NYO-10301

A TECHNICAL AND ECONOMIC EVALUATION OF SOLID-FUEL GASIFICATION USING NUCLEAR HEAT

Production of Pipeline Gas, Liquid Fuels, and Chemicals from Lignite and Bituminous Coal

By
L. J. Pieroni
R. P. Griffin
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November 30, 1962

Chemical Engineering Development Division
The M. W. Kellogg Company
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REACTOR TECHNOLOGY

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ABSTRACT

Results are reported of a study to determine the most economical method of utilizing nuclear heat to gasify a North Dakota lignite and a West Virginia bituminous coal so that the resulting gas is suitable as a source of synthesis gas for fuels and chemicals. Heat is transferred from a high-temperature, gas-cooled nuclear reactor to the solid-fuel gasification plant by a stream of 435-psig helium heated to a maximum temperature of 2500°F. There is no in-reactor processing involved, hence radioactivity in the processing plant is assumed to be negligible. Design of the nuclear reactor is beyond the scope of this study.

It is concluded that the gasification may be carried out more economically in a fluidized bed containing immersed heat-transfer tubes than in either a fixed-bed or an entrained-bed gasifier. If nuclear heat costs no more than 95 cents per million Btu, pipeline gas can be produced at lower cost than by any steam-oxygen gasification process, but still not cheaply enough to compete with natural gas at the present time. Production of gasoline is not economically attractive, but ammonia and methanol deserve more detailed study.

SUMMARY

This report has been prepared to present the results of a technical and economic evaluation of solid-fuel gasification using nuclear heat. The overall objective has been to determine the most economical method of utilizing nuclear heat to gasify a North Dakota lignite and a West Virginia bituminous coal so that the resulting gas will be suitable as a source of synthesis gas for fuels and chemicals.

Two plants are evaluated for each raw material. One produces 90,000,000 standard cubic feet per day (SCFD) of 930-Btu pipeline gas. The other produces a variety of fuels and chemicals, including the following:

A. Ammonia - 600 tons per day.
B. Methanol - 300 tons per day.
C. Hydrogen, 99 percent pure - 35,000,000 standard cubic feet per day.
D. Gasoline - 5000 to 5700 barrels per day.
E. A number of co-products of gasoline manufacture, including diesel oil, liquefied petroleum gas, ethylene, alcohols, and acetic acid.

Heat is transferred from a high-temperature, gas-cooled nuclear reactor to the solid-fuel gasification plant by a stream of 435 psig helium heated to a maximum temperature of 2500°F. A schematic diagram of the gasification section of the plant is presented as Figure 1. There is no in-reactor processing involved, hence radioactivity in the processing plant is assumed to be negligible. The design of the nuclear reactor is beyond the scope of this study. Investment figures quoted do not include investment in the nuclear portion of the plant; calculated product costs do include a charge for the nuclear heat consumed, however, and this is treated as a parameter in the evaluation.

It is concluded that steam gasification of solid fuels using nuclear heat transferred by the reactor coolant in a closed loop (no direct contact between reactor coolant and process streams) may be conducted more economically in a fluidized bed containing immersed heat-transfer tubes than in either a fixed-bed or an entrained-bed gasifier. Optimum gasification conditions for producing pipeline gas are believed to be as follows:

	Lignite	Bituminous Coal
Temperature, °F	1600	1800
Pressure (helium and gasifier), psia	450	450
Molar ratio, steam fed/carbon gasified	1.25	3.12
Carbon utilization (approx.), %	90	80

A plant producing 90,000,000 standard cubic feet per day of 930-Btu pipeline gas from lignite will require an investment of about

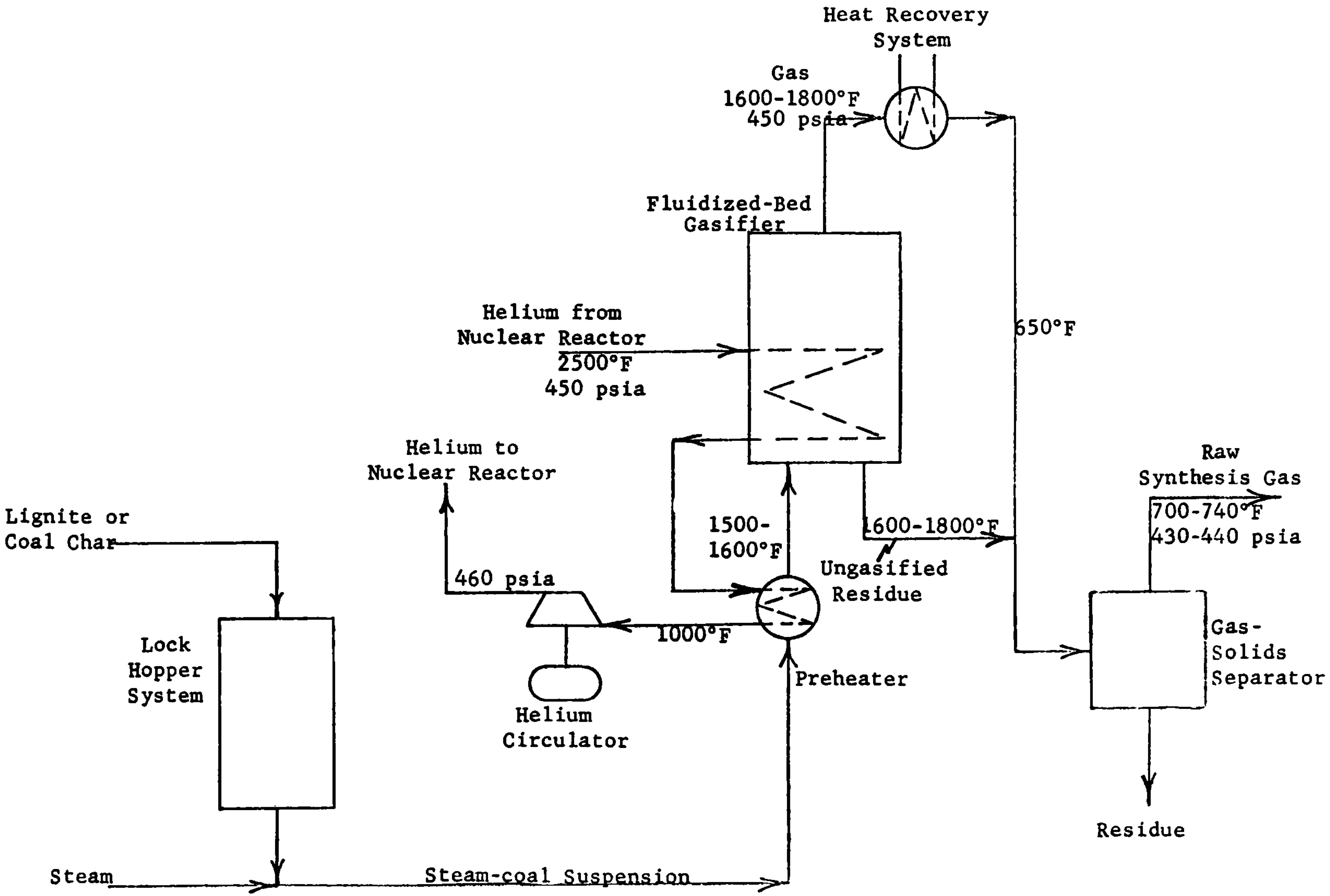


Figure 1

Schematic Flow Diagram of Gasification Section

$45,000,000, not including the nuclear reactor. If bituminous coal is used as raw material, investment will be about $70,000,000. Selling price of product gas will depend on cost of raw material, cost of nuclear heat and desired return on investment, among other factors. These effects are summarized in Table 1, where gross return on investment includes interest on bonded indebtedness, Federal income taxes, and net return to investors of equity capital.

Table 1
Effect of Raw Material Cost, Cost of Nuclear Heat, and Gross Return on Investment on the Selling Price of Pipeline Gas

Cost of Nuclear Heat ¢/MM Btu	Selling Price of Pipeline Gas, ¢/MSCF					
	Lignite			Bituminous Coal		
	$1.50/ ton	$2.00/ ton	$2.50/ ton	$4.00/ ton	$5.00/ ton	$6.00/ ton
	------------------Gross return = 12%----------------------					
50	71	75	79	108	113	118
80	78	82	86	119	124	129
	------------------Gross return = 20%----------------------					
50	83	87	91	127	132	137
80	90	94	98	138	143	148
	------------------Gross return = 30%----------------------					
50	98	102	106	151	156	161
80	105	109	113	162	167	172

The major difference between lignite and bituminous coal, yielding a difference in gas selling price of 40-60 cents per thousand standard cubic feet, is the higher reactivity of the lower-rank fuel, lignite. It is evident from Table 1 that desired return on investment is a more important factor in determining gas selling price than either the cost of raw material or the cost of nuclear heat.

The temperature at which helium is available from the nuclear reactor has a significant effect on the price one can afford to pay for nuclear heat. For example, with bituminous coal at $5 per ton, 2500°F helium at 50 cents per million Btu, and 20 percent gross return on investment, calculated gas selling price is $1.32 per MSCF (Table 1). If helium temperature is 2300°F instead, a charge of only 25 cents per million Btu for nuclear heat will yield the same gas selling price.

A comparison with the results of previous studies shows that nuclear-heated steam gasification of bituminous coal to produce pipeline gas can compete with the best steam-oxygen gasification process if nuclear heat costs no more than 95 cents per million Btu. The cost of nuclear heat must include all operating costs and return on investment associated with the nuclear portion of the plant.

The plant for making gasoline, ammonia, methanol, hydrogen, and other chemicals would require an investment of about $133,000,000 with lignite as raw material, or about $142,000,000 with bituminous coal (again not including the nuclear reactor). The sale of all products at their respective current market prices would yield a gross return on investment of only about 10 percent. Certain portions of the plant are economically more interesting than the whole, however. Ammonia is produced at a cost of $33 per ton and, if sold at the current selling price, would yield a 50 percent gross return. Methanol, with a production cost of about 10 cents per gallon, shows nearly the same result. The cost of nuclear heat has relatively little effect on the production cost of ammonia or methanol, and it is concluded that these areas are worthy of more detailed study.

Selling prices for the fuels and chemicals are nearly the same whether the raw material is lignite or bituminous coal, in contrast with the results for pipeline gas production. The reason is found in the appreciable quantity of methane produced when lignite is gasified. This is an advantage when the final product is pipeline gas, but is a distinct liability if chemicals or liquid fuels are to be made.

Because of the huge investments involved, the manufacture of pipeline gas, liquid fuels, or chemicals from solid fuels must be a base-load operation. Product costs have been calculated using an assumed stream efficiency of 90 per cent.

CONCLUSIONS

In the steam gasification of solid fuels using nuclear heat transferred by the reactor coolant in a closed loop (no direct contact between reactor coolant and process streams), use of a fluidized bed containing immersed heat-transfer tubes is more economical than either a fixed-bed or an entrained-bed gasifier. Easily the most expensive portion of the gasification system is the high-temperature tubular surface for transferring heat. With a fluidized-bed gasifier less heat-transfer surface is required, and the average tube metal temperature is lower than for either of the other two gasifiers.

Optimum gasification conditions for manufacturing pipeline gas from the two raw materials studied were concluded to be as follows:

Raw material	Lignite	Bituminous Coal
Temperature, °F	1600	1800
Pressure (helium and gasifier), psia	450	450
Molar ratio, steam fed/carbon gasified	1.25	3.12
Carbon utilization (approx.), %	90	80

Before a commercial plant is built, further experimental work is necessary in the areas of reaction kinetics in steam gasification of solid fuels and of the effect of operating conditions on effluent gas composition. Pilot-plant operation of the fluidized-bed gasifier containing tubular heat transfer surface will be necessary to demonstrate continued operability while injecting coal tar. Although immersed heat transfer surface has been used commercially with other bed materials, there is little experience concerning the amount of erosion that will occur. This effect should be investigated in a pilot plant. High-temperature materials of construction should be tested on bench-scale and pilot-plant units.

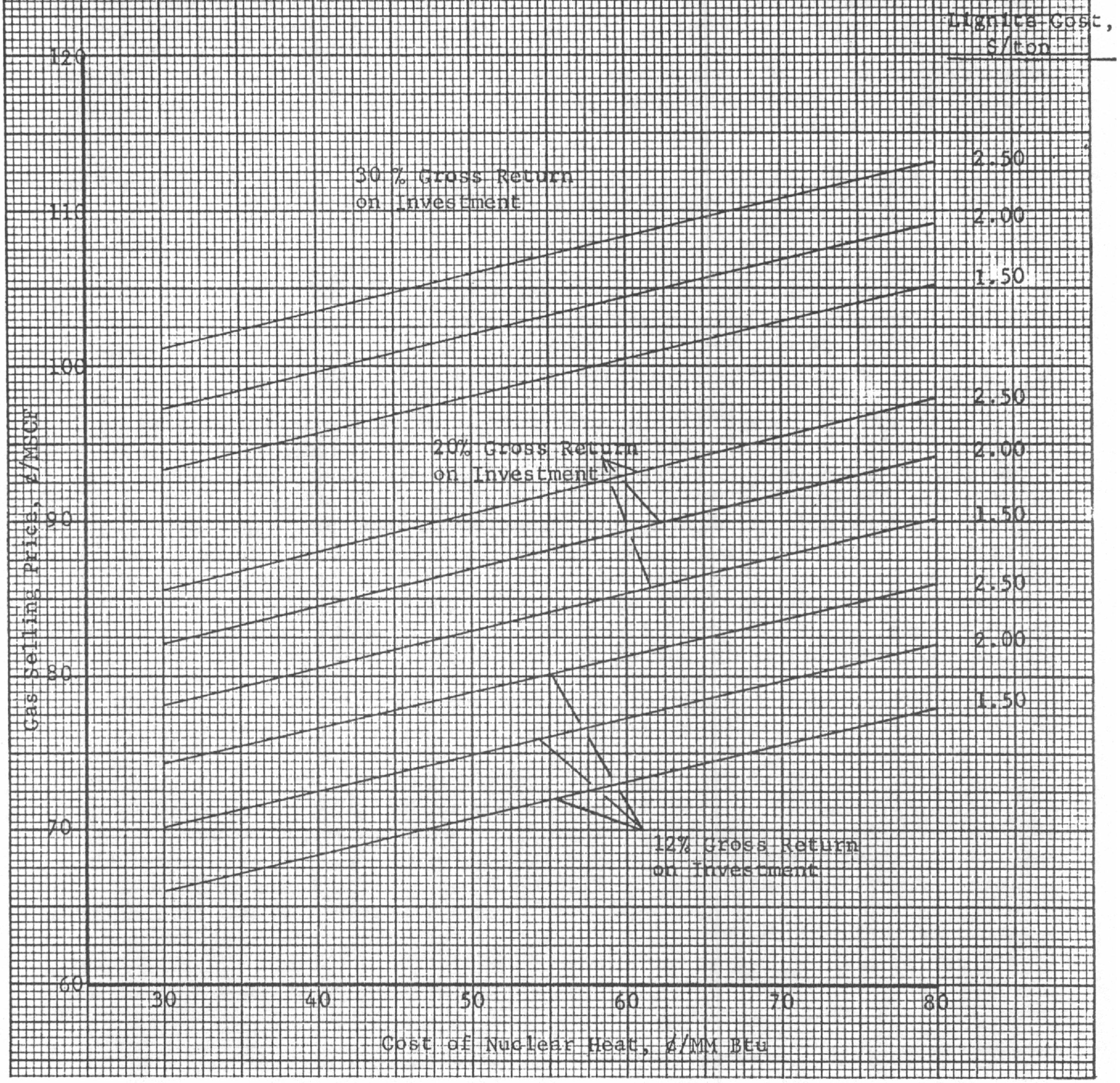
Production of pipeline gas from solid fuels is of interest because of the large potential market for the gas. A plant producing 90,000,000 SCFD of 930-Btu gas from lignite will require an investment of about $45,000,000 not including the 255-thermal-megawatt nuclear reactor. If bituminous coal is used as raw material, investment will be about $70,000,000 not including a 406 Mw(t) reactor. Selling price of product gas will depend on cost of raw material, cost of nuclear heat, and desired return on investment, among other factors. The effects are shown in Figures 2 and 3. The major difference between lignite and bituminous coal, yielding a difference in gas selling price of 40 - 60¢/MSCF, is the higher reactivity of the lower-rank fuel, lignite. It is evident from these graphs that desired return on investment is a much more important factor in determining gas selling price than either the cost of raw material or the cost of nuclear heat.

Pipeline gas selling price calculated for the present study is compared below with the results of other studies, all on bituminous coal, adjusted to the same basis of calculation where necessary:

Figure 2

PIPELINE GAS FROM LIGNITE

Effect of Lignite Cost, Cost of Nuclear Heat, and Return on Investment on the Selling Price of Gas



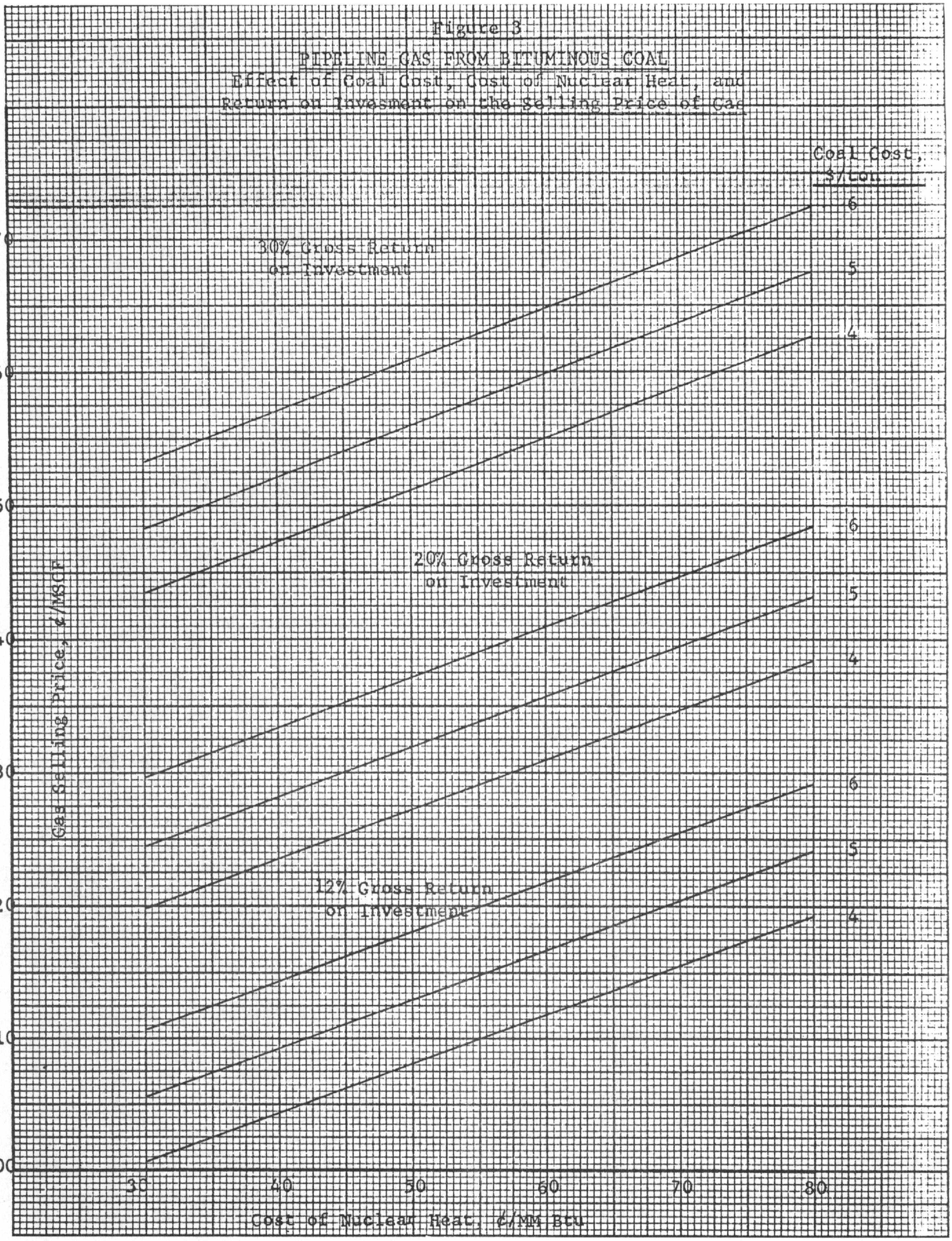


Figure 3

PIPELINE GAS FROM BITUMINOUS COAL

Effect of Coal Cost, Cost of Nuclear Heat, and Return on Invesment on the Selling Price of Gas

Source of Information	Process	Gas selling price at 20% gross return, 90% stream efficiency, $5/ton coal; $/MSCF
Present study	Fluid-bed gasification, nuclear heat at:	
	50¢/MM Btu	1.32
	80¢/MM Btu	1.43
	$1/MM Btu	1.51
Kellogg contract research studies	Lurgi steam-oxygen gasification	1.70
	Suspension gasification with steam and O_2	1.49
Bureau of Mines (11)	Suspension gasification with steam and oxygen	1.55

It is seen that nuclear heat at about 95¢/MM Btu can be competitive with the best alternative process for gasifying bituminous coal. Cost of nuclear heat, it should be made clear, must include all operating costs and return on investment associated with the nuclear portion of the plant.

Pipeline gas selling prices in Figures 2 and 3 are based on helium, the nuclear reactor coolant, being supplied to the processing plant at 2500°F. If this temperature is lower than 2500° the cost of the nuclear reactor will probably decline, but cost of the gasification plant will increase because of the lower temperature driving force, larger heat transfer surface requirement, and larger helium circulator. Figure 4 shows the cost of nuclear heat which, if helium is supplied at some temperature below 2500°F, will yield the same gas selling price as shown in Figures 2 and 3. It seems apparent that helium temperature should be at least 2500°F for gasifying bituminous coal. The optimum temperature will be lower for lignite than for bituminous, but it cannot be chosen until the effect of coolant temperature on nuclear reactor investment and operating costs is known.

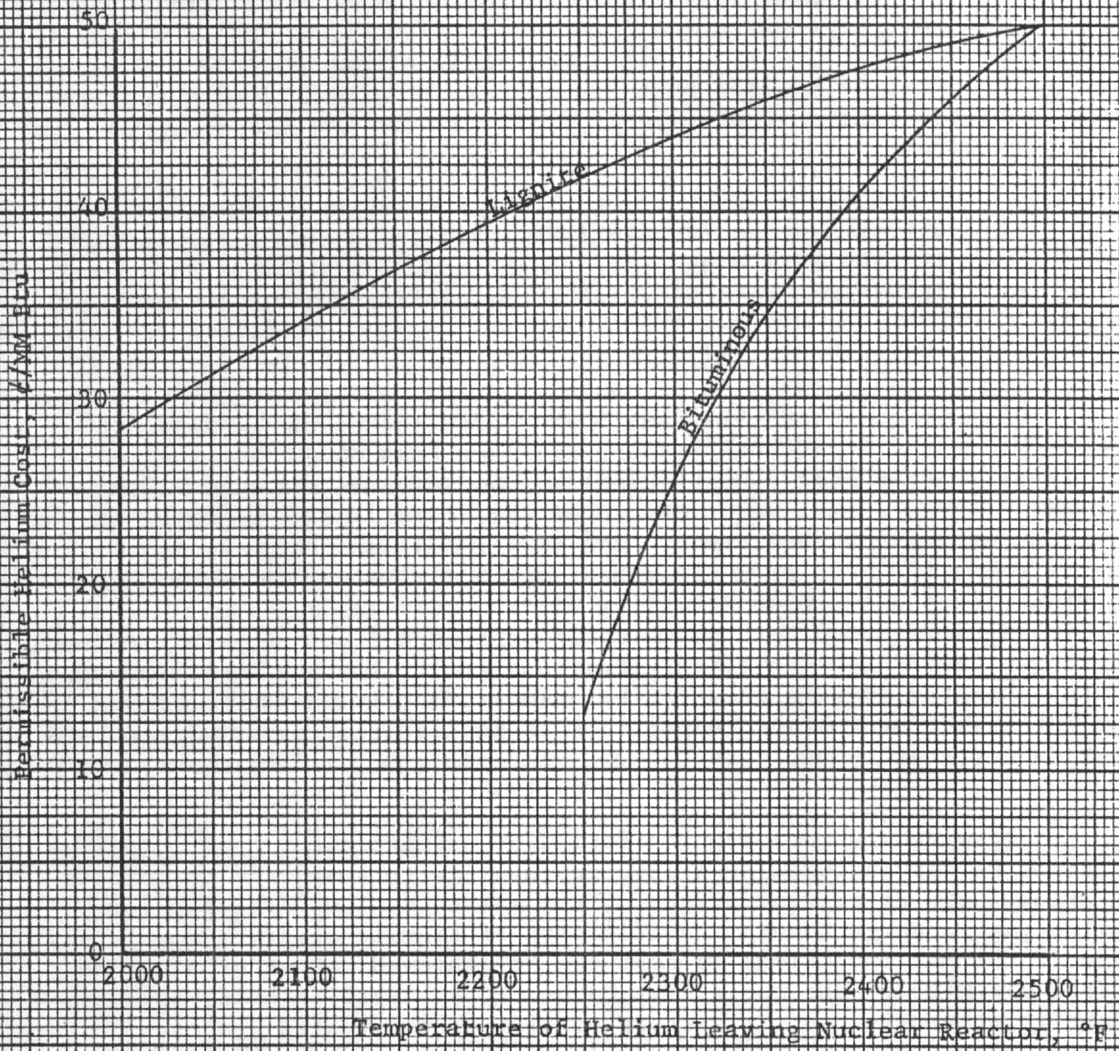
Despite the advantages afforded by using nuclear heat, pipeline gas cannot be manufactured at a low enough price to enable it to compete with natural gas at current prices. If a market can be developed for pipeline gas in North Dakota, gasification of lignite deserves more study to see if further reductions in cost can be achieved. Cost of pipeline gas from bituminous coal would have to be cut at least in half, however, to make it attractive in the eastern portion of the United States.

The manufacturing complex investigated for making gasoline, ammonia, methanol, hydrogen, and other chemicals shows a relatively small

Figure 4

Permissible Helium Cost
to Yield Same Pipeline Gas Selling Price
as 2500°F Helium at 50¢/MM Btu

Basis: 20% Gross Return on Investment



return on investment. The sale of all products at their current market prices would yield a gross return of only about 10 percent. On the other hand, certain portions of the plant are exceedingly interesting in this respect. Ammonia is produced at a cost of about \$33 per ton and, if sold at the current selling price, \$92 per ton, would yield a 50 percent gross return on investment. Methanol shows just about the same result, with a production cost of 10 cents per gallon and current selling price of 30 cents per gallon. Production costs for these two products are minimized by virtue of having synthesis gas supplied by a very large plant, with attendant lower unit costs. A gasification plant to provide synthesis gas for only ammonia and/or methanol production would be about 10 to 25 percent as large as designed for the present study. Unit costs would be higher, but it is felt that manufacture of ammonia and methanol would still yield a reasonable return. Cost of nuclear heat would also be higher from a much smaller reactor, but this has a relatively small effect on production cost. Increasing the cost of nuclear heat from 50¢/MM Btu to \$1.50/MM Btu, for example, increases the cost of ammonia by \$8 per ton and methanol by 3 cents per gallon.

Hydrogen production cost (before return on investment) is about 28¢/MSCF, with coal at \$5/ton and nuclear heat at 50¢/MM Btu. This compares with reported production costs of 37¢/MSCF for steam reforming of natural gas costing 35¢/MM Btu, and about 45¢/MSCF for partial oxidation of fuel oil costing \$2.00 per barrel (7).

Selling prices of the liquid fuels and chemicals studied are approximately the same whether the raw material is lignite or bituminous coal. This is in contrast to the pipeline gas results, in which lignite yielded much lower costs. The explanation is found in the appreciable quantity of methane produced when lignite is gasified. This is an advantage when the final product is pipeline gas, but is a distinct liability if chemicals or liquid fuels are to be made.

It appears that the best applications for nuclear heat will be found in trying to improve an existing commercial process, not in trying to make a currently uneconomic process look attractive. For example, gasoline and pipeline gas are not produced from synthesis gas in the United States today, and the advent of cheap nuclear heat will not change the picture significantly. On the other hand, ammonia and methanol are manufactured from synthesis gas on a commercial scale, and the use of nuclear heat may furnish a significant cost advantage, whether the raw material used to make the synthesis gas be coal, oil, or natural gas. This would seem to be a fruitful area for further study.

Because of the huge investments involved, the manufacture of pipeline gas, liquid fuels, or chemicals from solid fuels must be a base-load operation. For example, selling price of pipeline gas from bituminous coal at \$5/ton, with nuclear heat at 50¢/MM Btu and 20 percent gross return on investment, increases from \$1.32/MSCF at 90 percent stream efficiency to $1.75/MSCF at 60 percent. And this does not take into account the increased cost of nuclear heat at the lower load factor.

At the present stage of development, a caking bituminous coal must be pretreated to prevent agglomeration as it is gasified. Pretreatment was accomplished in this study by fluidized-bed low-temperature carbonization, and it increased the cost of pipeline gas by about 3¢/MSCF. Further development work may reduce or eliminate this cost.

RECOMMENDATIONS

The economics of using nuclear heat in the manufacture of ammonia and methanol should be investigated in more detail. In addition to the gasification of solid fuels considered here, the application to existing commercial processes which use natural gas or liquid hydrocarbons as raw material to make ammonia and methanol should also be studied.

Fluidized-bed steam gasification of solid fuels, using heat transferred through tubes immersed in the bed, should be demonstrated on a pilot-plant scale. Possible methods of introducing a caking bituminous coal (without pretreatment) into the fluidized-bed gasifier should be studied.

INTRODUCTION

I. Purpose and Scope of Project

Many studies have been made of the conversion of coal to synthesis gas for use in the manufacture of ammonia, methanol, synthetic pipeline gas, liquid fuels, or reducing gas for direct reduction of ores. While the experimental programs have shown that it is technically feasible to gasify coal, economic evaluations have shown that the costs of such gasification are too high to permit this route to compete with present technology using naphtha or natural gas as sources of synthesis gas. The major contributor to these high costs has been the cost of oxygen. The competitive position of coal gasification processes would be improved greatly if oxygen costs could be eliminated.

The reaction that takes place when coal is gasified with steam to produce synthesis gas is endothermic; that is, it absorbs heat. The only reason for using oxygen is to burn a portion of the coal to carbon dioxide, thus providing the required heat. Air cannot be used because the nitrogen would end up as a major impurity in the synthesis gas. Using a nuclear reactor to supply the heat of reaction is one method that has been suggested for eliminating oxygen. A corollary benefit would be the reduction of carbon dioxide concentration in the synthesis gas.

This report has been prepared to present the results of a technical and economic evaluation of solid-fuel gasification utilizing nuclear heat. The work was conducted under contract with the U. S. Atomic Energy Commission. Overall objective of the work has been to determine the most economical method of using nuclear heat to gasify a lignite and a bituminous coal so that the resulting gas can be used to compete successfully as a source of synthesis gas for ammonia, methanol, hydrogen, pipeline gas, liquid fuels, and other chemicals.

II. Background Information

The application of nuclear energy to solid-fuel gasification is receiving increasing attention in the technical literature. In 1958 Katell (10) compared nuclear-heated steam gasification of bituminous coal with steam-oxygen gasification. With feed coal at $5 per ton and depreciation at the rate of 5 percent, he estimated the following gas costs:

	¢/MSCF	
	Before return on investment	With 12% gross return
Steam-oxygen gasification	78	121
Nuclear-heated steam gasification	64	93

Dalzell and McGee (6) discuss the use of nuclear heat in the form of helium at 2700°F to gasify a low-rank coal. Gasification of higher-rank coals is also expected to be attractive if heat exchangers (gasifiers) capable of operating in the range of 2000° to 2500°F can be developed. Perry and McGee (24) conclude that nuclear heat at 71¢/MMBtu is competitive with steam-oxygen gasification to produce synthesis gas if coal costs $4 per ton, or at 75¢/MM Btu if coal costs $5 per ton.

Experimental work is also being done along these lines. In cooperation with the Atomic Energy Commission, the U. S. Bureau of Mines has built and operated at its Morgantown, West Virginia, station an experimental gas loop to evaluate component performance and point out areas where further development is needed. Hot gas is circulated between an electrically-heated simulated nuclear reactor and an exchanger-type gasifier where coal and steam react (22, 21, 23). A 1000-hour demonstration run using helium at 2500°F and 250 psig as the heat-transfer medium has been described by Coates, McGee, and Fasching (5). Design of the system, and experimental progress in this and related areas are described in a series of quarterly reports (3).

III. Basis of Evaluation

A. Raw Materials Considered

Two raw materials have been considered for gasification, a North Dakota lignite and a West Virginia bituminous coal. Both are representative of major U. S. fuel reserves. One -- the lignite -- is highly reactive, thus easy to gasify, but is unfortunately found in a section of the country where markets for the end products are presently limited. The other -- the bituminous coal -- is less reactive and presents more difficulty in processing because it tends to cake or agglomerate as it is gasified. Because the bituminous coal is mined near the great population centers of the East Coast, marketing of the end products would present no problem if they could be manufactured at competitive prices.

The following analyses were chosen as typical of these two raw materials:

Fuel type	Lignite	Caking Bituminous Coal
Description	North Dakota; Beulah Seam	West Virginia; Pittsburgh Seam; High-volatile A
Proximate analysis:	Weight Percent	Weight Percent
H_2O	34.8	6.0
Volatile matter	28.2	34.2
Fixed carbon	30.8	51.1
Ash	6.2	8.7
	100.0	100.0

Fuel type	Lignite	Caking Bituminous Coal
Ultimate analysis (dry):	Weight Percent	Weight Percent
C	65.0	75.4
H	4.3	5.1
O	17.5	5.7
N	2.6	1.4
S	1.1	3.2
Ash	9.5	9.2
	100.0	100.0
Higher heating value (as received), Btu/pound	7,210	12,930

B. Plant Location

To minimize raw material transportation costs, the processing plants would be located at, or very near, the coal mine and close to an adequate cooling water supply, such as a large river. For the purposes of this study, it is assumed that the plant using lignite is constructed in North Dakota, while the other, processing bituminous coal, is in northern West Virginia.

C. Products Studied

Two separate plants have been evaluated for each raw material. One would produce 90,000,000 standard cubic feet per day (SCFD) of "pipeline gas", a product that can be substituted for natural gas in the country's pipeline network when supplies of natural gas are no longer adequate to meet demand. Plant capacity, at 90,000,000 SCFD, is equal to about one-quarter of one percent of U. S. natural gas consumption. Gross heating value of pipeline gas is about 915-930 Btu/SCF.

The second plant would produce a variety of products, as listed below:

1. Ammonia - 600 tons per day (about 3 percent of present U. S. synthetic ammonia capacity).
2. Methanol - 300 tons per day (10 percent of present U. S. production capacity).
3. Hydrogen, 99 percent purity - 35,000,000 SCFD (enough to produce 600 to 1400 tons of steel per day, depending on the process used, by direct reduction of iron ore).

4. Gasoline - 5,000 to 5,700 barrels per day (about 0.1 percent of present U. S. consumption).

5. A number of co-products of gasoline manufacture, including principally diesel oil, liquefied petroleum gas (LPG), ethylene, alcohols, and acetic acid.

D. Utilization of Nuclear Heat

Several methods can be visualized for utilizing nuclear heat to carry out an endothermic chemical reaction, including at least the following:

1. Chemical reactants can be preheated individually in the nuclear reactor, then reacted elsewhere, so that the sensible heat of reactants provides the required heat of reaction.

2. The chemical reaction can be conducted inside the nuclear reactor.

3. Heat can be transferred from nuclear reactor to adjacent processing plant by an inert heat carrier; for example, by the reactor coolant.

The three systems have previously been discussed by Graham (8) and by Dalzell and McGee (6).

The present study is limited to alternate 3, above. Heat is transferred from a high-temperature, gas-cooled reactor to the coal (or lignite) gasification plant by a stream of helium, the reactor coolant, heated to 2500°F at 435 psig. Helium returns to the reactor at a temperature of 1000° to 1450°F. There is no in-reactor processing involved, hence radioactivity in the processing plant is assumed to be negligible. A slipstream of helium must be continuously withdrawn from the nuclear reactor and purified of fission products to limit radioactivity to a tolerable level.

Nuclear reactor capacity is about 255 thermal megawatts for the lignite gasification plant and about 406 thermal megawatts in the case of bituminous coal. Design or evaluation of the nuclear reactor is beyond the scope of this study. Investment figures quoted do not include investment in the nuclear portion of the plant; calculated product costs do include a charge for the nuclear heat consumed, however. Cost of nuclear heat is treated as a parameter in the evaluation, and must be understood to include operating costs, return on investment, and any special costs associated with the nuclear portion of the plant (except for the helium circulators, which are included in the process plant).

E. Method of Presentation

The "Discussion" portion of this report is divided into four major parts:

Section I.

Part A. Production of pipeline gas from lignite.

Part B. Production of liquid fuels and chemicals from lignite.

Section II.

Part A. Production of pipeline gas from bituminous coal.

Part B. Production of liquid fuels and chemicals from bituminous coal.

Each part contains a description of the final process design, an evaluation of economics, a discussion of process design considerations, optimization of process variables, and possible alternate processing schemes, and an appendix of related information.

DISCUSSION

I. Lignite

A. Production of Pipeline Gas

1. Description of Process

For convenience, the plant for production of pipeline gas from lignite has been functionally divided into the following sections:

Section 000 - Lignite Storage and Reclamation
Section 100 - Crushing
Section 200 - Drying and Gasification
Section 300 - Shift Conversion
Section 400 - Gas Purification
Section 500 - Methanation and Compression
Section 1100 - Offsite Facilities

An overall process material balance flowsheet, divided into the above sections, is presented as Drawing No. CE-1185-B.

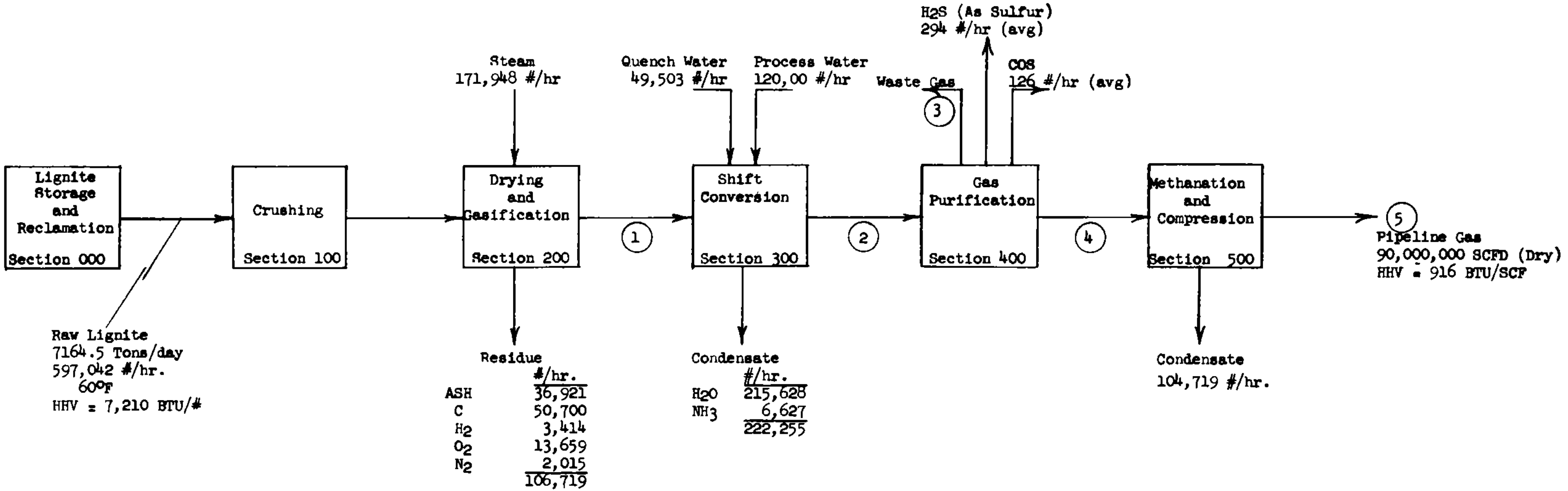
In the following pages Section 200 is described in detail and the other sections of the plant are discussed briefly. For further descriptive information on those sections only briefly discussed, reference may be made to previous coal gasification studies prepared by Kellogg for the U. S. Bureau of Mines (12) and the American Gas Association (13, 14).

a. Section 000 - Lignite Storage and Reclamation

Lignite is received by truck or conveyor belt from an adjacent mine during eight hours each day at an approximate rate of 900 tons per hour. The 4" x 0 lignite is transported to a distribution center, where about 300 tons per hour is dispatched to Section 100 for immediate processing while the remainder is conveyed to the storage area.

