if the BI-GAS
gasification system instead of Lurgi is used for the production of hydrogen,
the overall efficiency can be improved by 4%. The use of the catalytic cracking
unit instead of the high severity hydrocracking unit could improve the efficiency
of the refinery section. However, the use of the catalytic cracker produces
more distillate fuel.

Pollution

About 93% of the total sulfur is recovered as elemental sulfur
in this process by using a Claus plant with a Beavon tail-gas process. About
5% of the total sulfur is recovered as elemental sulfur by using iron oxide
towers. The reaction of iron oxide with hydrogen sulfide forms iron sulfide
and water. The sulfur from the iron sulfide can be recovered by using a
suitable solution. The sulfur balance around the system is given in Table B-7.

Table B-7. SULFUR BALANCE FOR CSF-PROCESS COAL-TO-
GASOLINE (50,000 bbl/Day) PLANT

	lb/hr (as sulfur)
Input	
Coal	89,482
Output	
Elemental Sulfur (By-product)	87,367
Sulfur in Liquid Products	24
Sulfur in Refinery Off-Gas	580
Sulfur Compounds to Atmosphere From Sulfur Recovery Plants	80
Sulfur in the Stack Gas	1,431
Total	89,482

The gas liquor, containing mainly ammonia and hydrogen sulfide, is treated in the Chevron waste-water-treating process. The 99.9%-pure hydrogen sulfide stream and the 99%-pure ammonia stream are recovered. The treated water contains less than 100 ppm ammonia and less than 20 ppm hydrogen sulfide. Some of the effluent water, which may contain phenols (about 30 mg/liter), is treated by biological oxidation. The process requires about 120,000-180,000 gpm of cooling water. Table B-8 lists the wastes, their sources, and the treatments required.

Table B-8. WASTES, SOURCES, AND TREATMENTS
FOR A COAL-TO-GASOLINE PLANT

Waste	Sources	Treatment
Coal Dust	Coal-crushing system, conveyor belt	Cyclone separators, bag filter
Ash	Gasifier	Scrubbing and various waste-water and solid treatments
Waste Water (Contains Ammonia, Hydrogen Sulfide)	Residue separation Solvent recovery Extract hydroconversion system Gasification system to produce hydrogen Refinery	Modified Chevron Process, Phenosolvan Process, biological treatments
Hydrogen Sulfide	Regenerator off-gas	Claus plant with Beavon tail-gas process, Stretford Process, etc.
Sulfur Dioxide	Gas and liquid fuels fired boilers, heaters, and incinerators	Lime treatment, Wellman-Lord, etc.

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Economic Analysis

The economic analysis is performed by using the DCF method. The investment and operating costs of a 50,000 bbl/day gasoline plant are given in Tables B-9 and B-10, respectively. The investment and operating costs in Ref. 2 are based on those in 1972; therefore, these numbers have been escalated to bring the costs to end-of-1973 levels. In this study, for the production of hydrogen, a Lurgi gasification system is used, but according to Ref. 2, a BI-GAS gasification system is somewhat cheaper compared with a Lurgi system. According to our estimate, the investment cost of a refinery for producing gasoline from naphtha and distillate is about $118 million. This cost can be reduced by using a catalytic cracking unit rather than a hydrocracking unit. However, the use of a catalytic cracking unit produces by-product distillate fuel, so the amount of the gasoline produced is reduced. The calculation method[8] for the unit cost of the product is presented in Table B-11. This financing method includes the following factors:

- A 25-year project (synthesis plant) life
- Depreciation calculated on a 16-year sum-of-the-digits formula
- 100% equity capital
- A 48% Federal Income Tax rate
- A 12% DCF
- Plant start-up costs as expenses in year zero.

For 30¢/million Btu coal, the production cost of the gasoline is about $13.85/bbl, or about $0.33/gal. This unit cost will vary depending on the accounting method used, feed cost, and variation in other financial factors; e.g., by using the utility method,[23] the unit cost of the product is $10.61/bbl, or $0.253/gal.

Table B-9. INVESTMENT COST FOR CSF-PROCESS COAL-TO-GASOLINE (50,000 bbl/Day) PLANT

Components	End-of-1973 Cost, $1000
Coal Preparation	14,024
Extraction	11,002
Residue Separation	4,953
Solvent Recovery	6,783
Low-Temperature Carbonization	11,938
Tar Distillation	2,758
Extract Hydroconversion and Distillation	52,693
Fuel Gas Compression and Treatment	9,354
Hydrogen Production System (using Lurgi gasification system)	130,425
Refinery and General Facilities for Refinery	117,480
Sulfur Recovery System	8,599
Waste-Water Treatment	1,739
Boiler Feed Water Treatment, Steam, and Power Generation System	28,377
Cooling Towers and Pumps	16,500
Initial Catalyst and Chemicals	1,813
Power Distribution	3,030
General Facilities	28,515
Total Direct Cost of Plant Including Contractor's Overhead and Profit	449,983
Contingencies (15%)	67,497
Total Plant Investment (I)	517,480
Interest During Construction (0.23676 X I)	122,518
Start-up Cost (20% of gross operating cost)	24,332
Working Capital	
Coal Inventory (60 days of feed at full rate)	11,390
Materials and Supplies (0.9% of total plant investment)	4,657
Net Receivables (1/24 X annual revenue received)	9,448
Total Capital Required	689,825

Table B-10. OPERATING COST FOR CSF-PROCESS COAL-TO-GASOLINE (50,000 bbl/Day) PLANT (90% Stream Factor)

Components	Annual Cost, $1000
Coal Feed (at 26,365 X 10^6 Btu/hr), 30¢/10^6 Btu	62,358
Other Direct Materials, Catalysts, and Chemicals	6,954
Purchased Utilities	
Raw-Water Cost (15,000 gpm X 30¢/1000 gal)	1,220
Electric Power (50,033 kWhr/hr X 0.9¢/kWhr)	3,550
Labor	
Process Operating Labor (131 men/shift at $5/hr and 8304 man-hr/yr)	5,439
Maintenance Labor (1.5% of total plant investment)	7,762
Supervision (15% of operating and maintenance labor)	1,967
Administration and General Overhead (60% of total labor, including supervision)	9,046
Supplies	
Operating (30% of process operating labor)	1,632
Maintenance (1.5% of total plant investment)	7,762
Local Taxes and Insurance (2.7% of total plant investment)	13,807
Total Gross Operating Cost	121,662
By-product Credit	
Butane (4499 bbl/day X 42 X 10¢/gal)	6,207
Fuel Gas (4054 X 10^6 Btu/hr X 24 X 50¢/10^6 Btu)	15,981
Sulfur (39 LT/hr X $10/LT X 24)	3,075
Ammonia (128.9 tons/day X $25/ton)	1,059
Total By-product Credit	26,322
Net Operating Cost	95,340

Table B-11.   CALCULATION FOR DETERMING UNIT PRODUCTION COST BY DCF METHOD FOR CSF-PROCESS COAL-TO-GASOLINE (50,000 bbl/Day) PLANT

Unit Cost of the Product

$$\frac{N + 0.23816\ I + 0.1275\ S + 0.23077\ W}{G}$$

where

N = Net Operating Cost = $95,340,000

I = Total Plant Investment = $517,480,000

S = Start-up Cost = $24,332,000

W = Working Capital = $25,495,000

G = Annual Product Production (49,840 bbl/day X 328.5 days/yr)

Unit Cost = $\dfrac{\$227,568,000}{16,372,440\ bbl}$ = $13.900/bbl

= $0.3309/gal

= $2.612/$10^6$ Btu (high heating value)

= $2.808/$10^6$ Btu (low-heating value)

Table B-12.   CALCULATION FOR DETERMINING UNIT PRODUCTION COST BY DCF METHOD FOR CSF-PROCESS COAL-TO-GASOLINE-PLUS-DISTILLATE-OIL (50,000 bbl/Day) PLANT

Unit Cost of the Product

$$\frac{N + 0.23816\ I + 0.1275\ S + 0.23077\ W}{G}$$

where

N = Net Operating Cost = $90,303,000

I = Total Plant Investment = $491,274,000

S = Start-up Cost = $23,325,000

W = Working Capital = $24,811,000

G = Annual Product Production (49,840 bbl/day X 328.5 days/yr)

Unit Cost = $\dfrac{\$216,005,000}{16,372,440\ bbl}$ = $13.193/bbl

= $0.3141/gal

= $2.356/$10^6$ Btu (high heating value)

= $2.511/$10^6$ Btu (low heating value)

If, instead of gasoline (primarily), the coal oil is refined to produce approximately a 50:50 product mix of gasoline and distillate oil (plus by-products of tar, sulfur, ammonia), the resulting costs are reduced somewhat. In this case, a catalytic cracking unit (instead of a hydrocracker) is used in the refinery, and other refinery investments costs could be reduced slightly. Certain operating costs also would be reduced.

The investment cost for the "Refinery and General Facilities for Refinery" becomes about $100 million; the "Total Direct Cost of Plant" becomes $427,195,000, and the "Total Plant Investment" becomes $491,274,000. Labor and supply costs also are reduced slightly and result in a "Total Gross Operating Cost" of $116,625,000, and the "Net Operating Cost" becomes $90,303,000. Table B-12 presents the corresponding DCF unit costs for a 50,000 bbl/day gasoline-plus-distillate-oil plant (from coal). The exact division of costs between gasoline and distillate oil has not been made (and would be arbitrary within the scope of this study).

If a 10% (instead of 12%) DCF financing model is used for the synthesis plant to produce gasoline and distillate oils from coal, the unit cost of the product becomes $2.27/million Btu (low heating value), rather than $2.51/million Btu.

Gasoline and Distillate Fuels From Oil Shale

The technology for the production of gasoline from oil shale exists today. The major steps required are as follows:

- Shale mining, crushing, and screening, and spent shale disposal
- Shale retorting to produce crude shale oil
- Crude shale oil upgrading to syncrude (to make it acceptable as a conventional petroleum refinery feedstock and to facilitate handling in pipelines)
- Refining of syncrude to produce gasoline and light distillates.

The final gasoline cost is strongly dependent on both the richness (organic matter content) of the oil shale and the type of mining employed. Flow sheets for the production of gasoline and light distillates from shale oil are given in Figure B-4 and Table B-13.

Description of Gas Combustion Process

Mining Step

The first and physically largest part of the process comprises mining, crushing, and screening and spent shale disposal. Both underground and strip or open-pit surface mining have been considered for mining oil shale in the Western United States. The areas considered — the Piceance Creek Basin of Colorado and Uinta Basin of Colorado and Utah — are shown in Figure B-5.

Two basic methods of underground mining are envisioned — shaft and adit entry. Adit entry would be used when the shale formation outcrops the surface. In this case, the formation can be mined by tunneling from the site of the outcropping. In the event of no outcropping, a shaft must be sunk. Because the number of areas with shale outcroppings is small compared with the number that would require shaft mines, adit mining was not considered here.[5] Nor was either method of surface mining (strip or pit) considered here, because we assumed that, to be economically attractive the first plants would be based on the high-quality resources for which underground mining generally is required.

Retorting Step

A number of processes for converting oil shale to oils have been studied on a rather large scale; Table B-14 lists those processes being seriously considered now.

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SHALE OIL AND GAS PRODUCTION | CRUDE SHALE UPGRADING TO SYN-CRUDE | SYN-CRUDE REFINING
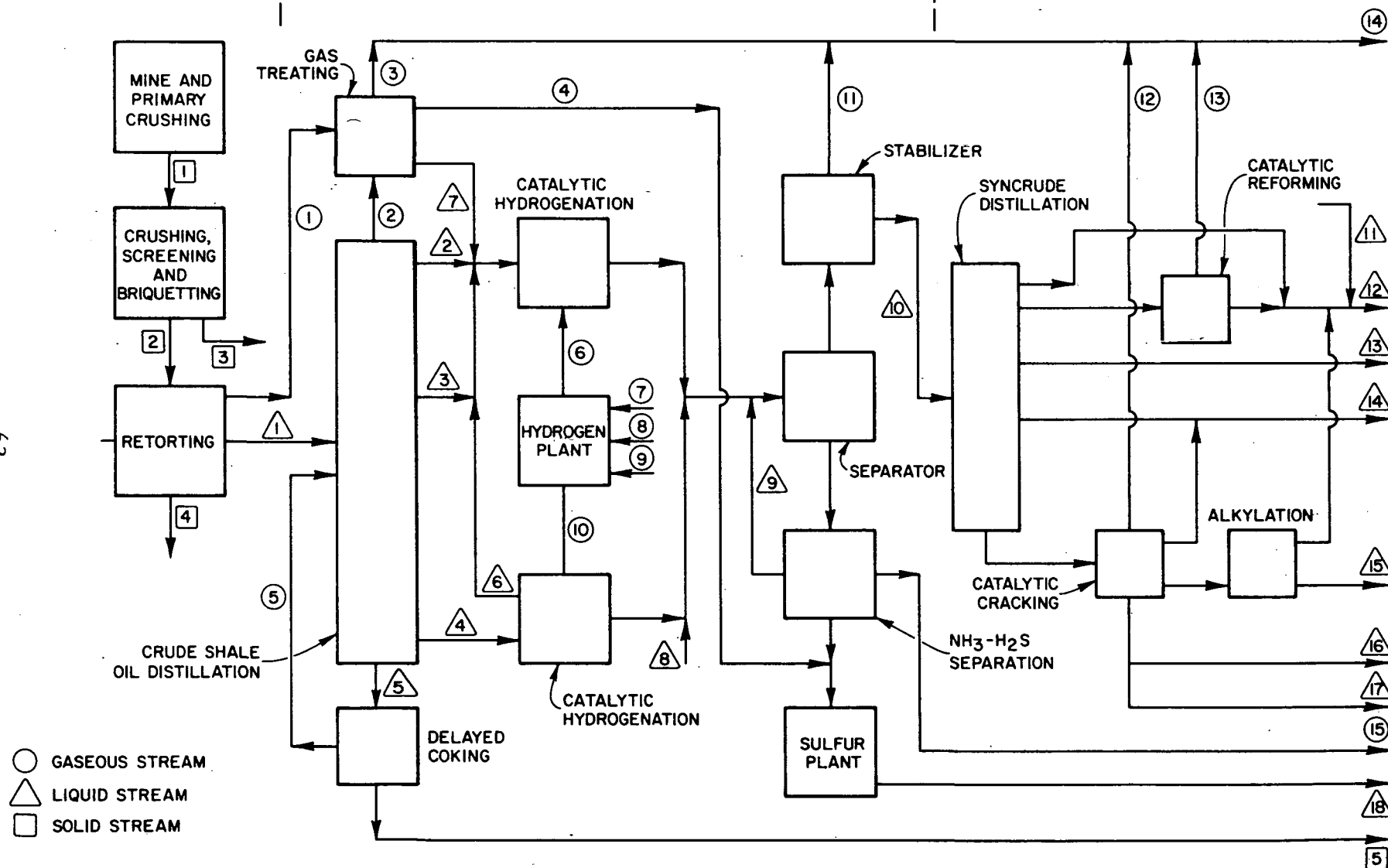
MINE AND PRIMARY CRUSHING