The lignite sent to storage is distributed over several piles which contain enough lignite to permit the plant to operate at normal capacity for 30 days in the event that the lignite supply is cut off. During the sixteen hours each day that the mine is not in operation, lignite for process use is reclaimed from the storage piles by gravity flow into underground tunnels and onto conveyor belts which transport the lignite to Section 100.

The equipment in this section has been spared in such a manner as to provide continuous operation at full capacity.



Proximate Analysis	WT. %
Water	34.8
Volatile Matter	28.2
Fixed Carbon	30.8
Ash	6.2
	100.0

Ultimate Analysis	WT. %
C	42.4
H_2	6.7
O_2	42.3
S	0.7
N_2	1.7
ASH	6.2
	100.0

Stream No.	①			②			③			④			⑤		
Temperature	735°F			100°F			—			80°F			100°F		
Pressure	428 PSIA			400 PSIA			—			380 PSIA			1000 PSIA		
Flow Rate	#/Hr	MPH	Mole %	#/Hr	MPH	Mole %	#/Hr	MPH	Mole %	#/Hr	MPH	Mole %	#/Hr	MPH	Mole %
CO	277,524	9,912	31.6	158,451	5,658.0	16.0	1,008	36.0	0.4	157,443	5,622.0	21.0	42	1.5	0.01
CO_2	169,776	3,858	12.3	356,940	8,112.0	23.0	345,000	7,840.5	96.3	11,940	271.5	1.0	8,157	185.4	1.9
CH_4	49,521	3,096	9.9	49,521	3,096.0	8.8	1,983	123.9	1.5	47,538	2,972.1	11.1	138,858	8,678.7	87.8
H_2	27,855	13,929	44.3	36,360	18,180.0	51.7	60	30.0	0.4	36,300	18,150.0	66.6	1,884	942.0	9.4
N_2	2,367	84.3	0.3	2,367	84.3	0.2	---	---	---	2,367	84.3	0.3	2,367	84.3	0.9
NH_3	6,627	390	1.2	---	---	---	---	---	---	---	---	---	---	---	---
H_2S	3,714	116.1	0.4	4,134	128.4	0.3	3,840	119.2	1.3	---	---	---	---	---	---
COS	861	14.4	0.05	126	2.1	0.01	---	---	0.1	---	---	---	---	---	---
Dry Gas	538,245	31,399.8	100.0	607,899	35,260.8	100.0	351,891	8,149.6	100.0	255,588	26,859.9	100.0	151,308	9,891.9	100.0
H_2O	124,026	6,891		1,620	90.0		1,011	56.1		609	33.9		170	9.5	
Wet Gas	662,271	38,290.8		609,519	35,350.8		352,902	8,205.7		256,197	26,893.8		151,478	9,901.4	

NO.	REVISION DESCRIPTION	DATE	BY	CHK.

ISSUED FOR FABRICATION
ISSUED FOR CONSTRUCTION

SCALE: NONE
DRAWN: RPG
CHECKED: RPG
APPROVED:
DATE:

THE M. W. KELLOGG COMPANY

PIPELINE GAS FROM LIGNITE
PROCESS STREAM BALANCE

CLASS	AREA	JOB NO.	DRAWING NO.
		4006	CE-1185-B

b. Section 100 - Crushing

The 4" x 0 lignite received from Section 000 at the rate of 300 tons per hour is crushed to 100 percent through a 1-1/4-inch screen in two Jeffrey Flextooth crushers. The lignite is then conveyed to two Pennsylvania reversible hammermills where it is milled to 98 percent minus 14 mesh. The milled lignite is transported to Section 200 by conveyor belt and bucket elevator. The equipment in the crushing section has also been judiciously spared in order to ensure continuous operation at full capacity.

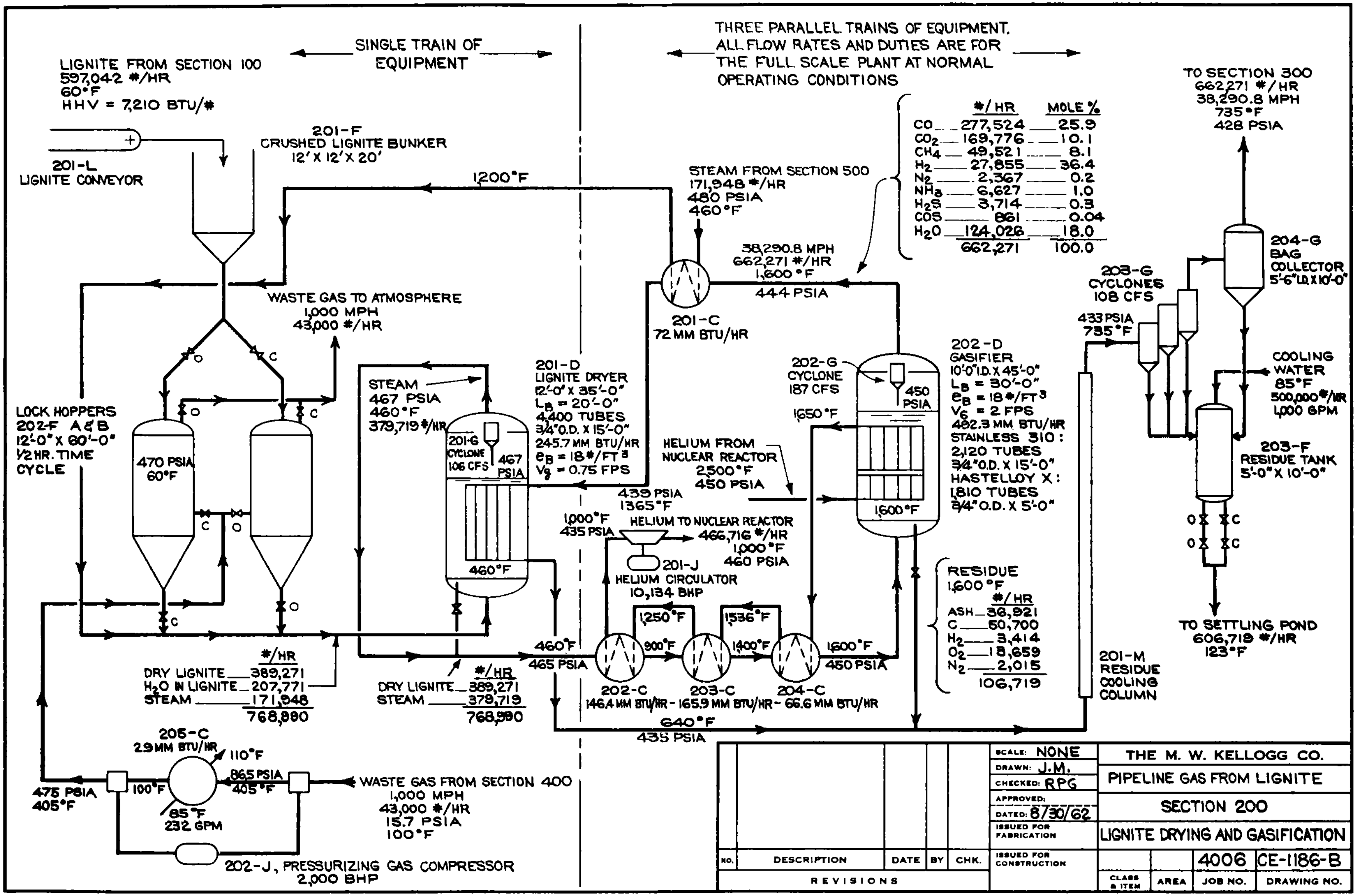
c. Section 200 - Drying and Gasification

The following description of the drying and gasification section may best be understood by referring to Drawing No. CE-1186-B, the process flowsheet for Section 200, which shows equipment sizes, flow rates, and operating conditions.

Lignite, sized to 98 percent minus 14 mesh and containing 34.8 weight percent moisture is received from Section 100 at the rate of 597,042 pounds per hour and is fed by elevated conveyor 201-L into the lignite bunker, 201-F. The lignite flows periodically from the bunker into lock hopper 202-F, which is alternately loaded and discharged on a one-half-hour time cycle. The hopper is pressurized to 470 psia with gas supplied by compressor 202-J. Steam, which has been superheated to 1200°F in exchanger 201-C, is used to convey the lignite from 202-F to the fluidized bed dryer 201-D. The 480 psia conveying steam is superheated to ensure that there is no condensation of water when the 60°F lignite discharging from the lock hopper mixes with the steam.

In dryer 201-D, which operates at 467 psia and 460°F, the moisture contained in the raw lignite is vaporized by heat supplied through heat exchange tubes immersed in the fluidized bed. The heat-supplying medium is raw synthesis gas, which enters at 1365°F and exits at 640°F, transferring 245.7 million Btu per hour. The fluidizing steam and vaporized moisture leave the dryer at a combined rate of 379,719 pounds per hour after passing through cyclone 201-G to remove entrained lignite particles. The dry lignite, which experiences a further degradation of size consist when its moisture content is reduced (4), is withdrawn from 201-D at a rate of 389,271 pounds per hour and is transported in three parallel streams by the effluent steam through heat exchangers 202-C, 203-C, and 204-C. In these exchangers the steam-lignite mixture is superheated by hot helium to a temperature of 1600°F. Commencing with these exchangers, Section 200 consists of three parallel trains of operating equipment.

The superheated mixture then enters fluidized-bed gasifier 202-D, which operates at 450 psia and 1600°F, where the steam reacts with the lignite to form raw synthesis gas. The 492.3 million Btu



	#/HR	MOLE %
CO	277,524	25.9
CO_2	169,776	10.1
CH_4	49,521	8.1
H_2	27,855	36.4
N_2	2,367	0.2
NH_3	6,627	1.0
H_2S	3,714	0.3
COS	861	0.04
H_2O	124,026	18.0
	662,271	100.0

RESIDUE
1,600°F

	#/HR
ASH	36,921
C	50,700
H_2	3,414
O_2	18,659
N_2	2,015
	106,719

	#/HR
DRY LIGNITE	389,271
H_2O IN LIGNITE	207,771
STEAM	171,948
	768,880

	#/HR
DRY LIGNITE	389,271
STEAM	379,719
	768,990

NO.	DESCRIPTION	DATE	BY	CHK.
REVISIONS				

SCALE: NONE
DRAWN: J.M.
CHECKED: RPG
APPROVED:
DATED: 8/30/62
ISSUED FOR FABRICATION
ISSUED FOR CONSTRUCTION

THE M. W. KELLOGG CO.
PIPELINE GAS FROM LIGNITE
SECTION 200
LIGNITE DRYING AND GASIFICATION

CLASS & ITEM	AREA	JOB NO.	DRAWING NO.
		4006	CE-1186-B

per hour endothermic heat of reaction is supplied through heat exchange tubes immersed in the fluidized bed by a 450 psia stream of hot helium which enters at 2500°F and exits at 1650°F. Eighty percent of the carbon in the lignite fed to 202-D is gasified, and the 1600°F raw synthesis gas leaves the reactor at a rate of 662,271 pounds per hour after passing through cyclone 202-G to remove entrained solid particles. Solid residue is withdrawn from the gasifier at a rate of 106,709 pounds per hour.

The overhead gasifier effluent is cooled to 640°F while superheating steam in 201-C and providing the heat of vaporization in lignite dryer 201-D. The gas then transports the gasifier residue through the cooling column 201-M, where heat is transferred from the solids to the gas. The residue is separated from the gas at 735°F in cyclones 203-G and bag collector 204-G, and is mixed with cooling water in residue tank 203-F. The 123°F slurry is then discharged from 203-F to a settling pond in Section 1100. The solids-free raw synthesis gas exits the bag collector at 735°F and 428 psia and proceeds to Section 300 for further processing.

As previously mentioned, the endothermic heat of gasification is supplied to 202-D by a stream of hot helium. This 466,716 pounds per hour stream is received from the nuclear reactor at 2500°F and leaves the gasifier heat transfer tubes at 1650°F. The helium is then cooled to 1000°F as it transfers heat to the steam-lignite mixture in exchangers 204-C, 203-C and 202-C. The 1000°F, 435 psia gas is compressed to 460 psia by helium circulator 201-J and is returned to the nuclear reactor.

d. Section 300 - Shift Conversion

Section 300 consists of three parallel operating trains of equipment. It is the purpose of this section to adjust the H_2/CO ratio in the raw gas to approximately 3/1 in preparation for synthesis of methane.

The 735°F, 428 psia raw synthesis gas stream is received from Section 200 and enters the shift converter where it is contacted with an iron oxide catalyst in order to promote the water-gas shift reaction:

$$CO + H_2O \rightleftharpoons CO_2 + H_2 + \text{heat} \qquad \text{(I-1)}$$

The reaction is mildly exothermic, therefore boiler feed water is injected into the converter to absorb heat.

The 800°F shift effluent is cooled to 100°F in a series of heat exchangers in which the waste heat is used to heat boiler feed water, generate steam, and superheat steam. The cooled synthesis gas is then countercurrently scrubbed with clean water to remove ammonia, and flows to Section 400 at the rate of 609,834 pounds per hour.

e. Section 400 - Gas Purification

Section 400 also consists of three parallel trains of operating equipment. This section has been designed to reduce the CO_2 concentration in the shifted synthesis gas to 1.0 mole percent and to reduce the total sulfur content to about 0.004 grains per 100 SCF of gas. The purification sequence consists of the "Fluor Solvent CO_2 Removal Process" for CO_2 and bulk H_2S removal, followed by sponge iron (iron oxide) and activated carbon for residual H_2S and organic sulfur removal, respectively. All organic sulfur is assumed to be carbonyl sulfide (COS).

The 100°F, 400 psia gas from Section 300 is countercurrently contacted with an organic solvent in an absorption tower in order to reduce the CO_2 concentration to the desired level. The CO_2-rich solvent from the bottom of the absorber is reconcentrated in a series of flash drums and then is recirculated to the absorption tower. Some flash gas, rich in methane, is recycled to the absorber, while the remainder is vented from the system and is subsequently used to pressurize the lock hoppers in Section 200.

The partially purified gas from the absorber is next treated for removal of residual H_2S. The gas, whose H_2S content has been reduced to approximately 20 grains per 100 SCF in the absorber, is contacted with finely divided iron oxide supported on wood chips (commonly called "sponge iron"), and the H_2S is removed according to the following reaction:

$$Fe_2O_3 \cdot 6\ H_2O + 3\ H_2S \longrightarrow Fe_2S_3 \cdot 6\ H_2O + 3\ H_2O \qquad \text{(I-2)}$$

Each train of equipment contains seven parallel iron oxide drums followed by an iron oxide guard chamber. Each drum is revivified with air every twelve days and the sponge iron is replaced every 65 days.

Synthesis gas leaving the iron oxide drums is finally treated for COS removal by adsorption on fixed beds of activated carbon. The activated carbon drums are arranged in six parallel trains, each train consisting of three drums which are manifolded for cyclic operation. Typically, for each train, gas flows through two vessels in series for twelve hours while the third is being regenerated with steam. During this period any COS leakage from the first drum is retained in the second. At the end of the period the first drum in line is taken off stream for regeneration, the second drum is moved into first position, and the freshly regenerated drum is placed into second position. This cycle is repeated every twelve hours.

The purified gas, containing about 1.0 mole percent CO_2 and an average of about 0.004 grains total sulfur per 100 SCF, proceeds to Section 500.

f. Section 500 - Methanation and Compression

Section 500 employs three parallel synthesis units and two parallel product gas compressors. In the Kellogg transport catalytic synthesis reactor, methane is formed from hydrogen and carbon monoxide according to the following reaction:

$$CO + 3\ H_2 \rightleftharpoons CH_4 + H_2O + 92{,}900\ \text{Btu} \qquad \text{(I-3)}$$

The very large heat of reaction is absorbed by generating 500 psig, saturated steam.

Effluent gas from the reactor, at 620°F is cooled, scrubbed with water to remove entrained catalyst fines, further cooled to 100°F, and then separated from condensed water.

The product gas is compressed to 1005 psia, cooled again to 100°F, and relieved of entrained condensate. Finally, the 1000 psia, 100°F product gas with a heating value of 916 Btu per SCF is delivered to the gas mains at the rate of 90 million standard cubic feet per day.

g. Section 1100 - Offsite Facilities

Section 1100 includes facilities for:

(1) generating steam for plant start-up
(2) generating electric power
(3) supplying cooling water, process water, and boiler feed water
(4) providing miscellaneous services necessary to make the plant completely self-sufficient.

Steam for start-up is generated at 500 psig, 550°F in a pulverized-lignite boiler capable of generating 150,000 pounds per hour. During normal operation the processing sections of the plant are capable of internally fulfilling their steam requirements, and thus the offsite boiler is not in operation.

Electric power is produced at 13,800 volts by turbogenerators using condensing steam turbine drives. An electric substation is provided to reduce the voltage to 4160, 440, and 110.

River water, destined for process water makeup and boiler feed water makeup, flows first to a feedwater treating system. Water enters precipitator-coagulators, where chemicals are injected by pumps to precipitate dissolved solids. A portion of this treated water is ready for use as process water throughout the plant, while the remainder is pumped through filters, cation exchangers, and anion exchangers to a condensate surge tank. Here the makeup water is mixed with the condensate returned from plant heaters and surface condensers. This entire stream is heated and deaerated and is finally ready for use throughout the plant as boiler feed water.

Further discussion of offsite facilities is contained in the Appendix as part of the Utilities Summary.

2. Economics

a. Economic Summary

The cost of producing 90,000,000 SCFD of pipeline gas from lignite according to the process sequence just described is calculated in Tables I-1 through I-4, assuming 90 percent stream efficiency.

Estimated operating labor for the complete plant, including offsite facilities, is 30 men per shift, as shown in Table I-1.

Estimated capital investment for the plant is summarized in Table I-2. Detailed costs for Section 200, Drying and Gasification, are presented in the Appendix. Interest during construction is calculated at 5.3 percent of the sum of total bare cost plus contractor's overhead and profit, assuming a design, engineering, and construction period equal to 24 months and the cost of money to be 6 percent per annum. The reason for using 5.3 percent is shown in Figure I-1, taken from the Nuclear Power Plant Cost Evaluation Handbook (1-A). Working capital includes 30 days' lignite inventory, 30 days' accounts receivable, and in-plant inventories of sponge iron and methanation catalyst. Shift catalyst and activated carbon are included in fixed investment because they have very long lifetimes. Total capital investment is about 45 million dollars.

Estimated annual operating costs are tabulated in Table I-3. Nuclear heat is charged at 50¢/MM Btu and lignite at $2 per ton. Makeup of Raney nickel methane synthesis catalyst constitutes more than 85 percent of the charge for catalysts and chemicals; that is, about 3¢/MSCF of product gas. Maintenance is charged at the rate of 4 percent of bare cost per year, and comprises approximately 70 percent labor and 30 percent material.

Estimated gas production cost, assuming 20-year straight-line depreciation and before any return whatever on invested capital, is about 56¢/MSCF. Of this total, about 28 percent is contributed by lignite, 21 percent by nuclear heat, and 20 percent by fixed costs.

The cost of operating and construction labor, $2.80 and $3.40 per hour, respectively, used in computing the preceding figures is low by current standards. It should probably be about 15 percent higher. These figures have been used here so that the results may be easily compared with those of previous reports on the manufacture of pipeline gas from coal (11, 12, 13, 14, 15). Use of the current figures would increase gas production cost by about 2¢/MSCF.

The effect on gas cost of adding gross return on investment (interest on bonded indebtedness, plus Federal income taxes, plus net return to investors of equity capital) is shown in Table I-4 and Figure I-2. If a gross return of 20 percent is desired, for example, about 30¢/MSCF must be added to the gas production cost. The effect of variations in the cost of lignite or nuclear heat is also shown. An increase of 25 percent in the cost of lignite or nuclear heat above the values used in Table I-3 results in an increase in gas selling price of only about 3 - 4c//MSCF.

The effect of stream efficiency on gas selling price, assuming a 20 percent gross return on investment, is shown in Figure I-3.

b. Temperature Level of Nuclear Heat

The preceding figures are based on helium being supplied to the process at 2500°F by a nuclear reactor. Reactor investment is not included in Table I-2. Instead, all costs associated with the nuclear side of the plant must be covered by the cost of nuclear heat shown in Tables I-3 and I-4, including ordinary operating costs, return on investment, and the special costs associated with nuclear operation.

If helium, the reactor coolant, is supplied to the process at a temperature below 2500°F, the cost of the nuclear reactor can reasonably be expected to decline. On the other hand, cost of the gasification plant will increase because of the lower temperature driving force, larger heat transfer surface requirement, and larger helium circulator. Figure I-4 shows the cost of nuclear heat which, if helium is supplied at some temperature other than 2500°F, will yield the same gas selling price as calculated in Table I-4. In the absence of specific information concerning the cost of the nuclear reactor as a function of helium temperature, it is not possible to choose the optimum temperature for lignite gasification.

c. Effect of Plant Capacity on Product Gas Cost

Forgetting the nuclear reactor for the moment, plant capacity might be reduced by one-third to one-half without increasing product cost significantly, because the plant consists largely of a number of parallel units. For the same reason, an increase in plant size will not decrease costs by much. The choice of economic plant size will probably depend on the nuclear reactor, which must be built in large capacity to achieve low unit costs, but this is beyond the scope of the present study. Nuclear reactor output for the plant studied here -- 90,000,000 SCFD of pipeline gas is 255 thermal megawatts.

d. Effect of Percentage Lignite Gasified on Product Gas Cost

When the decision was made to design for 80 percent lignite gasification, it was expected that the ungasified residue could be burned to generate steam. It turns out however, that all steam needed by the plant can be generated internally from waste heat, and the ungasified residue must be discarded. Subsequent studies should probably employ about 90 percent gasification, thus reducing product gas cost by about 2¢/MSCF.

Table I-1

Estimated Operating Labor
Pipeline Gas from Lignite

Basis: 90,000,000 SCFD of Pipeline Gas
90% Stream Efficiency

Section	Title	Men per Shift
000	Lignite Storage and Reclamation	1
100	Crushing	2
200	Drying and Gasification	4
300	Shift Conversion	1
400	Gas Purification	7
500	Methanation and Compression	8
1100	Offsite Facilities:	
	Power Plant	3
	Cooling Water Pumps	1
	Makeup Water Pumps	1
	Feedwater Treating System	2
	TOTAL OPERATING LABOR	30
	man-hours/day	720

Table I-2

Investment Summary
Pipeline Gas from Lignite

Basis: 90,000,000 SCFD of Pipeline Gas
90% Stream Efficiency

Section	Title	Material and Freight, Dollars	Bare Cost,* Dollars
000	Lignite Storage and Reclamation	1,245,000	2,116,500
100	Crushing	637,000	1,145,000
200	Drying and Gasification	7,321,130	11,764,130
300	Shift Conversion	2,067,000	3,006,000
400	Gas Purification	6,522,500	8,427,500
500	Methanation and Compression	1,858,000	2,800,000
1100	Offsite Facilities	4,890,000	7,000,000
	Total Material and Freight	24,540,630	
	Total Bare Cost		36,259,130
	Contractor's Overhead and Profit		3,810,000
	Interest during Construction @ 5.3%		2,120,000
	TOTAL FIXED INVESTMENT		42,189,130

Working Capital:		
30 days' lignite inventory (214,935 tons @ $2/ton)	429,870	
Accounts receivable (value of 30 days' production @ 90 ¢/MSCF)	2,430,000	
Catalyst inventory	230,000	
Total Working Capital		3,089,870
TOTAL CAPITAL INVESTMENT		45,279,000

* Bare cost includes materials, freight, construction labor, field administration and supervision, insurance during construction, cost of tools, field office expense, and cost of home office engineering and procurement.

Table I-3
Estimated Annual Operating Cost
Pipeline Gas from Lignite

Basis: 90,000,000 SCFD of Pipeline Gas
90% Stream Efficiency

Item	$/Year		¢/MSCF (2)
Direct Costs:			
Nuclear heat (871.2 MMBtu/hr x 7,884 hr/yr x 50¢/MMBtu)	3,430,000		11.6
Lignite to gasifiers (298.5 TPH x 7,884 hr/yr x $2/ton)	4,710,000		15.9
Operating labor (720 man-hrs/day x $2.80/man-hr x 365 days/yr)	736,000		2.5
Supervision @ 15% of operating labor	110,000		0.4
Catalysts and chemicals	983,000		3.3
Maintenance @ 4% of bare cost per year (1)	1,450,000		4.9
Supplies @ 15% of maintenance	218,000		0.7
Total Direct Costs		11,637,000	
Indirect Costs:			
Payroll overhead @ 20% of operating labor, maintenance labor, and supervision	372,000		1.3
General plant overhead @ 50% of operating labor, supervision, maintenance, and supplies	1,257,000		4.3
Total Indirect Costs		1,629,000	
Fixed Costs:			
Depreciation @ 5% of total fixed investment	2,110,000		7.1
Property taxes and insurance @ 3% of total fixed investment	1,266,000		4.3
Total Fixed Costs		3,376,000	
TOTAL ANNUAL OPERATING COST		16,642,000	
ESTIMATED GAS PRODUCTION COST			56.3

---

(1) Maintenance is 70% labor, 30% material

(2) Cents per thousand standard cubic feet of gas

Table I-4

Pipeline Gas from Lignite

Effect of Lignite Cost, Cost of Nuclear Heat, and Return on Investment on the Selling Price of Gas

A. Gross return on investment = 12%

Cost of nuclear heat, ¢/MMBtu	Gas selling price, ¢/MSCF Lignite at $1.50/ton	 Lignite at $2.00/ton	 Lignite at $2.50/ton
35	67	71	75
50	71	75	79
65	74	78	82
80	78	82	86

B. Gross return on investment = 20%

Cost of nuclear heat, ¢/MMBtu	Gas selling price, ¢/MSCF Lignite at $1.50/ton	 Lignite at $2.00/ton	 Lignite at $2.50/ton
35	79	83	87
50	83	87	91
65	86	90	94
80	90	94	98

C. Gross return on investment = 30%

Cost of nuclear heat, ¢/MMBtu	Gas selling price, ¢/MSCF Lignite at $1.50/ton	 Lignite at $2.00/ton	 Lignite at $2.50/ton
35	95	99	103
50	98	102	106
65	102	106	110
80	105	109	113

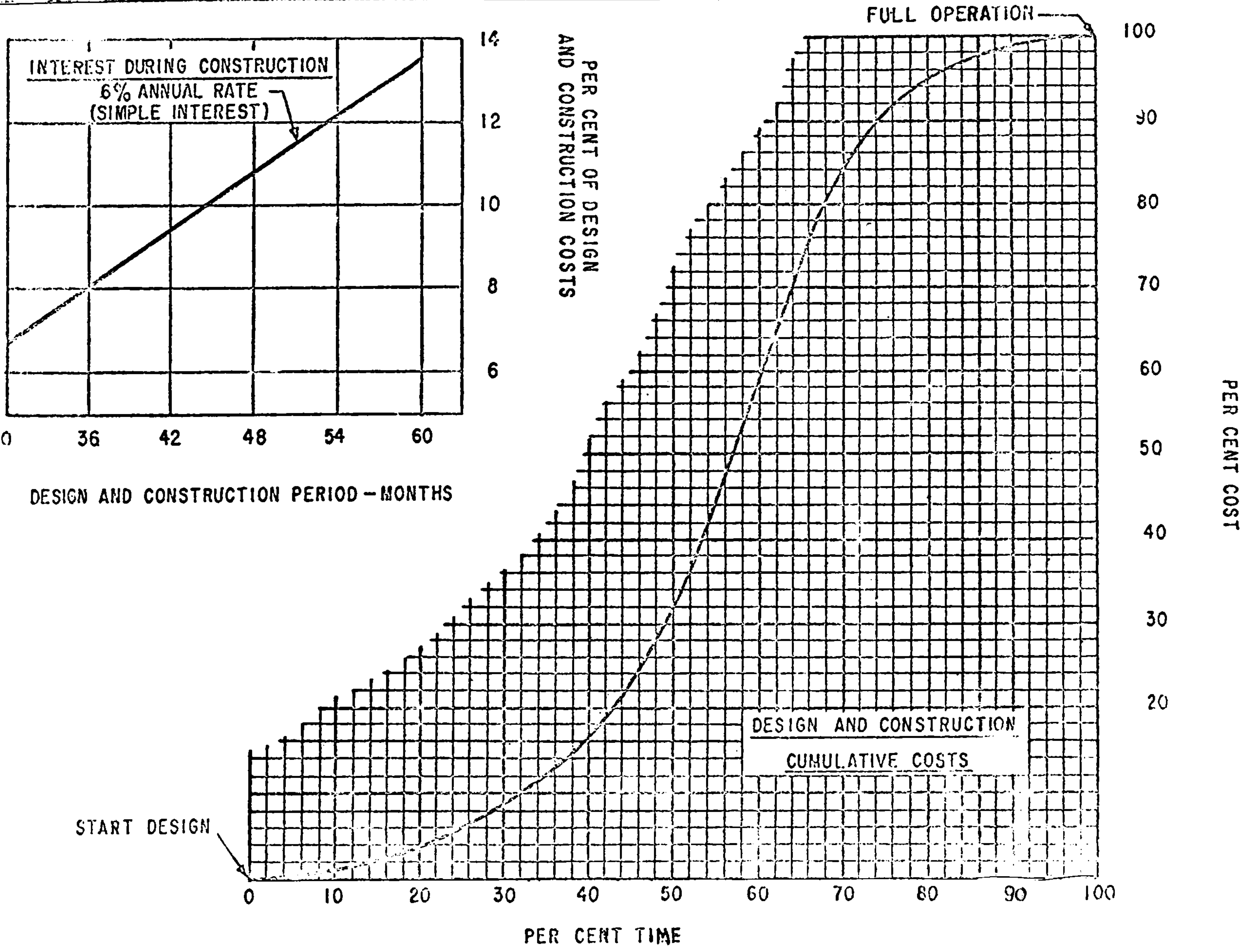

INTEREST DURING CONSTRUCTION
6% ANNUAL RATE
(SIMPLE INTEREST)
PER CENT OF DESIGN AND CONSTRUCTION COSTS
14
12
10
8
6
30
36
42
48
54
60
DESIGN AND CONSTRUCTION PERIOD - MONTHS
FULL OPERATION
START DESIGN
DESIGN AND CONSTRUCTION
CUMULATIVE COSTS
PER CENT COST
100
90
80
70
60
50
40
30
20
0
10
20
30
40
50
60
70
80
90
100
PER CENT TIME

FIGURE I-1

Figure 1-2

Effect of Lignite Cost, Cost of Nuclear Heat, and Return on Investment on the Selling Price of Gas

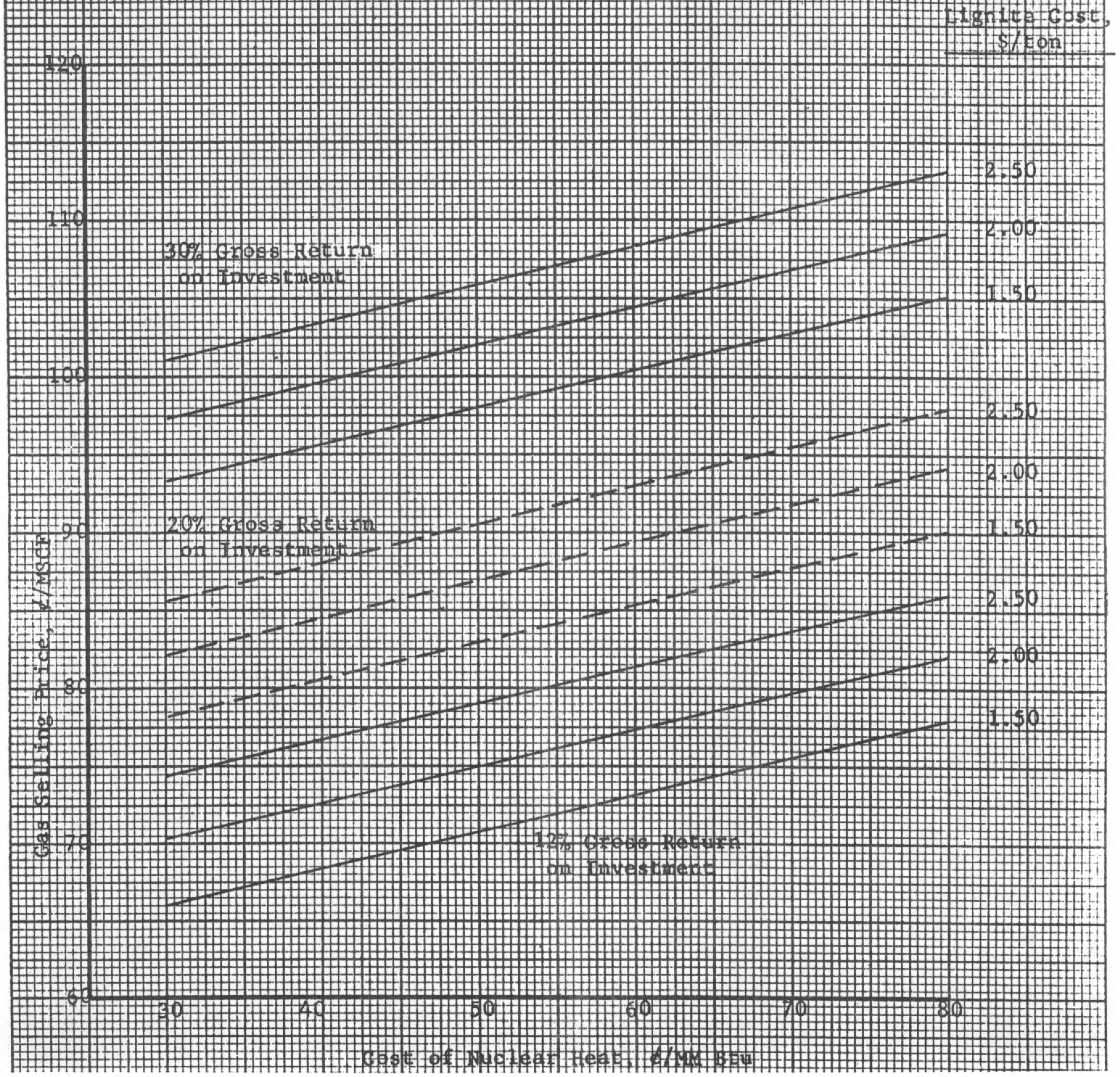


Figure I-3

Effect of Stream Efficiency on Pipeline Gas Selling Price

Basis: 20% Gross Return on Investment
Lignite at $2/ton
Nuclear heat at 50¢/MM Btu

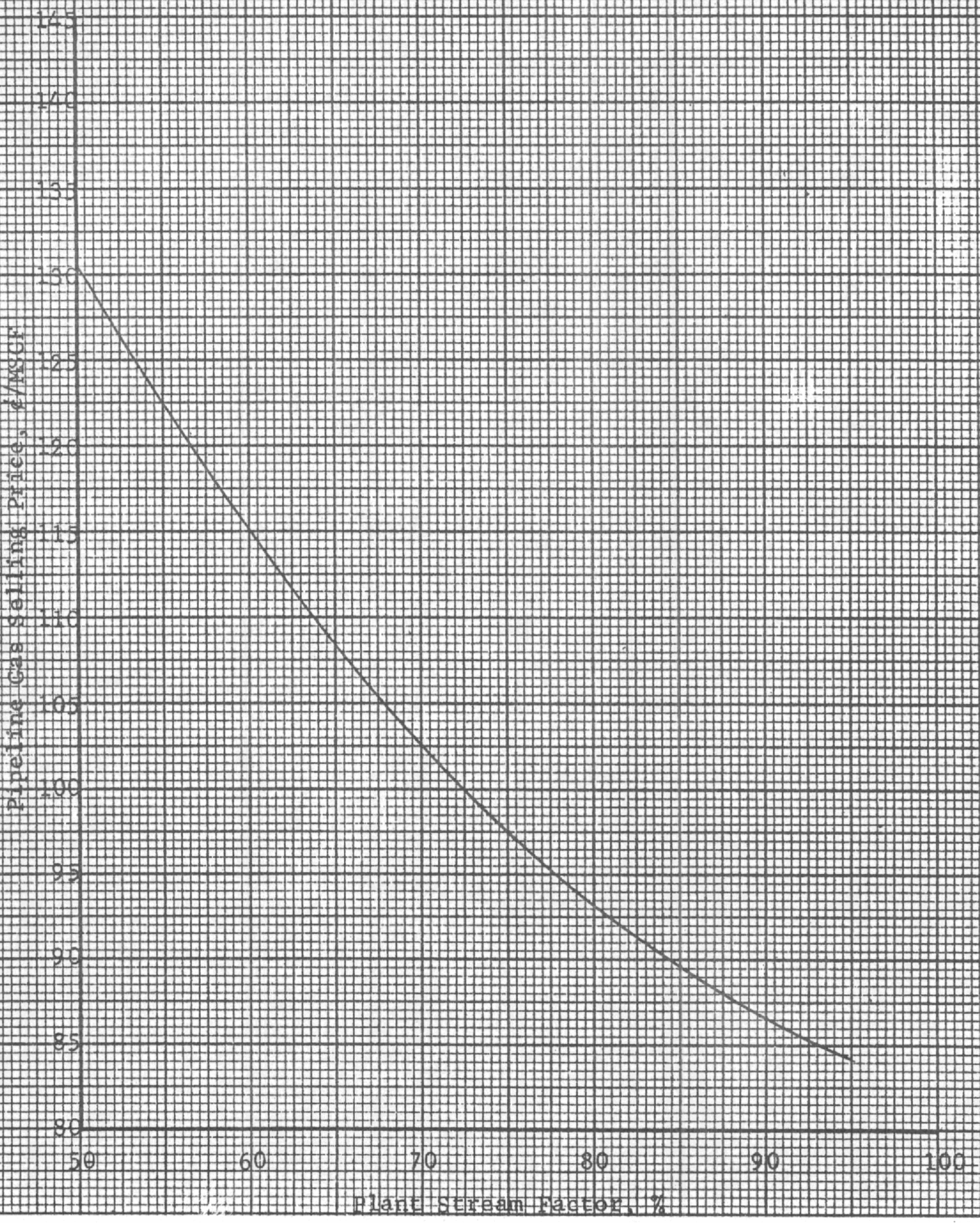
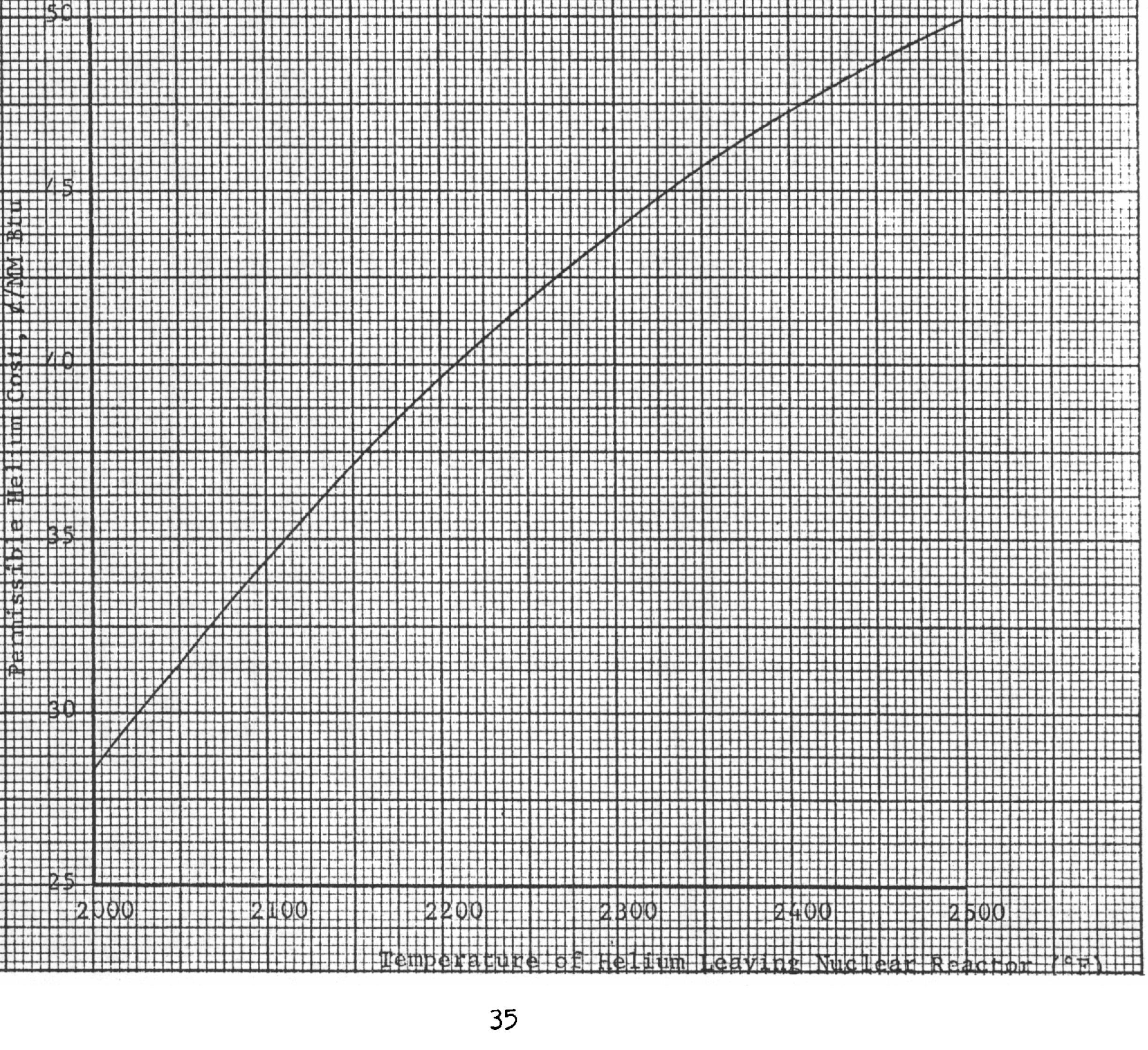


Figure I-4

Permissible Helium Cost
to Yield Same Pipeline Gas Selling Price
as 2500°F Helium at 50¢/MM Btu

Basis: 20% Gross Return on Investment



3. Design Considerations

This section of the report has several purposes. The process design of Section 200 will be discussed, showing in some detail how the design was made and citing sources of information. The designs and stages of development of the other processing sections will be discussed briefly and the reader will be referred to other reports for more detailed information. Important assumptions will be stated, and those areas where additonal experimental work is necessary to firm up the design will be pointed out.

a. Fluidized-Bed Drying

Since its caking properties are not severe, lignite requires no pretreatment to render it nonagglomerating. However, prior to heating to reaction temperature and gasifying, bound water contained in the lignite is evaporated in a fluidized-bed dryer. The vaporized water is subsequently used as a portion of the gasifying medium in the fluidized-bed gasifier.

Fluidized-bed drying of coal, with the heat of vaporization usually supplied by a hot gas fluidizing medium, is a common operation (25). The rate of vaporization is practically instantaneous and thus the beds are seldom greater than 12 inches deep. In the present design, the heat of vaporization is supplied through heat-exchange tubes which are immersed in the fluidized bed, and it was found that the dryer bed size was determined by the space necessary for the heat-transfer surface rather than by the rate of vaporization.

Gas-side and bed-side heat-transfer film coefficients were estimated by the Dittus-Boelter equation and Wender and Cooper (29) correlation, respectively. The overall coefficient calculated from these film coefficients is approximately 50 Btu per sq. ft-(hr)-(°F), and the heat-transfer surface area was estimated accordingly. In the final design, 20 percent excess area was specified as a safety factor.

Approximately 12 percent of the total cross-sectional area of the dryer is taken up by the heat-transfer tubes. These are not expected to cause any operating difficulties. Volk, Johnson, and Statler (28) have demonstrated experimentally that tubing vertically immersed in a fluidized bed of solids, taking up as much as 22 percent of the total cross-sectional area, did not adversely affect the bed, and in most instances the quality of fluidization was actually improved.

Pilot-plant demonstration of the operability of the fluidized bed dryer is not necessary to proper design of a commercial unit.

b. Fluidized Bed Gasification

(1) Reaction Kinetics

The major portion of kinetic information necessary for the design of the fluidized-bed gasifier was taken from the data of Konchesky, Stewart, and Sebastian (17). These Bureau of Mines investigators studied the effects of residence time, reaction temperature, steam/carbon ratio, and rank of coal on the percentage conversion of carbon in the coal fed to the reactor, which was a 112-foot-long alloy-steel tube. In this type of reactor, commonly called a "transport" or "entrained" gasifier, the steam-coal mixture flows through the tubing while reacting to form synthesis gas.

In addition, some kinetic information was obtained from the data of Maddox (18), who studied the effects of particle size and pressure in a similar reactor.

One of the coals studied by the investigators mentioned above was a Subbituminous C coal, which is of a rank comparable to the North Dakota lignite under study here. The effects of the major process variables on the steam gasification of this Subbituminous C coal are qualitatively summarized below.

(a) Reactor Residence Time

Carbon conversion was found to be a linear function of reactor residence time between 0.5 and 1.3 seconds at all temperatures and steam concentrations studied. The gas and coal flow through the transport reactor together, hence residence times of gas and solids are equal (assuming no "slippage").

(b) Reaction Temperature

Carbon conversion increased linearly as temperature was increased from 1660° to 1900°F. Over the 1570° to 1600°F range the change in carbon conversion was more abrupt.

(c) Steam/Carbon Ratio

Total carbon conversion also increased as steam concentration increased. This increase was rapid up to 4 pounds of steam per pound of dry coal, but above this ratio carbon conversion increased more slowly.

(d) Particle Size

On an ash-free basis, there was no noticeable effect on carbon conversion of particle size over the 0 to 35 micron range. For much larger particle sizes. however. it is reasonable to expect a definite effect.

(e) Pressure

Over the pressure range from 30 to 200 psig the carbon conversion decreased approximately 8 percent per 100 psig increase in pressure.

From the quantitative data on the effects of the above process variables it was estimated that the residence time of the 0.98 steam/dry lignite mixture in the 1600°F, 450 psia fluidized-bed reactor should be 7 seconds in order to attain 80 percent carbon conversion. The final gasifier design, which allows average gas and solids residence times of 14 seconds and 30 minutes, respectively, is thus quite conservative. It was felt that since many of the kinetic data were extrapolated to varying extents, some conservatism was necessary for safety of design.

(2) Heat Supply

As previously mentioned the 492.3 million Btu per hour endothermic heat of reaction is supplied to the steam-lignite mixture by a stream of hot helium flowing through heat-exchange tubes immersed in the fluidized bed. The bed-side heat transfer film coefficient was estimated by the Wender and Cooper correlation (29), while the helium-side film coefficient was obtained from extrapolation of Bureau of Mines data (2). The overall heat transfer coefficient calculated from these two film coefficients, approximately 80 Btu per sq. ft-(hr)-(°F), was used to estimate gasifier heat transfer surface. About 20 percent excess area was specified in the final design as a safety factor.

(3) High-Temperature Tube Materials

After a search of the literature and discussions with manufacturers' representatives, Hastelloy alloy X and Stainless 310 were chosen as the materials from which the gasifier heat transfer tubes would be fabricated. Hastelloy X and Stainless 310 have excellent strength and oxidation resistance up to temperatures of 2250° and 1800°F, respectively. Consequently, the reactor heat-transfer surface was divided into two sections, with Hastelloy X being used in the section where tube-wall temperature exceeded 1800°F and Stainless 310 being used in the section where tube-wall temperature was below 1800°F.

Refractory metals, such as columbium and tantalum, were also considered as possible materials of construction. However, tubing made of these materials would cost three to five times as much as the alloy tubing chosen and, moreover, these materials would require an oxidation-resistant coating.

The maximum percentage of total gasifier cross-sectional area taken up by heat-transfer tubing is approximately 8.5 percent and it is not expected that this heat-transfer surface will interfere with proper fluidization of the bed.

(4) Effluent Gas Composition

The composition of the gasifier effluent was calculated under the assumption that the following reactions were at thermodynamic equilibrium at reactor conditions:

$$CO + H_2O \rightleftharpoons CO_2 + H_2 \qquad (I\text{-}1)$$

$$CO + 3\ H_2 \rightleftharpoons CH_4 + H_2O \qquad (I\text{-}3)$$

Reaction (I-1), the water-gas shift reaction, is fast at the 1600°F gasification temperature, and thus it is reasonable to assume that it remains essentially in equilibrium in the product gas, which experiences a relatively long reactor residence time. At the high reactor pressure, long contact time, and high conversion of feed steam it is also reasonable to expect the relatively high concentration of methane predicted by steam-methane reaction equilibrium. Consequently, in the absence of any product gas composition data on the steam gasification of lignite in a fluidized bed reactor, the gasifier effluent composition calculated by the procedure outlined above is thought to be quite realistic.

The entire amount of sulfur contained by the lignite feed was assumed to be gasified and appears in the product gas -- 90percent as H_2S and 10 percent as COS. These sulfur compounds must be removed from the gas before synthesis of pipeline gas, thus it is evident that this is a conservative assumption. In addition, it was assumed that 80 percent of the H_2, O_2, and N_2 in the raw lignite would be gasified and, moreover, 70 percent of the N_2 gasified would appear in the product gas as NH_3.

Pilot-plant study and demonstration of reaction kinetics, product gas composition, and materials of construction are necessary before a commercial fluidized-bed gasifier may be designed with confidence.

c. Fluor Solvent CO_2 Removal Process

The Fluor Solvent CO_2 Removal Process (16) is a relatively new solution to the problem of purifying high-pressure gas streams rich in CO_2. Although Fluor apparently developed this process specifically for the purification of CO_2-rich natural gases, it is claimed to be equally applicable to high-pressure synthesis gases. In addition to CO_2, it is said to be effective for removal of substantial quantities of H_2S, mercaptans, water vapor, and some COS.

Over a period of several years the Fluor Corporation has acquired considerable bench-scale data, and in cooperation with El Paso Natural Gas Company, has field-tested a 1.5 MM SCFD pilot unit operating on CO_2-rich natural gas. In late 1960 Fluor built a 220 million SCFD commercial installation for El Paso Natural Gas, and there are now two additional plants in commercial operation.

The treated gas specifications for the Fluor unit are as follows:

CO_2	1.0 volume percent
H_2S	20 grains per 100 SCF
COS	30 grains per 100 SCF

In conjunction with a previous study (14), Fluor designed and estimated for M. W. Kellogg a similar unit to meet the above specifications.

Of the impurities which might be present in the synthesis gas, only NH_3 might be detrimental to the Fluor solvent. Consequently, CO_2-removal is preceded by a simple water prewash to remove NH_3.

d. Sulfur Removal

As previously mentioned, the sulfur removal scheme consists of sponge iron for H_2S, followed by activated carbon for COS.

The design of the sponge iron system is based on various commercial data compiled by Kellogg's Process Engineering Department and on recommendations of sponge iron suppliers. The system is described under heading I-A-1-e of this report.

The design of the activated carbon system is based on adsorption data from such sources as Pittsburgh Coke and Chemical Company and the U. S. Bureau of Mines (26). Lacking sufficient kinetic data, a three-tower system, operating on a twelve-hour time cycle, was used for maximum safety of design. This system is also described under heading I-A-1-e of this report.

It is anticipated that the above sulfur removal sequence will reduce total sulfur concentration to about 0.004 grains per 100 SCF of gas. At this concentration the loss of methane synthesis catalyst due to sulfur contamination is insignificant.

e. Methane Synthesis

Research into the synthesis of methane using nickel catalysts has been performed recently in the United States by the Bureau of Mines and the Institute of Gas Technology, and in England by the Gas Research Board. This research is discussed in detail in a previous Kellogg study for the Bureau of Mines (12).

The methane synthesis unit in Section 500 is quite different in design from the equipment used by the Bureau of Mines, I.G.T., or the Gas Research Board. It is similar in some respects to the Kellogg-designed synthesis unit operating with iron oxide catalyst at the Sasol plant in South Africa to make liquid products from coal. No data being available on the synthesis of methane in a unit of this type, several assumptions were made in order to design the synthesis reactor. These assumptions are also discussed in the earlier report (12).

Experimental work on the methane synthesis unit is definitely required before a commercial plant can be built.

4. Optimization of Gasification Process Variables

In this section the reasons for the final choice of gasifier operating conditions are discussed. The process variables considered are steam/carbon ratio, reaction temperature, helium temperature, pressure, and percentage carbon utilization. Effect on product gas cost was the basis for selection of optimum operating conditions. Other criteria, however, such as operability, were considered when necessary for proper selection. The optimization of each of the above variables is discussed in some detail in the following pages.

a. Steam/Carbon Ratio

As previously mentioned, the gasifier was sized on the basis of gas contact time necessary for desired conversion of carbon fed to the reactor. The kinetic data showed that as steam/carbon feed ratio increased, a decreased gas residence time was necessary for the desired conversion. It was found, however, that this decreased residence time was almost completely offset by the increased total volume throughput, and gasifier bed size was only slightly affected by increased steam/carbon feed ratio. Thus, gasifier cost remained almost constant while the steam/carbon ratio was varied over a wide range.

Large amounts of energy are required to generate and superheat steam to the reaction temperature, thus it is evident that a low steam feed concentration is desirable. Consequently, the final design calls for a stoichiometric steam/carbon feed ratio. Since 80 percent of the feed carbon is gasified, this corresponds to a steam feed/carbon converted mole ratio of 1.25/1.

b. Reaction Temperature

Kinetic data indicate that the lignite gasification rate increases substantially when reaction temperature is raised above 1600°F. In designing the fluidized bed gasifier, however, it was not possible to take advantage of this increased reaction rate because gasifier bed volumes at reaction temperatures above 1600°F were determined by the quantity of heat transfer surface which must be immersed in the fluidized bed, and not by the residence time necessary for the desired carbon conversion. Higher reaction temperatures decreased the average helium to fluidized bed heat transfer driving forces, hence the amount of gasifier heat transfer surface increased. Thus, temperatures above 1600°F resulted in higher gasification cost. Another disadvantage of higher temperature is the fact that it inhibits methane formation. The final product is pipeline gas, hence it is obvious that any methane formed in the gasifier will reduce the synthesis requirement of Section 500 and thus lead to a lower total plant cost.

In view of the above discussion, it may appear that a reaction temperature below 1600°F might allow further economic improvement. Gasifier designs much below this temperature, however, would involve extrapolation of the kinetic data, which seem to indicate that the steam-carbon reaction is extremely slow at temperatures much less than 1600°F. For this reason, 1600°F was chosen as the minimum realistic reaction temperature.

c. Helium Temperature

Inlet helium temperatures of 2000°, 2250° and 2500°F were studied. The nuclear reactor was not considered, hence this variable was optimized from a process viewpoint, with the contract limitation that 2500°F was to be maximum helium temperature.

Mainly because of increased helium circulation rate, it was found that the equipment cost increased appreciably as the inlet helium temperature was lowered. As was shown in Section I-A-2, the 2250°F and 2000°F helium streams must be available at approximately 8.5¢ and 22¢ per million Btu less than the 2500°F helium stream, respectively, in order to compensate for the increased equipment costs. On the basis of these economics, 2500°F was chosen as the optimum helium temperature for this study.

d. Pressure

An operating pressure of 450 psia was chosen for the fluidized-bed gasifier. This choice was made in view of the fact that 400 psia is approximately the optimum pressure for methane synthesis. If the methanation step were carried out at a higher pressure, the optimum gasification pressure would probably increase correspondingly. The design of a higher pressure methane synthesis unit, however, would involve extrapolation of available data and deviation from existing technology. Consequently, prior to optimization of gasification pressure, the methanation pressure was set at approximately 400 psia. Gasification pressures below this level would necessitate subsequent costly compression of synthesis gas. Higher gasification pressures, with recovery of power by an expander prior to methane synthesis, were also considered but proved to be uneconomical. Thus, it was concluded that the 450 psia level was the desirable gasification pressure.

In order to minimize creep in the gasifier heat-transfer tubing, the helium pressure was also set at 450 psia, which allowed the use of short-term strength data as the basis for design. For maximum safety the tube-wall thicknesses were specified to withstand the entire 450 psia pressure differential. At normal operating conditions, of course, the pressure differential across the heat-transfer surface is only a few pounds per square inch.

e. Carbon Utilization

The fluidized-bed gasifier has been designed to gasify 80 percent of the carbon contained in the lignite feed. With further extrapolation of the kinetic data, the reactor could have been designed for 90 percent conversion. Such a design, however, would have resulted in a cost saving of less than 2¢ per MSCF of pipeline gas.

5. Evaluation of Alternate Processing Schemes

In this section the alternate processing schemes considered in this study will be discussed, and reasons will be given for selection of the scheme described earlier as the optimum process sequence for production of pipeline gas from lignite.

a. Introduction of Lignite to Pressure Processing

The following two methods were considered for introducing the lignite to the 450 psia gasification process:

(1) Lignite-water slurry pump feeder.
(2) Lock hopper feeder.

Although method (1) would appear to be more economical, there is some question of the operability of a slurry pump in this particular application. The feed slurry would consist of approximately 75 percent raw lignite and only 25 percent water by weight. Continuous pumping of such a slurry has not been demonstrated on a commercial scale. Furthermore, the slurry would be heated almost to boiling before introduction to a fluid-bed dryer and there would be a definite danger of plugging the heat transfer tubes.

For the above reasons method (1) was eliminated from consideration and method (2) was chosen to introduce the lignite to the 450 psia process. With proper instrumentation it is anticipated that the lock hopper feeding system will introduce no operability difficulties.

b. Drying of Lignite

Consideration was given to drying the lignite both before and after hammermilling to the desired size consist. Since the hammermill operates at atmospheric pressure, the water evaporated from the raw lignite would not be available for use as feed steam to the high-pressure gasifier if the drying were carried out before size reduction. On the other hand, if the drying step is performed after hammermilling, the water contained by the raw lignite may be recovered as high-pressure steam and then may be utilized as a portion of the steam feed to the gasifier. Since the total water content of the lignite represents over one-half of the gasification steam requirement, recovery and utilization of this bound water is highly desirable. Milling, followed by drying at gasification pressure, was chosen as the more economical sequence. Raw lignite has been milled on a commercial scale, and thus this sequence should present no operating difficulties.

Having made the above determination, consideration was given to the following two methods of drying the fine lignite:

(1) Flash drying in heat-exchange tubes.
(2) Fluidized-bed drying.

Method (1) has been tried at the Bureau of Mines installation at Morgantown, West Virginia. Difficulty has been experienced, however, in feeding slurries of lignite and water through a preheater in which the water, both that contained by the lignite and that making up the slurry, is flashed to steam (1). This method has led to heavy deposits, and in some cases to plugging of the preheater.

Because of the above difficulty, method (1) was eliminated from consideration and method (2) was chosen to dry the lignite. Fluidized-bed drying of coal is a common operation, and it is expected that this method will present no operating difficulties.

c. Gasification of Lignite

(1) Method of Heat Supply

Helium from the nuclear reactor must not come in direct contact with any process streams, but must be contained in a closed loop, because any entrained particulate matter returned to the reactor would become radioactive. With this limitation, the following general methods of transferring heat from the helium to the gasification medium were considered:

(a) Intermediate heat carrier. Heat would be transferred in a heat exchanger from the helium to a liquid or gaseous heat carrier, and the carrier would, in turn, transfer this heat to the reactants by direct contact.

(b) Heat transfer surface. Heat would be transferred, using heat-exchange tubes, directly from the helium to the reactants.

No advantage could be found for method (a) unless the intermediate heat carrier would also catalyze the reaction. In order to heat the carrier to a reasonable temperature (at least 2200°F), large quantities of refractory-metal heat-transfer surface would be required. For this reason method (a) appeared highly uneconomical, and method (b) was chosen to transfer heat from the helium to the reactants.

(2) Gas-solids Contacting

Entrained, fixed-bed, and fluidized-bed gasifiers were considered in an effort to determine the optimum method of gas-solids contacting.

(a) Entrained gasifier

In an entrained gasifier the steam-coal mixture flows through the gasifier tubes while reacting to form synthesis gas. The endothermic heat of reaction is supplied through the tube surface by the stream of hot helium. In order to achieve the desired carbon conversion, the design of such a gasifier must provide for adequate residence time for the reactants. In addition, the design must provide for enough tubing surface to transfer the endothermic heat of reaction to the reacting mixture. To satisfy these design requirements, it was found that relatively large quantities of tubing were necessary and thus the entrained gasifier appeared to be much less economical than the type eventually chosen.

(b) Fixed-bed gasifier

A fixed-bed reactor, in which the endothermic heat of reaction is supplied by steam or recycled reactor effluent gas superheated against helium, also received consideration. It was found, however, that an excessive amount of heat-transfer surface was necessary. Moreover, because the steam or recycle gas must be heated to at least 2200°F, the heat-transfer surface must be fabricated of refractory metal with an oxidation-resistant coating. The cost of such a gasification scheme is astronomical in comparison with the cost of the fluidized-bed gasifier, which was eventually selected in our final design.

(c) Fluidized-bed gasifier

Primarily because of the higher overall heat transfer coefficient obtained, the fluidized-bed gasifier requires a relatively small quantity of heat-transfer surface. Heat transfer surface constitutes the major portion of gasifier cost, hence it is evident that the fluidized-bed gasifier is most economical. At the gasifier operating conditions chosen for final design, the heat-transfer surface required by both the entrained and fixed-bed schemes would cost substantially more than the heat-transfer surface required by the fluidized-bed scheme, as shown in the following table.

Scheme	Total Heat-Transfer Surface, sq. ft	Total Cost of Heat-Transfer Surface, dollars
Fluidized Bed	109,850	2,500,000
Entrained	209,000	4,900,000
Fixed Bed	460,000	18,000,000 - 25,000,000

On the basis of these relative economics, the fluidized-bed gasifier was selected for incorporation into the final flowsheet.

d. Gas Purification

Although alternate gas purification schemes were not specifically studied for this report, alternate schemes for the purification of a synthesis gas of similar composition were studied in another investigation (14) Various pressure levels, desulfurization methods, and CO_2-removal schemes were reviewed and the following conclusions were drawn:

(1) Pressure level. Purification at gasification pressure, followed by methanation and compression is most economical

(2) CO_2 removal. The most economical processing sequence was found to be synthesis gas shift to a H_2/CO ratio

of approximately 3.0/1, reduction of CO_2 concentration to 1.0 mole percent by the "Fluor Solvent CO_2 Removal Process", desulfurization, and methanation.

(3) Desulfurization. Sponge iron for H_2S removal, followed by activated carbon for COS removal, is preferable at the particular processing conditions encountered in this study; i.e., ambient gas temperature and low H_2S and COS concentrations following CO_2 removal.

e. Methane Synthesis

Design of the methane synthesis unit was also based on a previous study (14) in which the Kellogg transport reactor was found to be economically superior to both a fluidized-bed reactor and the Bureau of Mines hot-gas-recycle system.

Appendix

Pipeline Gas from Lignite

Material Balances

Section 200

Input:	Lb/Hr
Lignite from Section 100 to 201-L	597,042
Steam from Section 500 to 201-C	171,948
Waste gas from Section 400 to 202-J	43,000
Cooling water to 203-F	500,000
	1,311,990

Output:	Lb/Hr
Raw synthesis gas from 204-G to Section 300	662,271
Residue slurry from 203-F to settling pond	606,719
Waste gas from 202-F to atmosphere	43,000
	1,311,990

Section 300

Input:	Lb/Hr
Raw synthesis gas from Section 300	662,271
Boiler feed water to shift converter	49,503
Process water to water scrubber	120,000
	831,774

Output:	Lb/Hr
Shifted synthesis gas to Section 400	609,519
Condensate from knockout drums	5,616
Waste water from scrubbers	216,639
	831,774

Section 400

Input:	Lb/Hr
Shifted synthesis gas from Section 300	609,519
Regeneration steam to activated carbon drums	42,000
	651,519

Output:	Lb/Hr
Purified synthesis gas to Section 500	256,197
Waste gas from flash drums	352,902
H_2S from iron oxide drums	294
COS and steam from activated carbon drums	42,126
	651,519

Section 500

Input:	Lb/Hr
Purified synthesis gas from Section 400	256,197

Output:	Lb/Hr
Pipeline gas product	151,478
Condensate from scrubber and knockout drums	104,719
	256,197

Overall Energy Balance

Datum Temperature = 60°F

Input:	MM Btu/Hr	Percent of Total
Lignite feed heating value	4,300.0	83.2
Heat transferred from helium	871.2	16.8
	5,171.2	100.0

Output:	MM Btu/Hr	Percent of Total
Product gas heating value	3,440.0	66.6
Gasifier residue heating value	923.0	17.8
Losses to cooling water	456.4	8.8
Latent heat of excess steam generated	85.0	1.6
Latent heat of activated carbon regeneration steam	55.0	1.1
Heating value of waste gas from Section 400	54.3	1.1
Electric power consumption	26.8	0.5
Sensible heat of water purge streams	47.0	0.9
Sensible heat of gas streams	3.0	0.1
Convection losses, miscellaneous, etc.	80.7	1.5
	5,171.2	100.0

Utilities Summary

Steam

I. 500 psig, 550°F

Steam for plant start-up is generated in Section 1100 at 500 psig, 550°F in a pulverized-lignite boiler capable of generating 150,000 pounds per hour. However, during normal operation the processing sections of the plant are capable of internally fulfilling all steam requirements, and thus the offsite boiler is not on stream.

II. 500 psig, saturated

A. Generation

Section	Normal Generation, Lb/Hr
500	525,000

B. Consumption

Section	Normal Consumption, Lb/Hr
200	286,648
400	6,200
1100	162,000
Available	70,152
	525,000

C. Summary for Section 200

Item	Normal Consumption, Lb/Hr
201-C	171,948
201-J	90,000
202-J	24,700
	286,648

III. 40 psia, saturated

A. Generation

Section	Normal Generation, Lb/Hr
300	60,000

B. Consumption

Section	Normal Consumption, Lb/Hr
400	42,000
1100	18,000
	60,000

Power

A. Generation

Electric power is produced at 13,800 volts in Section 1100 by turbogenerators using condensing steam turbine drives. The steam is obtained from Section 500. An electric substation reduces the voltage to 4160, 440 and 110. Normal production is 9070 kilowatts, but the turbogenerators are capable of producting 10,000 kilowatts.

Section	Normal Generation	
	HP	Kw
1100	12,157	9,070

B. Consumption

Section	Normal Consumption	
	HP	Kw
000	700	522
100	1,040	775
200 - condensate pumps	50	37
300	-	-
400	8,450	6,300
500	27	20
1100	1,890	1,416
	12,157	9,070

Cooling Water

A. Generation

It is anticipated that the plant will be located near a river which may prove inadequate as a complete heat sink. Accordingly, the plant cooling system is based on tower cooling of recirculated water, with the necessary makeup water obtained from the river.

Cooling water is available throughout the plant at 40 psig and a maximum temperature of 85°F.

Section	GPM
1100	73,132

B. Consumption

Section	GPM
200	26,532
300	9,000
500	17,600
1100	20,000
	73,132

C. Summary for Section 200

Item	GPM
205-C	232
203-F	1,000
Surface Condensers	25,300
	26,532

Boiler Feed Water

A. Generation

Boiler feed water at temperature levels of 250° and 350°F is produced by the following sequence:

(1) River water is filtered and treated with chemicals to remove dissolved solids.

(2) Further purification is effected by passing the treated water over anion- and cation-exchange resins.

(3) Purified makeup water is mixed with return condensate and heated to about 180°F.

(4) The water is deaerated with low-pressure steam.

(5) Deaerated water is heated to temperature levels of 250° and 350°F in the desired quantities.

Section	Normal Generation, Lb/Hr
1100	652,212

B. Consumption

Section	Normal Consumption, Lb/Hr
300	111,512
500	540,700
	652,212

Process Water

A. Generation

Process water is produced by treating filtered river water to precipitate dissolved solids.

Section	Normal Generation, Lb/Hr
1100	120,000

B. Consumption

Section	Normal Consumption, Lb/Hr
300	120,000

Detailed Cost Estimate -

Section 200

SUMMARY OF CONTRACT OPERATIONS - MATERIAL, LABOR & SUBCONTRACT COSTS

ACCOUNT CLASSIFICATION	A/C'S 110 TO 140 MATERIAL & FREIGHT	A/C 310 & 320 CONSTR FORCE WAGES & FRINGE	IN-PLACE SUBCONTRACTS (ALL)	
A - SITE PREPARATION, FOUNDATIONS AND CONCRETE STRUCTURES	$ 55,000.			
B - FURNACES				
C - EXCHANGERS	1,749,000.			
D - CONVERTERS	1,267,230.			
E - TOWERS				
F - DRUMS AND TANKS	190,300.			
H - STEEL STRUCTURES AND PLATFORMS	160,000.			
J - PUMPS AND COMPRESSORS	1,647,000.			
K - BUILDINGS	70,000.			
L - SPECIAL EQUIPMENT	65,000.			
M - PIPING	1,350,000.			
N - ELECTRICAL	60,000.			
O - INSTRUMENTS	225,000.			
P - INSULATION AND PAINT	100,000.			
U - UTILITY EQUIPMENT				
V - TRANSPORTATION AND CONVEYING EQUIPMENT				
W - CHEMICALS & CATALYST				
Z - FIRE AND SAFETY EQUIPMENT	15,100.			
G - CYCLONES	167,500.			
FREIGHT - UNALLOCATED	200,000.			
EXPORT PACKING - UNALLOCATED				
SUBTOTAL: DIRECT MATERIAL	$7,321,130.			
S - TEMPORARY FACILITIES INCLUDING RIGGING				
T - TOOL HANDLING & TOOL REPAIRS & MAINTENANCE(CHARGEABLE TO JOB)				
Y - MISCELLANEOUS SUPPLIES & UNALLOCABLE LABOR				
SUBTOTAL: INDIRECT MATERIAL (S, T, Y)	180,000.			
TOTALS	$7,501,130.			

SUMMARY - MATERIAL, LABOR AND SUBCONTRACTS

DESCRIPTION: Pipeline Gas from Lignite
Section 200 Gasification - 3 Units

R.P.G. - J.E.R.

ESTIMATE NO. 4106-L JOB NO. ________ DATE Sept. 30, 1962
CLIENT U.S. Atomic Energy Commission
LOCATION North Dakota, U.S.A.
TYPE OF ESTIMATE OR APPRAISAL Budget Cost Estimate

ESTIMATE DETAILS

ITEM	UNIT	TOTAL UNITS	UNIT PRICE	MATERIAL COST	M.H. UNIT	TOTAL MANHOURS
C - 201 Steam Superheater	Sq.Ft.	1,100	20.00	$ 22,000.		
ABC Stainless 302 Tubes						
Insulated Shell with 302 liner						
C - 202 Steam - Coal Heater	Sq.Ft.	4,600	10.00	46,000.		
ABC 1 Cr 1/2 Moly Tubes						
C - 203 Steam - Coal Heater	Sq.Ft.	9,200	20.00	184,000.		
ABC Stainless 302 Tubes						
Insulated Shell with 302 Liner						
C - 204 Steam - Coal Heater	Sq.Ft.	9,700	30.00	291,00 .		
Stainless 310 Tubes						
Insulated Shell with 310 Liner						
Total for One Unit				543,000.		
Total for Three Units				1,629,000.		
C - 205 Surface Condensers for Helium						
Compressors - J201 - 3700 BHP	3	26,000		78,000.		
C - 206 Intercoders for CO_2 Compressors	2	8,000		16,000.		
J-202 - 1500 BHP						
C - 297 Surface Condensers for CO_2 Comp.	2	13,000		26,000.		
J-202 - 1500 BHP						
					TOT MH	
					RATE	$ /HR
TOTALS for Three Units				$ 1,749,000	TOT LAB	$

CLASS C - Exchangers   Sec 200 Gasification

EST. NO. 4006L   JOB NO. AEC   SHEET NO.   DATE Sept.30, 1962

ESTIMATE DETAILS - CLASS D & E VESSELS

SKETCH NO. ______ REV. ______ DATE ______ D.P. 500 TEMP. 650°F C.A. ______ DESIGN SPEC. ______

TYPE AND HEIGHT OF SUPPORT 6' Skirt MANHOLES ______ PLATE SPEC. A204 Or B FBQ (carbon 1/2 Moly)

FABRICATION SHOP FIELD ______ X RAY ______ SR ______

ITEM	UNIT	TOTAL UNITS	UNIT PRICE	MATERIAL COST	M.H. UNIT	TOTAL MANHOURS	
VESSEL SIZE 11' 8" x 45' x 2"	LBS						
Shell and Heads - 1/2 Moly	LBS	185,000					
EXTERNAL CLIPS (NO.)	LBS						
6' Skirt & Base - Carbon Steel	LBS	18,500					
INTERNALS & Misc. Clips							
GRATING (REMOVABLE - NON-REMOVABLE)	LBS						
SCREENS - ALLOY (SQ FT)	LBS						
Manholes & Nozzles		21,000					
PIPING	LBS						
SHROUD	LBS						
SHOP INSTALLED STUDS (NO.)	LBS						
OTHER - SPECIFY	LBS						
Water Jacket - 1/4" Carbon Steel		29,000					
SUB TOTAL		253,500	.30	$76,050.			
SEPARATORS DEMISTER Plenum Chamber, Cyclone	EACH						
PACKING Supports, Dip Leg & Vapor Stops	CU FT						
S.S. 310		5,000	4.00	20,000.			
WELDING ROD FOR FIELD FABRICATED VESSELS	LBS						
REMOVABLE TRAYS PANS OR BAFFLES							
ITEM TYPE DIA. TK. MATERIAL	EACH						
Heat Exchanger Sections							
Upper 2070 - 3/4" OD x .125 x 15' Stainless 310 Tubes	Sq.Ft.	6100	30.00	183,000.			
Lower 1770 - 3/4 OD x .120 x 5' Hastelloy Tubes	Sq.Ft.	1730	52.00	89,960.			
INTERNAL INSULATION INCLUDING FIELD INSTALLED STUDS	SQ FT						
Vessel Lining 7" Kaolite							
3" Purotab	Sq.Ft.	2300	18.00	41,400.			
2 Arched Brick Grids			6000	12,000.			
FREIGHT					TOT MH		
Total for One				422,410.	RATE	$ /HR	
TOTALS for Three				$1,267,230.	TOT LAB	$	

CLASS D - 202 A B & C

Sec 200 Gasification

DE

EST. NO. 4006L JOB NO. AEC SHEET NO. DATE Sept. 30, 1962

ESTIMATE DETAILS - PUMPS, COMPRESSORS AND DRIVERS

ITEM		NO. OF UNITS	TOTAL WEIGHT	UNIT PRICE	MATERIAL COST	M.H. UNIT	TOTAL MANHOURS	REMARKS
					EACH		TOTAL	
J-201 SERVICE Helium Circulator								
DESCRIPTION: ☐ HORIZ, ☐ VERT, ☐ CENTR, ☐ RECIP, ☐ PUMP,								
☐ COMPRESSOR, ☐ JET ARRANGED FOR ☐ MOTOR, ☐ TURBINE,								
☐ STEAM ☐ GAS ENGINE DRIVE								
MFR. ABC	CAP. EA. 56000#/Hr ~~GPM~~ CASE MTL							
SIZE	TEMP 1000 °F PRODUCT S.G.							
TYPE 3 - 33 1/3%	SUCT. 435 PSIA DISCH 460 PSIA							
Capacity	MOTOR HP RPM							
SPARE Units	TURBINE 3700 HP CONDENSING RPM	3		$130^{\underline{00}}$	481,000.		$ 1,443,000.	
J-202 SERVICE CO_2 Compressor								
DESCRIPTION: ☐ HORIZ, ☐ VERT, ☐ CENTR, ☐ RECIP, ☐ PUMP,								
☐ COMPRESSOR, ☐ JET ARRANGED FOR ☐ MOTOR, ☐ TURBINE,								
☐ STEAM ☐ GAS ENGINE DRIVE								2-65%
MFR.	CAP. EA. GPM CASE MTL							
SIZE 2 Stage	TEMP. °F PRODUCT S.G.	2		$65^{\underline{00}}$	97,500.		195,000.	Capacity
TYPE Reciprocating	SUCT. PSIG DISCH. 500# PSIG							Machines
	MOTOR HP RPM							
SPARE	TURBINE 1500 HP CONDENSING RPM							
J-203 SERVICE Condensation Pumps								
DESCRIPTION: ☐ HORIZ, ☐ VERT, ☐ CENTR, ☐ RECIP, ☐ PUMP,								
☐ COMPRESSOR, ☐ JET ARRANGED FOR ☐ MOTOR, ☐ TURBINE,								
☐ STEAM ☐ GAS ENGINE DRIVE								
MFR	CAP. EA. GPM CASE MTL	6			1,500.		9,000.	
SIZE For 5	TEMP °F PRODUCT S.G.							
TYPE	SUCT. PSIG DISCH. PSIG							
Surface	MOTOR HP 1 Spare RPM							
SPARE Condensers	TURBINE HP CONDENSING RPM							
SERVICE								
DESCRIPTION: ☐ HORIZ, ☐ VERT, ☐ CENTR, ☐ RECIP, ☐ PUMP,								
☐ COMPRESSOR, ☐ JET ARRANGED FOR ☐ MOTOR, ☐ TURBINE,								
☐ STEAM ☐ GAS ENGINE DRIVE								
MFR.	CAP. EA. GPM CASE MTL							
SIZE	TEMP °F PRODUCT S.G.							
TYPE	SUCT. PSIG DISCH. PSIG							
	MOTOR HP RPM							
SPARE	TURBINE HP CONDENSING RPM							
							$ 1,647,000.	

CLASS J PUMPS, COMPRESSORS AND DRIVERS Sec 200 Gasification J.

EST NO. 4006L JOB NO. AEC SHEET NO. DATE Sept. 30, 1962

B. Production of Liquid Fuels and Chemicals

1. Description of Process

The plant for production of liquid fuels and chemicals from lignite has been divided according to product and process functions. An overall process material balance flowsheet is presented as Drawing No. CE-1526-A.

These sections are briefly discussed in the following pages. Further descriptive information is presented in Section I-A of this report, as well as in other Kellogg reports cited there.

a. Preparation of Raw Synthesis Gas

(1) Section 000 - Lignite Storage and Reclamation

Section 000 receives and handles raw lignite according to the general scheme described previously for the pipeline-gas-from-lignite plant. The capacity is larger, however, and in addition to the 300 tons of lignite per hour dispatched to Section 100, approximately 105 tons per hour is transported by conveyor belt to Section 3000, where it is used to generate steam.

(2) Section 100 - Crushing

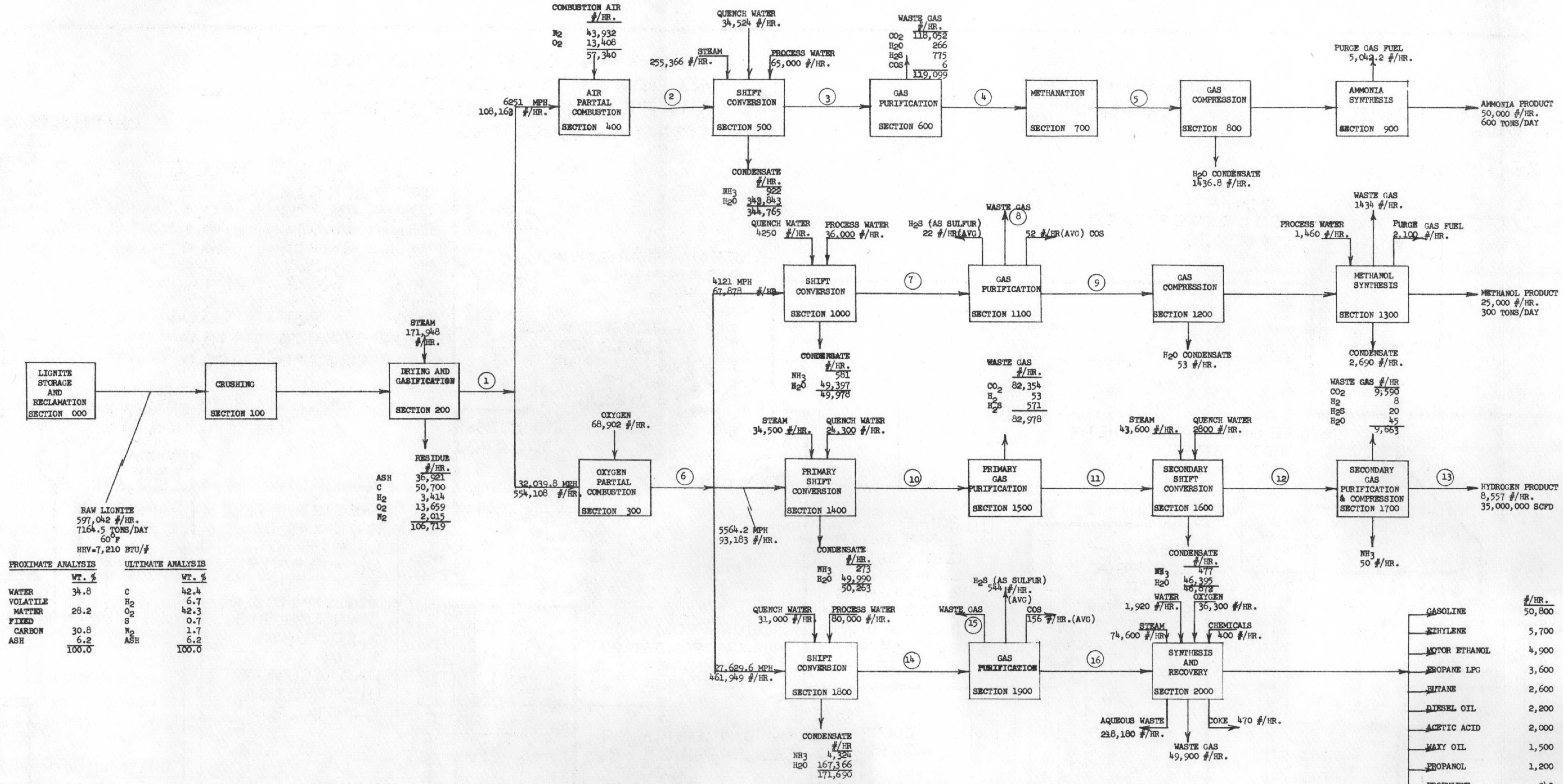
Lignite received from Section 000 at the rate of 300 tons per hour is crushed and milled to 98 percent minus 14 mesh according to the scheme discussed under heading I-A-1-b of this report.