CRUSHING, SCREENING AND BRIQUETTING

RETORTING

GAS TREATING

CATALYTIC HYDROGENATION

HYDROGEN PLANT

CRUDE SHALE OIL DISTILLATION

DELAYED COKING

CATALYTIC HYDROGENATION

STABILIZER

SEPARATOR

NH₃-H₂S SEPARATION

SULFUR PLANT

SYNCRUDE DISTILLATION

CATALYTIC REFORMING

CATALYTIC CRACKING

ALKYLATION

○ GASEOUS STREAM
△ LIQUID STREAM
□ SOLID STREAM

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Figure B-4. FLOW DIAGRAM FOR PRODUCTION OF GASOLINE AND
LIGHT DISTILLATE (50,000 bbl/Day) FROM OIL SHALE

D-54-820

Table B-13. PROCESS STREAMS FROM PRODUCTION OF GASOLINE AND
DISTILLATE FUELS FROM OIL SHALE*

Stream No.	Gaseous Streams		Liquid Streams		Solid Streams	
	Description	Flow, tons/CD[†]‡	Description	Flow, bbl/CD[†]	Description	Flow, tons/CD[†]
1	Fuel gas	508	Crude shale oil	57,083	Run-of-mine shale	85,780
2	Light gases	--	Naphtha	--	Sized Shale	84,650
3	Fuel gas	--	Light oil	--	Dust loss	1,130
4	Hydrogen sulfide	--	Heavy oil	--	Spent shale	65,713
5	Gas plus oil vapors	--	Resdual oil	--	Coke	932
6	Process hydrogen	--	Light oil and naphtha	--	--	--
7	Steam	--	Naphtha	--	--	--
8	Process gas	--	Water	--	--	--
9	Fuel gas	--	Water	--	--	--
10	Process hydrogen	--	Syncrude	54,521	--	--
11	Fuel gas	--	Gasoline additives	--	--	--
12	Fuel gas	--	Motor gasoline	25,193	--	--
13	Fuel gas	--	Jet fuel	4,471	--	--
14	Plant fuel gas	--	Distillate fuel	20,336	--	--
15	Ammonia	150	i-Butane to sales	1,613	--	--
16	--	--	Decant oil to sales	191	--	--
17	--	--	Decant oil to plant fuel	--	--	--
18	--	--	Sulfur, tons/CD	47	--	--
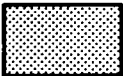
* See Figure B-4.
† CD = calendar-day.
‡ In 10^6 SCF/day.
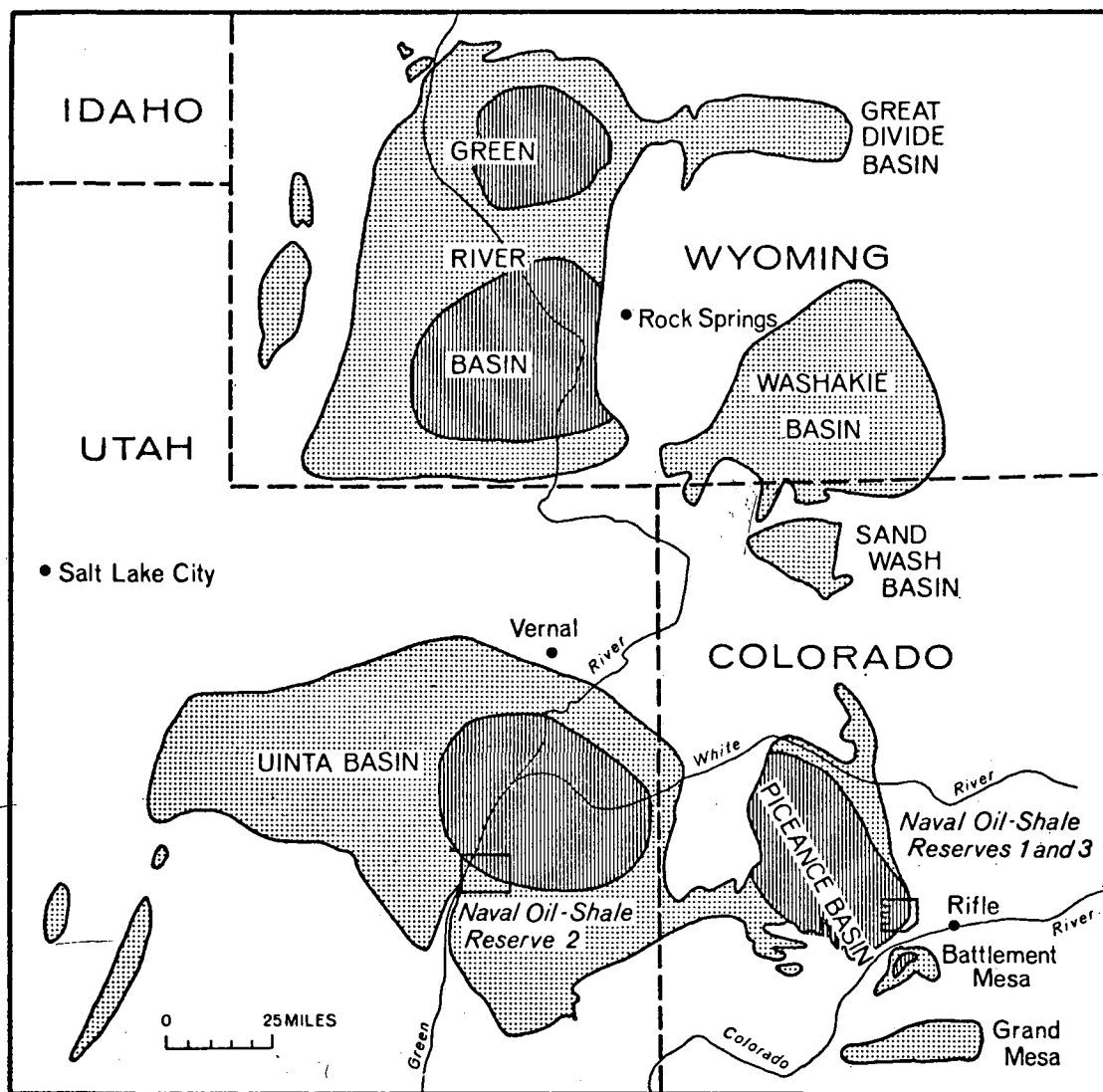
EXPLANATION

Area underlain by the Green
River Formation in which the
oil shale is unappraised or
low grade

Area underlain by oil shale
more than 10 feet thick, which
yields 25 gallons or more oil
per ton of shale

Figure B-5. GREEN RIVER OIL SHALE FORMATION
OF COLORADO, UTAH, AND WYOMING (Source: Ref. 5)*

Table B-14. CURRENT OIL-SHALE-RETORTING TECHNOLOGY

I. Processes Requiring Mining of Oil Shale

    A. Processes employing hot solids to supply the heat required for
       retorting
          1. TOSCO II Process (Colony Development Co.)
          2. Lurgi-Ruhrgas Process

    B. Processes employing internal gas combustion within the retort to
       supply heat
          1. Gas Combustion Process (U.S. Bureau of Mines)
          2. Paraho Process (Development Engineering)
          3. Rock-Pump Process (Union Oil Co.)

    C. Processes employing external heat generation
          1. Modified Paraho Process (Development Engineering)
          2. Modified Rock-Pump Process (Union Oil Co.)
          3. Petrosix Process (Petrobras)
          4. IGT Process

II. In Situ Retorting Processes

    A. U.S. Bureau of Mines

    B. Occidental Petroleum Company

Separate flow sheets and economic studies for each of these processes were not practical within the scope of this program. However, a number of excellent reviews recently have been published describing all the various oil shale conversion processes.[7,9]

We selected the Gas Combustion Process for study here; adequate engineering data are available for detailed assessments. This process was developed by the U.S. Bureau of Mines. Although operational difficulties were encountered with the equipment used in tests prior to 1955, work was done in a demonstration plant with a capacity as large as 150 tons/day. The Bureau's large-scale equipment also was operated later by six petroleum companies, and testing was continued until 1967 through a lease agreement with the Colorado School of Mines Research Foundation. In the latter tests, shale feed rates as high as 360 tons/day were achieved.

A simplified flow sheet of the process is given in Figure B-6. The heart of this process is the retort itself. Here, raw shale in the 1-1/4 to 3-inch size range is fed countercurrently to hot recycled product gas in a vertical, refractory-lined retort. At the top of the retort, where the raw shale enters, is a product cooling zone in which the raw shale is partially heated and the hot
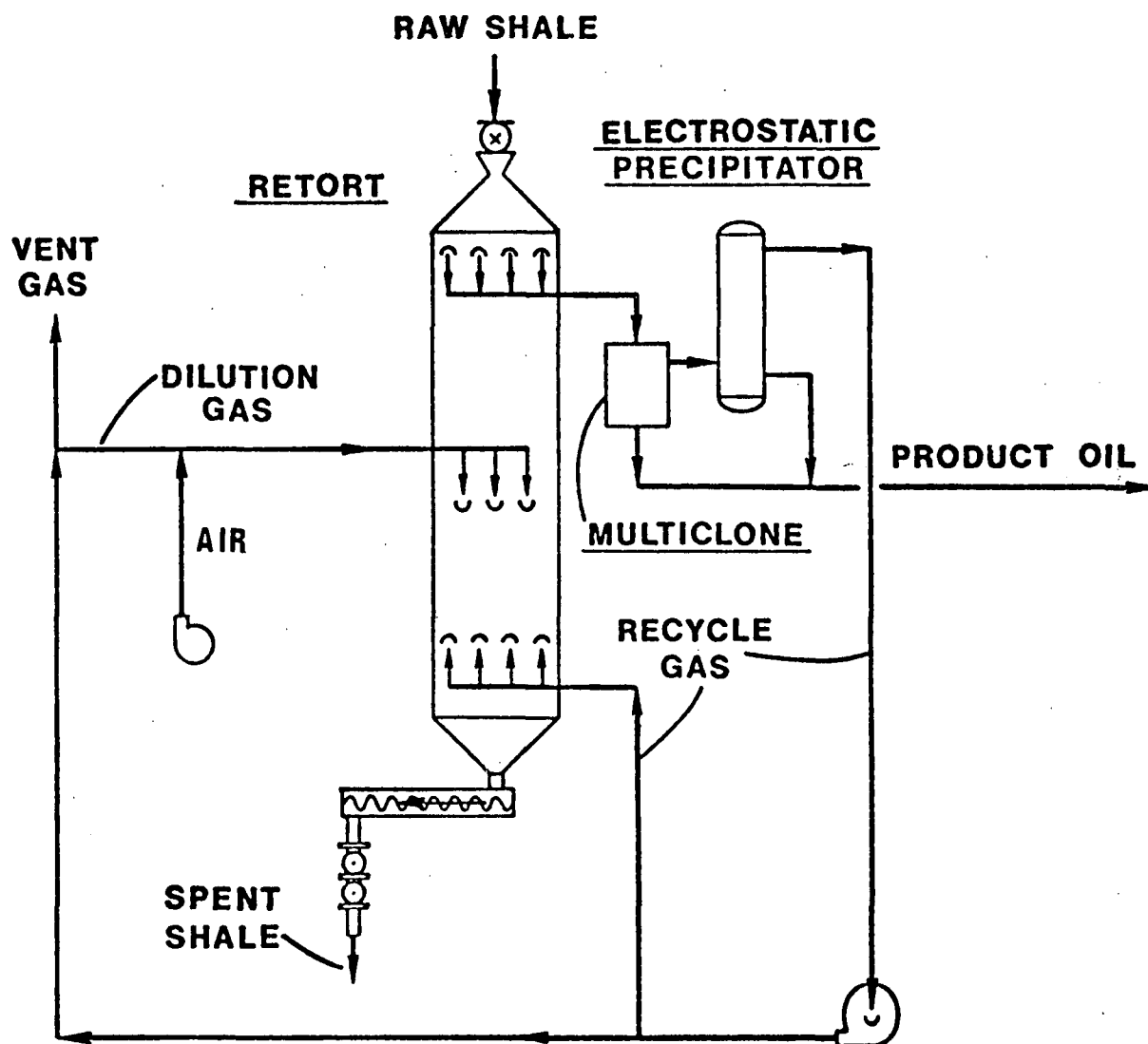
Figure B-6. FLOW DIAGRAM OF GAS COMBUSTION PROCESS DEVELOPED
BY U.S. BUREAU OF MINES (Source: Ref. 4)

gaseous and vaporized liquid products from retorting are cooled. The next zone of the retort is the retorting zone in which the shale is finally heated to retorting temperature by hot flue gas injected below this zone. This flue gas is generated by burning a portion of the recycle gas with air in a combustion zone. Some of the organic matter in the spent shale is also burned here to provide part of the heat requirement. The spent shale then travels downward through a heat recovery zone in which the hot solids transfer their heat to recycle gas flowing countercurrently upward. The primary products of this process are crude shale oil, a low-Btu product gas, and spent shale. Typical yields and product properties are given in Tables B-15 and B-16, respectively.