(3) Section 200 - Drying and Gasification

Drawing No. CE-1187-B, the process flowsheet for Section 200, shows equipment sizes, flow rates, and operating conditions. It is evident from this drawing that the processing scheme corresponds to that presented under heading I-A-1-c, with the following exception: The 662,271 pounds per hour gasifier effluent stream at 1600°F is divided and sent to Sections 300 and 400 at rates of 554,108 and 108,163 pounds per hour, respectively. After partial combustion with oxygen, the stream is returned from Section 300 at 1615°F, 429 psia and at a rate of 623,010 pounds per hour. This gas is then used to superheat steam in 201-C, to dry lignite in 201-D, etc., and eventually flows to Sections 1000, 1400, and 1800 for further processing.

(4) Section 300 - Oxygen Partial Combustion

Section 300 consists of three parallel trains of operating equipment. It is the purpose of this section to remove methane from the raw synthesis gas in order that the gas may be subsequently processed to make methanol, hydrogen, and liquid fuels.



PROXIMATE ANALYSIS	WT. %
WATER	34.8
VOLATILE MATTER	28.2
FIXED CARBON	30.8
ASH	6.2
	100.0

ULTIMATE ANALYSIS	WT. %
C	42.4
H_2	6.7
O_2	42.3
S	0.7
N_2	1.7
ASH	6.2
	100.0

	#/HR.	BBL/DAY	MM #/YR. (90% S.F)
GASOLINE	50,800	4,930	---
ETHYLENE	5,700	---	44.9
MOTOR ETHANOL	4,900	430	---
PROPANE LPG	3,600	480	---
BUTANE	2,600	300	---
DIESEL OIL	2,200	180	---
ACETIC ACID	2,000	---	15.8
WAXY OIL	1,500	114	---
PROPANOL	1,200	---	9.4
PROPYLENE	940	---	7.4
ACETONE	860	---	6.8
PROPIONIC ACID	450	---	3.5
N-BUTANOL	380	---	3.0
METHL ETHYL KETONE	290	---	2.3
ACETALDEHYDE	260	---	2.1
ISOBUTANOL	180	---	1.4
HEAVY ALCOHOLS	127	---	1.0
	77,987		

STREAM NO.	①			②			③			④			⑤			⑥			⑦			⑧		
TEMPERATURE	1600°F			700°F			100°F			100°F			100°F			700°F			100°F			---		
PRESSURE	444 PSIA			429 PSIA			400 PSIA			375 PSIA			355 PSIA			413 PSIA			385 PSIA			---		
FLOW RATE	#/HR.	MPH	MOL.%	#/HR.	MPH	MOL.%	#/HR.	MPH	MOL. %	#/HR.	MPH	MOL. %	#/HR.	MPH	MOL. %	#/HR.	MPH	MOL. %	#/HR.	MPH	MOL. %	#/HR.	MPH	MOL. %
CO	277,524	9,912	31.6	64,182	2,292.2	31.9	2,044	73.0	0.8	2,044	73.0	1.1	---	---	---	329,006	11,750.2	38.7	24,528	876.0	25.1	112	4.0	0.6
CO_2	169,776	3,858	12.3	20,398	463.6	6.5	118,052	2,683.0	28.7	---	---	---	---	---	---	104,539	2,375.9	8.6	28,899	656.8	18.8	27,667	628.8	97.2
CH_4	49,521	3,096	9.9	---	---	---	---	---	---	---	---	---	1,167	73.0	1.1	---	---	---	---	---	---	---	---	---
H_2	27,855	13,929	44.3	5,523	2,761.6	38.4	9,962	4,981.0	54.7	9,962	4,981.0	75.0	9,524	4,762.0	74.2	28,311	14,155.9	51.0	3,870	1,935.0	55.6	6	3.0	0.5
N_2	2,367	84.3	0.3	44,363	1,584.4	22.0	44,363	1,584.4	17.0	44,363	1,584.4	23.9	44,363	1,584.4	24.7	1,936	68.9	0.2	240	8.6	0.2	---	---	---
NH_3	6,627	390	1.2	922	61.5	0.9	---	---	---	---	---	---	---	---	---	5,705	328.5	1.2	---	---	---	---	---	---
H_2S	3,714	116.1	0.4	694	20.4	0.3	775	22.8	0.2	---	---	---	---	---	---	3,020	95.7	0.3	403	11.84	0.3	381	11.2	1.7
COS	861	14.4	0.05	150	2.5	0.03	6	0.1	0.01	---	---	---	---	---	---	711	11.9	0.04	52	0.86	0.02	---	---	---
DRY GAS	538,245	31,399.8	100.0	136,232	7,186.2	100.0	175,202	9,344.3	100.0	56,369	6,638.4	100.0	55,054	6,419.4	100.0	473,228	28,787.0	100.0	57,992	3,489.1	100.0	28,166	647.0	100.0
H_2O	124,026	6,891		29,271	1,648.1		426	23.7		160	8.9		1,475	81.9		149,782	8,527.8		158	8.8		93	5.2	
WET GAS	662,271	38,290.8		165,503	8,834.3		175,628	9,368.0		56,529	6,647.3		56,529	6,501.3		623,010	37,315.8		58,150	3,497.9		28,259	652.2	

STREAM NO.	⑨			⑩			⑪			⑫			⑬			⑭			⑮			⑯		
TEMPERATURE	80°F			220°F			220°F			100°F			100°F			100°F			---			80°F		
PRESSURE	365 PSIA			395 PSIA			390 PSIA			375 PSIA			450 PSIA			390 PSIA			---			375 PSIA		
FLOW RATE	#/HR.	MPH	MOL. %	#/HR.	MPH	MOL. %	#/HR.	MPH	MOL. %	#/HR.	MPH	MOL. %	#/HR.	MPH	MOL. %	#/HR.	MPH	MOL. %	#/HR.	MPH	MOL. %	#/HR.	MPH	MOL. %
CO	24,416	872.0	30.7	5,600	200.0	3.4	5,600	200.0	5.1	588	21.0	0.5	588	21.0	0.5	160,275	5,724.1	23.7	560	20.0	0.5	159,715	5,704.1	28.3
CO_2	1,232	28.0	1.0	84,070	1,910.6	32.7	1,716	39.0	1.0	9,590	218.0	5.3	---	---	---	209,378	4,758.6	19.7	167,578	3,808.6	97.6	41,800	950.0	4.7
CH_4	---	---	---	---	---	---	---	---	---	---	---	---	---	---	---	---	---	---	---	---	---	---	---	---
H_2	3,864	1,932.0	68.0	7,341	3,670.6	62.9	7,288	3,644.0	92.8	7,646	3,823.0	93.8	7,639	3,819.2	99.2	27,000	13,500.0	56.1	32	16.0	0.4	26,968	13,484.0	66.8
N_2	240	8.6	0.3	330	11.8	0.2	330	11.8	0.3	330	11.8	0.3	330	11.8	0.3	1,366	48.5	0.2	---	---	---	1,366	48.5	0.2
NH_3	---	---	---	527	31.0	0.5	527	31.0	0.8	50	3.0	0.1	---	---	---	---	---	---	---	---	---	---	---	---
H_2S	---	---	---	588	17.3	0.3	17	0.5	0.01	20	0.6	0.02	---	---	---	2,040	74.7	0.3	1,496	58.7	1.5	---	---	---
COS	---	---	---	6	0.1	0.02	6	0.1	0.002	---	---	---	---	---	---	156	2.6	0.01	---	---	---	---	---	---
DRY GAS	29,752	2,840.6	100.0	98,462	5,841.4	100.0	15,484	3,926.4	100.0	18,224	4,077.4	100.0	8,557	3,852.0	100.0	400,215	24,108.5	100.0	169,666	3,903.3	100.0	229,849	20,186.6	100.0
H_2O	65	3.6		3,258	181.0		3,258	181.0		45	2.0		---	---		1,044	58.0		576	32.0		468	26.0	
WET GAS	29,817	2,844.2		101,720	6,022.4		18,742	4,107.4		18,270	4,079.4		8,557	3,852.0		401,259	24,166.5		170,242	3,935.3		230,317	20,212.6	

DESCRIPTION	DATE	BY	CHECKED

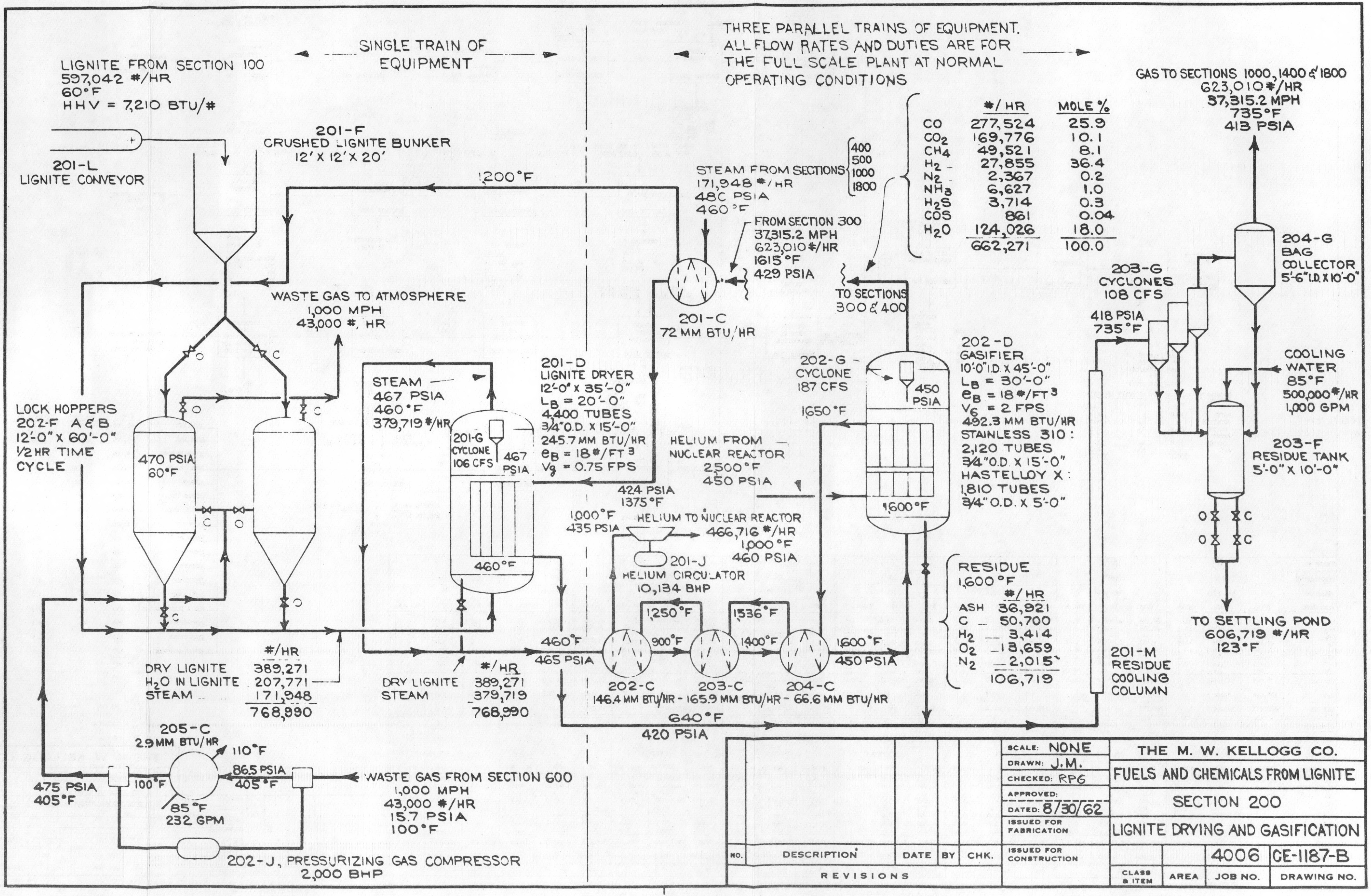
REVISIONS

SCALE: NONE
DRAWN: RPG
CHECKED: RPG
APPROVED:
DATED:
ISSUED FOR FABRICATION
ISSUED FOR CONSTRUCTION

THE M. W. KELLOGG CO.

U. S. ATOMIC ENERGY COMMISSION
LIQUID FUELS AND CHEMICALS FROM
LIGNITE PROCESS STREAM BALANCE

CLASS & ITEM	AREA	JOB NUMBER	DRAWING NUMBER
		4006	CE-1526-A



	#/HR	MOLE %
CO	277,524	25.9
CO_2	169,776	10.1
CH_4	49,521	8.1
H_2	27,855	36.4
N_2	2,367	0.2
NH_3	6,627	1.0
H_2S	3,714	0.3
COS	861	0.04
H_2O	124,026	18.0
	662,271	100.0

RESIDUE
1,600°F

	#/HR
ASH	36,921
C	50,700
H_2	3,414
O_2	13,659
N_2	2,015
	106,719

NO.	DESCRIPTION	DATE	BY	CHK.

REVISIONS

SCALE: NONE
DRAWN: J.M.
CHECKED: RPG
APPROVED:
DATED: 8/30/62
ISSUED FOR FABRICATION
ISSUED FOR CONSTRUCTION

THE M. W. KELLOGG CO.
FUELS AND CHEMICALS FROM LIGNITE
SECTION 200
LIGNITE DRYING AND GASIFICATION

CLASS & ITEM	AREA	JOB NO.	DRAWING NO.
		4006	CE-1187-B

The 1600°F, 444 psia gas stream is received from the gasifier in Section 200 at the rate of 554,108 pounds per hour and is superheated against effluent gas from the partial combustion furnace to 2050°F. The gas enters the partial combustion furnace where it reacts with oxygen, which is fed at 450 psia, 1000°F and at a rate of 68,902 pounds per hour. The methane-free gas leaves the furnace at 2350°F, passes through the raw synthesis gas superheater, the oxygen heater, and a waste-heat boiler, and is finally returned to Section 200 at 1615°F at the rate of 623,010 pounds per hour.

Section 300 includes a 900 tons per day oxygen plant, which supplies the 450 psia oxygen fed to the partial combustion furnace.

b. Production of Ammonia

(1) Section 400 - Air Partial Combustion

It is the purpose of Section 400 to remove methane from the raw synthesis gas and to add nitrogen in the proper proportion for subsequent synthesis of ammonia.

The 1600°F, 444 psia gas stream is received from the gasifier in Section 200 at the rate of 108,163 pounds per hour and is superheated to 2000°F against a portion of the effluent gas from the partial combustion furnace. This gas then enters the partial combustion furnace where it reacts with the oxygen from an air stream which is fed at 450 psia, 2000°F and at a rate of 57,340 pounds per hour. The 2350°F methane-free effluent from the furnace is employed to superheat raw synthesis gas, superheat air, and generate steam in a series of heat exchangers, and 700°F gas is sent to Section 500 for further processing at a rate of 165,503 pounds per hour.

(2) Section 500 - Shift Conversion

In Section 500 the CO-content of the methane-free gas from Section 400 is reduced to less than 1 mole percent in preparation for synthesis of ammonia.

The 700°F, 429 psia gas stream is received at a rate of 165,503 pounds per hour and enters two shift converters where it is contacted with an iron oxide catalyst in order to promote the water-gas shift reaction as follows:

$$CO + H_2O \rightleftharpoons CO_2 + H_2 + \text{heat} \qquad \text{(I-1)}$$

The reaction is mildly exothermic, therefore boiler feed water is injected into the converter to absorb heat. In addition, 700°F steam is fed in order that the reaction will proceed to the desired extent.

The 700°F shift effluent is cooled to 100°F in a series of heat exchangers in which the waste heat is used to generate steam and heat boiler feed water. The cooled synthesis gas is then countercurrently scrubbed with clean water to remove soluble impurities, and flows to Section 600 at the rate of 175,628 pounds per hour.

(3) Section 600 - Gas Purification

This section has been designed to reduce the carbon dioxide and sulfur concentrations in the synthesis gas to approximately 5 and 1 parts per million (by volume), respectively. The purification sequence consists of monoethanolamine (MEA) absorption for bulk removal of CO_2, H_2S, and COS, followed by caustic wash and activated carbon for removal of final traces.

The 100°F, 400 psia gas from Section 500 enters the MEA absorption tower where it is countercurrently scrubbed with a 20 percent MEA solution. This removes most of the CO_2 and H_2S and some of the COS. Rich solvent from the bottom of the absorber is heated and sent to the MEA stripper, where stripping gas is provided by a reboiler heated by low-pressure steam. Lean solvent from the bottom of this tower is cooled and returned to the top of the absorber. The overhead gas from the stripper is vented to the atmosphere.

Effluent gas from the MEA absorber enters the caustic scrubber where aqueous NaOH reduces the CO_2 concentration in the gas to approximately 5 ppm and the H_2S to about 1 ppm. Reacted solvent is purged from the system and replaced with makeup aqueous NaOH.

Overhead gas from the caustic scrubber is then treated for COS removal by adsorption on fixed beds of activated carbon. This scheme for COS removal was described in Section I-A-1-e of this report.

The synthesis gas, containing about 5 ppm CO_2 and 1 ppm total sulfur, flows to Section 700 for removal of carbon monoxide.

(4) Section 700 - Methanation

It is the purpose of this section to remove the remaining carbon monoxide from the synthesis gas.

The 100°F, 375 psia gas stream is received from Section 600 at the rate of 56,529 pounds per hour and is heated to 500°F with effluent gas from the methanator. The gas then enters the methanator where it is contacted with a nickel catalyst in order to promote the following reaction:

$$CO + 3\ H_2 \rightleftharpoons CH_4 + H_2O + 92{,}900\ \text{Btu} \qquad \text{(I-3)}$$

The 640°F, CO-free effluent gas is cooled to 100°F in the feed gas heater and water cooler and flows to Section 800. The purified gas contains approximately 10 ppm carbon oxides and 1 ppm sulfur.

(5) Section 800 - Gas Compression

It is the purpose of this section to compress the purified synthesis gas to the desired pressure for synthesis of ammonia.

The 355 psia gas from Section 700 is compressed to 4720 psia in a three-stage reciprocating compressor. Coolers and knockout drums are used to condense and remove water after each stage of compression. The 4720 psia, 100°F stream flows to Section 900 at the rate of 55,092.2 pounds per hour.

(6) Section 900 - Ammonia Synthesis

In this section ammonia is synthesized from the hydrogen-nitrogen gas mixture supplied from Section 800.

Feed gas to Section 900 is received at 100°F, 4720 psia and is mixed with a recycle stream. The resulting gas mixture is cooled with ammonia refrigeration and passed to a secondary separator where anhydrous ammonia (contained in the recycle stream) drops out. The gas is then passed through a heat exchanger and fed to the Kellogg catalytic ammonia converter at 90°F. The "quench-type" converter operates at 4,700 psia and the temperature is accurately and flexibly controlled to allow a catalyst temperature gradient giving a maximum yield of ammonia per pass.

Product gas exits the converter at 410°F and is cooled with water, converter feed gas, and ammonia refrigeration in a series of heat exchangers. Anhydrous liquid ammonia is then separated out in the primary separator, and the gas stream is recycled, after purging of impurities that have entered the synthesis loop with the feed gas.

The liquid streams from the primary and secondary separators are flashed to remove impurities, and anhydrous liquid ammonia product is recovered at a rate of 50,000 pounds per hour.

c. Production of Methanol

(1) Section 1000 - Shift Conversion

It is the purpose of Section 1000 to adjust the H_2/CO ratio in the methane-free raw synthesis gas to approximately 2.2/1 in preparation for synthesis of methanol.

Approximately 60 percent of the 67,878 pounds per hour gas stream received from Section 200 is fed to the shift converter, where it is contacted with an iron oxide catalyst in order to promote the water-gas shift reaction as follows:

$$CO + H_2O \rightleftharpoons CO_2 + H_2 + \text{heat} \qquad \text{(I-1)}$$

The reaction is mildly exothermic, therefore boiler feed water is injected into the converter to absorb heat.

The 800°F shift effluent is mixed with the 700°F by-passed gas and the combined stream is cooled to 100°F in a series of heat exchangers in which the waste heat is used to generate steam and heat boiler feed water. The cooled synthesis gas is countercurrently scrubbed with clean water to remove ammonia and flows to Section 1100 at the rate of 58,150 pounds per hour.

(2) Section 1100 - Gas Purification

This section has been designed to reduce the CO_2 concentration in the shifted synthesis gas to 1.0 mole percent and to reduce the total sulfur content to about 0.004 grains per 100 SCF of gas. The purification sequence consists of the "Fluor Solvent CO_2 Removal Process" for CO_2 and bulk H_2S removal, followed by sponge iron and activated carbon for residual H_2S and organic sulfur removal, respectively. The process scheme is essentially the same as that described in Section I-A-1-e of this report, except that flash gas from the "Fluor Solvent" reconcentrator is not recycled to the CO_2 absorption tower. The purified gas from Section 1100 flows to Section 1200 at the rate of 29,817 pounds per hour.

(3) Section 1200 - Gas Compression

It is the purpose of this section to compress the purified synthesis gas to the desired pressure for synthesis of methanol.

The 365 psia, 85°F gas from Section 1100 is compressed to 5145 psia in a three-stage reciprocating compressor. Coolers and separator drums are used to condense and remove water after each stage of compression. The 5145 psia, 100°F stream proceeds to Section 1300 at a rate of 29,764 pounds per hour.

(4) Section 1300 - Methanol Synthesis

In this section methanol is synthesized from the gas supplied by Section 1200. The 5145 psia, 100°F gas received at a rate of 29,764 pounds per hour is mixed with a recycle stream and is sent to the methanol converter. Within the converter a portion of the carbon monoxide and the hydrogen react in the presence of a catalyst to form methanol, according to the following reaction:

$$CO + 2\ H_2 \rightleftharpoons CH_3OH + heat \qquad (I-4)$$

Cold feed is injected between the catalyst beds as quench in order to dissipate the exothermic heat of reaction.

Effluent from the last catalyst bed passes through an effluent-feed exchanger, a fin-fan cooler, and a water cooler to a high-pressure flash drum from which the overhead gas is compressed and recycled to the methanol converter. The recycle gas is compressed in a single steam-driven circulator which exhausts steam to a barometric condenser. The liquid from the high-pressure separator is let down in two stages. Flash gases from the let-down drums are chilled with refrigeration in order to reduce methanol loss.

Methanol from the let-down drums is sent to a chemical treatment system where most of the impurities produced during the synthesis operation are removed. The methanol is next pumped to a fractionating tower which produces high-purity specification methanol. The light ends are removed overhead and are used as fuel in the offsite boilers. Water and any traces of heavy alcohols are removed from the bottom. The refined methanol product is taken off above the feed plate, cooled, and sent to storage at the rate of 25,000 pounds per hour.

d. Production of Hydrogen

(1) Section 1400 - Primary Shift Conversion

Section 1400 removes a large quantity of CO from the methane-free raw synthesis gas. Gas is received from Section 200 at 735°F, 413 psia and at a rate of 93,183 pounds per hour. After heating the feed gas to Section 1600, the 550°F raw synthesis gas enters two parallel shift converters where it is contacted with an iron oxide catalyst in order to promote the water-gas shift reaction as follows:

$$CO + H_2O \rightleftharpoons CO_2 + H_2 + \text{heat} \qquad \text{(I-1)}$$

Boiler feed water is injected into the converter to absorb the mildly exothermic heat of reaction and 700°F steam is added to the feed gas in order to drive the reaction to the desired degree of completion. Approximately 88 percent of the entering carbon monoxide is shifted in these converters.

The 700°F shift effluent is cooled to 220°F while performing a reboiling duty for Section 1500. Entrained condensate is removed in a knockout drum and the gas flows to the carbon dioxide absorber in Section 1500 at a rate of 101,720 pounds per hour.

(2) Section 1500 - Primary Gas Purification

It is the purpose of this section to remove the bulk of the CO_2 and H_2S from the shifted synthesis gas.

The 395 psia gas from Section 1400 enters an absorption tower where it is countercurrently scrubbed with a 30 percent potassium carbonate solution to remove about 98 percent of the entering CO_2 and H_2S. A small quantity of hydrogen is also absorbed, but the other components pass through the tower unaffected. The 220°F absorber effluent gas is heated to 570°F by the feed gas to the primary shift converters and then flows to Section 1600 at a rate of 18,742 pounds per hour.

The solute-rich carbonate solution from the bottom of the absorber flows to the top of a stripping column where absorbed CO_2, H_2S, and H_2 are removed at low pressure and vented to the stack. Stripping gas is supplied by a reboiler where heat is supplied by the effluent gas from the primary shift converters. Solvent cuts are taken from the middle and bottom of the stripper and fed to the middle and top of the absorber, respectively, to yield the optimum system design.

(3) Section 1600 - Secondary Shift Conversion

It is the purpose of this section to remove most of the carbon monoxide remaining in the synthesis gas. The 387 psia, 570°F gas received from Section 1500 enters the two parallel secondary shift converters where 90 percent of the remaining CO is removed by the water-gas shift reaction. A small quantity of boiler feed water is injected into the converters to absorb the exothermic heat of reaction and 700°F steam is added to the feed gas to drive the reaction to the desired degree of completion.

The 650°F shift effluent is cooled to 100°F in a series of heat exchangers in which the waste heat is used to generate steam and heat boiler feed water. After removal of entrained condensate in a knockout drum, the gas flows to Section 1700 at a rate of 18,269 pounds per hour.

(4) Section 1700 - Secondary Gas Purification

It is the purpose of this section to remove the remaining CO_2, H_2S, and COS from the synthesis gas. This is the final step in the production of 99 percent pure hydrogen.

The 100°F, 375 psia gas received from Section 1600 at the rate of 18,270 pounds per hour enters the absorption tower where it is countercurrently scrubbed with 20 percent MEA solution. The rich solvent from the bottom of the absorber is stripped of absorbed components and then returned to the top of the absorber.

Effluent gas from the MEA absorber is scrubbed with aqueous NaOH to remove residual CO_2 from the gas. Reacted solvent is purged from the system and replaced with makeup aqueous NaOH.

The purified gas then flows to an alumina dryer where water is removed and the dry gas is subsequently compressed to 450 psia. The 99 percent pure hydrogen gas is thus produced at a rate of 35 million standard cubic feet per day.

Further descriptive information on Sections 1400 through 1700 may be found in a previous Kellogg report for the Bureau of Mines (12).

Production of Liquid Fuels

(1) Section 1800 - Shift Conversion

Section 1800 consists of two parallel trains of operating equipment. It is the purpose of this section to adjust the H_2/CO ratio in the methane-free raw synthesis gas to approximately 2.34/1 in preparation for synthesis of liquid fuels.

The 413 psia, 735°F gas stream is received from Section 200 at the rate of 461,949 pounds per hour and enters the shift converter where an iron oxide catalyst promotes the water-gas shift reaction. Boiler feed water is injected into the converter to absorb the mildly exothermic heat of reaction.

The 780°F shift effluent is cooled to 100°F in a series of heat exchangers where the waste heat is used to generate steam and heat boiler feed water. The cooled synthesis gas is countercurrently scrubbed with clean water to remove ammonia and flows to Section 1900 at the rate of 401,259 pounds per hour.

(2) Section 1900 - Gas Purification

Section 1900 also consists of two parallel operating trains of equipment. This section has been designed to reduce the CO_2 concentration in the shifted synthesis gas to below 5.0 mole percent and to reduce the total sulfur content to about 0.004 grains per 100 SCF of gas. The purification sequence consists of the "Fluor Solvent CO_2 Removal Process" for CO_2 and bulk H_2S removal, followed by sponge iron and activated carbon for residual H_2S and COS removal, respectively. The process scheme is essentially the same as that described in Section I-A-1-e of this report, except that flash gas from the solvent reconcentrator is not recycled to the absorption tower. The purified gas is sent to Section 2000 for synthesis and recovery of liquid fuels.

(3) Section 2000 - Synthesis and Recovery

The purpose of this section is to synthesize and recover the various end products shown on Drawing No. CE-1526-A from the gas supplied by Section 1900.

Fresh feed gas enters Section 2000 at the rate of 230,317 pounds per hour at 80°F and 375 psia and is combined with a recycle stream of hydrogen and carbon monoxide. The gas mixture then enters the Kellogg synthesis reactor where the carbon monoxide is hydrogenated over an iron catalyst to produce hydrocarbons and water. Side reactions also produce some oxygenated compounds. Synthesis is carried out at 370 psia and about 650°F in a transport-type reactor. Catalyst for synthesis, a promoted iron, is prepared in the plant by reduction of iron ore.

Reactor effluent gas is quenched in a scrubbing tower, and the residue from this tower, a mixture of catalyst fines and heavy oil, is returned to the reactor. Heavy oil product is also taken off the bottom of the tower. The overhead gas from the scrubber is cooled and separated into a gas phase and two liquid phases -- water and oil.

The water phase from the separator is sent to a chemicals recovery plant where it undergoes distillation. All of the oxygenated products and their azeotropes, except the organic acids, are carried overhead.

The acid-water stream from the bottom of the distillation tower is sent to an acid recovery section, where the acids are extracted with an ethyl acetate solvent. This solvent-acid stream is sent to a distillation tower where the crude acids are separated from the solvent. These crude acids then flow to the acid purification system, where pure acetic and propionic acids are isolated and removed as products.

The overhead product from the water-phase distillation column, containing the non-acidic oxygenated compounds, is cooled and distilled to separate the carbonyls and methanol from the heavier alcohols. Further distillation of the light components yields acetone, acetaldehyde, and methyl ethyl ketone (MEK) as products, while distillation of the alcohols yields motor ethanol, 95 percent propanol, isobutanol, 1-butanol, and heavy alcohols as products.

The oil product from the synthesis unit is processed in an oil treating section where it undergoes fractionation. Three liquid streams and a gas stream result. The heavy liquid bottoms are sent to storage as a waxy oil product. The gas stream is combined with product gas from the synthesis unit, and the resulting mixture passes through an oil absorption tower where the bulk of the C_3 fraction and all of the C_4 and heavier hydrocarbons are removed. The overhead gas from the absorber, now largely hydrogen, methane, and CO_2, undergoes CO_2 removal and is then catalytically reformed with steam and oxygen. This hydrogen-rich gas is recycled and combined with the fresh feed entering the synthesis unit.

The hydrocarbon-rich bottoms are combined with the light oil product from the oil treating section and the mixture is sent to a de-ethanizer tower for further fractionation. The tail gas from this tower is sent to an ethylene plant for further processing to yield ethylene and propylene products. The light liquid product from the de-ethanizer flows to a catalytic polymerization section, where it is treated to produce propane LPG, butane, and polymer gasoline. The heavy liquid product from the de-ethanizer is combined with the medium oil product from the oil treating section, and this mixture is fed to a fractionation unit where straight-run gasoline and diesel oil are produced.

f. Plant Utilities

(1) Section 3000 - Offsite Facilities

Section 3000 includes facilities for:

(a) generating steam and electric power
(b) supplying cooling water, process water, and boiler feed water
(c) providing miscellaneous services necessary to make the plant completely self-sufficient.

These offsite facilities are discussed in the Utilities Summary contained in the Appendix. The offsite equipment was briefly described in Section I-A-1-g of this report.

2. Economics

In order to calculate production costs and selling prices for the several major products of this plant, it is first necessary to calculate the cost of dry, raw synthesis gas as it leaves the gasification section. This gas is then treated as the raw material for subsequent processes to produce ammonia, methanol, hydrogen, and gasoline.

The gasoline-manufacturing step produces a number of fuels and chemicals as co-products. To avoid further complexity, these co-products are credited against the cost of gasoline manufacture at 95 percent of their current selling price (allowing 5 percent for selling and distribution expenses). This method of calculation throws the complete burden of the unorthodox manufacturing process into the gasoline cost and makes it higher than it would be if all co-products shared the burden.

Utilities -- steam, power, cooling water, etc -- are produced in Section 3000 (with some steam also from waste-heat boilers in other sections) for use in all sections of the plant. With the method of calculation outlined above, it is necessary to calculate a unit production cost for each utility and then to charge each section for the number of units consumed. In calculating return on investment, the bare cost of the offsite facilities, Section 3000, is then prorated among the major product plants. This will become clearer as the economic tables are discussed.

Estimated operating labor for the complete process plant is 78 men per shift, as detailed in Table I-5. Operating labor for offsite facilities is not included here because it is included in the calculated cost of each utility.

Estimated capital investment for the plant is summarized in Table I-6. Interest during construction is calculated at 6.7 percent of the sum of total bare cost plus contractor's overhead and profit, assuming a design, engineering, and construction period equal to 30 months, and the cost of money to be 6 percent per annum. Details are presented in Section I-A-2-a. Working capital includes 30 days' lignite inventory, 30 days' accounts receivable. and in-plant inventories of catalysts with relatively short lives. Shift catalyst, ammonia synthesis catalyst, and activated carbon are included in fixed investment because they have long lifetimes. Total capital investment is about 133 million dollars.

Estimated annual operating cost for synthesis gas production is calculated in Table I-7. Fixed investment of $18,080,000 includes the bare cost of Sections 000, 100, and 200, and a pro-rata share of contractor's overhead and profit and interest during construction. Nuclear heat is charged at 50 cents per million Btu and lignite at $2 per ton. Operating labor is that for Sections 000,100, and 200, taken from Table I-5. Utilities costs are obtained from Table I-12.

Estimated synthesis gas production cost, assuming 20-year straight-line depreciation and not including any return on invested capital, is 12.0¢/MSCF.

Using synthesis gas at 12.0¢/MSCF as raw material, production costs for the major products are calculated in Tables I-8 through I-11 to be as follows:

ammonia	34.30 $/ton
methanol	9.5¢/gal
hydrogen (99%)	27.5¢/MSCF
gasoline	15.1¢/gal

Catalyst and chemicals charges are detailed in Table I-13, and co-product credits in the case of gasoline production are calculated in Table I-14.

The effect on product costs of adding gross return on investment is shown in Figures I-5 through I-8. If a gross return of 20 percent is desired, for example, the above production costs are increased as follows:

	Production Cost	20 Percent Gross Return	Selling Price
ammonia, $/ton	34.30	23.00	57.30
methanol, ¢/gal	9.5	6.4	15.9
hydrogen, ¢/MSCF	27.5	19.6	47.1
gasoline, ¢/gal	15.1	26.3	41.4

The effect of variations in the cost of lignite or nuclear heat is seen to be relatively insignificant for all four products. Investment was prorated in the following manner to calculate gross return. For example, total investment for methanol production is the sum of the following:

a. total bare cost of Sections 1000 to 1300

b. 9.20 percent of the bare cost of Sections 000 to 200, since that fraction of the raw synthesis gas is consumed in producing methanol

c. 11.0 percent of Section 300, since that fraction of the gas treated by Section 300 is used in producing methanol

d. 7.2 percent of the sum of offsite facilities, contractor's overhead and profit, interest during construction and working capital $\left(\frac{a + b + c}{\text{total bare cost of process sections}} = 0.072\right)$.

Investment for the other major products was calculated similarly.

The effect of stream efficiency on production costs is illustrated in Figure I-9.

Table I-5

Estimated Operating Labor
Liquid Fuels and Chemicals from Lignite

Section	Title	Men per Shift	
Synthesis Gas Production:			
000	Lignite Storage and Reclamation	1	
100	Crushing	2	
200	Drying and Gasification	4	
	Subtotal		7
Ammonia Production:			
400	Air Partial Combustion	2	
500	Shift Conversion	1	
600	Gas Purification	2	
700	Methanation	1	
800	Gas Compression		
900	Synthesis	1	
	Subtotal		7
Oxygen Partial Combustion (Section 300)			8
Methanol Production:			
1000	Shift Conversion	1	
1100	Gas Purification	3	
1200	Gas Compression	1	
1300	Synthesis	2	
	Subtotal		7
Hydrogen Production:			
1400	Primary Shift Conversion	1	
1500	Primary Gas Purification	1	
1600	Secondary Shift Conversion	1	
1700	Secondary Gas Purification and Compression	1	
	Subtotal		4
Gasoline and Chemicals Production:			
1800	Shift Conversion	1	
1900	Gas Purification	5	
2000	Synthesis and Recovery	39	
	Subtotal		45
	TOTAL OPERATING LABOR		78
	man-hours/day		1,872

Table I-6

Investment Summary
Liquid Fuels and Chemicals from Lignite

Section	Title	Materials & Freight, Dollars	Bare Cost,* Dollars	
Synthesis Gas Production:				
000	Lignite Storage & Reclamation	1,143,000	2,430,000	
100	Crushing	637,000	1,145,000	
200	Drying & Gasification	7,321,130	11,764,130	
	Subtotal	9,101,130		15,339,130
Ammonia Production:				
400	Air Partial Combustion	1,750,000	2,750,000	
500	Shift Conversion	760,000	1,185,000	
600	Gas Purification	1,475,000	2,750,000	
700	Methanation	266,000	450,000	
800	Gas Compression	1,100,000	1,650,000	
900	Synthesis	3,100,000	4,376,000	
	Subtotal	8,451,000		13,161,000
Oxygen Partial Combustion (Section 300)		8,630,000		12,200,000
Methanol Production:				
1000	Shift Conversion	271,000	451,000	
1100	Gas Purification	704,000	1,085,000	
1200	Gas Compression	660,000	1,125,000	
1300	Synthesis	660,000	1,200,000	
	Subtotal	2,295,000		3,861,000
Hydrogen Production:				
1400	Primary Shift Conversion	265,000	445,000	
1500	Primary Gas Purification	1,170,000	2,160,000	
1600	Secondary Shift Conversion	420,000	680,000	
1700	Secondary Purification	437,500	750,000	
	Subtotal	2,292,500		4,035,000
Gasoline and Chemicals Production:				
1800	Shift Conversion	1,400,000	2,100,000	
1900	Gas Purification	2,975,000	4,950,000	
2000	Synthesis and Recovery	23,000,000	36,000,000	
	Subtotal	27,375,000		43,050,000
Offsite Facilities (Section 3000)		10,000,000		16,180,000
	TOTAL BARE COST (Carried Forward)			107,826,130

Table I-6 (Continued)

Investment Summary

Liquid Fuels and Chemicals from Lignite

TOTAL BARE COST (Brought Forward)		107,826,130
Contractor's Overhead and Profit		11,320,000
Interest during Construction		7,980,000
TOTAL FIXED INVESTMENT		127,126,130
Working Capital:		
30 days' lignite inventory (291,000 tons @ $2/ton)	582,000	
Accounts receivable(value of 30 days' production)		
Ammonia @ $90/ton	1,620,000	
Methanol @ 30¢/gal	818,000	
Hydrogen @ 40¢/MSCF	420,000	
Gasoline @ 20¢/gal	1,240,000	
Co-products	1,000,000	
Catalyst inventory	115,000	
Total Working Capital		5,795,000
TOTAL CAPITAL INVESTMENT		132,921,130

* Bare cost includes materials, freight, construction labor, field administration and supervision, insurance during construction, cost of tools, field office expense, and cost of home office engineering and procurement.