Upgrading of Crude Shale Oil

Crude shale oil presents two major problems: First, its high viscosity and pour point make transport by pipeline difficult. Second, it has a very high nitrogen content, so if shale oil were a large fraction of the refinery feed, existing refinery processes could not use it directly. Therefore, crude shale oil probably would be upgraded at the production site before being refined.

Techniques for upgrading crude petroleum fractions are applicable to crude shale oils. The most likely process is catalytic hydrotreating, which converts the nitrogen compounds to ammonia and the sulfur compounds to hydrogen sulfide. In the process, not only is the sulfur content reduced to a very small value, but the specific gravity and viscosity are reduced. Because the crude shale oil must be distilled into various fractions before the catalytic hydrotreating step, the oil is effectively thermally treated; this sufficiently improves the pour point of the material so that it can be transported by pipeline.

The major steps in a typical upgrading stage are shown in Figure B-4. This is only one possible approach, however, and is based largely on the flow sheet given in the NPC study.[5] First, the crude shale oil must be distilled to remove the heavy end fractions that could not be upgraded sufficiently. These heavy end fractions are sent to a delayed coker with the coke (by-product) and gas plus oil vapors (recycled to process) produced. Because different conditions are required for hydrotreating the light and the heavy fractions, each is sent to a separate hydrotreating stage. After treating, these two streams are combined, and the water, hydrogen sulfide, and ammonia are separated. Water is added to wash out the ammonia and hydrogen sulfide. After fractionation,

Table B-16. PROPERTIES OF TYPICAL CRUDE SHALE OIL

Specific Gravity, °API	28.0
Pour Point, °F	75
Sulfur Content, wt %	0.8
Nitrogen Content, wt %	1.7
Viscosity, SSU at 100°F	120

Analysis of Fractions

Butanes + Butanes, vol % of total	4.6
C_5 to 350°F Naphtha	
Vol % of Total Oil	19.1
Specific Gravity, °API	50.0
Sulfur Content, wt %	0.70
Nitrogen Content, wt%	0.75
350°-550°F Distillate	
Vol % of Total Oil	17.3
Specific Gravity, °API	31.0
Sulfur Content, wt %	0.80
Nitrogen Content, wt %	1.35
550°-850°F Distillate	
Vol % of Total Oil	33.0
Specific Gravity, °API	21.0
Sulfur Content, wt %	0.80
Nitrogen Content, wt %	1.90
Above 850°F Residue	
Vol % of Total Oil	26.0
Specific Gravity, °API	12.0
Sulfur Content, wt %	1.0
Nitrogen Content, wt %	2.4

Table B-15. TYPICAL RETORTING PRODUCT YIELDS*

Component	Semiworks Plant	Fischer Assay Product Balance
	— Yield, lb/100 lb Fischer Assay Oil—	
Oil		
Butanes and Butenes	2.19	1.84
C_5-C_8 Hydrocarbons	4.06	2.30
Fischer Assay Oil	99.59	100.00
Total C_4 + Oil	105.84	104.14
Gas		
H_2, CO, C_1-C_3	10.78	11.72
CO_2	8.58	9.14
H_2S	1.34	1.14
Total Gas	20.70	22.00
Total Gas and Oil	126.54	126.14

* Based on TOSCO II Process data.

the water is recycled, the ammonia is liquefied for storage and sale, and the hydrogen sulfide is sent to a Claus-type sulfur plant, where it is converted to elemental sulfur and sold. Hydrogen for hydrotreating is made by catalytic steam reforming of natural gas or light naphtha. The properties of the final syncrude are given in Table B-17. As shown, there is no fraction boiling above 850°F.

Table B-17. PROPERTIES OF TYPICAL SYNCRUDE

Specific Gravity, ° API	46.2
Pour Point, °F	50
Sulfur Content, wt %	0.005
Nitrogen Content, wt %	0.035
Reid Vapor Pressure, psi	8
Viscosity, SSU at 100°F	40

Analysis of Fractions

Butanes and Butenes, vol % of total	9.0

C_5 to 350° F Naphtha
Vol % of Total	27.5
Specific Gravity, °API	54.5
Sulfur Content, wt %	<0.0001
Nitrogen Content, wt %	0.0001

350°-550°F Distillate
Vol % of Total Oil	41.0
Specific Gravity, ° API	38.3
Sulfur Content, wt %	0.0008
Nitrogen Content, wt %	0.0075

550°-850°F Distillate
Vol % of Total Oil	22.5
Specific Gravity, °API	33.1
Sulfur Content, wt %	< 0.01
Nitrogen Content, wt %	0.12

Syncrude Refining

Because at this point the shale oil is upgraded to the equivalent of a completely desulfurized crude, the next step is assumed to be a conventional petroleum refining.

The light overhead stream is the equivalent of straight-run gasoline and needs no further treatment other than the option of additives. The next heavier stream goes to catalytic reforming for upgrading to gasoline with a fuel-gas

by-product. The next two streams are jet fuel and distillate fuel. The bottoms product goes to catalytic cracking, which results in several products: more fuel gas, more distillate fuel, decant oil (one stream to sales and one stream to plant fuel), and C_3 and C_4 olefins to an alkylation unit. The products from the alkylation unit are isobutane and alkylate for more motor gasoline.

Overall Energy Balance and Efficiencies

The energy balance for the synthesis process is presented in Table B-18. The efficiency (including by-product heat credit except heat credit of coke) of the process is about 62.5%, and the efficiency of oil shale-to-liquid products (gasoline, jet fuel, and distillate fuel) conversion is about 60%. If the heat credit of coke is taken, the overall efficiency of the synthesis process is about 68%.

Table B-18. ENERGY BALANCE FOR PRODUCTION OF 50,000 bbl/DAY
OF GASOLINE AND LIGHT DISTILLATE FROM 30 gal/TON
COLORADO OIL SHALE

	10^6 Btu/day
Input	
Oil Shale (85,780 tons/day X 2540 Btu/lb X 2000)	452,918
Electricity (92,100 kWhr/day X 3413 Btu/kWhr)	314
Natural Gas (3,610,000 SCF/day X 1000 Btu/SCF)	3,610
Total Input	456,842
Output	
Motor Gasoline (25,193 bbl/day X 5.3 million Btu/bbl)	133,523
Jet Fuel (4471 bbl/day X 5.4 million Btu/bbl)	24,143
Distillate Fuel (20,336 bbl/day X 5.6 million Btu/bbl)	113,882
i-Butane (1613 bbl/day X 4.325 million Btu/bbl)	6,976
Decant Oil (191 bbl/day X 6 million Btu/bbl)	1,146
Coke (932 tons/day X 2000 X 14,000 Btu/lb)	26,096
Sulfur (47 tons/day X 2000 X 3983 Btu/lb)	374
Ammonia (150 tons/day X 2000 X 9675 Btu/lb)	2,902
Spent Shale (65.713 tons/day X 305 Btu/lb X 2000)	40,085
Cooling by Air and Water	64,852
Other (by Difference)*	42,863
Total Output	456,842

* Includes sensible heat of product streams, heating values of other unaccounted products, and heat lost to the atmosphere.

Pollution

The largest pollutants from this process are the spent shale, the dust from crushing the shale, the pollutants in the flue gases from the retorting of

the sulfur and nitrogen compounds removed from the crude shale oil, and the waste water used to wet the spent shale. The quality of the waste products from the upgrading and refining steps should not be significantly different from that of the products of present commercial processes. There is a considerable lack of information about the retorting and spent shale disposal.

In general, however, we can expect to be able to remove all gaseous sulfur compounds by conventional processing. This would involve a Claus-type plant with one of the many tail-gas cleanup processes in series (such as Scot, Beavon, etc.). Similarly, ammonia removal should not be a problem. The gas streams from retorting and catalytic cracking will contain dust, which might require electrostatic precipitators. Also, the gas from retorting may contain fine oil mist, which might require electrostatic precipitators.

The low-Btu stack gases probably will be used for plant fuel. The stack gases from burning this gas could be sent to a Wellman-Lord Process to recover sulfur dioxide, which, in turn, could then be fed to the Claus plant to produce more elemental sulfur. Table B-19 gives the sulfur balance around the system.

Table B-19. SULFUR BALANCE FOR PRODUCTION OF 50,000 bbl/DAY OF GASOLINE PLUS LIGHT DISTILLATE FROM 30 gal/TON COLORADO OIL SHALE

Analysis of Shale	Feed Shale	Spent Shale
	wt %	
Composition, dry basis		
Organic Carbon	13.61	4.94
Mineral Carbon Dioxide	15.87	14.74
Hydrogen	2.06	0.27
Nitrogen	0.46	0.28
Oxygen (by difference)	0.45	--
Sulfur	0.75	0.62
Ash	66.80	79.15
Total	100.00	100.00

	lb/hr (as sulfur)
Input	
Oil Shale to Retorting	52,906
Output	
Spent Shale Dust	33,952
Sulfur From Sulfur Plant	3,917
Stack Gases	13,987
Gaseous Effluent Sulfur Plant	1,050
Total	52,906

The liquid-phase pollutants would be primarily the quench water used to cool the spent shale and the resultant alkalinity. Because the Gas Combustion Process is carried out at relatively high temperatures, substantial mineral carbonate (primarily calcium and magnesium) decomposition and production of calcium and magnesium oxides would be expected. When contacted with water, these oxides will form basic calcium and magnesium hydroxides.

The process will use on the order of 600,000 gph of total makeup, cooling, and process water. The waste water of the process may be contaminated with hydrogen sulfide, other sulfide compounds, nitrogen compounds (principally amines and ammonia), and oil shale fines (both from raw shale and spent shale). There will also possibly be thermal pollution from the effluent cooling water.

The major solid waste environmental problem is the volume of the spent shale. The inability to return all the spent shale to the mine (because the crushed shale has on the order of 30% voids and some possible swelling during processing) creates a severe problem in spent shale disposal. Dust from the mining operation, as well as the spent shale, can be carried off by the wind and also cause environmental problems. Table B-20 lists wastes and their sources and required treatments.

Table B-20. WASTES, SOURCES, AND TREATMENTS FOR AN OIL-
SHALE-TO-GASOLINE PLANT

Waste	Sources	Treatment
Shale Dust and Fines	Shale crushing, spent shale disposal, retort off-gases, shale mining operations	Briquetting of fines, cyclone separators, bag filters, electrostatic precipitators, scrubbers, etc.
Spent Shale	Retorting	Watering down to cool and reduce dusting, compaction, and benching
Waste Water (containing dissolved sulfur and nitrogen compounds and traces of hydrocarbons)	Mining, retorting, and refining of crude shale oil	Conventional biological and chemical treatment. Also separations to remove oil droplets.
Hydrogen Sulfide	Retorting and refining off-gases	Typically, Claus Process plus recent tail-gas clean-up processes
Sulfur Dioxide	Flue gases from combustion of above retorting and refining off-gases	Wellman-Lord lime treatment, etc.

Economic Analysis

The economic analysis is performed by using the DCF method. The investment and operating costs of a 50,000 bbl/day plant are given in Tables B-21 and B-22, respectively. The investment and operating costs are taken from Refs. 3, 5, and 8, and these numbers have been escalated by an appropriate factor to bring them to December 1973 levels. According to our estimate, the investment cost of the refinery to produce gasoline and distillate fuel from syncrude is about $73 million. One-half of the labor and supplies costs is charged to mining to estimate the working capital for raw material inventory. The calculation method[8] for the average unit cost of the products is presented in Table B-23. This financing method includes the following factors:

- A 25-year project (synthesis plant) life
- Depreciation calculated on a 16-year sum-of-the-digits formula
- 100% equity capital
- A 48% Federal Income Tax rate
- A 12% DCF rate
- Plant start-up costs as expenses in year zero.

The average production cost of gasoline and distillate fuel from oil shale is about $10.33/bbl, or about $0.25/gal. The price differential between gasoline and distillate fuel will set the exact price for these two fuels. Because both the liquid products can be utilized as a motor fuel, the price differential could be very small. This unit cost depends on the accounting method used, feed cost, and variation in other financial factors; e. g., by using the utility method,[8] the average unit cost of the products is $7.39/bbl, or about $0.18/gal.

If a 10% (instead of 12%) DCF financing model is used for the synthesis plant to produce gasoline and distillate oils from oil shale, the unit product cost becomes $1.81/million Btu (low heating value), rather than $2.05/million Btu.

Table B-21. INVESTMENT COST FOR THE PRODUCTION OF 50,000 bbl/DAY
OF GASOLINE PLUS LIGHT DISTILLATE FROM 30 gal/TON COLORADO
OIL SHALE

Components	End-of-1973 Cost, $1000
Mining (initial investment and present worth of deferred investment)	53,772
Retorting (including crushing, screening, and briquetting)	89,899
Crude Shale Oil and Gas Upgrading	78,815
Syncrude Refining for Production of Gasoline and General Facilities for Refinery	73,415
Particulate Emission Control	5,740
Spent-Shale Disposal	17,000
Initial Catalyst	5,517
Utilities	28,837
General Facilities	21,486
Total	374,481
Contractor's Overhead and Profit (10%)	37,448
Total	411,929
Contingencies (15%)	61,789
Total Plant Investment (I)	473,718
Interest During Construction (0.23676 X I)	112,158
Start-up Cost (20% of gross operating cost)	11,155
Working Capital	
Raw Material Inventory (60 days of feed at full rate)	3,978
Material and Supplies (0.9% of total plant investment)	4,263
Net Receivables (1/24 X annual revenue received)	7,070
Total Capital Required	612,342

Table B-22. OPERATING COST FOR THE PRODUCTION OF 50,000 bbl/DAY
OF GASOLINE PLUS LIGHT DISTILLATE FROM 30 gas/TON COLORADO
OIL SHALE

Component	Annual Cost, $1000
Direct Material, Catalysts, and Chemicals	4,400
Purchased Utilities	
Raw Water Cost (9800 gpm X $0.3/1000 gal)	1,391
Electricity for Refinery	
(92,000 kWhr/day X $0.009/kWhr)	272
Natural Gas (3,610,000 SCF/day X $1.0/1000 SCF)	1,186
Labor	
Process Operating Labor (150 men/shift at $5/hr and 8304 man-hr/yr)	6,228
Maintenance Labor (1.5% of total plant investment)	6,516
Supervision (15% of operating and maintenance labor)	1,912
Administration and General Overhead (60% of total labor, including supervision)	8,794
Supplies	
Operating (30% of process operating labor)	1,868
Maintenance (1.5% of total plant investment)	6,516
Local Taxes and Insurance (2.7% of total plant investment)	11,729
Spent Shale Disposal (at $0.23/ton X 65,713 tons/day)	4,965
Total Gross Operating Cost	55,777
By-product Credit	
Sulfur (47 tons/day X 2000 tons/2240 LT X $10/LT X 328.5)	138
Ammonia (150 tons/day X $25/ton X 328.5)	1,232
i-Butane and Decant Oil (1804 bbl/day X 42 X $0.10/gal X 328.5)	2,489
Total By-product Credit	3,859
Net Operating Cost	51,918

Table B-23. CALCULATION FOR DETERMINING UNIT PRODUCTION
COST BY DCF METHOD FOR 50,000 bbl/DAY OF GASOLINE PLUS
LIGHT DISTILLATE FROM 30 gal/TON COLORADO OIL SHALE

Unit Cost of the Product

$$\frac{N + 0.23816\ I + 0.1275 S + 0.230777\ W}{G}$$

where

N = Operating Cost = $51,918,000

I = Total Plant Investment = $473,718,000

S = Start-up Cost = $11,155,000

W = Working Capital = $15,311,000

G = Annual Product Production [(29.664 bbl/day
     gasoline + 20,336 bbl/day distillate fuel) X
     328.5 days/yr] = $16,425,000 bbl/yr
  or

on Btu Basis: (29,664 bbl/day X 5.3 X 10^6 Btu/
  bbl + 20,336 bbl/day X 5.6 X 10^6 Btu/bbl) X
  328.5 days/yr = $89,056,550 X 10^6 Btu/yr (high heating value)

Unit Cost = $\dfrac{\$169,694,000}{16,425,000}$ = $10.33/bbl (average price of gasoline and
                                                        distillate fuel)

= $1.905/$10^6$ Btu (high heating value)

= $2.048/$10^6$ Btu (low heating value)

Methanol From Coal

The reactions occurring in the process for making methanol are as follows:

$$CO + 2H_2 \rightarrow CH_3OH \tag{B-3}$$

$$CO_2 + 3H_2 \rightarrow CH_3OH + H_2O \tag{B-4}$$

$$CO + 3H_2 \rightarrow CH_4 + H_2O \tag{B-5}$$

$$CH_3OH + CO \rightarrow HCOOCH_3 \tag{B-6}$$

$$2CH_3OH \rightarrow CH_3OCH_3 + H_2O \tag{B-7}$$
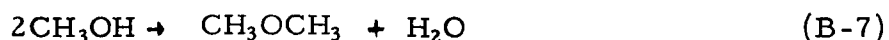
Reactions B-3 and B-4 are highly desirable, and the remaining reactions are unwanted side reactions. For the production of methanol, the synthesis gas (containing carbon monoxide, hydrogen, and carbon dioxide) is required. The synthesis gas can be produced from coal, naphtha, natural gas, or heavy oil and water. Its manufacturing cost is highly sensitive to the price of raw material (cents/million Btu), and in the future coal may be the most attractive raw material for the production of methanol, compared with other conventional fuels, because of the faster increase in the prices of conventional fuels.

For the production of synthesis gas (carbon monoxide, carbon dioxide, and hydrogen) from coal, many gasifiers are available (e.g., Lurgi, Koppers-Totzek, Winkler, Wellman-Galusha). On the basis of energy requirements, a high-pressure gasifier is more desirable than a low-pressure gasifier. In high-pressure operation, only oxygen has to be compressed, not the entire amount of synthesis gas. The amount of oxygen is only one-third or one-fourth of the synthesis gas produced; as a result, the compression cost is reduced in the high-pressure process. However, high-pressure processes make more methane, compared with low-pressure processes; consequently, more gas has to be purged from the methanol synthesis loop. However, this purge gas can be utilized as a fuel, so this is not a major disadvantage. Also, at high-pressure operation, oxygen consumption per ton of coal is less compared with low-pressure operation. In this study a Koppers-Totzek gasifier, which is a low-pressure gasifier, is used to produce synthesis gas. This process is selected because more information is available; it does not produce liquid products such as tars; and Lurgi is the only commercially available high-pressure process, but it also produces tars, phenols, and other liquid products.

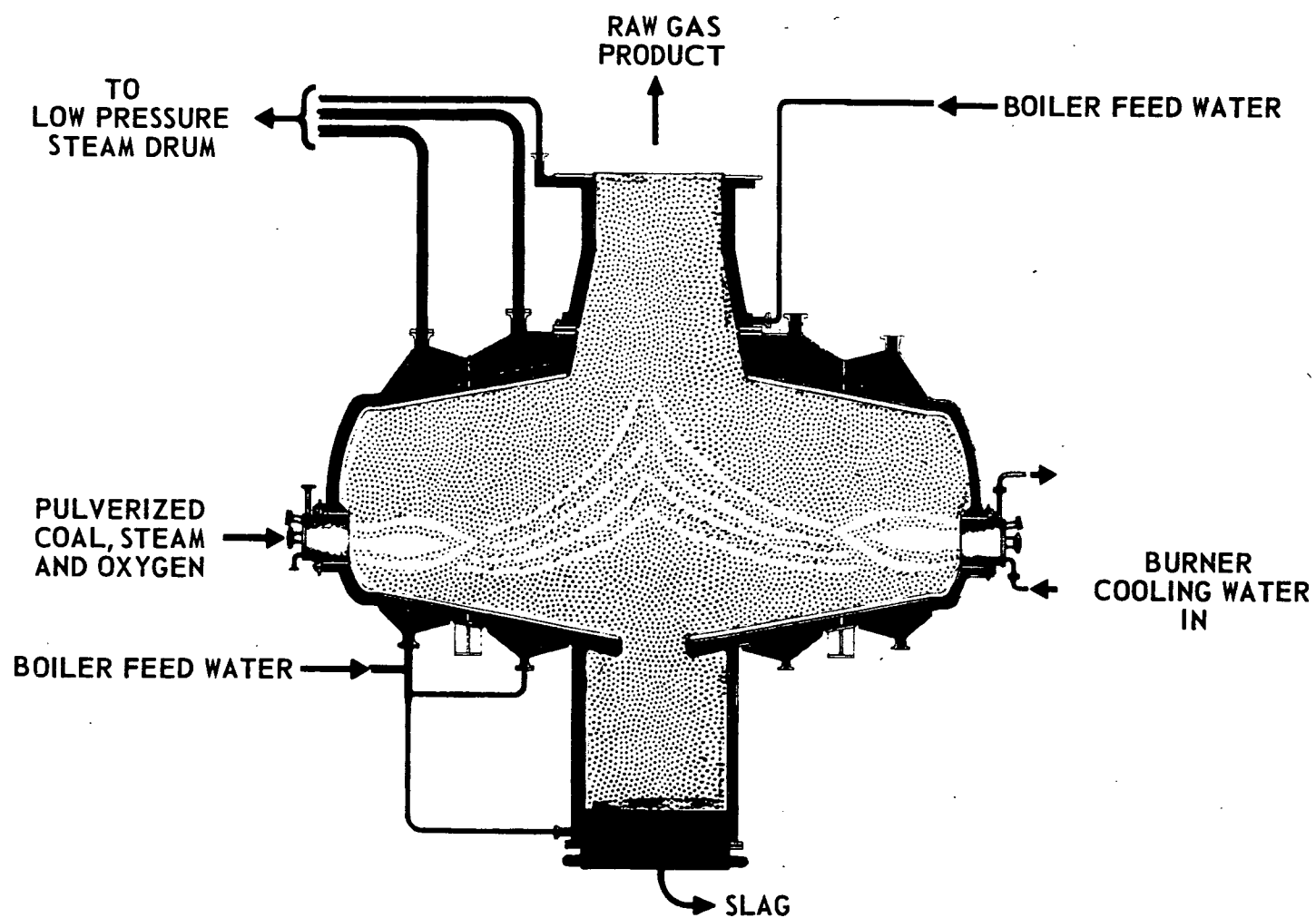## Description of Koppers-Totzek Gasifier and ICI Synthesis

The Koppers-Totzek gasifier can be operated on all types of coal without pretreatment. Coal is dried and pulverized (70% through 200 mesh). A homogenous mixture of oxygen and pulverized coal is introduced to the gasifier through coaxial burners at each end. The gasifier is a refractory-lined, horizontal, cylindrical vessel with conical ends, as shown in Figure B-7. Oxygen, steam, and coal react at about atmospheric pressure and about 3300°F in the gasifier. Fixed carbon and volatile matter are gasified to produce off-gas containing carbon monoxide and hydrogen. Coal ash is converted into molten slag and some of this drops into a water-quench tank, while the remainder is carried by the gas. Low-pressure steam is circulated around burners and refractory to protect them from excessive temperature.

Gas leaving the gasifier is quenched with water to solidfy entrained molten ash and prevent it from solidifying on the walls of the waste-heat boiler. After passing through the waste-heat boiler, the gas is scrubbed in a high-energy water-scrubbing system, which reduces its solids loading to 0.002-0.005 grains/SCF and lowers it stmperature to about 100°F. For this kink of scrubbing, venturi scrubbers are used. Finally, prior to compression, particulates are removed by electrostatic precipitation.

The gas composition does not vary much with type of coal used. The extremely high temperatures ensure that any high-molecular-weight hydrocarbons pyrolyzed from the original coal will be destroyed. The product gas contains about 0.1% methane; 80% of the sulfur in the coal is converted to H_2S, and the rest appears as carbonyl sulfide (6-12%) and as sulfur in the fly ash. The concentration of nitrogen oxides is about 5 ppm. Steam and oxygen consumption in this case are 0.248 ton/ton and 0.798 ton/ton of coal, respectively. The synthesis gas yield is about 70,400 SCF (on wet basis) and 61,4000 SCF (on dry basis) per ton of coal fed to the gasifier.

The Koppers-Totzek gasifier has been commercially available since 1952, 20 plants using a total of 52 Koppers-Totzek gasifiers have been ordered. It can be set up for alternate firing of coal and heavy oil. The most common use for the gas has been for ammonia synthesis.

The scrubbed gas is compressed to 450 psig and passed through the shift reactor to adjust the hydrogen-to-carbon monoxide ratio. In this case, the gas is not purified before it enters the shift reactor (i.e., it contains hydrogen sulfide

RAW GAS
PRODUCT

TO
LOW PRESSURE
STEAM DRUM

BOILER FEED WATER

PULVERIZED
COAL, STEAM
AND OXYGEN

BURNER
COOLING WATER
IN

BOILER FEED WATER

SLAG

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A-94-1627

Figure B-7. KOPPERS TOTZEK LOW-PRESSURE GASIFIER

and carbonyl sulfide) so a catalyst tolerating sulfides has to be used in the reactor. The shifted gas goes to the purification system after waste-heat recovery. In this case, a hot-carbonate-scrubbing system removes hydrogen sulfide and carbonyl sulfide down to 10 ppm and carbon dioxide down to about 7%, so that a ratio of $H_2/(CO + 1.5\ CO_2) = 2.05$ can be achieved. Regenerator off-gas goes to the sulfur recovery system (Claus and Wellman-Lord Processes), and about 97% of the sulfur is recovered as an elemental sulfur.

The purified gas is passed through an iron sponge drum and a sulfur guard drum to remove traces of sulfur. The gas containing no sulfur is compressed from about 385 to about 1500 psia for the manufacture of methanol.

Methanol can be produced from synthesis gas either by using a high-pressure process (e. g., Japan Gas Chemical Company Process), a low-pressure process (e. g., Imperial Chemical Industries low-pressure process, Lurgi low-pressure process), or an intermediate-pressure process (e. g., Nissui-Topsoe Process).