Table I-7

Estimated Annual Operating Cost
Synthesis Gas from Lignite

Basis: 286,000,000 SCFD of Dry Synthesis Gas
90% Stream Efficiency
Fixed Investment = $18,080,000 (Sections 000, 100, 200)

Item	$/year		¢/MSCF
Direct Costs:			
Nuclear heat (871.2 MMBtu/hr x 7,884 hr/yr x 50¢/MMBtu)	3,430,000		3.7
Lignite to gasifiers (298.5 TPH x 7,884 hr/yr x $2/ton)	4,707,000		5.0
Operating labor (168 man-hr/day x $2.80/man-hr x 365 days/yr)	172,000		0.2
Supervision @ 15% of operating labor	26,000		0.03
Utilities (Table I-12)	183,000		0.2
Maintenance @ 4% of bare cost per year (1)	613,000		0.6
Supplies @ 15% of maintenance	92,000		0.1
Total Direct Costs		9,223,000	
Indirect Costs:			
Payroll overhead @ 20% of operating labor, maintenance labor, and supervision	125,000		0.1
General plant overhead @ 50% of operating labor, supervision maintenance, and supplies	452,000		0.5
Total Indirect Costs		577,000	
Fixed Costs:			
Depreciation @ 5% of fixed investment	904,000		1.0
Property taxes and insurance @ 3% of fixed investment	542,000		0.6
Total Fixed Costs		1,446,000	
TOTAL ANNUAL OPERATING COST		11,246,000	
ESTIMATED SYNTHESIS GAS PRODUCTION COST			12.0

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(1) Maintenance is 70% labor, 30% material

Table I-8

Estimated Annual Operating Cost
Ammonia from Lignite

Basis: 600 Tons/Day of Ammonia
90 % Stream Efficiency
Fixed Investment = $15,500,000 (Sections 400 through 900)

Item	$/Year		$/Ton NH_3
Direct Costs:			
Synthesis gas (1,946 MSCF/hr x 7,884 hr/yr x 12.0¢/MSCF)	1,841,000		9.30
Operating labor (168 man-hr/day x $2.80/man-hr x 365 days/yr)	172,000		0.90
Supervision @ 15% of operating labor	26,000		0.10
Utilities (Table I-12)	2,254,000		11.40
Catalysts and chemicals (Excerpted from Table I-13)	119,000		0.60
Maintenance @ 4% of bare cost per year (1)	526,000		2.70
Supplies @ 15% of maintenance	79,000		0.40
Total Direct Costs		5,017,000	
Indirect Costs:			
Payroll overhead @ 20% of operating labor, maintenance labor, and supervision	113,000		0.60
General plant overhead @ 50% of operating labor, supervision, maintenance, and supplies	402,000		2.00
Total Indirect Costs		515,000	
Fixed Costs:			
Depreciation @ 5% of fixed investment	775,000		3.90
Property taxes and insurance @ 3% of fixed investment	465,000		2.40
Total Fixed Costs		1,240,000	
TOTAL ANNUAL OPERATING COSTS		6,772,000	
TOTAL AMMONIA PRODUCTION COST			34.30

(1) Maintenance is 70% labor, 30% material

Table I-9

Estimated Annual Operating Cost
Methanol from Lignite

Basis: 300 Tons/Day of Methanol
90% Stream Efficiency
Fixed Investment = $6, 130, 000 (Sections 1000 - 1300 plus 11% of Section 300)

Item	$/Year		¢/gal CH_3OH
Direct Costs:			
Synthesis gas (1, 096 MSCF/hr x 7, 884 hr/yr x 12.0¢/MSCF)	1, 037, 000		3.5
Operating labor (190 man-hr/day x $2.80/man-hr x 365 days/yr)	194, 000		0.6
Supervision @ 15% of operating labor	29, 000		0.1
Utilities (Table I-12)	532, 000		1.8
Catalysts and chemicals (Excerpted from Table I-13)	22, 000		0.1
Maintenance @ 4% of bare cost per year (1)	208, 000		0.7
Supplies @ 15% of maintenance	31, 000		0.1
Total Direct Costs		2, 053, 000	
Indirect Costs:			
Payroll overhead @ 20% of operating labor, maintenance labor, and supervision	74, 000		0.2
General plant overhead @ 50% of operating labor, supervision, maintenance, and supplies	231, 000		0.8
Total Indirect Costs		305, 000	
Fixed Costs:			
Depreciation @ 5% of fixed investment	306, 000		1.0
Property taxes and insurance @ 3% of fixed investment	184, 000		0.6
Total Fixed Costs		490, 000	
TOTAL ANNUAL OPERATING COST		2, 848, 000	
ESTIMATED METHANOL PRODUCTION COST			9.5

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(1) Maintenance is 70% labor, 30% material

Table I-10

Estimated Annual Operating Cost
Hydrogen from Lignite

Basis: 35,000,000 SCFD of 99.2% H_2
90% Stream Efficiency
Fixed Investment = \$6,920,000 (Sections 1400 - 1700 plus 14.9% of Section 300)

Item	$/Year		¢/MSCF H_2
Direct Costs:			
Synthesis gas (1,487 MSCF/hr x 7,884 hr/yr x 12.0¢/MSCF)	1,407,000		12.2
Operating labor (125 man-hr/day x \$2.80/man-hr x 365 days/yr)	128,000		1.1
Supervision @ 15% of operating labor	19,000		0.2
Utilities (Table I-12)	471,000		4.1
Catalysts and chemicals (Excerpted from Table I-13)	46,000		0.4
Maintenance @ 4% of bare cost per year (1)	234,000		2.1
Supplies @ 15% of maintenance	35,000		0.3
Total Direct Costs		2,340,000	
Indirect Costs:			
Payroll overhead @ 20% of operating labor, maintenance labor, and supervision	62,000		0.5
General plant overhead @ 50% of operating labor, supervision, maintenance, and supplies	208,000		1.8
Total Indirect Costs		270,000	
Fixed Costs:			
Depreciation @ 5% of fixed investment	346,000		3.0
Property taxes and insurance @ 3% of fixed investment	208,000		1.8
Total Fixed Costs		554,000	
TOTAL ANNUAL OPERATING COST		3,164,000	
ESTIMATED HYDROGEN PRODUCTION COST			27.5

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(1) Maintenance is 70% labor, 30% material

Table I-11

Estimated Annual Operating Cost
Gasoline from Lignite

Basis: 4,930 Bbl/Day of Gasoline
90% Stream Efficiency
Fixed Investment = $61,420,000 (Sections 1800 - 2000 plus 74.1% of Section 300)

Item	$/Year		¢/gal gasoline
Direct Costs:			
Synthesis gas (7,390 MSCF/hr x 7,884 hr/yr x 12.0¢/MSCF)	6,990,000		10.3
Operating labor (1,222 man-hr/day x $2.80/man-hr x 365 days/yr)	1,249,000		1.8
Supervision @ 15% of operating labor	187,000		0.3
Utilities (Table I-12)	2,388,000		3.5
Catalysts and chemicals (Excerpted from Table I-13)	587,000		0.9
Maintenance @ 4% of bare cost per year (1)	2,080,000		3.1
Supplies @ 15% of maintenance	312,000		0.5
Total Direct Costs		13,793,000	
Indirect Costs:			
Payroll overhead @ 20% of operating labor, maintenance labor, and supervision	578,000		0.8
General plant overhead @ 50% of operating labor, supervision, maintenance, and supplies	1,914,000		2.8
Total Indirect Costs		2,492,000	
Fixed Costs:			
Depreciation @ 5% of fixed investment	3,071,000		4.5
Property taxes and insurance @ 3% of fixed investment	1,843,000		2.7
Total Fixed Costs		4,914,000	
Co-product credits (Table I-14)		(10,947,000) CR.	(16.1) CR.
TOTAL ANNUAL OPERATING COST		10,252,000	
ESTIMATED GASOLINE PRODUCTION COST			15.1

(1) Maintenance is 70% labor, 30% material

Table I-12

Utilities Costs
Liquid Fuels and Chemicals from Lignite

Basis: Stream Efficiency = 90%
Unit Costs:

Steam:	¢/Mlb.		
1000 psig, 850° F	40	Electricity	0.7¢/kwh
500 psig, 550° F	31	Cooling Water	0.7¢/M gal
500 psig, saturated	29	Boiler Feed Water	20. ¢/M gal
300 psig, saturated	27	Process Water	15. ¢/M gal
40 psig, saturated	19	Fuel	13. ¢/MM Btu

Plant Section	Raw Synthesis Gas		Ammonia		Methanol		Hydrogen		Gasoline	
Product Rate	286,000,000 SCFD		600 T/D		300 T/D		35,000,000 SCFD		4,930 Bbl/Day	
Utilities Consumption:	Rate	$/year	Rate (1)	$/year	Rate (2)	$/year	Rate (3)	$/year	Rate (4)	$/year
Steam, M#/hr:										
1000 psig, 850° F	89.	281,000	154.	486,000	47.	148,000	9.6	30,000	405.	1,277,000
500 psig, 550° F	--	--	255.	623,000	--	--	78.1	215,000	75.	207,000
500 psig, sat'd.	172.	393,000	--	--	--	--	--	--	--	--
300 psig, sat'd.	--	--	8.5	18,000	--	--	13.0	27,000	--	--
40 psig, sat'd.	--	--	3.0	5,000	37.1	56,000	82.6	124,000	125.	187,000
Electricity, kw	1,557.	86,000	24,680.	1,362,000	6,820.	376,000	2,423.	134,000	24,050.	1,327,000
Cooling water, MGPM	16.0	53,000	43.0	142,000	14.5	48,000	12.6	42,000	53.9	178,000
Boiler feed water, M#/hr	--	--	205.	39,000	36.4	7,000	47.7	9,000	237.	45,000
Process water, M#/hr	--	--	65.	9,000	37.5	5,000	--	--	81.9	12,000
Total Costs		813,000		2,684,000		640,000		581,000		3,233,000
Utilities Production:										
Steam:										
500 psig, sat'd.	--	--	120.	274,000	10.7	24,000	--	--	41.2	94,000
300 psig, sat'd.	--	--	--	--	--	--	--	--	21.5	46,000
40 psig, sat'd.	--	--	45.	67,000	20.6	31,000	60.8	91,000	138.	207,000
Boiler feed water										
Heating, MM Btu/hr	--	--	30.	31,000	17.	17,000	19.	19,000	475.	487,000
Fuel, MM Btu/hr	615.	630,000	57.	58,000	35.	36,000	--	--	11.	11,000
Total Credits		630,000		430,000		108,000		110,000		845,000
NET UTILITIES COST		183,000		2,254,000		532,000		471,000		2,388,000

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(1) Sections 400 through 900
(2) Sections 1000 through 1300 plus 11.0% of Section 300
(3) Sections 1400 through 1700 plus 14.9% of Section 300
(4) Sections 1800, 1900, & 2000 plus 74.1% of Section 300

Table I-13

Catalysts and Chemicals Consumption
Liquid Fuels and Chemicals from Lignite

Basis: 90% Stream Efficiency

Section	Chemical	Quantity, lb./hr.	Unit Cost, cents/lb.	$/year
600	MEA	45	28	99,000
	NaOH	68	3.8	20,000
1100	Sponge iron	49	3	12,000
1300	Synthesis catalyst			10,000
1500	K_2CO_3	33	10	26,000
1700	MEA	3	28	7,000
	NaOH	42	3.8	13,000
1900	Sponge iron	350	3	83,000
2000	Makeup Chemicals	400	15	472,000
	Iron ore for making synthesis catalyst	8(tons/day)	12($/ton)	32,000
	TOTAL			774,000

Table I-14

Co-product Credits
Gasoline from Lignite

Basis: 4,930 Bbl/Day of Gasoline
90% Stream Efficiency

Co-product	Unit	Production Rate, units/hour	Unit Value (1)	Credit, $/year
Ethylene	pound	5,700	4.5	2,020,000
Motor ethanol	gallon	747	45.	2,650,000
Propane LPG	gallon	845	7.6	506,000
Butane	gallon	528	8.8	366,000
Diesel oil	gallon	319	9.5	239,000
Acetic acid	pound	2,000	9.5	1,497,000
Waxy oil	gallon	199	5.0	78,000
Propanol	pound	1,200	11.	1,040,000
Propylene	pound	940	3.3	245,000
Acetone	pound	860	6.2	420,000
Propionic acid	pound	450	20.	708,000
1-Butanol	pound	380	13.	389,000
Methyl Ethyl Ketone	pound	290	12.	274,000
Acetaldehyde	pound	260	9.5	195,000
Isobutanol	pound	180	12.	170,000
Heavy alcohols	pound	127	15	150,000
TOTAL CO-PRODUCT CREDITS				10,947,000

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(1) Credit taken at 95% of current selling price to allow for selling and distribution costs

Figure I-5
Effect of Lignite Cost, Cost of Nuclear Heat, and
Return on Investment on the Selling Price of Ammonia

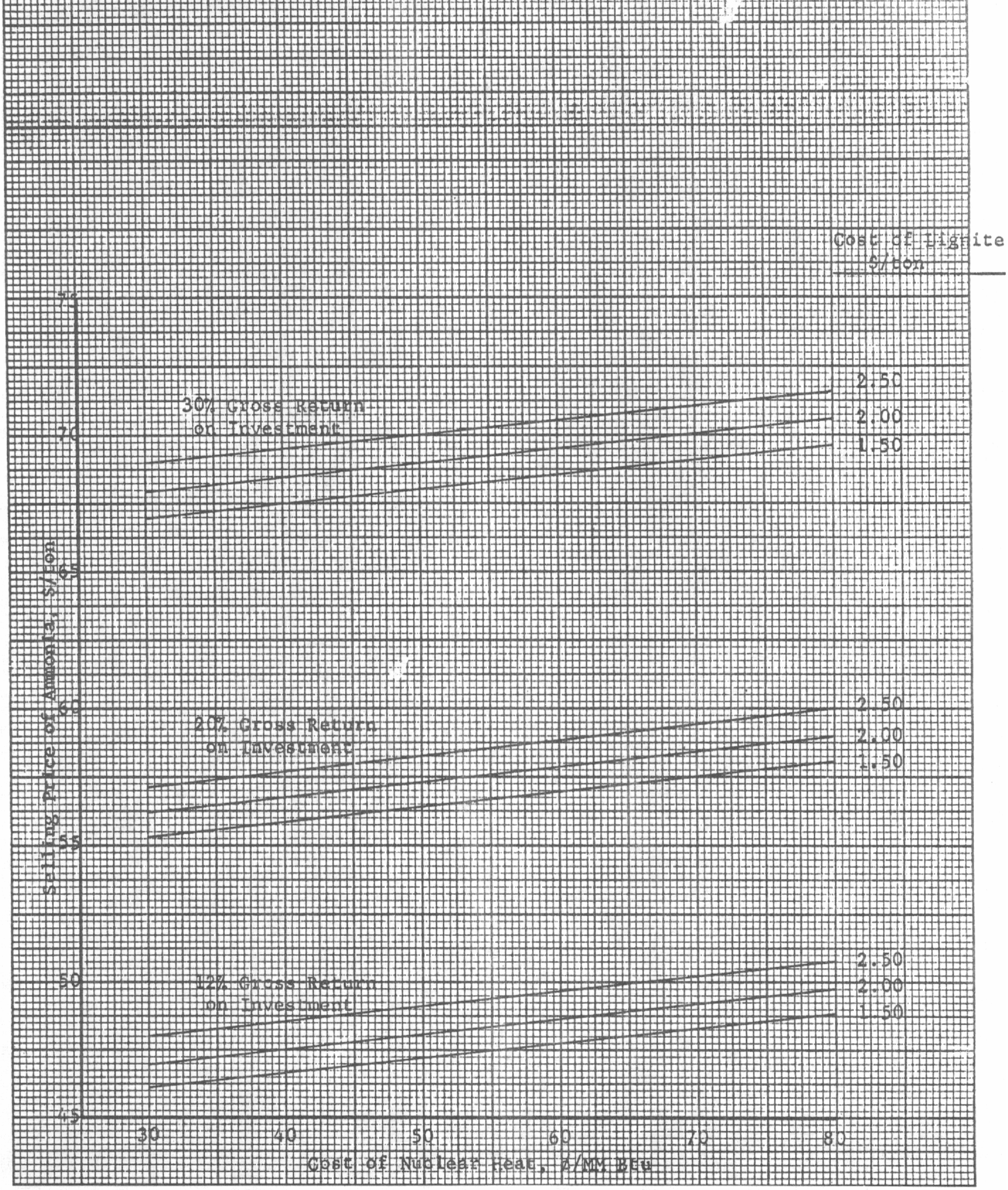


Figure I-6
Effect of Lignite Cost, Cost of Nuclear Heat, and Return on Investment on the Selling Price of Methanol

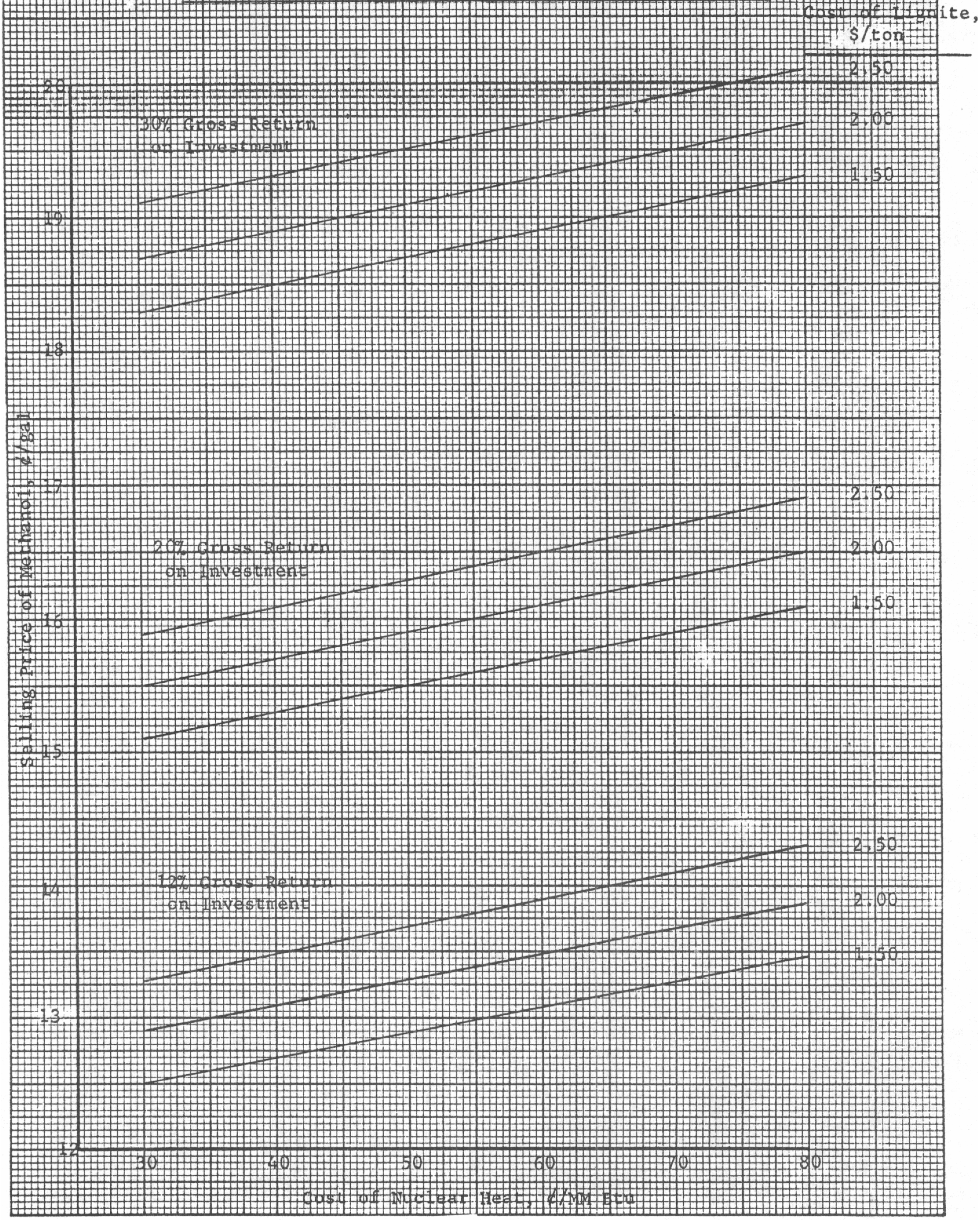


Figure I-7

Effect of Lignite Cost, Cost of Nuclear Heat, and Return on Investment on the Selling Price of Hydrogen

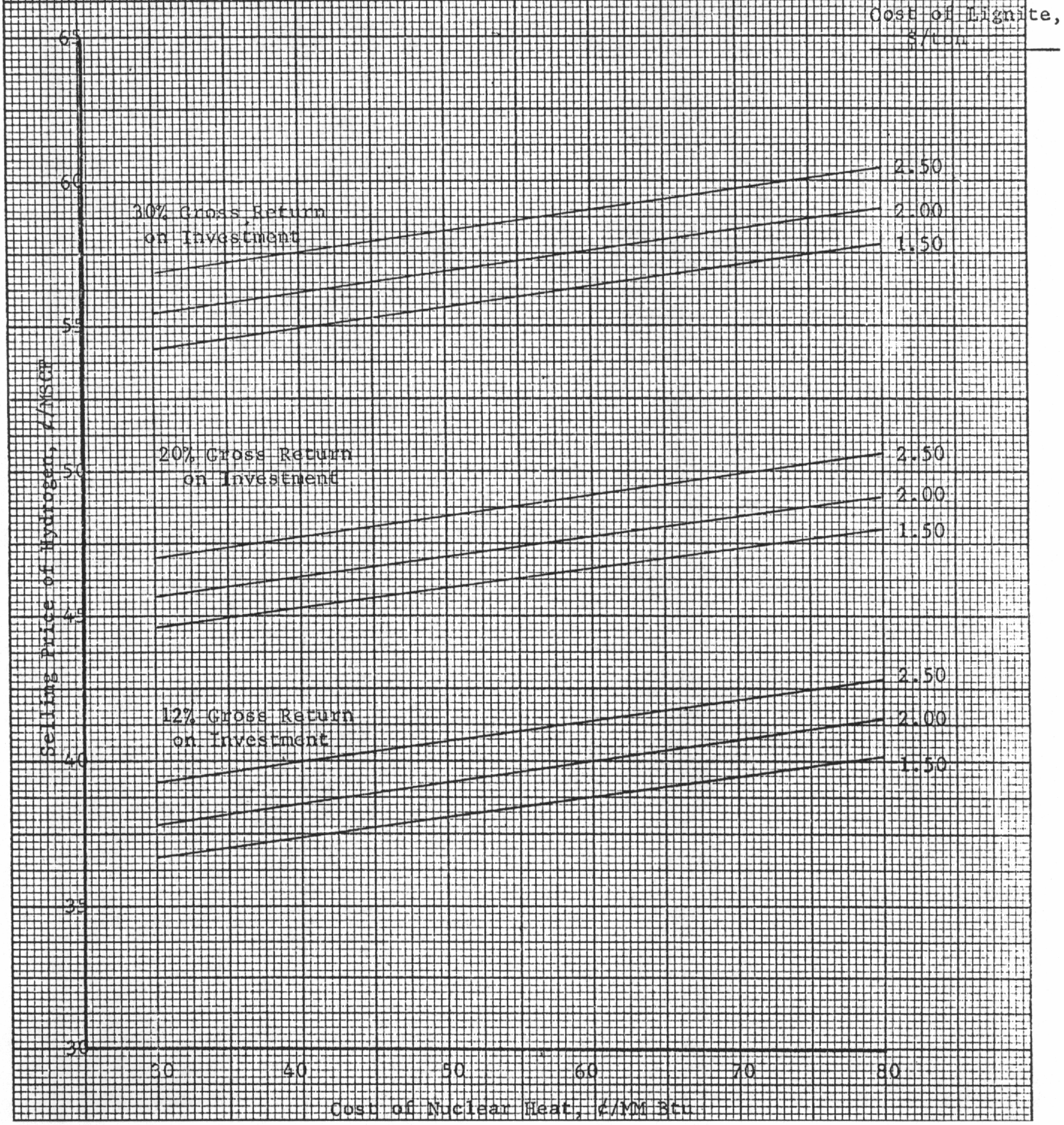


Figure I-8

Effect of Lignite Cost, Cost of Nuclear Heat, and Return on Investment on the Selling Price of Gasoline

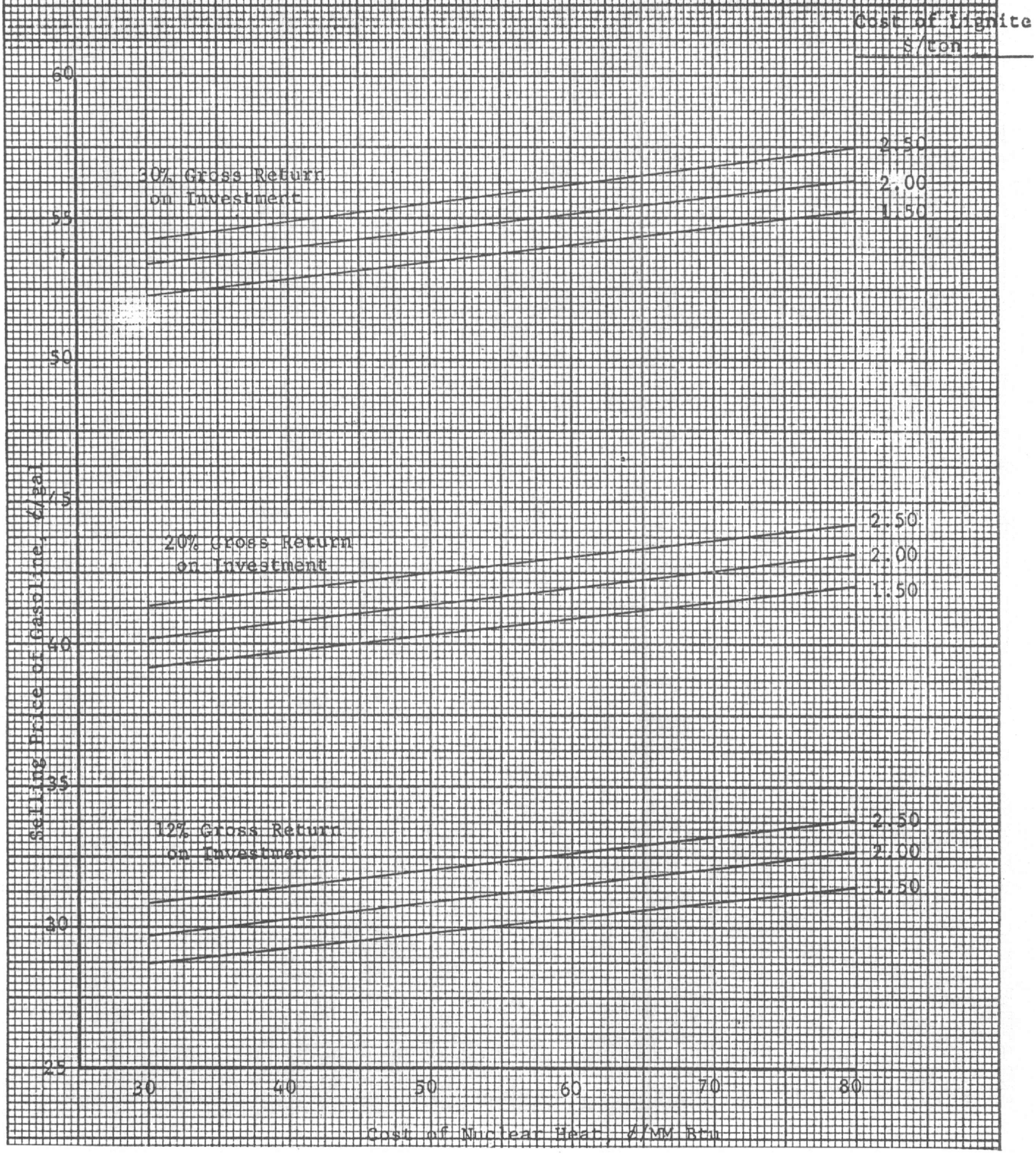
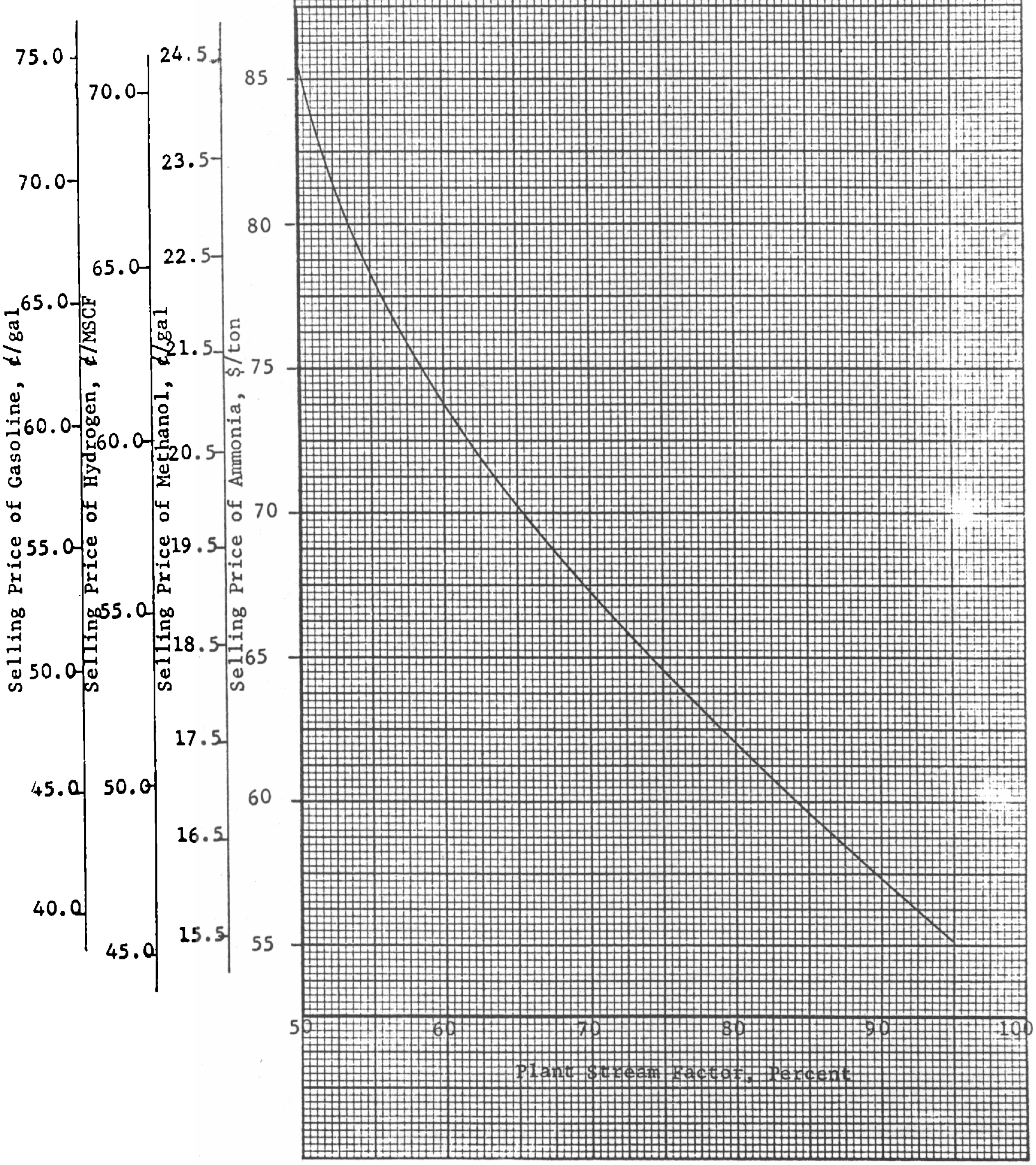


Figure L-9

Effect of Stream Efficiency on the Selling Prices of Ammonia, Methanol, Hydrogen, and Gasoline

Basis: 20% Gross Return on Investment
Lignite at \$2.00/ton
Nuclear Heat at 50¢/MM Btu



3. Design Considerations

In this section the design and stage of development of the various processing operations in the plant for production of fuels and chemicals from lignite will be discussed briefly. Those areas where additional experimental work is necessary to firm up the design will be pointed out.

a. Fluidized-Bed Drying

The design of the fluidized-bed dryer has been discussed under heading I-A-3-a of this report. Fluidized-bed drying of coal being a common operation, pilot plant demonstration of the operability of the fluidized-bed lignite dryer is probably not necessary to proper design of a commercial plant.

b. Fluidized-Bed Gasification

The design of the fluidized-bed gasifier has previously been discussed under heading I-A-3-b of this report. It should, however, be repeated here that pilot-plant study and demonstration of reaction kinetics, product gas composition, and materials of construction are necessary before a commercial fluidized-bed reactor may be designed with confidence.

c. Methane Removal by Partial Combustion

Methane is removed from the raw synthesis gas in Sections 300 and 400 by partial combustion with oxygen. The refractory-lined partial combustion furnaces operate at 2350°F, which is the minimum temperature yielding the proper reaction rates to approach equilibrium closely (20). The methane concentration in the partial combustor effluent, as predicted by water-gas shift and steam-methane reaction equilibrium, is essentially negligible.

The exothermic partial combustion reaction yields a temperature rise of approximately 350°F for the total gas stream, therefore the feed gas must be preheated to about 2000°F. This preheating duty is performed by the 2350°F effluent gas from the partial combustion furnace in a Hastelloy X heat exchanger. Hastelloy X has excellent strength and oxidation resistance up to a temperature of 2250°F, and it is expected that it will be able to withstand the conditions present in this heat exchanger. A long-term demonstration at these conditions is necessary, however, before the unit could be safely designed for commercial use.

d. Ammonia Production

As previously mentioned, the gas purification scheme used in the ammonia plant consists of an MEA scrubber to remove essentially all the CO_2 and H_2S, caustic wash to remove traces of these compounds, and activated carbon for COS removal. This processing sequence has been utilized in several commercial ammonia plants with completely satisfactory results. No further testing is required.

The activated carbon adsorption scheme has previously been discussed in Section I-A-3-d, and no further discussion need be given here.

Design of the methanator employed data available from commercial operation. Several of these units are operating successfully in commercial plants, and no further experimental work is necessary.

The ammonia synthesis scheme used in this design is the standard Kellogg ammonia process. This process represents well-developed technology, and Kellogg has designed and built a number of these units, which are at present in commercial operation.

e. Methanol Production

The gas purification sequence, which consists of the "Fluor Solvent CO_2 Removal Process" for CO_2 and bulk H_2S removal, followed by sponge iron and activated carbon for residual H_2S and COS removal, respectively, has already been discussed in section I-A of this report. The CO_2 concentration in the shifted synthesis gas is reduced to 1.0 mole percent and the sulfur content is reduced to approximately 0.004 grains per 100 standard cubic feet of gas.

Adequate technical data are available for the design of the methanol synthesis process and many of these units are in commercial operation.

f. Hydrogen Production

The process scheme chosen for production of hydrogen from methane-free raw synthesis gas is technically ready for commercial use. The hot potassium carbonate primary gas purification and MEA-caustic wash secondary gas purification units have been designed to reduce the product gas concentrations of carbon oxides and total sulfur to 10 and 1 ppm, respectively.

g. Liquid Fuels Production

The previously described Fluor Solvent, sponge iron, activated carbon gas purification sequence was also employed in the liquid fuels plant. The unit has been designed to reduce the CO_2 concentration in the shifted synthesis gas to below 5.0 mole percent and to reduce the total sulfur content to about 0.004 grains per 100 SCF of gas. Adequate technical data are available for commercial design of this purification scheme.

The world's largest oil-from-coal plant, built for the South African Coal Oil and Gas Corporation Ltd., served as a model for the design of the liquid fuels synthesis and recovery plant for this study. All units in this plant represent commercial installations, which have operated satisfactorily for a number of years.

4. Optimization of Gasification Process Variables

The gasifier operating conditions, which were optimized in conjunction with the design of the plant for production of pipeline gas from lignite and discussed under heading I-A-4 of this report, are as follows:

a. Steam feed/carbon converted mole ratio = 1.25/1.
b. Gasification temperature = 1600°F.
c. Inlet helium temperature = 2500°F.
d. Gasification and helium pressures = 450 psia.
e. Carbon utilization = 80 percent.

The ammonia and methanol synthesis units operate at relatively high pressures (around 5000 psia), therefore some consideration was given to operation of the gasifier at pressures above 450 psia. While this might have resulted in slightly more economical production of ammonia and methanol, the cost of producing hydrogen and liquid fuels would have risen. Since the high-pressure products account for only about 25 percent of total production, it was concluded that gasification at 450 psia would result in the most attractive overall economics. Moreover, operation at pressures much above 450 psia would involve a departure from existing technology for many of the process units. For the above reasons, the gasification pressure was again set at 450 psia.

5. Evaluation of Alternate Processing Schemes

a. Drying and Gasification

The reasons for selection of both fluidized-bed drying and fluidized-bed gasification have already been discussed under heading I-A-5 of this report.

b. Methane Removal

The following methods were considered for removal of methane from the raw synthesis gas:

(1) Catalytic partial combustion.
(2) Propane wash followed by catalytic partial combustion.
(3) Noncatalytic partial combustion.

The nickel reforming catalyst used for catalytic partial combustion is highly susceptible to poisoning by sulfur compounds, hence the H_2S and COS would have to be removed from the raw synthesis gas before method (1) could be employed. Catalytic partial combustion is carried out at 1600°F, while sulfur removal is carried out at about 100°F. Thus, the 1600°F raw synthesis gas from the gasifier would be cooled to 100°F, relieved of sulfur, and then reheated to 1600°F for removal of methane by partial combustion. Obviously, this is an expensive method of methane removal and was quickly eliminated from consideration.

A portion of the expense associated with method (1) may be eliminated by using a propane wash to absorb the methane from the bulk gas and then partially combusting the smaller, concentrated methane stream. Thus, only a portion of the total raw synthesis gas has to be reheated to the 1600°F partial combustion temperature if method (2) is used.

Method (3) involves no cooling of raw synthesis gas until after the methane is removed in a 2350°F partial combustion furnace and it was found to be the least expensive of the three methods considered. Consequently, methane removal by noncatalytic partial combustion was selected for use.

Appendix

Liquid Fuels and Chemicals from Lignite

Material Balances

Section 200

Input:	Lb/Hr
Lignite from Section 100 to 201-L	597,042
Steam to 201-C	171,948
Waste gas to 202-J	43,000
Cooling water to 203-F	500,000
	1,311,990

Output:	Lb/Hr
Raw synthesis gas from 202-D	662,271
Residue slurry from 203-F	606,719
Waste gas from 202-F	43,000
	1,311,990

Section 300

Input:	Lb/Hr
Raw synthesis gas from Section 200	554,108
Oxygen to partial combustion furnace	68,902
	623,010

Output:	Lb/Hr
Methane-free synthesis gas	623,010

Section 400

Input:	Lb/Hr
Raw synthesis gas from Section 200	108,163
Air to partial combustion furnace	57,340
	165,503

Output:	Lb/Hr
Methane-free synthesis gas to Section 500	165,503

Section 500

Input:	Lb/Hr
Methane-free synthesis gas from Section 400	165,503
Steam to shift converters	255,366
Boiler feed water to shift converters	34,524
Process water to water scrubber	65,000
	520,393

Output:	Lb/Hr
Shifted synthesis gas to Section 600	175,628
Condensate and waste water from scrubber	344,765
	520,393

Section 600

Input:	Lb/Hr
Shifted synthesis gas from Section 500	175,628
Regeneration steam to activated carbon drums	5,000
	180,628

Output:	Lb/Hr
Purified synthesis gas to Section 700	56,529
Waste gas from stripper	119,093
COS and steam from activated carbon drums	5,006
	180,628

Section 700

Input:	Lb/Hr
Purified synthesis gas from Section 600	56,529

Output:	Lb/Hr
CO-free synthesis gas to Section 800	56,529

Section 800

Input:	Lb/Hr
CO-free synthesis gas from Section 700	56,529

Output:	Lb/Hr
Compressed synthesis gas to Section 900	55,092.2
Condensate from knockout drums	1,436.8
	56,529

Section 900

Input:	Lb/Hr
Compressed synthesis gas from Section 800	55,092.2

Output:	Lb/Hr
Anhydrous liquid ammonia product	50,000.0
Purge gas from ammonia synthesis loop	5,092.2
	55,092.2

Section 1000

Input:	Lb/Hr
Methane-free synthesis gas to shift converter	67,878
Boiler feed water to shift converter	4,250
Process water to water scrubber	36,000
	108,128

Output:	Lb/Hr
Shifted synthesis gas to Section 1100	58,150
Condensate and waste water from scrubber	49,978
	108,128

Section 1100

Input:	Lb/Hr
Shifted synthesis gas from Section 1000	58,150
Regeneration steam to activated carbon drums	4,600
	62,750

Output:	Lb/Hr
Purified synthesis gas to Section 1200	29,817
Waste gas to atmosphere	28,259
H_2S from iron oxide drums	22
COS and steam from activated carbon drums	4,652
	62,750

Section 1200

Input:	Lb/Hr
Purified synthesis gas from Section 1100	29,817

Output:	Lb/Hr
Compressed synthesis gas to Section 1300	29,764
Condensate from knockout drums	53
	29,817

Section 1300

Input:	Lb/Hr
Compressed synthesis gas from Section 1200	29,764
Process water	1,460
	31,224

Output:	Lb/Hr
Pure liquid methanol product	25,000
Condensate	2,690
Waste gas	1,434
Purge gas	2,100
	31,224

Section 1400

Input:	Lb/Hr
Methane-free synthesis gas to shift converters	93,183
Steam to shift converters	34,500
Boiler feed water to shift converters	24,300
	151,983

Output:	Lb/Hr
Shifted synthesis gas to Section 1500	101,720
Condensate from knockout drums	50,263
	151,983

Section 1500

Input:	Lb/Hr
Shifted synthesis gas from Section 1400	101,720

Output:	Lb/Hr
Purified synthesis gas to Section 1600	18,742
Waste gas from stripper	82,978
	101,720

Section 1600

Input:	Lb/Hr
Purified synthesis gas to shift converters	18,742
Steam to shift converters	43,600
Quench water to shift converters	2,800
	65,142

Output:	Lb/Hr
Shifted synthesis gas to Section 1700	18,270
Condensate from knockout drums	46,872
	65,142

Section 1700

Input:	Lb/Hr
Shifted synthesis gas from Section 1600	18,270

Output:	Lb/Hr
Hydrogen product	8,557
Waste gas to atmosphere	9,663
Ammonia in waste solvent	50
	18,270

Section 1800

Input:	Lb/Hr
Methane-free synthesis gas to shift converters	461,949
Quench water to shift converters	31,000
Process water to water scrubbers	80,000
	572,949

Output:	Lb/Hr
Shifted synthesis gas to Section 1900	401,259
Condensate and waste water from scrubbers	171,690
	572,949

Section 1900

Input:	Lb/Hr
Shifted synthesis gas from Section 1800	401,259
Regeneration steam to activated carbon drums	28,000
	429,259

Output:	Lb/Hr
Purified synthesis gas to Section 2000	230,317
Waste gas from flash drums	170.242
H_2S from iron oxide drums	544
COS plus steam from activated carbon drums	28,156
	429,259

Section 2000

Input:	Lb/Hr
Purified synthesis gas from Section 1900	230,317
Steam	74,600
Water	1,920
Oxygen	36,300
Chemicals	400
	343,537

Output:	Lb/Hr
Gasoline	50,800
Diesel Oil	2,200
Waxy Oil	1,500
Coke	470
Carbon Dioxide	20,200
Propane LPG	3,600
Butane	2,600
Acetaldehyde	260
Acetone	860
MEK (methyl ethyl ketone)	290
Motor Ethanol	4,900
Propanol	1,200
Isobutanol	180
1-butanol	380
Heavy alcohols	127
Acetic Acid	2,000
Propionic Acid	450
Ethylene	5,700
Propylene	940
Waste Gas	26,700
Aqueous Waste	218,180
	343,537

Overall Energy Balance

Datum Temperature = 60°F

Input:	MM Btu/Hr	Percent of Total
Heating value of lignite feed to gasifier	4,300.0	76.3
Heating value of lignite feed to boilers	1,514.1	15.0
Heat transferred from helium	871.2	8.7
	6,685.3	100.0

Output:	MM Btu/Hr	Percent of Total
Heating value of gasoline product	1,025.0	15.3
Heating value of hydrogen product	470.0	7.0
Heating value of ammonia product	442.0	6.6
Heating value of methanol product	240.0	3.6
Heating value of miscellaneous fuel and chemical products	452.4	6.8
Losses to cooling water	2,756.0	41.2
Losses from offsite boilers	435.0	6.5
Latent heat of excess steam generated	150.0	2.2
Electric power consumption	210.0	3.2
Sensible and latent heat of waste gas streams	45.0	0.7
Sensible heat of aqueous waste streams	70.0	1.0
Latent heat of regeneration steam for activated carbon drums	40.0	0.6
Convection losses, miscellaneous, etc.	349.9	5.3
	6,685.3	100.0

Thermal Efficiency, (percent of total heat input appearing in combined product streams) = 39.3%

Utilities Summary

Steam

I. 1000 psig, 850°F

A. Generation

Steam is generated in Section 3000 at 1000 psig, 850°F in two boilers capable of generating a total of 1,800,000 pounds per hour. Since normal consumption is only 1,663,000 pounds per hour, there is about 8 percent excess capacity available if needed.

Fuel for the boilers consists of 190,000 pounds per hour of lignite from Section 000 and 106,719 pounds per hour of gasifier residue from the settling ponds.

Section	Normal Generation, Lb/Hr
3000	1,800,000

B. Consumption

Section	Normal Consumption, Lb/Hr
200 - compressor drives	89,000
400	49,000
800	105,000
1100	300
1200	46,700
1600	9,600
1900	2,500
2000	402,500
3000	958,466
Available	136,934
	1,800,000

II. 500 psig, 550°F

A. Generation

Steam at 500 psig, 550°F is obtained as steam turbine exhaust in Section 3000 at the rate of 408,466 pounds per hour.

Section	Normal Generation, Lb/Hr
3000	408,466

B. Consumption

Section	Normal Consumption, Lb/Hr
500	255,366
1400	34,500
1600	43,600
2000	75,000
	408,466

III. 500 psig, saturated

A. Generation

Section	Normal Generation, Lb/Hr
400	90,000
500	30,000
1000	10,700
1800	41,248
	171,948

B. Consumption

Section	Normal Consumption, Lb/Hr
200 - to 201-C	171,948

IV. 300 psig, saturated

A. Generation

Section	Normal Generation, Lb/Hr
1800	21,500

B. Consumption

Section	Normal Consumption, Lb/Hr
600	8,500
1500	13,000
	21,500

V. 40 psig, saturated

A. Generation

Section	Normal Generation, Lb/Hr
300	113,000
500	45,000
1000	8,100
1600	44,000
1800	54,000
3000	73,600
	337,700

B. Consumption

Section	Normal Consumption, Lb/Hr
600	3,000
1100	4,600
1300	32,500
1500	60,000
1700	22,600
1900	28,000
2000	97,000
3000	90,000
	337,700

Power

A. Generation

Electric power is produced at 13,800 volts in Section 3000 by turbogenerators using both condensing and noncondensing steam turbine drives. About 10 percent excess capacity is available during normal operation. An electric substation reduces the voltage to 4160, 440, and 110.

Section	Normal Generation	
	HP	Kw
3000	85,238	63,427

B. Consumption

Section	Normal Consumption	
	HP	Kw
000	1,000	745
100	1,040	775
200	50	37
300	20,900	15,550
400	7,200	5,350
600	67	50
800	11,040	8,220
900	14,900	11,060
1100	403	300
1200	4,950	3,690
1300	1,500	1,120
1500	4	3
1700	134	100
1900	2,580	1,927
2000	14,100	10,500
3000	5,370	4,000
	85,238	63,427

Cooling Water

A. Generation

It is anticipated that the plant will be located near a river which may prove to be inadequate as a complete heat sink. Accordingly, the plant cooling system is based on tower cooling of recirculated water, with the necessary makeup water obtained from the river.

Cooling water is available throughout the plant at 40 psig and a maximum temperature of 85°F.

Section	GPM
3000	240,096

B. Consumption

Section	GPM
200	16,100
300	1,000
400	8,100
500	425
600	10,000
700	620
800	21,000
900	2,840
1000	108
1100	250
1200	9,318
1300	4,700
1500	9,080
1600	125
1700	3,240
1800	740
1900	2,450
2000	50,000
3000	100,000
	240,096

C. Summary for Section 200

Item	GPM
205-C	232
203-F	1,000
Surface Condensers	14,868
	16,100

Boiler Feed Water

A. Generation

Boiler feed water at temperature levels of 250°F and 350°F is produced by the following sequence:

(1) River water is filtered and treated with chemicals to remove dissolved solids.

(2) Further purification is effected by passing the treated water over anion- and cation-exchange resins.

(3) Purified makeup water is mixed with return condensate and heated to about 180°F.

(4) The water is deaerated with low-pressure steam.

(5) Deaerated water is heated to temperature levels of 250° and 350°F in the desired quantities.

Section	Normal Generation, Lb/Hr
3000	2,476,202

B. Consumption

Section	Normal Consumption, Lb/Hr
300	116,000
400	92,700
500	112,024
1000	23,600
1400	24,300
1600	2,800
1700	3,730
1800	151,048
3000	1,950,000
	2,476,202

Process Water

A. Generation

Process water is produced by treating filtered river water to precipitate dissolved solids.

Section	Normal Generation, Lb/Hr
3000	184,380

B. Consumption

Section	Normal Consumption, Lb/Hr
500	65,000
1000	36,000
1300	1,460
1800	80,000
2000	1,920
	184,380

Fuels

A. Generation

Section	Item	MM Btu/Hr	Lb/Hr
000	Raw lignite	1,514	210,000
200	Gasifier residue	615	106,719
900	Purge gas	57	5,092
1300	Purge gas	35	2,100
2000	Purge gas	11	26,700
		2,232	350,611

B. Consumption

Section	Item	MM Btu/Hr	Lb/Hr
3000		2,232	350,611

II. Bituminous Coal

A. Production of Pipeline Gas

1. Description of Process

For convenience, the pipeline gas-from-coal plant has been divided into functional sections, as follows:

Section 000 - Coal Storage and Reclamation
Section 100 - Coal Grinding
Section 150 - Coal Pretreatment
Section 200 - Gasification and Dust Removal
Section 300 - Shift Conversion and Heat Recovery
Section 400 - Gas Purification
Section 500 - Methane Synthesis and Gas Compression
Section 1100 - Offsite Facilities

A process stream balance, showing the above breakdown according to plant section, is presented as Drawing No. CE-1183-B.

a. Section 000 - Coal Storage and Reclamation

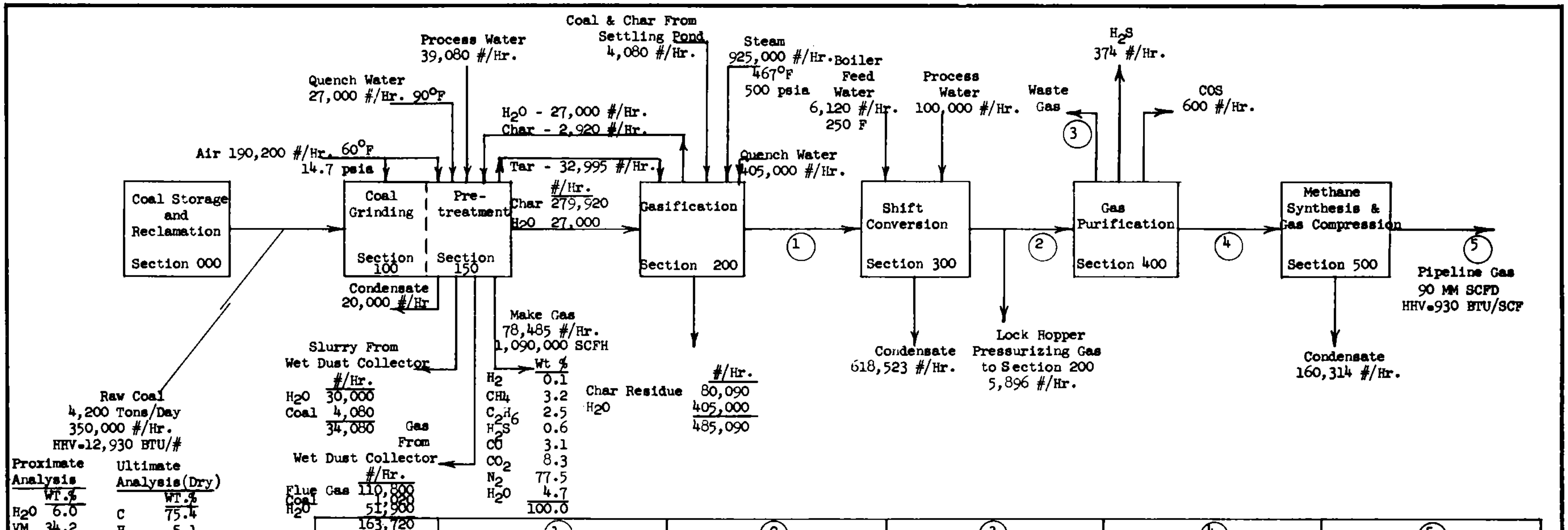
During eight hours each day, coal is received by truck or conveyor belt from an adjacent coal mine at the rate of 535 tons per hour. The 4" x 0 coal is conveyed to a coal distributing center, where about 175 tons per hour is dispatched for immediate use and the remainder is conveyed to the storage area.

Coal is distributed to several storage piles by a traveling stacker. These storage piles contain enough coal to permit the plant to operate for 30 days at normal capacity in the event the coal supply is cut off.

During the 16 hours each day that the mine is not operating, coal is reclaimed from storage by gravity flow into two underground tunnels onto conveyor belts. The equipment in Section 000 has been spared in such a manner as to permit continuous operation at full capacity.

b. Section 100 - Coal Grinding

Coal from Section 000 is conveyed at the rate of 350,000 pounds per hour by belt conveyor to a coal bunker. The coal contains about 6.0 percent moisture and is in the size range 4" x 0. Coal flows by gravity to crushers where most of it is reduced in size to minus 1/4 inch. Particles larger than 1/4 inch are separated by screens and recycled by conveyors and an elevator back to the coal bunker. Particles smaller than 1/4 inch pass through the screens and are transported to the coal dryer feed bins.



Proximate Analysis

	WT.%
H_2O	6.0
VM	34.2
FC	51.1
Ash	8.7
	100.0

Ultimate Analysis (Dry)

	WT.%
C	75.4
H	5.1
O	5.7
N	1.4
S	3.2
Ash	9.2
	100.0

Steam No.	①			②			③			④			⑤		
Temperature, °F,	700			100			---			80			100		
Pressure, PSIA	440			420			---			400			1000		
Flow Rate	#/Hr.	MPH	Mol.%(Dry)	#/Hr.	MPH	Mol.%(Dry)	#/Hr.	MPH	Mol.%(Dry)	#/Hr.	MPH	Mol.%(Dry)	#/Hr.	MPH	Mol.%(Dry)
CO2	289,000	6,571	15.14	327,000	7,443	16.95	311,000	7,080	92.71	16,000	363	1.00	8,500	193	1.95
CO	269,000	9,620	22.16	242,530	8,676	19.77	3,530	126	1.65	239,000	8,550	23.59	14	0.5	---
H_2	53,080	26,540	61.15	54,558	27,279	62.14	258	129	1.69	54,300	27,150	74.92	1,600	800	8.09
CH_4	2,460	154	0.35	2,447	153	0.35	192	12	0.16	2,255	141	0.39	141,775	8,861	89.58
N_2	1,065	38	0.09	1,065	38	0.09	---	---	---	1,065	38	0.10	1,065	38	0.38
H_2S	9,500	279	0.64	9,624	283	0.64	9,250	272	3.56	---	---	---	---	---	---
COS	1,920	32	0.07	1,680	28	0.06	1,080	18	0.23	---	---	---	---	---	---
NH_3	2,960	174	0.40	---	---	---	---	---	---	---	---	---	---	---	---
Dry Gas	628,985	43,403	100.00	638,904	43,900	100.00	325,310	7,637	100.00	312,620	36,242	100.00	152,954	9,892	100.00
H_2O	530,000	29,460		1,782	99		972	54		810	45		162	9	
Total Gas	1,158,985	72,868		640,686	43,999		326,282	7,691		313,430	36,287		153,116	9,901	

NO.	REVISION DESCRIPTION	DATE	BY	CHK.

ISSUED FOR FABRICATION

ISSUED FOR CONSTRUCTION

SCALE:
DRAWN: LJP
CHECKED:
APPROVED:
DATE: 9-12-62

THE M. W. KELLOGG COMPANY

PIPELINE GAS FROM BITUMINOUS COAL
PROCESS STREAM BALANCE

CLASS	AREA	JOB NO.	DRAWING NO.
		4006	CE-1183-B

Coal is introduced by screw feeders into drying columns where hot flue gases generated in dryer furnaces pick up the coal fines and convey them to dryer cyclones, meanwhile heating the coal to about 190°F and evaporating much of its contained moisture. Flue gases and water vapor are separated from the coal particles in cyclones at about 270°F and atmospheric pressure. The gases flow through secondary dust collectors for further recovery of coal fines and finally to a wet dust collector. Flue gas from this collector is vented to a stack and recovered dust is pumped as a water slurry to the plant settling pond.

Crushed coal, now at 190°F and containing about 1.5 percent moisture, flows by gravity from the dryer cyclones through hammer mill feeders to the hammer mills, where it is ground so that about 93 percent will pass through a 14-mesh screen (Tyler), and about 22 percent through a 200-mesh screen. Ground coal is carried by conveyor belt and elevator to the rod mill feed bin. From here coal flows by gravity to the rod mill screw feeders and the rod mills. Rod mill product, now about 98 percent minus 14 mesh and 30 percent minus 200 mesh, and still containing about 1.5 percent moisture, is carried by a conveyor belt and an elevator to Section 150.

Continuous operation of the plant is assured by judicious sparing of equipment in Section 100. Items such as elevators, crushers, hammer mills, and rod mills, which might require frequent maintenance, are provided with complete spares.

c. Section 150 - Coal Pretreatment

The purpose of this section is to destroy the caking tendencies of the bituminous coal so that it will not agglomerate in the gasifier.

Coal from Section 100 is conveyed at the rate of 332,500 pounds per hour to a coal surge drum where it is fluidized with air. Overhead gas, containing some entrained coal fines, joins hot flue gases from the coal dryers and flows through the dust recovery system. Fine coal is withdrawn from the surge drum in the fluidized state and conveyed with air from air compressors to the pretreater. Here coal is partially oxidized in a fluidized bed at 760°F. This treatment is designed to permit subsequent carbonization at 890°F in the carbonizer without agglomerating difficulties. A relatively severe preoxidation (to "fix" the carbon in the solid) and mild carbonization (to minimize gas and liquid yields) are employed.

Carbonizer overhead gas passes through 3-stage cyclones to a quench tower, where it is cooled to about 160°F. Condensed tar and liquor (water containing a variety of dissolved carbonization products) are separated in a tar decanter. The heavy tar from the bottom of the quench tower is cooled to about 120°F; some of it is sprayed into the tower as quench while the remainder is added to the overhead liquor, resulting in a tar product of 32,995 pounds per hour. This tar is then sent to Section 200 as part of the feed to the gasifier.

A gas having a net heating value of about 113 Btu/SCF is produced as another by-product of carbonization. About 22 percent of this gas is burned in the dryer furnaces in Section 100; the rest is vented to a stack.

Char at 890°F leaves the carbonizer via a standpipe and is quenched to 500°F with process water at 90°F. The resultant steam-char mixture then flows to Section 200.

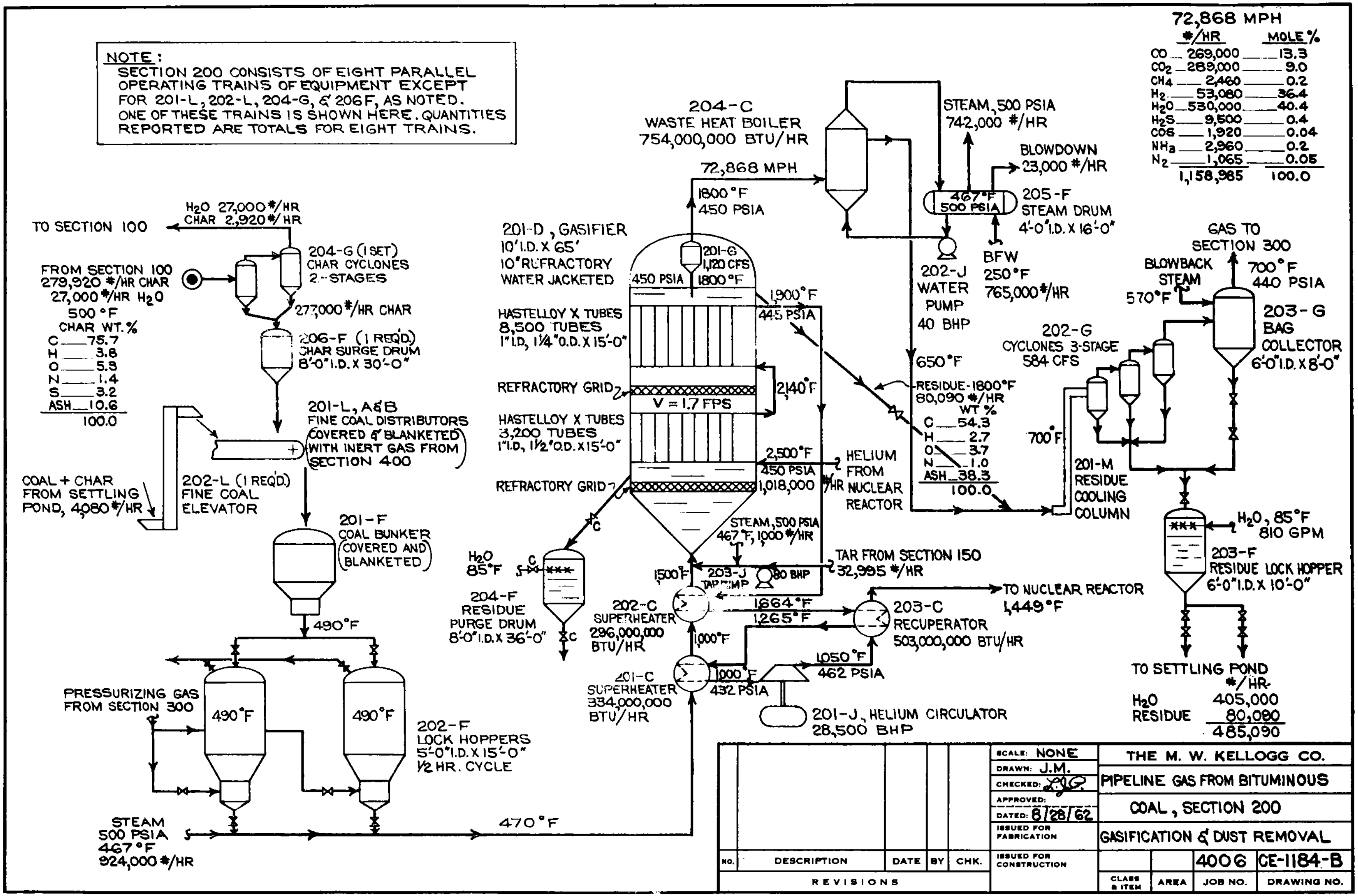
d. Section 200 - Gasification and Dust Removal

Section 200 is the focal point of this study. Here char (pretreated coal) and steam are contacted in a fluidized bed to produce a raw synthesis gas to be used to make high-Btu pipeline gas. Heat for the endothermic char-steam reaction is supplied by a stream of helium at 2500°F, heated to that temperature in a nuclear reactor. To aid the reader in understanding the ensuing description of Section 200, reference is made to the process flowsheet, Drawing No. CE-1184-B. Because of the large volume of gas produced, Section 200 is subdivided into eight parallel operating units. All flow rates and duties shown on Drawing No. CE-1184-B and mentioned in this description are total quantities for the eight units.

Char (pretreated coal) from Sections 100-150 is separated from conveying steam by 2-stage cyclone 204-G. The steam containing some entrained fines is returned to the dust recovery system in Section 100. The char flows by gravity at the rate of 277,000 pounds per hour to a char surge drum, 206-F. From here it flows by gravity to fine coal distributor 201-L, which is covered and blanketed with inert gas from Section 400 to prevent ignition of the hot char particles upon contact with air. Dust recovered in Section 100 and dried in the settling pond is fed at the rate of 4,080 pounds per hour by elevator 202-L to coal distributor 201-L, where it is combined with the fresh char and distributed among eight parallel operating units. The char now flows into coal bunker 201-F, which is also covered and blanketed with inert gas.

Char from 201-F flows by gravity to a set of lock hoppers, 202-F, whose purpose is to receive the char at atmospheric pressure and to deliver it to the process at the operating pressure of about 450 psia. Each of the lock hoppers operates on a 30-minute cycle comprising the following steps:

(1) filling with char at atmospheric pressure
(2) pressurizing to 500 psia with shift effluent gas from Section 300
(3) discharging the char into the steam line
(4) de-pressurizing.



NOTE:
SECTION 200 CONSISTS OF EIGHT PARALLEL OPERATING TRAINS OF EQUIPMENT EXCEPT FOR 201-L, 202-L, 204-G, & 206F, AS NOTED. ONE OF THESE TRAINS IS SHOWN HERE. QUANTITIES REPORTED ARE TOTALS FOR EIGHT TRAINS.

72,868 MPH

	#/HR	MOLE %
CO	269,000	13.3
CO_2	289,000	9.0
CH_4	2,460	0.2
H_2	53,080	36.4
H_2O	530,000	40.4
H_2S	9,500	0.4
COS	1,920	0.04
NH_3	2,960	0.2
N_2	1,065	0.05
	1,158,985	100.0

NO.	DESCRIPTION	DATE	BY	CHK.
	REVISIONS			

SCALE: NONE
DRAWN: J.M.
CHECKED: [illegible]
APPROVED:
DATED: 8/28/62
ISSUED FOR FABRICATION
ISSUED FOR CONSTRUCTION

THE M. W. KELLOGG CO.
PIPELINE GAS FROM BITUMINOUS COAL, SECTION 200
GASIFICATION & DUST REMOVAL

CLASS & ITEM	AREA	JOB NO.	DRAWING NO.
		4006	CE-1184-B

The lock hopper system is designed so that while one hopper is being filled and pressurized, the other is discharging char into the process stream. This provides a continuous flow of char to the gasifier. The gas released when 202-F is de-pressurized is vented to the atmosphere.

Char from 202-F flows by gravity at the rate of 281,080 pounds per hour into the steam line, where it is mixed with 924,000 pounds per hour of 500 psia, saturated steam, generated in waste heat boiler 204-C and in waste heat boilers in Section 500. This char-steam mixture is superheated to 1500°F against hot helium in superheaters 201-C and 202-C. Tar from Section 150 is pumped up to gasification pressure in tar pump 203-J, is atomized with 1000 pounds per hour of 500 psia, saturated steam, and is fed into the transfer line with the char and steam. It is felt that agglomeration can be avoided by introducing this tar into the dilute feed stream (about 0.3 pound solids per pound steam) traveling at about 30 feet per second.

The gasifier, 201-D, is a vertical cylindrical vessel with an inside diameter of 10 feet and a length of 65 feet. Normal operating conditions are 1800°F, 450 psia. The 2-inch-thick, low-alloy steel shell has a 10-inch internal refractory lining. For further protection the entire vessel is enclosed in an atmospheric pressure water jacket. A fluidized bed about 50 feet deep is maintained in the gasifier. The bed is divided into two zones by horizontal refractory grids in order to obtain uniform gas distribution and to minimize gas by-passing and short-circuiting of solids in the bed. Superficial gas velocity varies from about 1.3 feet per second at the bottom of the gasifier to about 2.2 feet per second at the top. Bed density will differ in the two zones because of variation in gas velocity, but it is expected to be in the range of 18 to 24 pounds per cubic foot. Heat for the endothermic reaction is supplied by circulating 2500°F helium from a nuclear reactor through tubes immersed in the fluidized bed. This tubing is designed in two bundles. The bottom bundle consists of a total of 3200 1-inch I.D. x 1-1/2-inch O.D. x 15-feet long Hastelloy alloy X tubes, while the top bundle consists of a total of 8,500 1-inch I.D. x 1-1/4-inch O.D. x 15-feet long Hastelloy X tubes.

As the char-steam-tar mixture enters the gasifier, its temperature is raised almost instantaneously to the 1800°F reaction temperature. The principal reactions which occur are the decomposition of tar and of volatile matter still present in the char (reaction II-1), reaction of steam and carbon to form hydrogen and carbon monoxide (reaction II-2), and the water-gas shift reaction (reaction II-3).

$$C_xH_y \longrightarrow C_n H_m + C + H_2 \qquad \text{(II-1)}$$

$$C + H_2O \rightleftharpoons CO + H_2 \qquad \text{(II-2)}$$

$$CO + H_2O \rightleftharpoons CO_2 + H_2 \qquad \text{(II-3)}$$

Char flows upward in the fluidized state from the lower bed to the upper bed. Because the grid is designed to yield relatively high gas velocity through its perforations, there is expected to be little, if any, backmixing of solids from the upper bed to the lower one. Average particle residence time in the gasifier is about 7-1/2 hours.

Gasifier effluent at 1800°F and 450 psia flows through single-stage cyclone 201-G. This gas is cooled to 650°F in waste heat boiler 204-C to generate 742,000 pounds per hour of saturated steam at 500 psia for use as feed to the gasifier. Char residue at 1800°F, which is continuously withdrawn from the top of the gasifier at the rate of 80,090 pounds per hour, is mixed with the raw synthesis gas in residue cooling column 201-M, raising the gas temperature to 700°F. The residue-gas mixture is fed to 3-stage cyclone 202-G, in which about 99.9 percent of the entrained solids are removed. Overhead gas is then fed to bag collector 203-G, where the remainder of the solids are separated from the gas. The dust is blown off the bags with steam and is combined with the dust removed in 202-G. The residue flows by gravity into residue lock hopper 203-F, where it is quenched to 150°F with cooling water at 85°F. This lock hopper is designed to operate continuously at about 440 psia. This is done by keeping the vessel about three-fourths full of water at all times while purging the residue slurry through reducing valves to a settling pond.

Clean gas from 203-G flows to section 300 at the rate of 72,868 moles per hour.

A residue purge hopper, 204-F, is provided for two reasons:

(1) To permit the removal of agglomerates which may form in the gasifier and become too large to pass through the grid perforations.

(2) To permit complete removal of solids from the gasifier when the vessel is taken off-stream.

The solids are quenched in cold water as they flow into the pressurized hopper, and are then discharged at atmospheric pressure to a settling pond.

The heat supply for the gasifiers, a stream of hot helium at about 450 psia, is heated to 2500°F in a nuclear reactor where it is assumed to have picked up negligible radioactivity. It enters the tubes in the bottom of the gasifier at the rate of 1,018,000 pounds per hour and flows upward to the top bundle of tubes, which it enters at about 2140°F. It leaves the top of the gasifier at 1900°F and is cooled to 1664°F in superheater 202-C against the feed stream. It is then cooled to 1265°F in recuperator 203-C against itself, and is further cooled to 1000°F in superheater 201-C against the feed stream. The helium is next compressed to 462 psia by turbine-driven helium circulator 201-J in order to overcome pressure losses in the circuit. Effluent helium at 1050°F flows to recuperator 203-C, where it is reheated to 1449°F. This stream is returned to the nuclear reactor, reheated to 2500°F, and recirculated to the gasifier.

e. Section 300 - Shift Conversion and Heat Recovery

Section 300 consists of three parallel trains of operating equipment. All flow rates mentioned in this description are total quantities for the three trains. The purpose of this section is to adjust the H_2/CO ratio to approximately 3/1 in preparation for synthesis of methane.

Clean gas from Section 200 at 700°F and 440 psia enters Section 300 at the rate of 72,868 moles per hour. Of this, 10,570 moles per hour are fed to a shift converter where reaction II-3 occurs.

$$CO + H_2O \rightleftharpoons CO_2 + H_2 \qquad \text{(II-3)}$$

The reaction is mildly exothermic; therefore, boiler feed water at 250°F is injected between catalyst beds in the converter to absorb the bulk of the heat liberated. Shift effluent at 780°F is mixed with the 62,298 moles per hour of by-passed gas. About 350 moles per hour of the gas mixture is compressed to 500 psia and is sent to lock hoppers 202-F as pressurizing gas. The remainder of the mixed gas, now at 710°F, is cooled to 650°F in a steam superheater where 459,200 pounds per hour of 500 psia, saturated steam is superheated to 570°F to run turbine drives throughout the plant. The gas is further cooled to 345°F in three exchangers where the recovered heat is used to generate low-pressure steam (45 psia), to heat deaerated boiler feed water from 214° to 250°F, and to heat condensate from 95° to 200°F. The gas is finally cooled to 100°F against cooling water.

The cooled synthesis gas plus entrained condensate enter a water scrubber where the gas is countercurrently scrubbed with clean water to remove trace amounts of ammonia which might be present.

Overhead gas from the water scrubber, now having a H_2/CO ratio of 3.15/1, flows to Section 400.

f. Section 400 - Gas Purification

The gas purification section is designed to reduce the CO_2 concentration in the synthesis gas to 1.0 mole percent, and to reduce the total sulfur concentration to about 0.004 grains/100 SCF of gas. The purification sequence consists of the "Fluor Solvent CO_2 Removal Process" for CO_2 and bulk H_2S removal, followed by sponge iron (iron oxide) and activated carbon for residual H_2S and organic sulfur removal, respectively. All organic sulfur is assumed to be carbonyl sulfide (COS).

The Fluor system is designed as three parallel operating trains. The iron oxide drums are also arranged in three parallel trains, each train consisting of seven parallel drums followed by one guard chamber. The activated carbon drums are arranged in six parallel trains, each train consisting of three drums which are manifolded for cyclic operation. A more detailed description of this gas purification sequence may be found in Section I-A-1-e of this report.

Purified synthesis gas, now containing about 1.0 mole percent CO_2 and about 0.004 grains total sulfur/100 SCF, flows to Section 500.

g. Section 500 - Methane Synthesis and Gas Compression

Section 500 is designed as three parallel operating units. Methane is formed from hydrogen and carbon monoxide, according to reaction (II-4), in a Kellogg transport reactor.

$$CO + 3\ H_2 \rightleftharpoons CH_4 + H_2O + 92{,}900\ \text{Btu/hr} \qquad \text{(II-4)}$$

Heat of reaction is removed by generating saturated steam at 500 psia. Methanator effluent gas is assumed to be at equilibrium at 650°F, a 30°F approach to the actual effluent temperature of 620°F. The result of this assumption is a conversion of about 97.8 percent of the incoming H_2 + CO and a product gas with a heating value of 930 Btu/SCF.

Product gas, which is about 89.6 percent methane, flows at 100°F and 375 psia to the product gas compressors.

The turbine-driven centrifugal product gas compressors are designed as two parallel operating machines, each capable of carrying 65 percent of total plant capacity. The gas is compressed to 1005 psia and cooled to condense water. Product gas at 100°F and 1000 psia flows to the gas mains at the rate of 90,000,000 SCFD.

h. Section 1100 - Offsite Facilities

Section 1100 includes facilities for:

(1) generating steam and electric power
(2) supplying cooling water, process water, and boiler feed water
(3) providing miscellaneous services necessary to make this a completely self-sufficient plant.

Steam generation facilities consist solely of a start-up boiler capable of producing 450,000 pounds per hour of 500 psia, 570°F steam. Once the plant is in full operation, enough steam is generated by waste heat in the process to provide all the high- and low-pressure steam needed.

Electric power is produced at 13,800 volts by turbogenerators using condensing steam turbine drives. An electric substation is provided to reduce the voltage to 4160, 440 and 110 volts.

Brief descriptions of the means of supplying cooling water, boiler feed water, and process water are included in the Utilities Summary in the Appendix.

2. Economics

a. Economic Summary

The cost of producing 90,000,000 SCFD of pipeline gas from bituminous coal according to the process sequence just described is calculated in Tables II-1 to II-4, assuming 90 percent stream efficiency.

Estimated operating labor for the complete plant, including offsite facilities, is 39 men per shift, as shown in Table II-1.

Estimated capital investment for the plant is summarized in Table II-2. Detailed costs for Section 200, Gasification, are presented in the Appendix. Interest during construction is calculated at 5.3 percent of the sum of total bare cost plus contractor's overhead and profit, as discussed in Section I-A-2. Working capital includes 30 days' coal inventory, 30 days' accounts receivable, and in-plant inventories of sponge iron and methanation catalyst. Shift catalyst and activated carbon are included in fixed investment because they have very long lifetimes. Total capital investment is about 70 million dollars, as compared with 45 million for pipeline gas from lignite. The chief difference is the added cost of the gasification section due to the lower reactivity of bituminous coal.

Estimated annual operating costs are tabulated in Table II-3. Nuclear heat is charged at 50¢/MM Btu and coal at $5 per ton. Makeup of Raney nickel methane synthesis catalyst constitutes more than 85 percent of the charge for catalysts and chemicals, or about 4.4¢/MSCF of product gas. Gas production cost, assuming 20-year straight-line depreciation and before any return whatever on invested capital, is 85¢/MSCF. Of the total, about 27 percent is contributed by coal, 22 percent by nuclear heat, and 21 percent by fixed costs. The gas production cost of 85¢/MSCF compares with 56¢/MSCF for pipeline gas from lignite, and is a reflection of the greater cost and lower reactivity of the raw material.

The effect on gas cost of adding gross return on investment is shown in Table II-4 and Figure II-1. If a gross return of 20 percent is desired, for example, about 47¢/MSCF must be added to the gas production cost. The effects of variations in the cost of coal or nuclear heat are also shown. An increase of 20 percent in the cost of coal or nuclear heat above the values used in Table II-3 results in an increase in gas cost of about 5¢/MSCF.

The effect of stream efficiency on gas selling price, assuming a 20 percent gross return on investment, is shown in Figure II-2.

b. Temperature Level of Nuclear Heat

The preceding figures are based on helium being supplied to the process at 2500°F by a nuclear reactor. Reactor investment is not included in Table II-2. Instead, all costs associated with the nuclear side of the plant must be covered by the cost of nuclear heat shown in Tables II-3 and II-4, including ordinary operating costs, return on investment, and the special costs associated with nuclear operation.

If helium, the reactor coolant, is supplied to the process at a temperature below 2500°F, the cost of the nuclear reactor can reasonably be expected to decline. On the other hand, cost of the gasification plant will increase because of the lower temperature driving force, larger heat transfer surface requirement, and larger helium circulator. Figure II-3 shows the cost of nuclear heat which, if helium is supplied at some temperature other than 2500°F, will yield the same gas selling price as calculated in Table II-4. The curve demonstrates that, for gasification of a high-volatile bituminous coal, nuclear reactor coolant should probably be heated to at least 2500°F.

c. Effect of Plant Capacity on Product Gas Cost

Statements made in section I-A-2-c relating to pipeline gas from lignite also apply here. In this case, however, the nuclear reactor output equivalent to 90,000,000 SCFD of pipeline gas is 406 thermal megawatts.

d. Effect of Percent Coal Gasified on Product Gas Cost

The plant has been designed to gasify 80 percent of the coal feed. Since all steam needed by the plant can be generated internally from waste heat, there is no use for the ungasified residue and it must be discarded. If the plant were designed to gasify 90 percent of the coal, larger gasifiers would be needed and plant cost would be increased by about 4 million dollars. Less coal would be required, but the savings would not offset the increased investment, as demonstrated in section II-A-4-f. The optimum point is somewhere between 80 and 90 percent gasified, but operating at the optimum would reduce gas cost by only about 1¢/MSCF from the figures presented in Tables II-3 and II-4.

e. Effect on Product Gas Cost of Eliminating Pretreatment

The approach taken toward pretreatment has been a conservative one; that is, the pretreatment is perhaps more severe and the process perhaps more costly than would be necessary. The optimistic approach would be to assume that no pretreatment at all is needed, that the caking bituminous coal could somehow be gasified with steam in a fluidized bed without agglomerating. If this were possible, gas production cost would be reduced by about 3¢/MSCF.