In the high-pressure process, the synthesis gas is compressed to about 4300 psi. The compressed gas is combined with recycle gas and passed to the methanol catalytic (zinc-chromium oxide catalyst) converter. The synthesis gas entering the converter is preheated to the reaction temperature by heat exchange with the product gas.

In the low-pressure process, the synthesis gas is compressed with re-cycle gas and passed to the catalytic (highly active copper catalyst) converter. The synthesis gas entering the converter is heated to 480°-570°F by heat exchange with the product gas.

In the intermediate-pressure process, the synthesis gas is compressed to about 2300 psi. The compressed gas is combined with recycle gas and passed to the catalytic (similar to Cu-Zn-Cr catalyst) converter. The synthesis gas entering the converter is heated to 460°-540° F by heat exchange with the product gas. In all the processes, the catalysts are vulnerable to sulfur poisoning, so careful removal of sulfur compounds from the synthesis gas is very essential.

The crude methanol is condensed and separated from the untreated gas, which is recycled to the converter. The crude methanol is then let down to

a lower pressure and dissolved gases are flashed off. Some of the flash gas is purged to control the concentration of inerts and nonreacting components, and the remaining gas is recycled. If the concentration of inerts and non-reacting components is very high, the high-pressure gas has to be purged. The purged gas is used as a fuel in the methanol plant. The crude methanol then is purified by distillation. Usually, two fractionation towers are required — one to remove light end fractions such as dimethyl ether and methyl formate, the other to remove high-boiling components such as water and higher alcohols. The product may have purity as high as 99.95% methanol. Crude methanol contains about 30 compounds with normal boiling points from −23.7°C to 174°C, as shown in Table B-24. The purified methanol (99.85%) contains about 900 ppm ethanol and about 500 ppm water. This 99.9% pure methanol is known as chemical-grade methanol. The fuel-grade methanol need not be 99.9% pure. Usually, fuel-grade methanol is 98% pure containing about 2% impurities such as water, ethanol, and higher alcohols.

Table B-24. COMPONENTS EXPECTED IN CRUDE METHANOL[10]

Components	Normal Boiling Point, °C
Dimethyl Ether	− 23.7
Acetaldehyde	+ 20.2
Methyl Formate	31.8
Diethyl Ether	34.6
n-Pentane	36.4
Propionaldehyde	48.0
Methyl Acetate	54.1
Acetone	56.5
Methanol	64.7
Isopropyl Ether	67.5
n-Hexane	69.0
Methyl Propionate	78.0
Ethanol	78.4
Methyl Ethyl Ketone	79.6
t-Butyl Alcohol	83.0
n-Propanol	97.0
n-Heptane	98.0
Water	100.0
Methyl Isopropyl Ketone	101.7
Acetal	103.0
Isobutanol	107.0
n-Butyl Alcohol	117.7
Isobutyl Ether	122.3
Diisopropyl Ketone	123.7
n-Octane	125.0
Isoamyl Alcohol	130.0
4-Methyl Amyl Alcohol	131.0
n-Amyl Alcohol	138.0
n-Nonane	150.7
n-Decane	174.0

The high-pressure, low-pressure, and intermediate-pressure methanol synthesis processes are commercially available. However, compared with the high-pressure process, the low-pressure process has lower operating and capital costs. The biggest saving in the low-pressure process compared with

Table B-25. COMPOSITION OF GASEOUS STREAMS FROM A
COAL-TO-METHANOL (5000 Ton/Day) PLANT (Figure B-8)

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
								mol %							
Temperature, °F	2728	350	115	280	625	220	100	100	100	340	500	120	150	250	450
Pressure, psia	20	18	16	400	400	390	385	20	385	1500	1500	20	30	30	400
CO	51.23	49.74	50.94	55.71	55.71	16.30	23.43	0.06	23.43	23.43	17.16	16.92	--	--	--
CO_2	5.87	7.36	7.54	8.24	8.24	29.86	6.71	95.73	6.71	6.71	4.35	4.26	--	--	--
H_2	28.00	29.49	30.20	33.03	33.03	47.75	68.54	0.43	68.54	68.54	66.35	66.25	--	--	--
H_2O	12.83	11.34	9.20	0.70	0.70	4.41	0.25	0.25	0.25	0.25	0.13	0.13	--	100.00	100.00
CH_4	0.11	0.11	0.11	0.12	0.12	0.09	0.13	--	0.13	0.13	1.37	1.42	--	--	--
N_2	0.43	0.43	0.44	0.48	0.48	0.35	0.50	--	0.50	0.50	5.36	5.55	0.07	--	--
Ar	0.38	0.38	0.39	0.43	0.43	0.31	0.44	--	0.44	0.44	4.74	4.91	1.43	--	--
H_2S	1.10	1.10	1.13	1.23	1.23	0.89	10 ppmv	3.38	--	--	--	--	--	--	--
COS	0.05	0.05	0.05	0.06	0.06	0.04		0.15	--	--	--	--	--	--	--
CH_3OH	--	--	--	--	--	--	--	--	--	--	0.54	0.56	--	--	--
O_2	--	--	--	--	--	--	--	--	--	--	--	--	98.50	--	--
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Moles/hr	55,311.5	55,311.5	54,006.5	49,382.5	49,382.5	68,415.0	47,551.5	18,008.5	47,551.0	47,551.0	1,251,702.5	4283.0	14,841.5	8196.5	24,176.5
10^6 SCF/hr	20.95	20.95	20.45	18.70	18.70	25.95	18.00	6.85	18.00	18.00	474.5	1.6	5.6	3.1	9.15

the high-pressure process is power cost. However, this difference is substantially reduced for a high-capacity methanol plant. In this study, calculations are based on ICI low-pressure methanol.

The compressed gas is mixed with recycled gas and heated to 500°F by heat exchange with the product gas. The composition of the synthesis gas satisfies the condition of $H_2/(CO + 1.5\ CO_2) = 2.05$. The heated gas is passed through a fixed-bed catalytic (highly active copper catalyst) converter. The gas coming out is at about 580°F and is used in heating the feed gas. Then the product gas is cooled down to about 120°F by heating the boiler feed water and is sent to the separator for separation into methanol and gas. The portion of the recycled gas is purged to control the concentration of the inerts and unreactive components to about 10%. The purged gas is used as a fuel, and recycled gas is compressed to 1500 psia. The pressure drop in the loop is about 200 psi, and the conversion of carbon monoxide and carbon dioxide per pass is about 5%. The crude methanol is let down to lower pressure, and dissolved gases are flashed off. The flash gases are used as a fuel. The crude methanol is purified to make fuel-grade or chemical-grade methanol.

Figure B-8 is detailed flow diagram for producing methanol from coal. Tables B-25, B-26, and B-27 present the material balance around the system and the composition of the important streams for a 5000 ton/day methanol plant. The streams enumerated in Tables B-25, B-26, and B-27 are those denoted by the flow diagram (Figure B-8).

Table B-26. COMPOSITION OF SOLID STREAMS FROM A COAL-TO-
METHANOL (5000 Ton/Day) PLANT

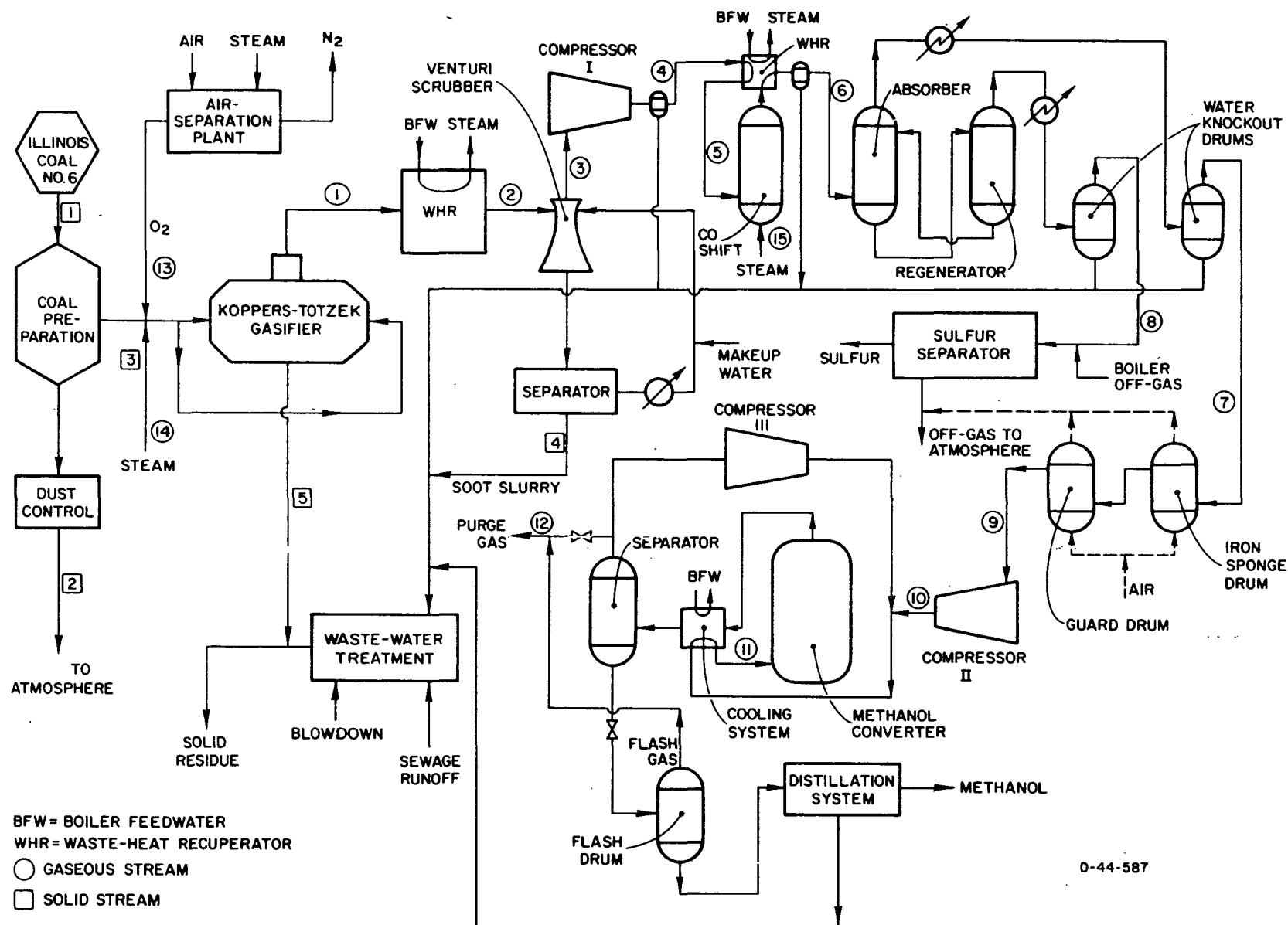|		Stream No.			
	1	2	3	4	5
Components, wt %					
C	67.30	67.30	67.30	15.64	
H	4.68	4.68	4.68	0.02	
O	9.43	9.43	9.43	--	
N	1.05	1.05	1.05	--	
S	3.84	3.84	3.84	1.80	
H_2O	4.00	4.00	4.00	60.00	
Ash	9.70	9.70	9.70	22.54	100.00
Total	100.00	100.00	100.00	100.00	100.00
lb/hr	594,958.5	30	594,928.5	138,549.5	26,478.0
tons/hr	297.48	0.015	297.465	69.27	13.24

Figure B-8. FLOW DIAGRAM OF PRODUCTION OF METHANOL FROM COAL

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Table B-27. COMPOSITION OF PRODUCT FROM A COAL-TO-METHANOL (5000 Ton/Day) PLANT

Methanol	5000 tons/day
Higher Alcohol	30 tons/day
Components, wt%	
Methanol	98.0
Ethanol	0.1
Higher Alcohols	0.5
Water	1.4
Total	100.0
Density, lb/gal at 60°F	6.64
Btu/lb	9760
Btu/gal	64,800
Heat of Vaporization, Btu/lb	473

Overall Energy Balance and Efficiencies

The overall energy balance is presented in Table B-28.

Table B-28. ENERGY BALANCE FOR A COAL-TO-METHANOL (5000 Ton/Day) PLANT

	10^6 Btu/hr
Input	
Coal to Gasifier (297.5 tons/hr X 2000 X 12,120 Btu/lb)	7,211.5
Coal to Boiler (125.85 tons/hr X 2000 X 12,120 Btu/lb)	3,050.5
Total Input	10,262.0
Output	
Methanol [5000 tons/day X (2000/24) X 9760 Btu/lb]	4,066.5
Sulfur (14.56 tons/hr X 2000 X 3983 Btu/lb)	116.0
Isobutanol and Higher Alcohols (assuming mainly isobutanol, 1.23 tons/hr X 2000 X 15,500 Btu/lb)	38.1
Cooling by Air and Water	3,100.0
Other (by difference)*	2,941.4
Total Output	10,260.0

* Includes sensible heat of product streams, heating values of other unaccounted products, and heat lost to the atmosphere.

The overall efficiency (including by-product heat credit) of the process is about 41%, and the coal-to-methanol efficiency is about 40%, which can be increased by an additional 5% by using a high-pressure gasifier. In this low-pressure gasification process, 150,000 hp is required to compress the gas from atmospheric

pressure to about 1500 psia, which is about 1300 million Btu/hr. In a high-pressure gasifier, oxygen has to be compressed to the required pressure, but its amount is one-third or one-quarter of the synthesis gas produced. Consequently, about 900 million Btu/hr can be saved, which amounts to 5% of the total energy required. Therefore, overall efficiency of the process could be 46%. However, the investment cost in a high-pressure operation is higher than that in a low-pressure operation.

The efficiency of the methanol loop in this case is about 73% and that of the low-pressure gasification system is about 56%, making the overall efficiency about 41%. The efficiency of synthesis gas production with the high-pressure gasification system is about 60-65%, which makes the overall efficiency of the coal-to-methanol process about 46%.

Pollution

Sulfur is the biggest pollutant resulting from the process. However, 90% of the total sulfur can be recovered as elemental sulfur balance by using suitable processes. The sulfur balance is reported in Table B-29.

Table B-29. SULFUR BALANCE FOR A COAL-TO-METHANOL
(5000 Ton/Day) PLANT

	lb/hr (as sulfur)
Input	
Coal to Gasifier	22,846.5
Coal to Boiler	9,665.0
Total Input	32,511.5
Output	
Elemental Sulfur (by-product)	29,120.0
Sulfur Compounds to Atmosphere From Sulfur-Recovery Plant	881.0
Sulfides to Atmosphere From From Iron Sponge	16.0
Sulfur to Atmosphere With Coal Dust	1.0
Sulfur With Soot in Waste-Water Recovery	2,493.5
Total Output	32,511.5

The stack gas from the boiler containing sulfur dioxide can be fed to the Wellman-Lord Process to recover sulfur dioxide. This sulfur dioxide is mixed with the hydrogen sulfide removed from synthesis gas and fed to the Claus plant to recover elemental sulfur.

About 2300 gpm of waste water requires treatment; it may contain ammonia, sulfur compounds, traces of hydrogen cyanide, thiocyanate, and ash. The process requires 100,000-200,000 gpm of cooling water, which is the biggest source of heat pollution. Table B-30 lists wastes, their sources, and possible treatments.

Table B-30. WASTES, SOURCES, AND TREATMENTS FOR A COAL-TO-METHANOL PLANT

Waste	Sources	Treatment
Coal Dust	Coal-crushing system, conveyor belts	Cyclone separators, bag filters, scrubbing, etc.
Soot and Ash	Gasifier	Scrubbing and various waste-water and solid treatments
Waste Water (containing alcohols, ammonia, hydrogen sulfide, hydrogen cyanide)	Gasifier, compression, gas-cooling system after shift, purification, etc.; methanol distillation system, boiler blowdown, sewage run-off	Biological treatments, Phenosolvan, and modified Chevron to remove hydrogen sulfide, ammonia, etc.
Hydrogen Sulfide	Regenerator off-gas	Claus Process or any suitable sulfur recovery process
Sulfur Dioxide	Boiler flue gas	Wellman-Lord lime treatment, etc.

Economic Analysis

The economic analysis is performed by using the DCF method. The investment and operating costs of a 5000 ton/day methanol-from-coal plant are estimated in Tables B-31 and B-32, respectively. The calculation method[8] for the unit production cost of the product is presented in Table B-33. This financing method includes the following factors:

- A 25-year project (synthesis plant) life
- Depreciation calculated on a 16-year sum-of-the-digits formula
- 100% equity capital
- A 48% Federal Income Tax rate
- A 12% DCF rate
- Plant start-up costs as expenses in year zero.

For 30¢/million Btu coal, the cost of methanol is about $71/ton, or $0.234/gal. This unit cost depends on the accounting method used, material costs, and variations in other financial factors (e.g., by using a utility method,[8] the unit cost of the product is $54/ton, or $0.179/gal).

If a 10% (instead of 12%) DCF financing model is used for the synthesis plant to produce methanol from coal, the unit product cost becomes $3.51/million Btu (low heating value).

Table B-31. INVESTMENT COST FOR COAL-TO-METHANOL (5000 Ton/Day) PLANT USING KOPPERS-TOTZEK GASIFICATION AND ICI METHANOL PROCESSES

Components	End-of-1973 Cost, $1000
Coal Storage	1,900
Syngas Train	33,000
Syngas Compressor I	11,100
Carbon Monoxide	4,730
Carbon Monoxide-Shift Waste-Heat Recovery	1,430
Hot Carbonate System	13,500
Trace Hydrogen Sulfide Removal	820
Syngas Compressor II	3,000
Methanol Loop	30,170
Air Separation Plant	37,200
Oxygen Compressor	1,100
Steam Generation and Boiler Feed Water Pumps	18,380
Boiler Feed Water Treatment	7,600
Cooling Tower and Pumps	6,100
Waste-Water Treatment	10,300
Particulate Emission Control	1,740
Sulfur Recovery	5,080
Wellman-Lord Stack-Gas Cleanup	12,270
Turbo Generator	3,330
Power Distribution System	4,000
General Facilities	5,000
Total	211,750
Contractor's Overhead and Profits (10%)	21,175
Total	232,925
Contingencies (15%)	34,900
Total Plant Investment (I)	267,825
Interest During Construction (0.23676 X I)	63,410
Start-up Cost (20% of gross operating cost)	10,002
Working Capital	
Coal Inventory (60 days of feed at full rate)	4,433
Materials and Supplies (0.9% of total plant investment)	2,410
Net Receivables (1/24 X annual revenue received)	4,888
Total Capital Required	352,968

Table B-32. OPERATING COST FOR COAL-TO-METHANOL (5000 Ton/Day) PLANT USING KOPPERS-TOTZEK GASIFICATION AND ICI METHANOL PROCESSES

Components	Annual Cost, $1000
Coal Feed (at 246,288 X 10^6 Btu/day), 30¢/10^6 Btu	24,272
Other Direct Materials, Catalysts, and Chemicals	1,947
Purchased Utilities	
Raw-Water Cost (5000 gpm X 30¢/1000 gal)	710
Labor	
Process Operating Labor (50 men/shift at $5/hr and 8304 man-hr/yr)	2,076
Maintenance Labor (1.5% of total plant investment)	4,017
Supervision (15% of operating and maintenance labor)	914
Administration and general overhead (60% of total labor, including supervision)	4,204
Supplies	
Operating (30% of process operating labor)	623
Maintenance (1.5% of total plant investment)	4,017
Local Taxes and Insurance (2.7% of total plant investment)	7,231
Total Gross Operating Cost	50,011
By-product Credit	
Sulfur (310 LT/day X $10/LT X 328.5)	1,018
Total Net Operating Cost	48,993

Table B-33. CALCULATION FOR DETERMINING UNIT PRODUCTION COST BY DCF METHOD FOR A COAL-TO-METHANOL (5000 Ton/Day) PLANT

Unit Cost of the Product

$$\frac{N + 0.238161\,I + 0.1275\,S + 0.230777\,W}{G}$$

where

N = Net Operating Cost = $48,993,000

I = Total Plant Investment = $267,825,000

S = Start-up Cost = $10,002,000

W = Working Capital = $11,731,000

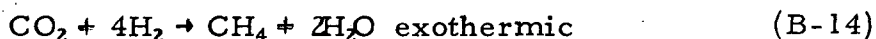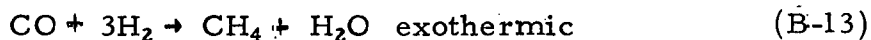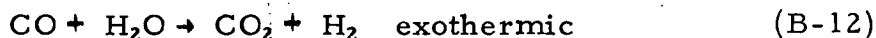G = Annual Production (5000 tons/day X 328.5 days/yr + 30 ton/day higher alcohol X 328.5 days/yr)
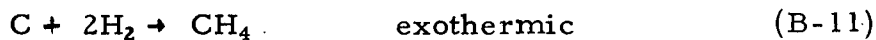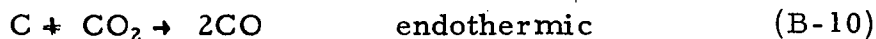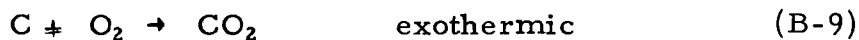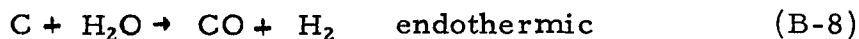
$$\text{Unit cost} = \frac{\$116,760,000}{5030 \times 328.5} = \$70.66/\text{ton}$$

= $0.2342/gal

= $3.881/million Btu (low heating value)

SNG From Coal

The reactions occurring in the process for making SNG are as follows:

$$C + H_2O \rightarrow CO + H_2 \qquad \text{endothermic} \qquad \text{(B-8)}$$
$$C + O_2 \rightarrow CO_2 \qquad \text{exothermic} \qquad \text{(B-9)}$$
$$C + CO_2 \rightarrow 2CO \qquad \text{endothermic} \qquad \text{(B-10)}$$
$$C + 2H_2 \rightarrow CH_4 \qquad \text{exothermic} \qquad \text{(B-11)}$$
$$CO + H_2O \rightarrow CO_2 + H_2 \qquad \text{exothermic} \qquad \text{(B-12)}$$
$$CO + 3H_2 \rightarrow CH_4 + H_2O \qquad \text{exothermic} \qquad \text{(B-13)}$$
$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O \qquad \text{exothermic} \qquad \text{(B-14)}$$

Methane is produced by Reactions B-11, B-13, and B-14. The main components of these reactions are carbon and hydrogen. The hydrogen is produced by the highly endothermic reaction between carbon and steam. The heat required for the reaction is supplied by combustion of a portion of the coal with oxygen by other exothermic reactions or by some other means. The reaction between carbon monoxide and hydrogen is highly exothermic, and the reaction kinetics are highly sensitive to the partial pressure of hydrogen; e.g., if the partial pressure of hydrogen is doubled, Reaction B-13 goes 8 times faster. Therefore, Reactions B-13 and B-14 are highly favored at high pressure and supply part of the heat required for endothermic Reaction B-8. Consequently, the oxygen requirement is reduced in the high-pressure process. Another advantage in the high-pressure process is some saving in compression of the gas. In the high-pressure gasification process, only oxygen has to be compressed to the required pressure, not the entire amount of synthesis gas. And the amount of oxygen is only one-third or one-quarter of the synthesis gas produced. However, the investment cost of the high-pressure operation is higher than that of the low-pressure operation.

Many processes exist for gasifying coal. Some of the processes are in commercial production (e.g., Lurgi, Koppers-Totzek, Winkler, and Wellman-Galusha); some are on a pilot-plant scale (e.g., HYGAS, CO_2-Acceptor, BI-GAS, and Synthane); and some are in the development stage (e.g., ATGAS and Exxon). In this study, the Lurgi gasification process, a medium-pressure (about 450 psia) process, is used. This process is selected because it is commercially available and because it is operated at higher pressures than other commercially available processes.

The Lurgi gasifier was developed by the Lurgi Mineralötechnik GmbH of Frankfurt, West Germany. It is currently limited to noncaking coals. Recent gasifier research has been directed toward mechanical modifications to allow the use of mildly caking coals. Sixteen commercial Lurgi plants, producing a gas of about 400-450 Btu/CF, have been built during the last 30 years, and some are still in operation. A plant with a capacity of about 288 million CF/day of SNG has been designed for the El Paso Natural Gas Company on a site near Farmington, N.M. Process design work has been completed and an environmental impact statement has been filed. To a major extent, the process setup and data required for this study have been taken from this filing. However, this study should not be considered representative of El Paso's Lurgi plant because some (minor) modifications are made in this presentation.

Description of Lurgi Process

Crushed (1/2 to 1-1/4 inch) and dried coal is fed to a moving-bed gasifier in which gasification of coal takes place at 350-450 psi. Devolatilization occurs initially and is accompanied by gasification in the temperature range of 1150°-1400°F. The nominal residence time of the coal is about 1 hour. Steam is the source of the hydrogen. Combustion of a portion of the char with oxygen supplies the heat required for the carbon-steam (endothermic) reaction. A revolving grate at the base of the reactor supports the fuel bed, removes the ash, and introduces the steam and oxygen mixture. Crude gas leaves the gasifier at temperatures between 700° and 1100°F (depending on the type of coal) and contains tar, oil, naphtha, phenols, ammonia plus coal, and ash dust.