Table II-1

Estimated Operating Labor
Pipeline Gas from Bituminous Coal

Basis: 90,000,000 SCFD of Pipeline Gas
90% Stream Efficiency

Section	Title	Men per Shift
000	Coal Storage and Reclamation	1
100	Coal Grinding	5
150	Pretreatment	2
200	Gasification	8
300	Shift Conversion	1
400	Gas Purification	7
500	Methane Synthesis and Gas Compression	8
1100	Offsite Facilities:	
	Power Plant	3
	Cooling Water Pumps	1
	Makeup Water Pumps	1
	Feedwater Treating System	2
	TOTAL OPERATING LABOR, men/shift	39
	man-hours/day	936

Table II-2

Investment Summary
Pipeline Gas from Bituminous Coal

Basis: 90,000,000 SCFD of Pipeline Gas
90% Stream Efficiency

Section	Title	Material and Freight	Bare Cost*
000	Coal Storage and Reclamation	$ 870,000	$ 1,500,000
100	Coal Grinding	2,213,500	3,593,000
150	Pretreatment	1,177,300	2,573,000
200	Gasification	15,101,000	23,320,000
300	Shift Conversion	2,375,000	3,465,000
400	Gas Purification	7,282,180	9,495,680
500	Methane Synthesis and Gas Compression	2,342,250	3,494,200
1100	Offsite Facilities	6,285,500	8,857,000
	Total Material and Freight	$37,646,730	
	Total Bare Cost		$56,297,880
	Contractor's Overhead and Profit		5,910,000
	Interest during Construction at 5.3%		3,300,000
	TOTAL FIXED INVESTMENT		$65,507,880

Working Capital:

30 days' coal inventory (126,000 tons at $5/ton)	$ 630,000	
Accounts receivable (value of 30 days' production at $1.30/MSCF)	3,510,000	
Catalyst Inventory	340,000	
Total working capital		$ 4,480,000
TOTAL CAPITAL INVESTMENT		$69,987,880

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* Bare cost includes materials, freight, construction labor, field administration and supervision, insurance during construction, cost of tools, field office expense, and cost of home office engineering and procurement.

Table II-3

Estimated Annual Operating Cost
Pipeline Gas from Bituminous Coal

Basis: 90,000,000 SCFD of Pipeline Gas
90% Stream Efficiency

Item	$/Year		¢/MSCF (2)
Direct Costs:			
Nuclear heat (1,385 MMBtu/hr x 7,884 hr/yr x 50¢/MMBtu)	5,460,000		18.5
Bituminous coal to gasifiers (175 TPH x 7,884 hr/yr x $5/ton)	6,900,000		23.3
Operating labor (936 man-hours/day x $2.80/man-hr x 365 days/yr)	956,000		3.2
Supervision at 15% of operating labor	143,000		0.5
Catalysts and chemicals	1,510,000		5.1
Maintenance at 4% of bare cost per year (1)	2,250,000		7.6
Supplies at 15% of maintenance	338,000		1.1
Total Direct Costs:		17,557,000	
Indirect Costs:			
Payroll overhead at 20% of operating labor, maintenance labor, and supervision	535,000		1.8
General plant overhead at 50% of operating labor, supervision, maintenance, and supplies	1,844,000		6.2
Total Indirect Costs		2,379,000	
Fixed Costs:			
Depreciation at 5% of total fixed investment	3,280,000		11.1
Property taxes and insurance at 3% of total fixed investment	1,965,000		6.6
Total Fixed Costs		5,245,000	
TOTAL ANNUAL OPERATING COST		$25,181,000	
ESTIMATED GAS PRODUCTION COST			85.0

---

(1) Maintenance is 70% labor, 30% material
(2) Cents per thousand standard cubic feet of gas

Table II-4

Pipeline Gas from Bituminous Coal

Effect of Coal Cost, Cost of Nuclear Heat, and Return on Investment on the Selling Price of Gas

A. Gross return on investment = 12%

Cost of nuclear heat, ¢/MMBtu	Gas selling price, ¢/MSCF		
	Coal at $4/ton	Coal at $5/ton	Coal at $6/ton
35	102	107	112
50	108	113	118
65	114	119	124
80	119	124	129

B. Gross return on investment = 20%

Cost of nuclear heat, ¢/MMBtu	Gas selling price, ¢/MSCF		
	Coal at $4/ton	Coal at $5/ton	Coal at $6/ton
35	121	126	131
50	127	132	137
65	133	138	143
80	138	143	148

C. Gross return on investment = 30%

Cost of nuclear heat, ¢/MMBtu	Gas selling price, ¢/MSCF		
	Coal at $4/ton	Coal at $5/ton	Coal at $6/ton
35	145	150	155
50	151	156	161
65	157	162	167
80	162	167	172

Figure II-1
Effect of Coal Cost, Cost of Nuclear Heat, and Return on Investment on the Selling Price of Gas

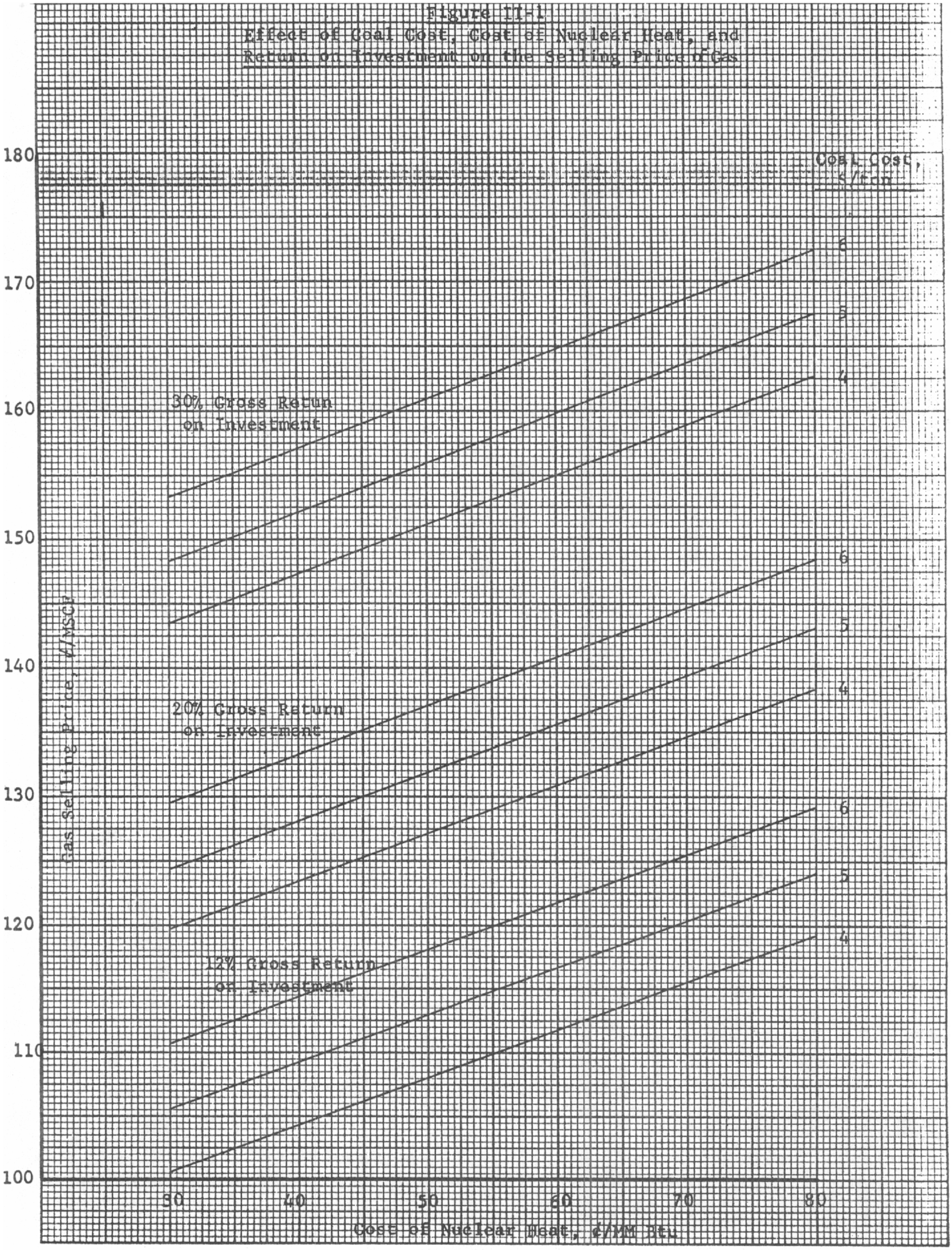


Figure II-2

Effect of Stream Efficiency on Pipeline Gas Selling Price

Basis: Nuclear Heat at 50¢/MM Btu
Coal at $5/ton
20% Gross Return on Investment

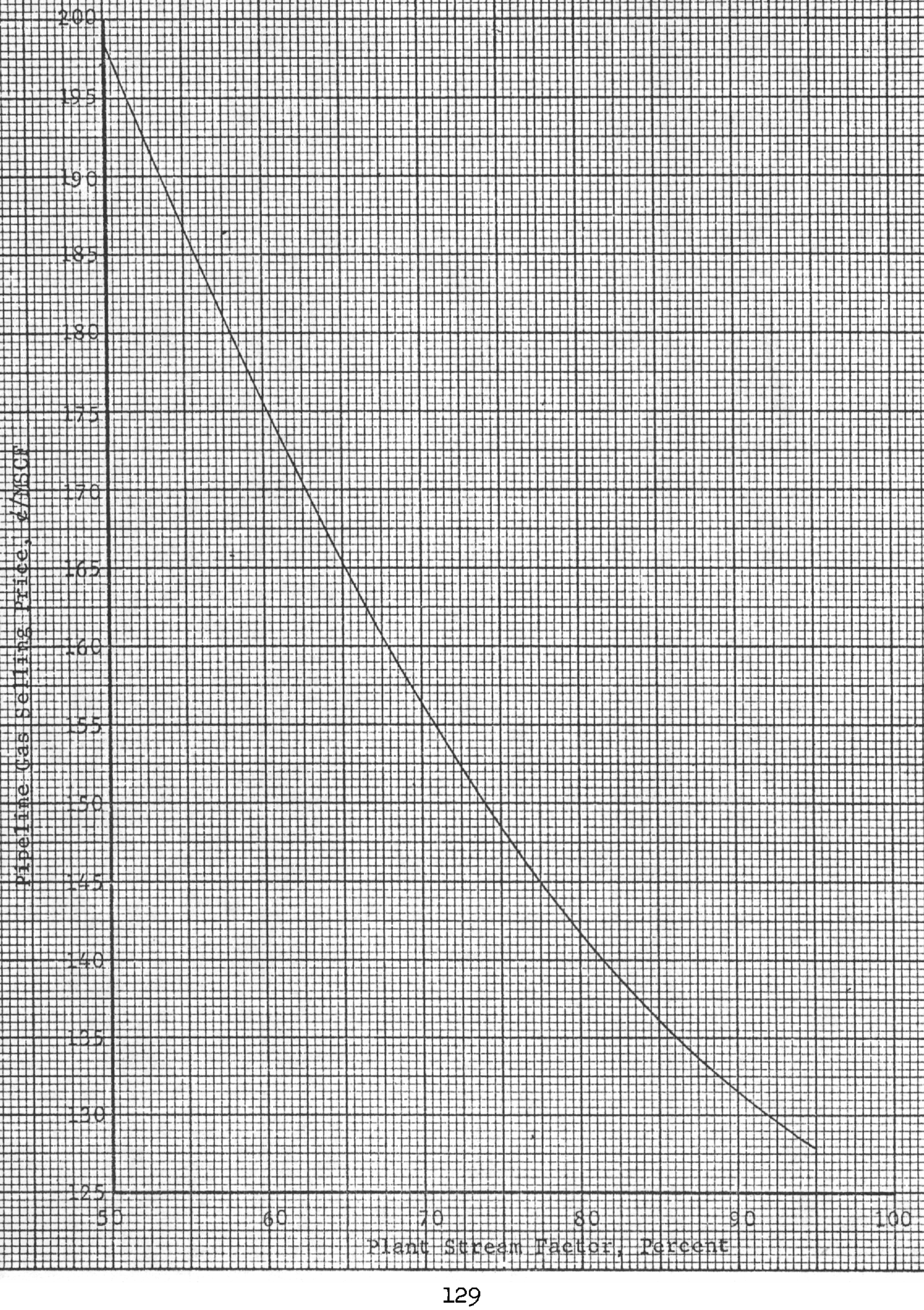
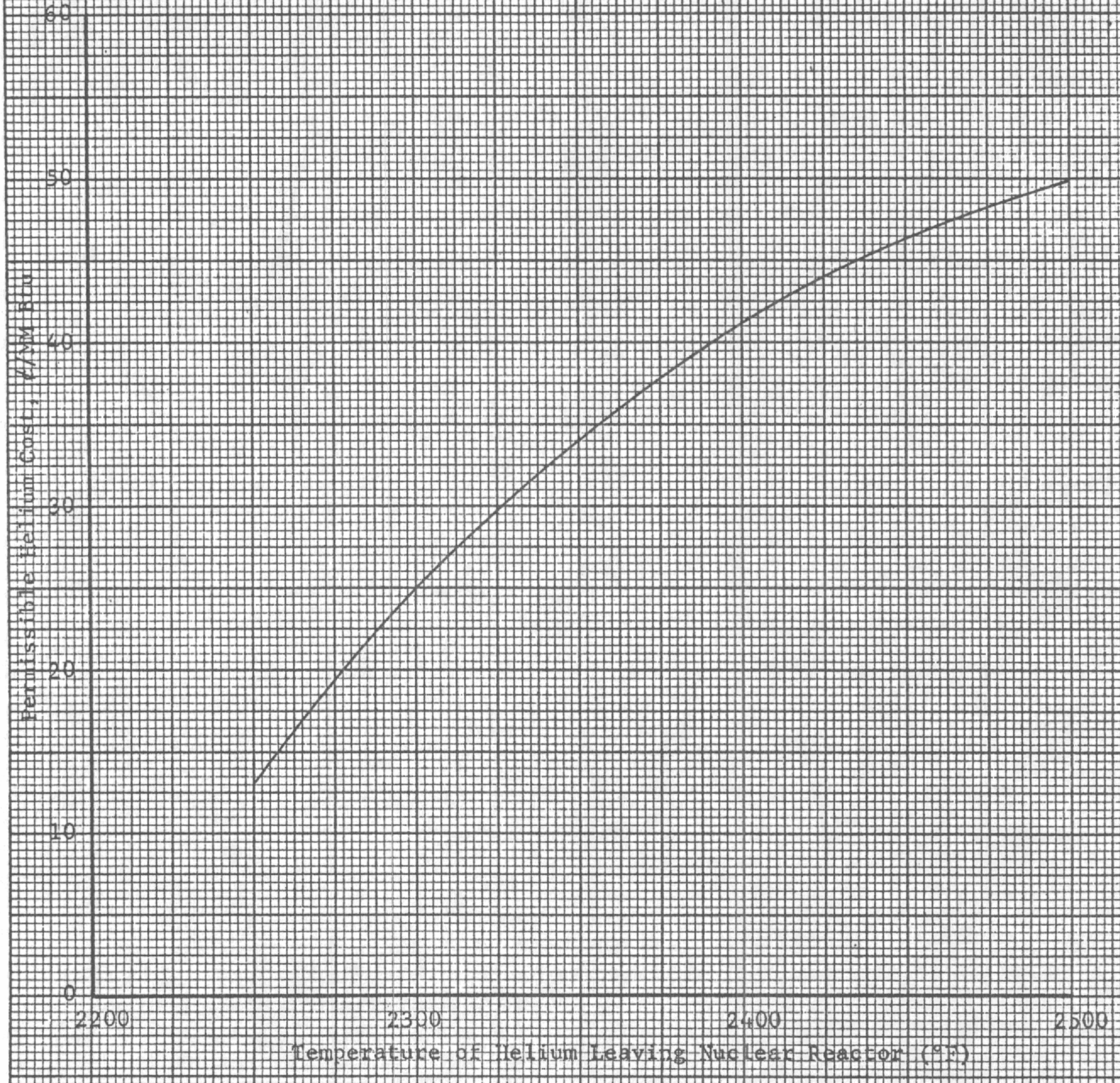


Figure 11-3
Permissible Helium Cost
to Yield Same Pipeline Gas Selling Price
as 2500°F Helium at 50¢/MM Btu

Basis: 20% Gross Return on Investment



3. Design Considerations

The purpose of this section is to elaborate on the process design of Section 200, showing in detail how the design was prepared. Design bases for the remainder of the plant are also given. Important assumptions are stated, and those areas where experimental work is essential to firm up a design are pointed out.

a. Coal Pretreatment Process

Fluidized-bed carbonization of Pittsburgh Seam coals, having properties quite similar to the coal chosen for this study, has been demonstrated in laboratory and pilot-plant units at the Research and Development Division of the Consolidation Coal Company at Library, Pennsylvania. Under the operating conditions used in this study -- that is, maximum preoxidation to "fix" the carbon, and a relatively low carbonization temperature to minimize liquid and gas yields -- the caking tendency of the coal is effectively destroyed. Excellent operability of the process is expected. The quenching of carbonizer effluent gas to condense tar and liquor has also been tested in the pilot plant.

This fluidized-bed low-temperature carbonization process is "ready to go", with no further experimentation required. Design data used for this process are the property of the Consolidation Coal Company.

b. Fluidized Bed Gasification

(1) Reactivity of Bituminous Char

A considerable amount of time and effort was expended in studying kinetic data for the steam gasification of bituminous char at elevated temperatures and pressures. The data of May et al. (19) for steam gasification of a low-temperature char of high volatile bituminous coal in a fluidized bed have been used as the basis for this design. Batch experiments were run by May in a 3.75-inch I.D. reactor capable of maintaining a 7-foot fluidized bed. Pressures studied ranged from atmospheric to 140 psia, while temperature was varied from 1600° to 1800°F. Reaction rates were measured and reported as pounds of carbon gasified per hour per pound of carbon inventory. These data have been correlated by Squires (27) by plotting measured reaction rate against specific steam rate (lb steam fed per hour per lb C inventory) on a log-log plot. Such a plot has been used in the present design, after extrapolating May's data to account for differences in operating conditions as discussed below.

Several other sources of fluid-bed data were considered. The first of these was the work of Zielke and Gorin (30) on the gasification of the same char used by May and his co-workers. However, these data were not directly comparable with the May data because a mixture of steam and hydrogen was used as the gasifying medium. Nevertheless, a qualitative comparison indicates that the data used in the present study should yield a conservative gasifier design.

The steam-carbon gasification data of Jolley et al. (9) were also considered, but proved to be of limited value. This work was done using a low-rank bituminous coal and a char from this coal. The measured gasification rates were considerably higher than those for the high-rank bituminous char used by both May and Zielke, hence further correlation was not attempted.

The following process variables are factors which were considered in determining the manner of extrapolating the May data.

(a) Temperature

The rate of carbon-steam reaction increases with increasing temperature. May et al. measured gasification rates in the range of 1800°F, the operating temperature chosen for this study, therefore no extrapolation of the data was necessary.

(b) Pressure

For a given specific steam rate an increase in operating pressure will theoretically result in an increased gasification rate because the contact time between steam and char has been increased in direct proportion to the increase in pressure. May (19) indicates, however, that this effect will be practically negligible in increasing pressure from 10 to 30 atmospheres, because the order of the steam-carbon reaction is reduced practically to zero when the steam pressure exceeds about 5 atmospheres. Squires (27) agrees with this and notes that "Calculated gasification rate at 125 psig is increased only 7 to 10 per cent if pressure is increased to infinity (overlooking fact that equilibrium for the steam-carbon reaction would then be exceeded)". For these reasons, no pressure correction was made to May's 140 psia data, yielding slightly more conservative results than might possibly be attained in practice.

(c) Steam/Carbon Ratio

Increasing steam/carbon ratio increases the rate of gasification. May's data (19) fully support this, and Squires' correlation (27) shows the extent of the effect. Our gasifier design utilizes a specific steam rate (effective steam/carbon rates for a batch operation) in the range studied by May, hence no correction of rate data had to be made for steam/carbon ratio.

(d) Particle Size

The effect of char particle size upon gasification rate has not been studied in much detail. The data of Maddox (18) are the only source of information uncovered on the effect of particle size. He did not work with bituminous coal or its char, however, and therefore there is no reason for using his results as a basis for the present design. No correction has been made for particle size because the char particles are roughly only 1-1/2 times as large as those used by May (19).

(e) Carbon Conversion

There is some disagreement as to the effect of carbon conversion on gasification rate. Jolley et al. (9) conducted extensive tests over a wide range of carbon levels in the bed. They concluded that reaction rate declined with increased carbon burn-off. The gasification rate measured for a char with 80 percent of its carbon reacted was only about 70 percent of the rate for similar tests with 20 to 50 percent burn-off.

Zielke and Gorin (30) observed similar results in their work with hydrogen-steam gasification. For a 25/75 hydrogen/steam feed at 30 atmospheres, they found the rate at 80 percent carbon conversion was about 63 percent of that measured at 20 percent carbon conversion.

May et al. (19) concluded that carbon reaction rate increased at low carbon levels. Only a few runs were used as the basis for their conclusion, however, and in view of all the conflicting evidence it is quite possible that this conclusion is incorrect. In order to adjust May's low carbon conversion data for use in the present design which converts 80 percent of the carbon, a 30 percent reduction in the measured gasification rates has therefore been assumed.

With May's data adjusted in accordance with the preceding assumptions, it was concluded that 80 percent of the carbon in the char could be gasified at 1800°F and 450 psia with a steam/carbon ratio of 3.75 pounds per pound by designing for about 1.9 pounds of residue inventory per pound of char fed per hour. This corresponds to a residue residence time of about 7-1/2 hours.

(2) Materials of Construction

A thorough survey was made of the capabilities and limitations of available materials of construction for use at the high temperatures and in the oxidizing atmosphere encountered in the gasifier.

The material chosen for use in the gasifier tubes is Hastelloy X. This metal has excellent strength properties and oxidation resistance up to 2200°F. The maximum tube-wall temperature encountered in the gasifier being about 2150°F, it is felt that the choice of Hastelloy X will comply with safe design procedures, even though it has not yet been approved for use by the ASME Pressure Vessel Code.

Refractory metals were at first thought to be the answer to the high-temperature problem, but it was soon learned that these materials possess absolutely no oxidation resistance at elevated temperatures. The technique of coating them with some nonmetallic refractory material is at present being developed, but the cost of such tubes -- if they were available -- is estimated to be three to five times that of tubes made of Hastelloy X.

(3) Heat Transfer

In order to determine the amount of heat transfer surface required, calculations were made first for the film coefficients both inside and outside the gasifier tubes.

Helium film coefficients, that is, coefficients inside the tubes, were estimated by extrapolating the data of the Bureau of Mines (2) for Reynolds numbers up to about 50,000.

Outside film coefficients were estimated by the correlation of Wender and Cooper (29) for heat transfer in fludidized beds. Here no extrapolation was necessary, since the data used in deriving the correlation included conditions close to those in the gasifier design.

The results of these calculations yielded an overall heat transfer coefficient of about 70 Btu per square foot per hour per °F for the top tube bundle and about 90 Btu per square foot per hour per °F for the bottom tube bundle.

The amount of heat transfer surface calculated using these correlations was found to occupy about 11 percent of the gasifier cross-section. Volk et al. (28) studied the effect of vertical surfaces in fluidized beds and found they were able to operate with up to about 20 percent of the reactor cross-section occupied by tubes. In fact, they found that the internals promoted better fluidization than was obtained in runs using fewer tubes in the reactor.

(4) Design Assumptions

No experimental data on gas composition for the conditions chosen for this design were available , therefore it was necessary to assume a composition. The assumption made was that of equilibrium of both the shift and the steam-methane reactions. The assumption of shift equilibrium is a good one for high steam conversions. May's data (19) indicate that for steam conversions above about 40 percent. shift equilibrium will be attained to within about 10 percent.

The steam-methane reaction plays a much smaller part in determining gas composition than does the water-gas shift. Even if the actual methane concentration was ten times the equilibrium amount, the concentration of hydrogen, carbon monoxide, and carbon dioxide would remain essentially unchanged, and no modification of the process design would be necessary. The gasifier effluent composition is therefore thought to be a reasonable one in the absence of any more detailed data.

All of the sulfur in the char is assumed to appear in the gasifier effluent, about 10 percent as COS and the remainder as H_2S. The exact ratio is established in the shift converter which determines how much of each sulfur compound must be removed in subsequent purification steps.

It is further assumed that 80 percent of the nitrogen, hydrogen, and oxygen in the char is present in the gasifier effluent; the remaining 20 percent being retained in the ash. Finally 70 percent of the nitrogen in the gas is assumed to be present as ammonia, while 30 percent appears as free nitrogen.

The proposed method of injecting coal tar into the gasifier is based upon the experiences of Dent (<u>31</u>). Direct injection of residual oil into a fluidized bed of coke in a hydrogen atmosphere at 1000 psig and about 1500°F resulted in agglomeration near the oil inlet. On the other hand, atomization of the oil into a high-velocity, dilute-phase transfer line carrying hydrogen and coke presented no such difficulties, and this method was adopted by Dent for pilot plant use. It is felt that atomization of coal tar into a steam-char transfer line should be no more difficult.

It is felt that the above considerations have yielded a reasonable and conservative design based upon the information available. Pilot plant tests, however, would certainly have to be conducted at the operating conditions chosen here before a commercial installation could be built.

c. Gas Purification

Design considerations for the "Fluor Solvent CO_2 Removal Process", as well as those for sponge iron removal of H_2S and activated carbon removal of COS, have been discussed in Section I-A-3 of this report and need not be repeated here.

d. Methane Synthesis

Design of the methane synthesis unit is essentially the same as the one in Section I-A-3-e of this report. A more detailed description of design considerations and assumptions is presented in a previous report (14).

Experimental work on this unit is definitely necessary before a commercial plant can be built.

4. Optimization of Gasification Process Variables

a. Gasification Temperature

The operating temperature chosen for the gasifiers is 1800°F. Temperatures much below this proved to be unattractive because of extremely low reaction rates. For example, an operating temperature of 1750°F would increase the required char residence time to a point where 18 gasifiers would be needed to contain the necessary char inventory. This would result in an increase of about $5,000,000 in the investment of the gasification section over that at 1800°F.

Temperatures much higher than 1800°F were unattractive because Hastelloy X could no longer be used as heat transfer surface. As was mentioned previously, the only materials capable of withstanding the conditions encountered in this temperature range would be refractory metals with an oxidation-resistant coating. The cost of such materials would be three to five times that of Hastelloy X, so there are no obvious advantages to operating at higher temperatures.

b. Gasification Pressure

The gasification pressure chosen for this study is 450 psia. This choice represents no departure from the established technology of coal gasification and gas treating. Also, all of the pilot-plant work on pipeline gas synthesis has been carried out at pressures up to 30 atmospheres, and extrapolation of these data could prove to be dangerous. There might be some economic incentive in gasifying at higher pressures if a higher pressure synthesis process were to be developed, but it is doubtful that it would represent a significant advantage.

c. Steam/Carbon Ratio

The optimum steam/carbon ratio was found to be about 3.75 pounds steam per pound of carbon fed. Gasification rate was again the determining factor in making this choice.

A ratio of 3 pounds of steam per pound of carbon required 14 gasifiers to contain the necessary char inventory. Since all of the steam fed to the gasifiers is generated by waste heat in the various process streams, a reduction in its consumption would require throwing away some of this available heat to cooling water. Thus, there is no saving in the cost of the steam to balance the cost of the additional 6 gasifiers.

Increasing the steam/carbon ratio to 4.5 pounds per pound increases the char gasification rate by about 35 percent, reducing the required char residence time to a point where 6 gasifiers could be used. This smaller number of gasifiers together with the increased gas rate, however, yields high superficial gas velocities which would lead to excessive entrainment of solids from the bed. In order to reduce entrainment, 8 gasifiers would therefore still be needed, based upon limitations on gas velocity rather than on reaction rates, thus no savings would be realized.

d. Helium Temperature

The optimum temperature of helium entering the gasifiers is 2500°F, the highest that the Atomic Energy Commission wished to consider. The economics of using helium temperatures lower than 2500°F has already been thoroughly discussed in Section II-A-2.

Taking 2500°F helium at 50 cents per million Btu as the basis, calculations were made to determine the allowable cost of helium at lower temperatures which would yield the same gas cost. The results of these calculations have been presented as Figure II-3. It seems obvious that helium cannot be produced cheaply enough at any temperature below 2500°F to make it economically attractive.

e. Helium Pressure

The pressure of the helium loop was set at about 450 psia in order to balance pressures inside and outside the gasifier tubes. In this manner, creep, or long-term deformation of the tubes, was eliminated, and metal rupture was the controlling factor in the tube design. In accordance with safe design procedure, the tubes were sized to withstand the total operating pressure, assuming that their exposure to such a stress would be for only a matter of seconds. Tubes sized in this manner, however, still have considerably thinner walls than those designed to withstand the total pressure for long periods of time.

f. Percent Carbon Utilization

The gasifiers are designed to utilize 80 percent of the carbon in the coal. Carbon conversions below this are unattractive because this would mean throwing away a greater amount of coal in the form of residue.

Increasing carbon utilization to 90 percent will decrease the reaction rate to a point where 16 gasifiers are necessary. This will result in an estimated increase of about $4,000,000 in the investment of the gasification section. If fixed charges are calculated at 30 percent of investment per year (including maintenance and return on investment), this increase in investment corresponds to about $3,300 per calendar day. On the other hand, increasing carbon utilization to 90 percent will decrease coal consumption by about 420 tons per calendar day. At $5 per ton, this represents a saving of $2,100 per calendar day. This is not enough to offset the increased plant cost, therefore operation at 80 percent carbon utilization is more attractive than at 90 percent. The optimum point is probably somewhere between 80 and 90 percent utilization, but the reduction in gas cost achieved by operating at the optimum will be very small.

5. Evaluation of Alternate Processing Schemes

a. Coal Pretreatment

One alternate method of coal pretreatment was considered. Basically, the coal is fed to the top of a "stovepipe" pretreater along with superheated steam. The dispersed particles are allowed to fall freely through the vessel while being heated to about 1600°F by the steam. At this temperature most of the volatile matter in the coal will have been driven off, and the particles at the bottom of the pretreater will be nonagglomerating. This scheme is based upon the hydrogasification apparatus used by R. Hiteshue at the Bruceton, Pa., station of the Bureau of Mines, in which coal is treated with hydrogen rather than steam.

Lacking design data for this process, the Consolidation Coal Company fluidized-bed, low temperature carbonization process was chosen instead. This represents a conservative approach in that the "pretreatment" is probably more severe than necessary to ensure operability of the gasifier. On the other hand, we have also considered the optimistic approach -- that no pretreatment at all is required for fluidized-bed steam gasification at 1800°F. The effect on gas cost of eliminating pretreatment is discussed in Section II-A-2.

b. Gasification

(1) Possible Methods of Heat Transfer

There are three general methods of transferring heat from the helium to the coal. They are:

(a) Direct thermal contact of coal and helium
(b) Direct thermal contact of coal and an inert heat carrier
(c) Indirect thermal contact through heat transfer surface.

Direct thermal contact of coal and helium was immediately discarded. This was done because the helium stream had to be kept free of all impurities (coal particles, for example) since these fragments would become radioactive in the nuclear reactor and would contaminate the process stream with fission products.

Direct thermal contact of coal and an inert heat carrier which has been heated by the helium was also unattractive. Because of the fact that the helium may not directly contact the inert carrier, it must first be heated through a heat transfer surface. This results in a degradation of the temperature level of the gasifier heat source. As this primary heat exchange must take place indirectly (through surface), there appeared to be no advantage to using an intermediate carrier and incur the additional thermal inefficiency of several stages of heat transfer.

Therefore, it was decided that the most efficient scheme would be the indirect transfer of heat from the helium at its highest temperature level to the gasifier.

(2) Possible Methods of Gas-Solids Contacting

In selecting the optimum method of gas-solids contacting, two gasifier designs other than the fluidized-bed type were considered: the transport reactor, and the fixed-bed reactor.

(a) Transport Reactor

The transport, or entrained solids, reactor has been described in Section I-A-5-c of this report. The average residence time of the coal particles in the reaction zone being much shorter than in a fluidized bed, higher gasification temperatures are required to attain a given degree of conversion. In fact, only temperatures above 2000°F give reaction rates high enough to yield a reasonable reactor design. For example, at 2000°F the coal residence time requirement could be fulfilled by using approximately 14,000 2" O.D. tubes, each one about 120 feet long. This represents a surface area of about 900,000 square feet compared to 60,000 square feet for a fluidized bed. Also, since the reaction temperature is so high, the tubes would have to be fabricated of a refractory metal costing three to five times as much as Hastelloy X. It was therefore quite evident that there was no advantage in using a transport system for the gasifiers.

(b) Fixed-Bed Reactor

A description of fixed-bed gasification has been given in Section I-A-5-c of this report. In this scheme heat is supplied to the coal in the form of sensible heat in a recycled product gas stream which has been superheated to about 2400°F by the helium. There are two major disadvantages in using this method. First, the fine char particles from the pretreatment step cannot be used in the bed because of the huge pressure drop the gas stream would incur in passing through it. A costly briquetting step would therefore have to be included after pretreatment. Also, because of the poor heat transfer obtained in the gas-gas superheater, more surface would be needed than for the dense-phase fluidized bed. It is estimated that for a gasification temperature of 1800°F the helium exchanger would have to contain about 325,000 square feet of surface. The extremely high tube-wall temperatures would require that the tubes be made of a refractory metal with an oxidation-resistant coating.

For these reasons, the idea of fixed-bed gasification was discarded, and a fluidized bed was deemed optimum.

c. Gas Purification

No alternate gas purification schemes were considered. A previous report (14) did, however, study several alternate purification processes for pipeline gas production, including the Fluor, Rectisol, and hot carbonate processes, and the reader is referred to that report for further details. It was concluded in that study that the most economical purification sequence was the Fluor CO_2 Removal Process, followed by sulfur removal with sponge iron and activated carbon. The same sequence was therefore used in the present design.

d. Methanation

No alternate methanation systems were considered. A previous study (14), however, compared the costs of the Kellogg transport reactor, a fluidized bed reactor, and the Bureau of Mines hot-gas-recycle reactor (12), and no further discussion need be given. The results of this comparison indicated that the transport reactor yielded lower investment and operating costs than either of the other two systems; it has, therefore, been chosen for the final plant design.

Appendix

Pipeline Gas from Bituminous Coal

Material Balances

Sections 100 and 150

Input:	Lb/Hr
Raw coal from Section 000	350,000
Air to compressors	190,200
Char quench water	27,000
Process water to wet dust collector	39,080
Char and steam from cyclones 204-G	29,920
	636,200

Output:	Lb/Hr
Char and steam to cyclones 204-G	306,920
Tar product to pump 203-J	32,995
Condensate from separator	20,000
Waste gas from wet dust collector	163,720
Slurry from wet dust collector	34,080
Make gas from tar recovery section	78,485
	636,200

Section 200

Input:	Lb/Hr
Char and steam from Section 150 to 204-G	306,920
Tar product from Section 150 to 203-J	32,995
Coal and char from settling pond to 202-L	4,080
Steam to 201-D	925,000
Residue quench water to 203-F	405,000
	1,673,995

Output:	Lb/Hr
Raw synthesis gas to Section 300	1,158,985
Residue slurry from 203-F	485,090
Char and steam from 204-G to Section 150	29,920
	1,673,995

Section 300

Input:	Lb/Hr
Raw synthesis gas from Section 200	1,158,985
Boiler feed water to shift converters	6,120
Process water to gas scrubbers	100,000
	1,265,105

Output:	Lb/Hr
Synthesis gas to Section 400	640,686
Lock hopper pressurizing gas to 202-F	5,896
Condensate	618,523
	1,265,105

Section 400

Input:	Lb/Hr
Synthesis gas from Section 300	640,686

Output:	Lb/Hr
Clean synthesis gas to Section 500	313,430
Waste gas from flash drums	326,282
H_2S from iron oxide drums	374
COS from activated carbon drums	600
	640,686

Section 500

Input:	Lb/Hr
Clean synthesis gas from Section 400	313,430

Output:	Lb/Hr
Pipeline gas product	153,116
Condensate from scrubber and separators	160,314
	313,430

Overall Energy Balance

Datum Temperature = 60°F

Input:	MM Btu/Hr	Percent of Total
Coal to process	4,530	76.6
Heat transferred from helium	1,385	23.4
	5,915	100.0

Output:	MM Btu/Hr	Percent of Total
Product gas heating value	3,500	59.2
Losses to cooling water	1,130	19.2
Combustible in gasifier residue	611	10.3
Heating value of make gas from Section 150	123	2.1
Sensible and latent heat of waste gas from Section 150	58	1.0
Electric power consumption	31	0.5
Latent heat of regeneration steam from activated carbon drums	50	0.8
Latent heat of miscellaneous heating steam	43	0.7
Sensible heat of water purge streams	100	1.7
Heating value of waste gas from Fluor Process	31	0.5
Convection losses, miscellaneous, etc.	238	4.0
	5,915	100.0

Utilities Summary

Steam

I. 500 psia, 570°F

Steam for plant start-up is generated in Section 1100 at 500 psia, 570°F in a coal-fired boiler capable of generating 450,000 pounds per hour. However, during normal operation the processing sections of the plant are capable of internally fulfilling all steam requirements, and thus the offsite boiler is not on stream.

A. Generation

Section	Normal Generation, Lb/Hr
300	459,200

B. Consumption

Section	Normal Consumption, Lb/Hr
100 and 150	22,500
200 - Turbine drives on helium circulators, 201-J	230,000
400	6,000
500	54,500
1100	146,200
	459,200

II. 500 psia, saturated

A. Generation

Section	Normal Generation, Lb/Hr
200 - Waste Heat Boilers, 204-C	742,000
500	758,000
	1,500,000

B. Consumption

Section	Normal Consumption, Lb/Hr
200 - Gasifiers, 201-D	925,000
300	459,200
Available	115,800
	1,500,000

III. 45 psia, saturated

A. Generation

Section	Normal Generation, Lb/Hr
150	29,500
300	130,800
	160,300

B. Consumption

Section	Normal Consumption, Lb/Hr
400	55,000
1100	105,300
	160,300

Power

A. Generation

Electric power is produced at 13,800 volts by two 5,000-kw turbogenerators with condensing steam turbine drive. About 10 percent excess capacity is available during normal operation. An electric substation is provided to reduce the voltage to 4160, 440 and 110 volts.

Section	Normal Generation, kw
1100	9,000

B. Consumption

Section	Normal Consumption, kw
000	400
100	1,500
200	200
300	30
400	6,000
500	20
1100	850
	9,000

C. Summary for Section 200

Item	Normal Consumption, kw
202-J, Water Pumps	40
203-J, Tar Pumps	50
201-L, Coal Distributors	100
202-L, Coal Elevator	10
	200

Cooling Water

A. Generation

Cooling water is available from the nearby Monongahela River. During the warm summer months, however, when the river is normally low, it may prove inadequate as a heat sink. Accordingly, the plant cooling system is based on tower cooling of recirculated water, with the necessary makeup water obtained from the river.

Cooling water is available throughout the plant at 40 psig and a maximum temperature of 85°F.

Section	GPM
1100	123,270

B. Consumption

Section	GPM
150	6,750
200	41,810
300	32,500
400	5,000
500	19,910
1100	17,300
	123,270

C. Summary for Section 200

Item	GPM
201-J, Surface condensers on steam turbines	41,000
203-F, Residue quench	810
	41,810

Boiler Feed Water

A. Generation

Boiler feed water at 250°F is produced by the following sequence:

(1) River water is filtered and treated with chemicals to remove dissolved solids.

(2) Further purification is effected by passing the treated water over anion- and cation-exchange resins.

(3) Purified makeup water is mixed with returning condensate and heated to about 200°F in condensate heaters in Section 300.

(4) Low-pressure steam is used for deaeration.

(5) Deaerated water is heated to about 250°F in boiler feed water heaters in Section 300

Section	Normal Generation, Lb/Hr
300	1,716,120

B. Consumption

Section	Normal Consumption, Lb/Hr
150	29,500
200 - Waste Heat Boilers, 204-C	765,000
300	140,620
500	781,000
	1,716,120

Process Water

A. Generation

Process water at 90°F is produced by filtering river water and treating it with chemicals to precipitate dissolved solids.