A typical Lurgi pressure gasifier is shown in Figure B-9. The process coal is fed through a lock hopper that holds about 6 tons of coal and that is cycled once every 15 minutes when the gasifier is operating at full capacity. The lock is pressurized with raw, cooled, product gas to feed the coal to the reactor, and depressurization releases a fuel gas that is collected in surge storage tanks, recompressed, and added to the main gas stream. Coal passes down through the bed, moving through zones of increasing temperature in which different types of chemical reactions occur; eventually, the ash is forced through the water-cooled revolving grate (which also acts as a distributor for oxygen and steam) into the ash lock hopper. The ash, which ranges from
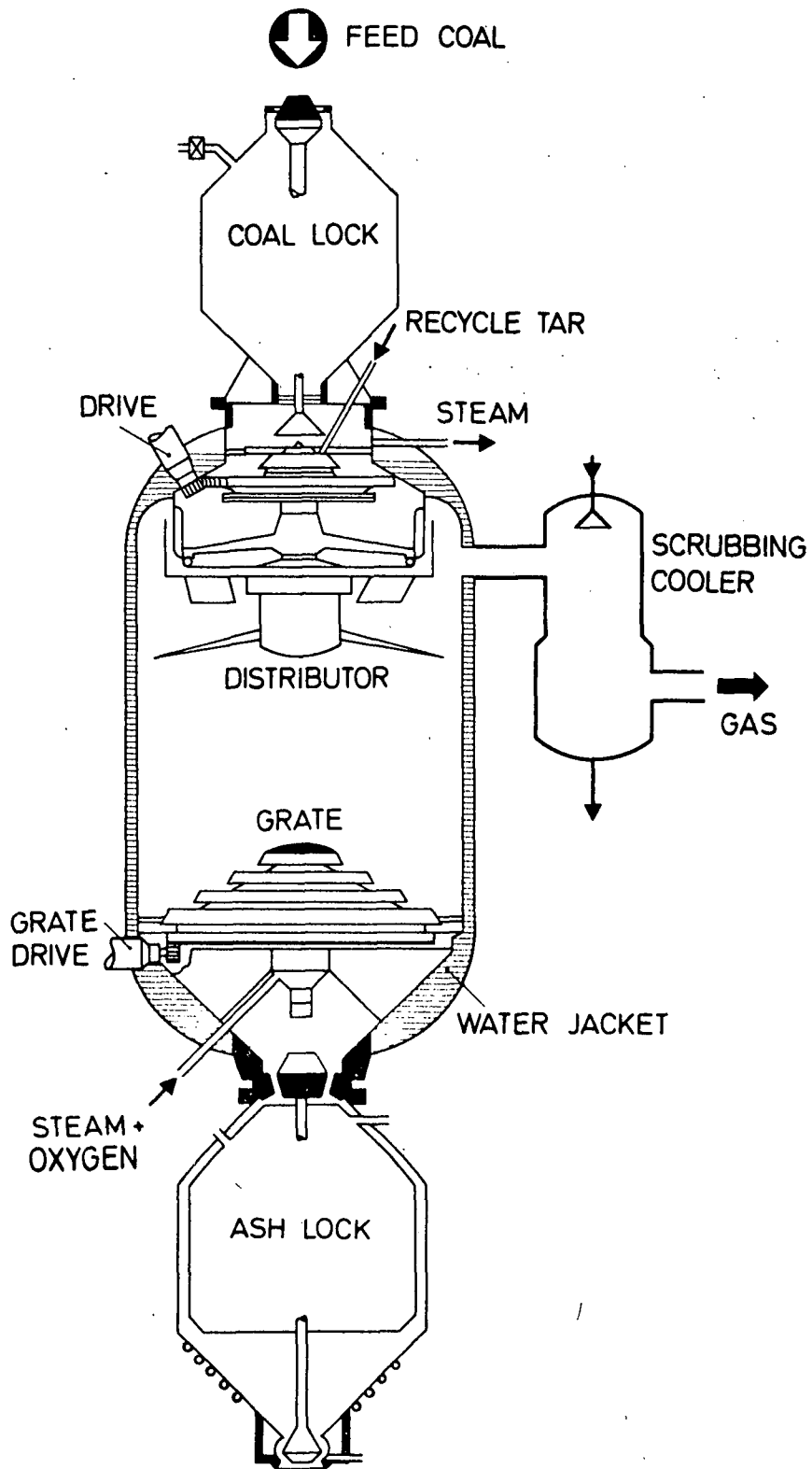
Figure B-9. LURGI PRESSURE GASIFIER (Source: Ref. 6)

very fine particles to 4-inch lumps, is discharged from the ash hopper once every 30 minutes. The carbon content of the ash is about 5%.

The solids and gas are contacted countercurrently in the gasifier; as a result, relatively large quantities of liquid by-products are formed. The raw product gas at 700°-1100°F leaves the gasifier and passes through a scrubber, in which it is washed by recirculating gas liquor and cooled to saturation. Tars are condensed, and the wash water contains tars that pick up particulates from the gas. The saturated gas is passed through a waste-heat boiler in which waste heat is recovered at a temperature of about 360°F. Some of the gas liquor condensed in the boiler is pumped to the scrubber, and some is routed to a tar-gas liquor separator. The separated tars can be recycled to the gasifier, hydrotreated to produce light hydrocarbon liquids, or stored. The separated gas liquor is sent to the Phenosolvan Process for treatment and recovery of ammonia and phenols.

Gasifier operating data and detailed stream compositions for the Navajo steam coal (New Mexico) are given in Figure B-10 and in Tables B-34, B-35, and B-36. The crude gas from the waste-heat-recovery system has a hydrogen-to-carbon monoxide ratio of about 1:93. It contains about 11% methane. Ninety-five percent of the sulfur in the coal is converted to (H_2S + COS + CS_2), 2-3% of the sulfur goes with the by-products (tars, tar oil, naphtha, etc.), and the rest goes to ash. About 60% of the nitrogen fed to the gasifier is converted to ammonia. Steam and oxygen consumption in this case are 0.92 ton/ton and 0.243 ton/ton of coal (as received), respectively. The crude gas yield is about 42,300 SCF (dry basis) per ton of coal (as received) fed to the gasifier.

About 55% of the total crude gas goes to a two-stage, carbon monoxide-shift reactors system, and the remaining amount of gas bypasses the shift and goes directly to the gas-cooling system. In the shift reactor, carbon monoxide and steam react in the presence of a nickel catalyst, producing carbon dioxide and hydrogen. The hydrogen-to-carbon monoxide ratio in the shift product gas is about 9:3, and the hydrogen-to-carbon monoxide ratio of the combined stream is about 3:7. The hot shift product gas is cooled in countercurrent heat exchangers with the shift feed gas. Then the converted gas, together

NAVAJO COAL

COAL FOR FUEL GAS

③ 208 TONS/hr

COAL PREPARATION

① 969 TONS/hr

DUST CONTROL EQUIPMENT

TO ATMOSPHERE

STEAM ①

OXYGEN ②

AIR

STEAM

AIR SEPARATION PLANT

N₂

COAL LOCK

LURGI GASIFIER 450 psia

ASH LOCK WATER

WASTEHEAT RECOVERY AND GAS WASH

④

③

SHIFT

⑤

WASTEHEAT RECOVERY AND NAPHTHA SEPARATION

⑥

RECTISOL SYSTEM

COLD METHANOL WASH

ABSORBER

REGENERATOR

NAPHTHA EXTRACTOR

⑦ SOUR FUEL GAS

⑧

SULFUR RECOVERY SYSTEM

SULFUR 15,582 lb/hr

OFF-GAS TO ATMOSPHERE SWEET FUEL GAS

NAPHTHA 20,005 lb/hr

TO ATMOSPHERE

IRON-SPONGE DRUM

GUARD DRUM

AIR

WASTES FROM FUEL GAS PRODUCTION

BOILER BLOWDOWN

SEWAGE RUNOFF

WASTEWATER TREATMENT

②

ASH 238 TONS/hr

GAS LIQUOR FROM AIR GASIFIER

GAS LIQUOR SEPARATION PHENOL EXTRACTION GAS LIQUOR STRIPPING

DEHYDRATION AND COMPRESSION

CONDENSATE

WASTEHEAT RECOVERY AND CONDENSATE SEPARATION

METHANATOR

⑨

⑩

SNG (288.6 million SCF/DAY)

TAR OIL 48,588 lb/hr

TAR 88,824 lb/hr

PHENOLS 11,271 lb/hr

AMMONIA SOLUTION(20%) 21,422 lb/hr OF AMMONIA

○ = GASEOUS STREAM
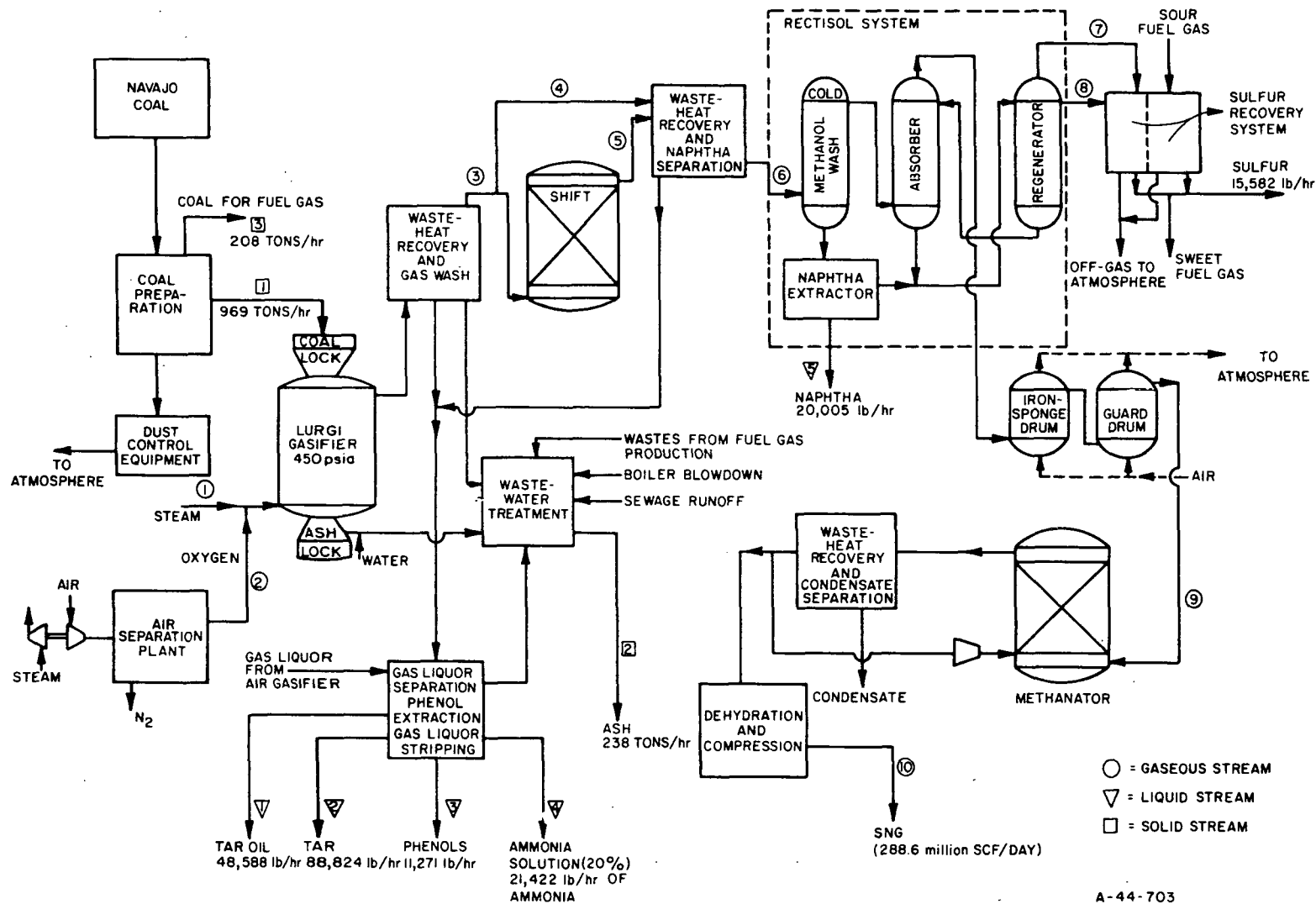▽ = LIQUID STREAM
□ = SOLID STREAM

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Figure B-10.   FLOW DIAGRAM OF LURGI-PROCESS PRODUCTION OF SNG (288.6 Million SCF/Day) FROM COAL

Table B-34. COMPOSITION OF GASEOUS STREAMS FROM COAL-TO-SNG (288.6 Million SCF/Day) PLANT*

| Components | Stream Number | | | | | | | | | |
	1	2	3	4	5	6	7	8	9	10
					mol %					
CO_2	--	--	28.03	28.03	36.95	32.36	97.53	86.17	3.10	1.81
H_2S	--	--	0.37	0.37	0.32	0.34	0.75	13.82	--	--
C_2H_4	--	--	0.40	0.40	0.36	0.39	0.24	--	0.45	--
CO	--	--	20.20	20.20	5.03	11.70	0.17	--	16.91	0.01
H_2	--	--	38.95	38.95	46.80	43.63	0.43	--	63.48	4.16
CH_4	--	--	11.13	11.13	9.75	10.70	0.56	--	14.94	92.93
C_2H_6	--	--	0.61	0.61	0.53	0.59	0.32	0.01	0.69	--
N_2 + Ar	--	2.0	0.31	0.31	0.27	0.29	--	--	0.43	1.09
O_2	--	98.0	--	--	--	--	--	--	--	--
Total (dry gas)	--	100.0	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Moles/hr (dry gas)	--	14,680.0	108,091.9	48,987.3	67,451.2	118,822.4	55,651.7	888.6	80,874.0	31,762.1
10^6 SCF/day	--	133.4	982.2	445.1	612.9	1,079.7	324.0	8.1	734.9	288.6
					lb/hr					
Water	1,783,540	--	1,394,960	632,196	357,765	2,680	--	--	--	66
Naphtha	--	--	20,005	9,066	10,939	20,005	--	--	--	--
Tar Oil	--	--	28,007	12,693	15,314	--	--	--	--	--
Tar	--	--	7,314	3,315	3,999	--	--	--	--	--
Crude Phenols	--	--	9,127	4,136	4,991	--	--	--	--	--
NH_3	--	--	17,629	7,989	9,640	--	--	--	--	--
Total	1,783,540	--	1,477,042	669,395	402,648	22,685	--	--	--	66

*See Figure B-10.

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Table B-35. COMPOSITION OF SOLID STREAMS FROM A COAL-TO-SNG
(288.6 Million SCF/Day) PLANT*

| | | Stream No. | |
	1	2[†]	3
Components. wt %			
C	49.19	5.00	49.19
H	3.60	--	3.60
N	0.85	--	0.85
S	0.69	--	0.69
O	10.15	--	10.15
Trace Compounds	0.02	--	0.02
Moisture	16.25	--	16.25
Ash	19.25	95.00	19.25
Total	100.00	100.00	100.00
lb/hr	1,938,480	477,080	415,587

* See Figure B-10.

† Dry ash.

Table B-36. COMPOSITION OF LIQUID STREAMS FROM A COAL-TO-SNG
(288.6 Million SCF/Day) PLANT*

Steam	Component	lb/hr	gal/day	
1	Tar Oil	48,588	157,370	Contains 0.27 wt % sulfur
2	Tar	88,824	239,250	Contains 0.15 wt % sulfur
3	Phenols	11,271	32,470	--
4	Ammonia	107,110	332,550	20 wt % ammonia, and solution contains 0.01 wt % sulfur
5	Naphtha	20,005	74,900	Contains 0.20 wt %

* See Figure B-10.

with the bypass gas, flow to the gas-cooling area for waste-heat recovery and cooling. Some of the gas liquor, tars, and tar oil are condensed from cooling the gas streams. The condensate is sent to separators to recover tar and tar oil. The separated gas liquor is sent to the Phenosolvan Process for recovery of ammonia and phenols. This gas liquor contains hydrogen cyanide, which is generated during gasification. The cyanide is withdrawn and sent to the sulfur recovery stage for conversion to thiocyanate and then discarded.

The gas from the gas-cooling system goes to the purification system to produce a gas free of impurities that may be harmful to the methanation catalyst. Naphtha also is recovered almost completely in this section. The sulfur compounds are reduced to a total concentration of less than 0.1 ppmv, and carbon dioxide is reduced from 33 to 3%. The gas from the gas-cooling system is chilled and then washed by cold methanol to remove naphtha and water. The naphtha-free gas enters the absorber where carbon dioxide, hydrogen sulfide, and carbonyl sulfide are removed. Then the gas is passed through an iron sponge to remove trace sulfur. The sulfur-free gas from the gas purification system goes to the methane synthesis system. A lean hydrogen sulfide acid gas stream is removed from methanol by multiflash in the flash regenerator. The remaining acid gases are stripped from the methanol in the hot regenerator, producing a hydrogen sulfide-rich gas stream. The purified methanol is recycled. The lean and rich hydrogen sulfide gas streams go to the Stretford Process for sulfur recovery. Hydrogen sulfide is removed by a Stretford solution, and then the solution is regenerated by contact with air, which also produces elemental sulfur as follows:

$$2H_2S + O_2 \rightarrow 2H_2O + 2S \qquad \text{(B-15)}$$
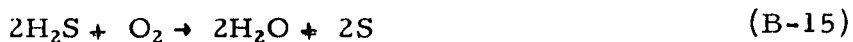
The gaseous stream from the Stretford unit contains hydrogen sulfide (about 10 ppm) and carbonyl sulfide, some of which are oxidized to sulfur dioxide and vented to the atmosphere.

The purified gas is converted to methane-rich gas in a two-stage catalytic methanator. Carbon monoxide and some of the carbon dioxide react with hydrogen (Reactions B-13 and B-14) to produce methane. These reactions are exothermic, and the heat of reaction is removed by generating process steam. To control the temperature of the reactor, a portion of the product gas is compressed and recycled. Then the product gas is cooled, and condensed water is separated in the separator. The gas then is compressed to the required pressure and dehydrated by a glycol solution. The product gas leaving the

plant contains about 93% methane and has a higher heating value of 954.0 Btu/SCF.

The oxygen required for the process is manufactured onsite by El Paso by using an air-separation plant. About 18% of the total coal feedstock supplied for the process is used to produce a low-Btu (195 Btu/CF) fuel gas that is used for the gas turbines and boilers for steam and electric power generation. The low-Btu gas is produced in a Lurgi Gasifier by using air rather than oxygen. Hydrogen sulfide from the fuel gas is removed, and sulfur is recovered by using the pressure Stretford Process.

Overall Energy Balance and Efficiencies

The energy balance for SNG-from-coal production is presented in Table B-37. The efficiency (including by-product heat credit, e.g., for tar, tar oil, sulfur, ammonia, etc.) of the process is about 70%, and the coal-to-SNG efficiency is about 56%. If tar and tar oil are hydrotreated to manufacture light oil, the overall efficiency is less than 70%.

Table B-37. ENERGY BALANCE FOR A COAL-TO-SNG
(288.6 Million SCF/Day) PLANT

	10^6 Btu/hr
Input	
Coal to Gasifier (969.24 tons/hr X 2000 X 8664 Btu/lb)	16,795.0
Coal for Fuel Gas (207.7935 tons/hr X 2000 X 8664 Btu/lb)	3,600.6
Total Input	20,395.6
Output	
Product Gas (288.6 X 10^6 SCF/day X 1/24 X 954 Btu/SCF)	11,471.9
Tar (88,824 lb/hr X 16,670 Btu/lb)	1,480.7
Tar Oil (48,588 lb/hr X 17,300 Btu/lb)	840.6
Phenols (11,271 lb/hr X 14,021 Btu/lb)	158.0
Naphtha (20,005 lb/hr X 18,400 Btu/lb)	368.1
Ammonia (21,422 lb/hr X 9598 Btu/lb)	205.6
Sulfur (15,582 lb/hr X 3983.4 Btu/lb)	62.1
Carbon in Ash (477,080 lb/hr X 704.3 Btu/lb)	336.0
Cooling Water	1,206.0
Other by Difference*	4,266.6
Total Output	20,395.6

* Includes sensible heat of product streams, heating values of other unaccounted products, and heat lost to the atmosphere.

Table B-38. SULFUR BALANCE FOR A COAL-TO-SNG
(288.6 Million SCF/Day) PLANT

	lb/hr (as sulfur)
Input	
Coal to Gasifier	13,378
Coal for Fuel Gas	2,868
Total Input	16,246
Output	
Elemental Sulfur (by-product)	15,582
Sulfur Compounds to Atmosphere From Sulfur Recovery Plant	133
From Turbine, Boiler, and Heater Effluents (sulfur dioxide discharged to atmosphere)	167
Sulfur Goes With By-products (i.e., with tar, tar oil, naphtha, ammonia solution)	364
Total Sulfur	16,246

Pollution

About 95% of the sulfur in the coal goes with gaseous streams, mainly as a hydrogen sulfide and some small amount as carbonyl sulfide and carbon disulfide. Theis sulfur is recovered as an elemental sulfur by using the Stretford Process. Sulfur from the fuel gas is recovered by using the pressure Stretford Process. The small amount of sulfur dioxide emitted by gas-fired turbines, boilers, heaters, and incinerators does not require any treatment because it is below the maximum allowable pollution limit. Overall sulfur recovery in this process is about 95%. The sulfur balance around the system is given in Table B-38. About 238 tons/hr of hot ash is quenched with water, then dewatered, and disposed of in the mine area.

The gas liquor containing tar, tar oil, phenol, and ammonia is treated in three stages. First, tar and tar oil are separated from the gas liquor; then the gas liquor is passed through the phenol extraction area for the extraction of phenol. Then in the gas liquor-stripping area, ammonia and other dissolved acid gases are stripped out. The acid gas is passed through the sulfur recovery area to convert hydrogen cyanide to thiocyanate. About 161,000 gpm of cooling water is required in the process, which is the biggest source of heat pollution. About 5600 gpm of makeup water is required in the process. Table B-39 lists wastes, their sources, and treatments required in the process.

Economic Analysis

The economic analysis is done by using a DCF method. The investment costs of a 288.6 million SCF/day plant are given in Table B-40, and much of the required source data were taken from the El Paso filing. The operating costs of the plant are given in Table B-41. The calculation method[8] for the unit cost of the product is presented in Table B-42. This financing method includes the following factors:

- A 25-year project (synthesis plant) life
- Depreciation calculated on a 16-year sum-of-the-digits formula
- 100% equity capital
- A 48% Federal Income Tax rate
- A 12% DCF rate
- Plant start-up costs as expenses in year zero.

For 30¢/million Btu coal, the cost of SNG is about $1.93/million Btu (high heating value), or about $2.14/million Btu (lower heating value). This unit cost depends on the accounting method used, the feed cost, and variation in other financial factors; e.g., by using the utility method,[8] the unit cost of the product is $1.45/million Btu (high heating value) or $1.61/million Btu (low heating value). If a 10% (instead of 12%) DCF financing model is used to produce SNG from coal, the unit product cost becomes $1.93/million Btu (low heating value), rather than $2.14/million Btu.

Table B-39. WASTES, SOURCES, AND TREATMENTS FOR
A COAL-TO-SNG PLANT

Waste	Sources	Treatment
Coal Dust	Coal-crushing system, conveyor belts, lock hoppers	Cyclone separators, bag filters, scrubbing, etc.
Soot and Ash	Gasifier and lock hopper	Scrubbing and various waste-water and solid treatments
Waste water, (containing phenols, ammonia, hydrogen cyanide, hydrogen sulfide, and oils)	Quench system, gas-cooling system, shift converter, purification system, compression, boiler blowdown, sewage run-off	Biological treatments, Phenosolvan, and modified Chevron to remove hydrogen sulfide, ammonia, etc.
Hydrogen Sulfide Carbonyl Sulfide Carbon Disulfide	Rectisol regenerator	Stretford Process or any suitable sulfur-recovery process
Sulfur Dioxide	Gas-fired turbines, boilers, heaters, incinerators	In this case, amount of sulfur dioxide below the allowable pollution standard. Otherwise, Wellman-Lord lime treatment, etc.

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Table B-40. INVESTMENT COST FOR LURGI-PROCESS COAL-TO-SNG
(288.6 Million SCF/Day) PLANT

Components	End-of-1973 Cost, $1000
Gas Production System (including lock gas storage and compression)	66,370
Carbon Monoxide	7,680
Gas Cooling and Heat Recovery	8,080
Gas Purification (including refrigeration)	40,640
Methane Synthesis	18,550
Product-Gas Compression and Dehydration	5,660
Gas-Liquor Treatment and By-product Recovery	18,650
Sulfur Recovery System	8,160
Fuel Gas Production System	21,620
Fuel Gas Cooling and Treatment	5,320
Air Compression	20,470
Steam and Power Generation	30,200
Oxygen Production and Compression	28,930
Cooling-Water System	5,800
Raw Water Treatment System and Miscellaneous Plant Utility Systems	11,940
Ash Dewatering and Transfer	6,320
Raw-Water Storage, Pumping, and Pipeline and River-Water Pumping	14,210
Initial Catalyst and Chemicals	4,010
General Facilities	34,720
Total Direct Cost of Plant Including Contractor Engineer Fees, Licenses, and State Taxes	357,780
Contingencies (15%)	53,667
Total Plant Investment (I)	411,447
Interest During Construction (0.23676 X I)	97,414
Start-up Cost (20% of gross operating cost)	17,234
Working Capital	
Coal inventory (60 days of feed at full rate)	8,811
Materials and Supplies (0.9% of total plant investment)	3,703
Net Receivables (1/24 X annual revenue received)	7,273
Total Capital Required	545,882

Table B-41. OPERATING COST FOR LURGI-PROCESS COAL-TO-SNG
(288.6 Million SCF/Day PLANT (90% Stream Factor)

Component	Annual Cost, $1000
Coal Feed (at 489,495.3 X 10^6 Btu/day), 30¢/10^6 Btu	48,240
Other Direct Materials, Catalysts, and Chemicals	3,520
Purchased Utilities	
Raw-Water Cost (at 7300 gpm X 30¢/1000 gal)	1,036
Labor	
Process Operating Labor (62 men, shift at $5/hr and 8304 man-hr/yr)	2,574
Maintenance Labor (1.5% of total plant investment)	6,172
Supervision (15% of operating and maintenance labor)	1,312
Administration and General Overhead (60% of total labor, including supervision)	6,035
Supplies	
Operating (30% of process operating labor)	772
Maintenance (1.5% of total plant investment)	6,172
Local Taxes and Insurance (2.7% of total plant investment)	11,109
Total Gross Operating Cost	86,942
By-product Credit	
Tar Oil (48,588 lb/hr X 24 X 17,300 Btu/lb X $ 0.5/10^6 Btu X 328.5)	3,314
Tar (88.824 lb/hr X 24 X 16,670 Btu/lb X $0.5/$10^6$ Btu X 328.5)	5,837
Phenols (11,271 lb/hr X 24 X $0.04/lb X 328.5)	3,554
Ammonia (21,422 lb/hr X 24/2000 X $25/short ton X 328.5)	2,111
Naphtha (20,005 lb/hr X 24 X 18,400 Btu/lb X $0.5/$10^6$ Btu X 328.5)	1,451
Sulfur (167 LT/day X $10/LT X 328.5)	549
	16,816
Total Net Operating Cost	70,126

Table B-42. CALCULATION FOR DETERMINING UNIT PRODUCTION COST
BY DCF METHOD FOR A LURGI-PROCESS COAL-TO-SNG
(288.6 Million SCF/Day) PLANT

Unit Cost of the Product

$$\frac{N + 0.23816\,I + 0.1275\,S + 0.230777\,W}{G}$$

where

N = Net Operating Cost = $70,126,000

I = Total Plant Investment = $411,447,000

S = Start-up Cost = $17,234,000

W = Working Capital = $19,787,000

G = Annual Product Production (288.6 million SCF/day
X 954 Btu/SCF X 328.5 days/yr) = 90,444,065.4 X 10^6 Btu/yr

$$\text{Unit cost} = \frac{\$174,879,000}{90,444,065.4} = \$1.934/\text{million Btu (high heating value)}$$

$$= \frac{\$174,879,000}{81,515,730} = \$2.145/\text{million Btu (low heating value)}$$

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