Section	Normal Generation, Lb/Hr
1100	156,080

B. Consumption

Section	Normal Consumption, Lb/Hr
150	56,080
300	100,000
	156,080

SUMMARY OF CONTRACT OPERATIONS - MATERIAL, LABOR & SUBCONTRACT COSTS

ACCOUNT CLASSIFICATION	A/C'S 110 TO 140 MATERIAL & FREIGHT	A/C 310 & 320 CONSTR FORCE WAGES & FRINGE	IN-PLACE SUBCONTRACTS (ALL)	
A - SITE PREPARATION, FOUNDATIONS AND CONCRETE STRUCTURES	$ 110,000.			
B - FURNACES				
C - EXCHANGERS	2,986,000.			
D - CONVERTERS	4,090,400.			
E - TOWERS				
F - DRUMS AND TANKS	463,400.			
H - STEEL STRUCTURES AND PLATFORMS	370,000.			
J - PUMPS AND COMPRESSORS	2,559,500.			
K - BUILDINGS	95,000.			
L - SPECIAL EQUIPMENT	105,000.			
M - PIPING	2,800,000.			
N - ELECTRICAL	130,000.			
O - INSTRUMENTS	500,000.			
P - INSULATION AND PAINT	210,000.			
U - UTILITY EQUIPMENT				
V - TRANSPORTATION AND CONVEYING EQUIPMENT				
W - CHEMICALS & CATALYST				
Z - FIRE AND SAFETY EQUIPMENT	24,700.			
G - CYCLONES	357,000.			
FREIGHT - UNALLOCATED	300,000.			
EXPORT PACKING - UNALLOCATED				
SUBTOTAL: DIRECT MATERIAL	$15,101,000			
S - TEMPORARY FACILITIES INCLUDING RIGGING				
T - TOOL HANDLING & TOOL REPAIRS & MAINTENANCE (CHARGEABLE TO JOB)				
Y - MISCELLANEOUS SUPPLIES & UNALLOCABLE LABOR				
SUBTOTAL: INDIRECT MATERIAL (S, T, Y)	380,000.			
TOTALS	$15,481,000.			

SUMMARY - MATERIAL, LABOR AND SUBCONTRACTS

DESCRIPTION: Pipeline Gas from Bituminous Coal
Section 200 - Gasification and Dust Removal - 8 Units

L.J.P. - J.E.R.

ESTIMATE NO. 4006-B **JOB NO.** ____ **DATE** Sept. 30, 1962

CLIENT U.S. Atomic Energy Commission

LOCATION West Virginia, U.S.A. **TYPE OF ESTIMATE OR APPRAISAL** Budget Cost Estimate

ESTIMATE DETAILS

ITEM	UNIT	TOTAL UNITS	UNIT PRICE	MATERIAL COST	M.H. UNIT	TOTAL MANHOURS	
C-201 Superheater	Sq.Ft.	5,050	10.00	$ 50,500.			
1 Cr - 1/2 Mo Tubes							
1 1/4 Cr Shell							
C-202 Superheater	Sq.Ft.	3,200	30.00	96,000.			
Stainless 310 Tubes							
Insulated Shell with 310 Liner							
C-203 Recuperator	Sq.Ft.	4,850	30.00	145,500.			
Stainless 310 Tubes							
Insulated Shell with 310 Liner							
Helium to Helium							
C-204 Waste Heat Boiler	Sq.Ft.	12,400	5.00	62,000.			
Carbon Tubes							
Carbon Shell							
Total for One Unit				354,000.			
Total for Eight Units				2,832,000.			
C-205 Surface Condensers for							
Helium Compressors	4	38,500		154,000.			
J-201 - 7000 HP							
					TOT MH		
					RATE	$ /HR	
TOTALS For 8 Units				$2,986,000.	TOT LAB	$	

CLASS C - Exchangers Sec 200 - Gasification and Dust Removal

EST. NO. 4006-B JOB NO. AEC SHEET NO. DATE Sept. 30, 1962

ESTIMATE DETAILS - CLASS D & E VESSELS

SKETCH NO. ______ REV. ______ DATE ______ D.P. 500 TEMP. 650°F C.A. ______ DESIGN SPEC. ______
TYPE AND HEIGHT OF SUPPORT 6' Skirt MANHOLES 1-48" 2-30" PLATE SPEC. A 204 Gr 13 FDQ (carbon 1/2 Moly)
FABRICATION SHOP FIELD ______ X RAY ______ SR ______

ITEM	UNIT	TOTAL UNITS	UNIT PRICE	MATERIAL COST	M.H. UNIT	TOTAL MANHOURS	
VESSEL SIZE 11 1/8" OD x 65' x 2"	LBS		.	.			
Shell & Heads - 1/2 Moly	LBS	250,000					
EXTERNAL CLIPS (NO.)	LBS		.	.			
6' Skirt & Base- carbon steel	LBS	19,000					
INTERNALS & Misc. clips							
GRATING (REMOVABLE - NON-REMOVABLE)	LBS		.	.			
SCREENS - ALLOY (SQ FT)	LBS		.	.			
Manholes & Nozzles		21,000					
PIPING	LBS		.	.			
SHROUD	LBS		.	.			
SHOP INSTALLED STUDS (NO.)	LBS		.	.			
OTHER - SPECIFY	LBS		.	.			
Water Jacket - 1/4" Carbon Steel		38,000					
SUB TOTAL		328,000	.30	98,400.			
SEPARATORS DEMISTER Plenum Chamber, Cyclone	EACH						
PACKING Supports, Dip Leg & Vapor Stops	CU FT						
S.S. 310	LBS.	5,000	4.00	20,000.			
WELDING ROD FOR FIELD FABRICATED VESSELS	LBS						
REMOVABLE TRAYS PANS OR BAFFLES							
ITEM TYPE DIA. TK. MATERIAL	EACH						
Heat Exchanger Sections							
Upper - 1065 -1 1/4 x .120 x 15' Hastelloy X Tubes	Sq.Ft.	5,200	37.00	192,400.			
Lower - 400 -1 1/2 x .25 x 15' Do	Sq.Ft.	2,380	55.00	130,900.			
INTERNAL INSULATION INCLUDING FIELD INSTALLED STUDS	SQ FT						
Vessel Lining 7" Kaolite							
3" Purotab	Sq.Ft.	3,200	18^{00}	57,600.			
2 - Arched Brick Grids			6000	12,000.			
FREIGHT					TOT MH		
Total for One Unit				511,300.	RATE	$ /HR	
TOTALS for Eight Units				4,090,400.	TOT LAB	$	

CLASS D-201 Gasifier Sec. 200 Gasification and Dust Removal

EST. NO. 4006-B JOB NO. A.E.C. SHEET NO. DATE Sept. 30, 1962

DE

ESTIMATE DETAILS - PUMPS, COMPRESSORS AND DRIVERS

ITEM		NO. OF UNITS	TOTAL WEIGHT	UNIT PRICE	MATERIAL COST	M.H. UNIT	TOTAL MANHOURS	REMARKS
					EACH		TOTAL	
J - 201 SERVICE Helium Circulator								
DESCRIPTION: ☐ HORIZ, ☐ VERT, ☐ CENTR, ☐ RECIP, ☐ PUMP,								
☐ COMPRESSOR, ☐ JET ARRANGED FOR ☐ MOTOR, ☐ TURBINE,								
☐ STEAM ☐ GAS ENGINE DRIVE								
MFR. ABCD	CAP. EA. 450,000 #/Hr ~~GPM~~ CASE MTL							
SIZE 4-25%	TEMP 1000 °F PRODUCT S.G.							
TYPE Capacity	SUCT. 432 PSIA DISCH 462 PSIA							
Units	MOTOR HP RPM							
SPARE	TURBINE 7000 HP CONDENSING RPM	4		$90^{\underline{00}}$	630,000.		2,520,000.	
J - 202 SERVICE Water Pump								
DESCRIPTION: ☐ HORIZ, ☐ VERT, ☐ CENTR, ☐ RECIP, ☐ PUMP,								
☐ COMPRESSOR, ☐ JET ARRANGED FOR ☐ MOTOR, ☐ TURBINE,								
☐ STEAM ☐ GAS ENGINE DRIVE								
MFR. A to H	CAP. EA. 250 GPM CASE MTL							
SIZE 8 Units	TEMP. 458 °F PRODUCT S.G.	8			2,500.		20,000.	
TYPE	SUCT. 460 PSIA DISCH. 470 PSIA							
	MOTOR 5 HP RPM							
SPARE	TURBINE HP CONDENSING RPM							
J - 203 SERVICE Tar Pump								
DESCRIPTION: ☐ HORIZ, ☐ VERT, ☐ CENTR, ☐ RECIP, ☐ PUMP,								
☐ COMPRESSOR, ☐ JET ARRANGED FOR ☐ MOTOR, ☐ TURBINE,								
☐ STEAM ☐ GAS ENGINE DRIVE								
MFR A to H	CAP. EA. 10 GPM CASE MTL							
SIZE	TEMP °F PRODUCT S.G.							
TYPE	SUCT. PSIG DISCH. PSIG							
	MOTOR 10 HP RPM	8			1,500.		12,000.	
SPARE	TURBINE HP CONDENSING RPM							
J - 204 SERVICE Condensate Pumps								
DESCRIPTION: ☐ HORIZ, ☐ VERT, ☐ CENTR, ☐ RECIP, ☐ PUMP,								
☐ COMPRESSOR, ☐ JET ARRANGED FOR ☐ MOTOR, ☐ TURBINE,								
☐ STEAM ☐ GAS ENGINE DRIVE								
MFR.	CAP. EA. GPM CASE MTL							
SIZE 1 Spare	TEMP °F PRODUCT S.G.	5			1,500.		7,500.	
TYPE	SUCT. PSIG DISCH. PSIG							
	MOTOR HP RPM							
SPARE	TURBINE HP CONDENSING RPM							
							$2,559,500.	

CLASS J PUMPS, COMPRESSORS AND DRIVERS Sec. 200 Gasification & Dust Removal J

EST NO. 4006B JOB NO. AEC SHEET NO. DATE Sept. 30, 1962

B. Production of Chemicals and Liquid Fuels

1. Description of Process

The plant for chemicals and liquid fuels from coal has been designed to produce ammonia, methanol, hydrogen, and gasoline as its major products and to recover the valuable co-products produced in gasoline synthesis. A process stream balance, showing the breakdown of this plant according to section, is presented as Drawing No. CE-1525-A.

a. Production of Raw Synthesis Gas

(1) Section 000 - Coal Storage and Reclamation

During eight hours each day, coal is received by truck or conveyor belt from an adjacent coal mine at the rate of 730 tons per hour. The 4" x 0 coal is conveyed to a coal distributing center, where about 239 tons per hour is dispatched for immediate use (175 tons per hour to process and 64 tons per hour to boiler plant) and the remainder is conveyed to the storage area.

Coal is distributed to several storage piles by a traveling stacker. These storage piles contain enough coal to permit the plant to operate for 30 days at normal capacity in the event the coal supply is cut off.

During the 16 hours each day that the mine is not operating, coal is reclaimed from storage by gravity flow into two underground tunnels onto conveyor belts. The equipment in Section 000 has been spared in such a manner as to permit continuous operation at full capacity.

(2) Section 100 - Coal Grinding

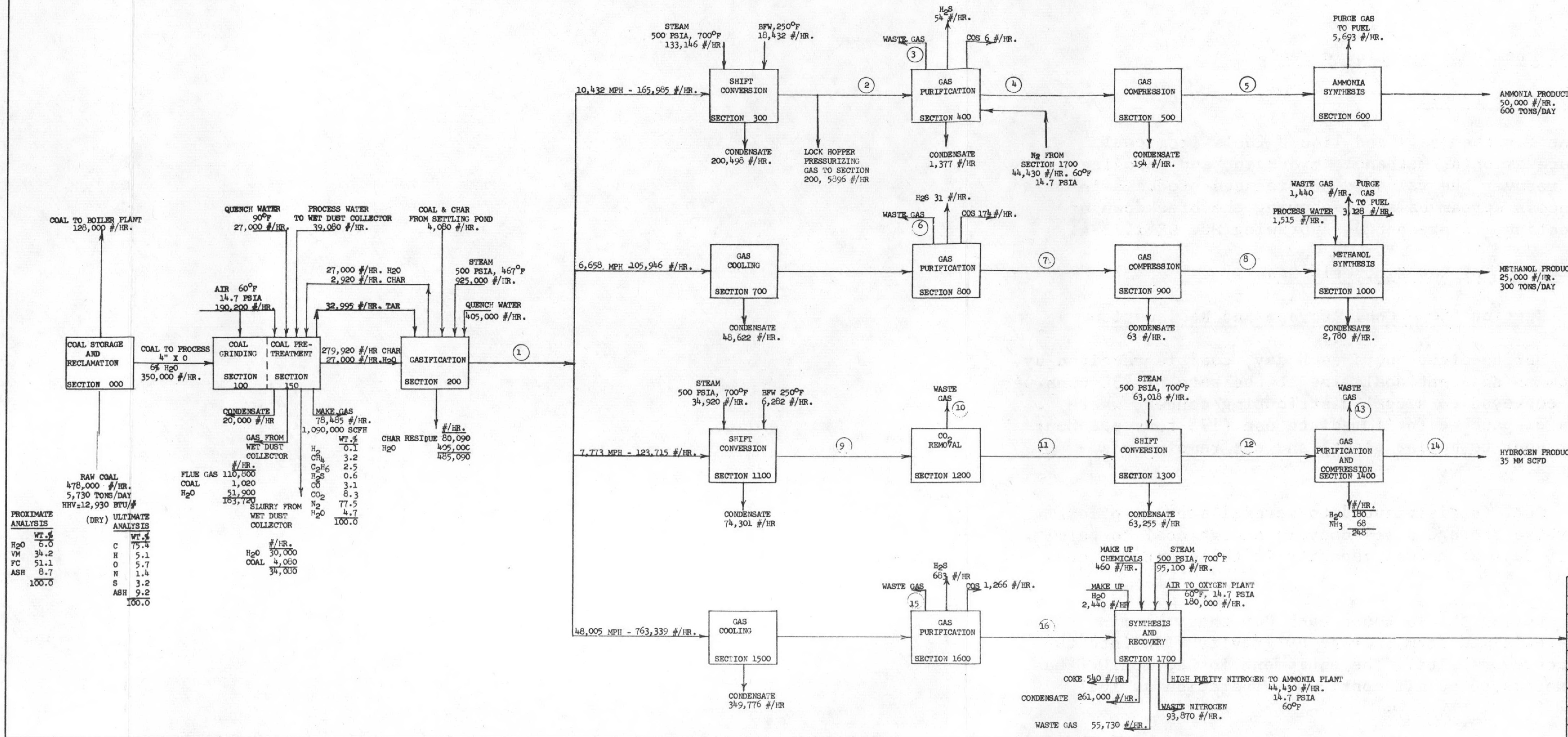
Section 100 consists of exactly the same process scheme as Section 100 of the Pipeline Gas Plant. The reader is referred to Section II-A-1-b of this report for a more detailed description of this section.

(3) Section 150 - Coal Pretreatment

Section 150 is the same as Section 150 of the Pipeline Gas Plant; namely, Consolidation Coal Company's low-temperature carbonization process. A complete process description has already been given in Section II-A-1-c of this report.

(4) Section 200 - Gasification and Dust Removal

Section 200 of this plant is also exactly the same as Section 200 of the Pipeline Gas Plant. A detailed description is presented in Section II-A-1-d of this report, and a process flowsheet is shown as Drawing No. CE-1184-B



	#/HR.	BBL/DAY	MM#/YR. (90% S.F.)
GASOLINE	58,200	5,650	---
ETHYLENE	6,600	---	52.0
MOTOR ETHANOL	5,600	490	---
PROPANE LPG	4,200	560	---
BUTANE	3,000	350	---
DIESEL OIL	2,500	210	---
ACETIC ACID	2,300	---	18.1
WAXY OIL	1,700	130	---
PROBANOL	1,350	---	10.6
PROPYLENE	1,100	---	8.6
ACETONE	1,000	---	7.9
PROPIONIC ACID	510	---	4.0
1-BUTANOL	440	---	3.5
MEK	330	---	2.6
ACETALDEHYDE	300	---	2.4
ISOBUTANOL	210	---	1.7
HEAVY ALCOHOLS	142	---	1.1
	89,482		

STREAM NO.		①			②			③	
TEMP. (°F)		700			100			---	
PRESS. (PSIA)		440			420			---	
FLOW RATE	#/HR.	MPH	MOL %(DRY)	#/HR.	MPH	MOL %(DRY)	#/HR.	MPH	MOL %(DRY)
CO_2	289,000	6,571	15.1	96,800	2,200	30.0	96,800	2,200	98.2
CO	269,000	9,620	22.2	2,086	74.5	1.0	---	---	---
H_2	53,080	26,540	61.2	10,000	5,000	68.1	---	---	---
CH_4	2,460	154	0.3	347	21.7	0.3	---	---	---
N_2	1,065	38	0.1	148	5.3	0.07	---	---	---
H_2S	9,500	279	0.6	1,482	43.5	0.6	1,428	41.9	1.8
COS	1,920	32	0.07	6	0.1	---	---	---	---
NH_3	2,960	174	0.4	---	---	---	---	---	---
DRY GAS	628,985	43,408	100.0	110,869	7,345.1	100.0	98,228	2,241.9	100.0
H_2O	530,000	29,460		300	16.6		47	2.6	
TOTAL GAS	1,158,985	72,868		111,169	7,361.7		98,275	2,244.5	

STREAM NO.		4			5			6	
TEMP. (°F)		100			100			---	
PRESS. (PSIA)		387			4715			---	
FLOW RATE	#/HR.	MPH	MOL %(DRY)	#/HR.	MPH	MOL %(DRY)	#/HR.	MPH	MOL %(DRY)
CO_2	---	---	---	---	---	---	24,935	566.7	92.2
CO	---	---	---	---	---	---	336	12.0	2.0
H_2	9,553	4,776.5	73.9	9,553	4,776.5	73.9	20	10.0	1.6
CH_4	1,539	96.2	1.5	1,539	96.2	1.5	16	1.0	0.2
N_2	44,578	1,592.1	24.6	44,578	1,592.1	24.6	---	---	---
H_2S	---	---	---	---	---	---	833	24.5	4.0
COS	---	---	---	---	---	---	---	---	---
NH_3	---	---	---	---	---	---	---	---	---
DRY GAS	55,670	6,464.8	100.0	55,670	6,464.8	100.0	26,140	614.2	100.0
H_2O	217	12.0		23	1.3		83	4.6	
TOTAL GAS	59,887	6,476.8		55,693	6,466.1		26,223	618.8	

STREAM NO.		7			⑧			⑨	
TEMP. (°F)		80			100			255	
PRESS. (PSIA)		415			5000			430	
FLOW RATE	#/HR.	MPH	MOL % (DRY)	#/HR.	MPH	MOL % (DRY)	#/HR.	MPH	MOL %(DRY)
CO_2	1,465	33.3	1.0	1,465	33.3	1.0	68,815	1,563	28.6
CO	24,224	865.0	26.0	24,224	865.0	26.0	4,592	164	3.0
H_2	4,830	2,415.0	72.5	4,830	2,415.0	72.5	7,386	3,693	67.2
CH_4	208	13.0	0.4	208	13.0	0.4	262	16.4	0.3
N_2	95	3.4	0.1	95	3.4	0.1	112	4.0	0.07
H_2S	---	---	---	---	---	---	1,125	33.1	0.6
COS	---	---	---	---	---	---	6	0.1	---
NH_3	---	---	---	---	---	---	211	12.4	0.2
DRY GAS	30,822	3,329.7	100.0	30,822	3,320.7	100.0	82,509	5,486.0	100.0
H_2O	74	4.1		11	0.6		8,107	448.0	
TOTAL GAS	30,896	3,333.8		30,833	3,330.3		90,616	5,934.0	

STREAM NO.		10			11			12	
TEMP. (°F)		---			680			100	
PRESS. (PSIA)		---			405			383	
FLOW RATE	#/HR.	MPH	MOL %(DRY)	#/HR.	MPH	MOL %(DRY)	#/HR.	MPH	MOL %(DRY)
CO_2	67,451	1,532	96.6	1,364	31	0.8	7,744	176	4.4
CO	---	---	---	4,592	164	4.2	532	19	0.5
H_2	46	23	1.4	7,340	3,670	94.2	7,630	3,815	94.4
CH_4	---	---	---	262	16.4	0.4	262	16.4	0.4
N_2	---	---	---	112	4.0	0.1	112	4.0	0.1
H_2S	1,101	32.4	2.0	24	0.7	---	27	0.79	---
COS	---	---	---	6	0.1	---	1	0.01	---
NH_3	---	---	---	211	12.4	0.3	138	8.1	0.2
DRY GAS	68,598	1,587.4	100.0	13,911	3,898.6	100.0	16,446	4,039.3	100.0
H_2O	5,155	284		2,952	164.0		180	10.0	
TOTAL GAS	73,753	1,871.4		16,863	4,062.6		16,626	4,049.3	

STREAM NO.		13			14			15			16	
TEMP. (°F)		---			100			---			80	
PRESS. (PSIA)		---			450			---			405	
FLOW RATE	#/HR.	MPH	MOL %(DRY)	#/HR.	MPH	MOL %(DRY)	#/HR.	MPH	MOL %(DRY)	#/HR.	MPH	MOL %(DRY)
CO_2	7,744	176	96.6	---	---	---	135,440	3,081.0	90.2	55,000	1,250.0	5.0
CO	---	---	---	532	19	0.5	2,674	95.5	2.8	174,485	6,245.0	24.9
H_2	4	2	1.1	7,626	3,813	99.0	138	69.0	2.0	34,832	17,416.0	69.6
CH_4	---	---	---	262	16.4	0.4	120	7.5	0.2	1,500	94.0	0.4
N_2	---	---	---	112	4.0	0.1	---	---	---	710	25.3	0.1
H_2S	27	0.79	0.1	---	---	---	5,583	163.8	4.8	---	---	---
COS	1	0.01	---	---	---	---	---	---	---	---	---	---
NH_3	70	4.1	2.2	---	---	---	---	---	---	---	---	---
DRY GAS	7,846	182.9	100.0	8,532	3,852.4	100.0	143,955	3,416.8	100.0	266,527	25,030.3	100.0
H_2O	---	---		---	---		607	33.7		525	29.2	
TOTAL GAS	7,846	182.9		8,532	3,852.4		144,562	3,450.5		267,052	25,059.5	

THE M. W. KELLOGG CO.
U. S. ATOMIC ENERGY COMMISSION
FUELS AND CHEMICALS FROM BITUMINOUS COAL
PROCESS STREAM BALANCE

NO.	DESCRIPTION	DATE	BY	CHECKED
REVISIONS				

SCALE:
DRAWN: L.J.PIERONI
CHECKED: LJP
APPROVED:
DATED: 9-12-62
ISSUED FOR FABRICATION
ISSUED FOR CONSTRUCTION

CLASS & ITEM	AREA	JOB NUMBER	DRAWING NUMBER
		4006	CE-1525-A

DRAWING No.

b. Ammonia Production

(1) Section 300 - Shift Conversion and Heat Recovery

Section 300 consists of one train of operating equipment. The purpose of this section is to reduce the CO content of the gas to about 1.0 mole percent in preparation for ammonia synthesis.

Clean gas from Section 200 at 700°F and 440 psia enters Section 300 at the rate of 10,432 moles per hour and is fed to two parallel shift converters. Steam at 700°F is also fed to the converters at the rate of 133,146 pounds per hour to promote conversion of CO. The reaction is mildly exothermic, and therefore boiler feed water at 250°F is injected between catalyst beds in the converters to absorb the heat which is liberated. About 350 moles per hour of the shift effluent at 700°F is compressed to 500 psia and is sent to lock hoppers 202-F as pressurizing gas. The remainder of the gas is cooled to 500°F in a waste heat boiler making 500 psia saturated steam for the process. It is further cooled to 260°F in the CO_2 stripper reboiler of Section 400, and is finally cooled to 100°F against cooling water. The synthesis gas is then fed to a water scrubber where it is countercurrently scrubbed with clean water to remove trace amounts of ammonia which might be present.

The overhead gas from the water scrubber, now containing 1.0 percent CO, flows to Section 400.

(2) Section 400 - Gas Purification

The gas purification section is designed to reduce the concentration of carbon oxides in the synthesis gas to 10 parts per million and to reduce the sulfur content to about 1 part per million in preparation for ammonia synthesis. The purification sequence consists of monoethanolamine (MEA) scrubbing for CO_2 and bulk H_2S removal, followed by caustic wash and activated carbon adsorption for residual H_2S and COS removal, respectively. Carbon monoxide is removed by reacting it with hydrogen over a nickel catalyst to form methane and water.

A more detailed description of the MEA process, caustic wash, and activated carbon sequence is presented in Section I-B-1-b of this report, and no further discussion will be given here.

Synthesis gas at 100°F leaves the activated carbon system at the rate of 5,117 moles per hour. It is heated to 500°F against the effluent gas from the methanator and is then fed to the methanator. The synthesis gas absorbs the exothermic heat of reaction and leaves the methanator at 685°F. It is cooled to 255°F against the feed gas and finally to 100°F in a cooler. The entrained water in the gas stream is removed in a vapor-liquid separator, and clean synthesis gas, containing about 98 percent hydrogen, leaves at 100°F and 387 psia. To this gas is added 1,587 moles per hour of high-purity nitrogen from the oxygen plant in Section 1700 after it has first been compressed to 387 psia. The resulting gas, now having a H_2/N_2 ratio of 3/1 and ready to be compressed to synthesis pressure, flows to Section 500.

(3) Section 500 - Gas Compression

Purified synthesis gas at 100°F from Section 400 enters Section 500 at the rate of 6,477 moles per hour. It enters a three-stage, steam-driven, reciprocating compressor designed to compress the gas from 387 psia to the ammonia synthesis pressure of 4,715 psia. It is cooled in intercoolers and aftercooler to 100°F and flows to Section 600 at the rate of 6,466 moles per hour.

(4) Section 600 - Ammonia Synthesis

In Section 600 ammonia is synthesized at the rate of 600 tons per day from the gas mixture supplied from Section 500. Gas is purged from the synthesis loop (to remove inerts) at the rate of 5,693 pounds per hour and is sent to Section 3000 as boiler fuel. Details of this process can be found in Section I-B-1-b of this report

c. Methanol Production

(1) Section 700 - Gas Cooling

Section 700 consists of one train of operating equipment. The purpose of this section is to cool the gas to about 100°F in preparation for CO_2 and sulfur removal.

Clean gas from Section 200 at 700°F and 440 psia enters Section 700 at the rate of 6,658 moles per hour. It is first cooled to 350°F in a waste heat boiler generating 45 psia steam, and finally to 100°F in a cooler. It then enters a water scrubber where it is countercurrently scrubbed with clean water to remove trace amounts of ammonia which are present.

The synthesis gas from the overhead of the water scrubber, now at 100°F and 435 psia, flows to Section 800.

(2) Section 800 - Gas Purification

The gas purification section is designed to reduce the CO_2 concentration in the synthesis gas to 1.0 mole percent and to reduce the total sulfur concentration to about 0.004 grains/100 SCF of gas. The purification sequence consists of the "Fluor Solvent CO_2 Removal Process" for CO_2 and bulk H_2S removal, followed by sponge iron (iron oxide) and activated carbon for residual H_2S and COS removal, respectively.

The Fluor system is designed as one train of operating equipment. The iron oxide drums are also designed as one train of equipment consisting of seven parallel drums followed by one guard chamber. The activated carbon drums are arranged in two parallel trains, each train consisting of three drums which are manifolded for cyclic operation. A more detailed description of the methanol synthesis gas purification sequence may be found in Section I-B-1-c of this report.

Purified synthesis gas, now containing about 1.0 mole percent CO_2 and about 0.004 grains total sulfur/100 SCF, flows to Section 900.

(3) Section 900 - Gas Compression

The purpose of Section 900 is to compress the synthesis gas from Section 800 to 5,000 psia in preparation for methanol synthesis.

Purified synthesis gas at 80°F enters Section 900 at the rate of 3,334 moles per hour. It is fed to a three-stage, steam-driven reciprocating compressor designed to compress the gas from 415 psia to 5,000 psia. It is cooled in intercoolers and an aftercooler to 100°F and flows to Section 1000 at the rate of 3,330 moles per hour.

(4) Section 1000 - Methanol Synthesis

In Section 1000 methanol is synthesized at the rate of 300 tons per day from the H_2-CO mixture supplied from Section 900. Gas is purged from the synthesis loop at the rate of 3,128 pounds per hour to remove methane and argon and is sent to Section 3000 as boiler fuel. For a detailed description of this process, refer to Section I-B-1-c of this report.

d. Hydrogen Production

(1) Section 1100 - Primary Shift Conversion and Heat Recovery

Section 1100 consists of one train of operating equipment. The purpose of this section is to reduce the CO content of the synthesis gas to about 3.0 mole percent.

Clean gas from Section 200 at 700°F and 440 psia enters Section 1100 at the rate of 7,773 moles per hour and flows to the shift converter. Steam at 700°F is also fed to the converter at the rate of 34,920 pounds per hour to push the shift equilibrium in the direction of hydrogen production. The reaction is mildly exothermic, and boiler feed water at 250°F is injected between catalyst beds in the converter to absorb the heat which is liberated. The effluent gas from the converter, now at 780°F, is cooled to 625°F against feed gas to Section 1300. It is further cooled to 376°F in a boiler feed water heater and finally to 255°F in the CO_2 stripper reboiler in Section 1200. Entrained water in the gas stream is removed in a vapor-liquid separator, and the gas flows on to Section 1200.

(2) Section 1200 - Primary Gas Purification

Section 1200 is designed to reduce CO_2 concentration in the synthesis gas to about 0.8 mole percent and to reduce the total sulfur concentration to about 80 parts per million. The purification scheme used for this is the "Hot Carbonate Process".

In this process, CO_2 and H_2S are removed by countercurrent scrubbing with hot potassium carbonate solution. A more detailed description of this primary purification sequence has already been given in Section I-B-1-d of this report.

Purified synthesis gas, now at 220°F is heated to 680°F against the shift effluent gas in Section 1100 and is sent to Section 1300.

(3) Section 1300 - Secondary Shift Conversion and Heat Recovery

This section consists of one train of equipment. Its purpose is to reduce the CO content of the synthesis gas to about 0.5 mole percent by reaction with steam.

Gas from Section 1200 at 680°F and 405 psia enters Section 1300 at the rate of 4,063 moles per hour and is fed to the shift converter together with 63,018 pounds per hour of 700°F steam. The heat of reaction is absorbed by the gas stream, which leaves the vessel at 730°F. The gas is cooled to 350°F in a waste heat boiler generating low-pressure (45 psia) steam, to 316°F in the CO_2 stripper reboiler in Section 1400, and finally to 100°F in a cooler. Entrained water is removed in a vapor-liquid separator, and the gas, now at 100°F and 383 psia, flows to Section 1400.

(4) Section 1400 - Secondary Gas Purification and Gas Compression

The purpose of this section is to remove virtually all of the CO_2, sulfur, ammonia, and water in the hydrogen stream, and to compress it to its final pressure of 450 psia.

Gas from Section 1300 at the rate of 4,049 moles per hour flows to an MEA absorber where it is countercurrently scrubbed with a 20 percent MEA solution to reduce the CO_2 concentration to 50 parts per million and to remove essentially all the H_2S. It is then fed to a caustic scrubber to remove trace quantities of these impurities, and finally to an alumina dryer where water is removed to a dew point of -80°F at 360 psia and where the remainder of the ammonia is adsorbed. A more detailed description of the hydrogen purification sequence is presented in Section I-B-1-d of this report.

Purified gas from the dryer is compressed to 450 psia by a single-stage, turbine-driven centrifugal compressor. The product gas, now containing 99 mole percent hydrogen, is cooled to 100°F in an aftercooler and flows from the plant at the rate of 35,000,000 standard cubic feet per day.

e. Liquid Fuels Production

(1) Section 1500 - Gas Cooling

Section 1500 consists of three parallel trains of operating equipment. All flow rates mentioned in this description are total quantities for the three trains. The purpose of this section is to cool the raw synthesis gas to about 100°F in preparation for CO_2 and sulfur removal.

Clean gas from Section 200 at 700°F and 440 psia enters Section 1500 at the rate of 48,005 moles per hour. It is first cooled to 335°F in waste heat boilers making both high-pressure (500 psia) and low-pressure (45 psia) steam. It is then cooled to 100°F against cooling water. This gas now flows to a water scrubber where it is countercurrently scrubbed with clean water to remove trace amounts of ammonia which might be present.

The synthesis gas from the overhead of the water scrubber, now at 100°F and 430 psia, flows to Section 1600.

(2) Section 1600 - Gas Purification

The purpose of this section is to reduce the CO_2 content of the synthesis gas to 5.0 mole percent and to reduce the total sulfur concentration to about 0.004 grains/100 SCF of gas. The purification sequence consists of the "Fluor Solvent CO_2 Removal Process" for CO_2 and bulk H_2S removal, followed by sponge iron (iron oxide) and activated carbon for residual H_2S and COS removal, respectively.

The Fluor system is designed as one train of operating equipment. The iron oxide drums are arranged in two parallel trains, each train consisting of eight parallel drums followed by one guard chamber. The activated carbon drums are arranged in four parallel trains, each train consisting of three drums which are manifolded for cyclic operation. A more detailed description of this gas purification sequence may be found in Section I of this report.

Purified synthesis gas, now containing 5.0 mole percent CO_2 and about 0.004 grains total sulfur/100 SCF flows at the rate of 25,060 moles per hour to Section 1700 at 80°F and 405 psia.

(3) Section 1700 - Synthesis and Recovery

The purpose of Section 1700 is to synthesize and recover the various fuels and chemicals shown in Drawing No. CE-1525-A. The details of this process sequence have already been discussed in Section I-B-1-e of this report, and no further description need be given here.

f. Plant Utilities

(1) Section 3000 - Offsite Facilities

Section 3000 includes facilities for:

(a) generating steam and electric power
(b) supplying cooling water, process water, and boiler feed water
(c) providing miscellaneous services necessary to make this a completely self-sufficient plant.

Steam generation facilities consist of two boilers capable of producing a total of 1,750,000 pounds per hour of 1015 psia, 850°F steam and one boiler capable of producing 160,000 pounds per hour of 500 psia, 470°F steam.

Electric power is produced at 13,800 volts by turbogenerators using condensing steam turbine drives. An electric substation is provided to reduce the voltage to 4,160, 440 and 110 volts.

Brief descriptions of the means of supplying cooling water, boiler feed water, and process water are included in the Utilities Summary in the Appendix.

2. Economics

The general philosophy of calculating production costs and selling prices for this multi-product plant is discussed in detail in Section I-B-2.

Estimated operating labor for the complete process plant is 77 men per shift, as detailed in Table II-5.

Estimated capital investment for the plant is summarized in Table II-6. Interest during construction is again calculated at 6.7 percent, assuming a design, engineering, and construction period of 30 months. Total investment is about 142 million dollars.

Estimated annual operating cost for synthesis gas production, using nuclear heat at 50¢/MM Btu and coal at $5 per ton is about 14.5¢/MSCF (Table II-7). Using this synthesis gas as raw material, production costs for the major products are calculated in Tables II-8 to II-11 to be as follows:

ammonia	$32.70/ton
methanol	10.4¢/gal
hydrogen (99%)	28.1¢/MSCF
gasoline	16.8¢/gal

These are approximately the same production costs as calculated when lignite is used as raw material (section I-B-2). This is in marked contrast to the pipeline gas comparison, in which the product gas could be produced from lignite at about 65 percent of the cost when using bituminous coal. The explanation is that the methane formed when lignite is gasified is an asset in the production of pipeline gas, which, after all, is largely methane, but it is a liability when liquid fuels and chemicals are produced.

The effect on product costs of adding gross return on investment is shown in Figures II-4 to II-7. If a gross return of 20 percent is desired, for example, the above production costs are increased as follows:

	Production Cost	20 Percent Gross Return	Selling Price
ammonia, \$/ton	32.70	23.90	56.60
methanol, ¢/gal	10.4	6.6	17.0
hydrogen, ¢/MSCF	28.1	19.1	47.2
gasoline, ¢/gal	16.8	25.0	41.8

Variations in the cost of coal and nuclear heat are seen to be relatively insignificant.

The effect of stream efficiency on the various selling prices is illustrated in Figure II-8.

Table II-5

Estimated Operating Labor
Liquid Fuels and Chemicals from Bituminous Coal

Section	Title	Men per Shift	
Synthesis Gas Production:			
000	Coal Storage and Reclamation	1	
100	Coal Grinding	5	
150	Coal Pretreatment	2	
200	Gasification	8	
	Subtotal		16
Ammonia Production:			
300	Shift Conversion	1	
400	Gas Purification	2	
500	Gas Compression	1	
600	Synthesis	1	
	Subtotal		5
Methanol Production:			
700	Gas Cooling	1	
800	Gas Purification	3	
900	Gas Compression	1	
1000	Synthesis	2	
	Subtotal		7
Hydrogen Production:			
1100	Primary Shift Conversion	1	
1200	CO_2 Removal	1	
1300	Secondary Shift Conversion	1	
1400	Gas Purification and Compression	1	
	Subtotal		4
Gasoline and Chemicals Production:			
1500	Gas Cooling	1	
1600	Gas Purification	5	
1700	Synthesis and Recovery	39	
	Subtotal		45
	TOTAL OPERATING LABOR		77
	man-hours/day		1,848

Table II-6

Investment Summary
Liquid Fuels and Chemicals from Bituminous Coal

Section	Title	Material & Freight, Dollars	Bare Cost,* Dollars	
Synthesis Gas Production:				
000	Coal Storage & Reclamation	1,070,000	1,825,000	
100	Coal Grinding	2,213,500	3,593,000	
150	Coal Pretreatment	1,177,300	2,573,000	
200	Gasification	15,101,000	23,320,000	
	Subtotal	19,561,800		31,311,000
Ammonia Production:				
300	Shift Conversion	800,000	1,250,000	
400	Gas Purification	2,268,000	4,050,000	
500	Gas Compression	1,150,000	1,750,000	
600	Synthesis	3,100,000	4,376,000	
	Subtotal	7,318,000		11,426,000
Methanol Production:				
700	Gas Cooling	244,000	410,000	
800	Gas Purification	490,000	898,000	
900	Gas Compression	750,000	1,275,000	
1000	Synthesis	660,000	1,200,000	
	Subtotal	2,144,000		3,783,000
Hydrogen Production				
1100	Primary Shift Conversion	325,000	550,000	
1200	CO_2 Removal	1,060,000	1,960,000	
1300	Secondary Shift Conversion	530,000	820,000	
1400	Gas Purification and Compression	435,000	734,000	
	Subtotal	2,350,000		4,064,000
Gasoline and Chemicals Production:				
1500	Gas Cooling	1,675,000	2,480,000	
1600	Gas Purification	2,090,000	3,625,000	
1700	Synthesis and Recovery	25,000,000	39,000,000	
	Subtotal	28,765,000		45,105,000
Offsite Facilities (Section 3000)		12,000,000	19,280,000	
				19,280,000
	TOTAL BARE COST			114,969,000
Contractor's Overhead and Profit				12,070,000
Interest during Construction at 6.7%				8,510,000
	TOTAL FIXED INVESTMENT			135,549,000

Table II-6 (cont'd)

TOTAL FIXED INVESTMENT (Brought Forward)		$135,549,000
Working Capital:		
30 days' coal inventory (172,000 tons at $5/ton)	$ 860,000	
Accounts receivable (value of 30 days' production):		
Ammonia at $90/ton	1,620,000	
Methanol at 30¢/gallon	818,000	
Hydrogen at 40¢/MSCF	420,000	
Gasoline at 20¢/gallon	1,420,000	
Co-products	1,150,000	
Catalyst inventory	140,000	
Total Working Capital		6,428,000
TOTAL CAPITAL INVESTMENT		$141,977,000

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* Bare cost includes materials, freight, construction labor, field administration and supervision, insurance during construction, cost of tools, field office expense, and cost of home office engineering and procurement.

Table II-7

Estimated Annual Operating Cost
Synthesis Gas from Bituminous Coal

Basis: 395,000,000 SCFD of Dry Synthesis Gas
90% Stream Efficiency
Fixed Investment = $36,900,000 (Sections 000 to 200)

Item	$/Year		¢/MSCF
Direct Costs:			
Nuclear heat (1,385 MM Btu/hr x 7,884 hr/yr x 50¢/MM Btu)	5,460,000		4.2
Coal to gasifiers (175 TPH x 7,884 hr/yr x $5/ton)	6,900,000		5.3
Operating labor (384 man-hr/day x $2.80/man-hour x 365 days/yr)	393,000		0.3
Supervision at 15% of operating labor	59,000		0.1
Utilities (Table II-12)	408,000		0.3
Maintenance at 4% of bare cost per year (1)	1,252,000		1.0
Supplies at 15% of maintenance	188,000		0.1
Total Direct Costs		14,660,000	
Indirect Costs:			
Payroll overhead at 20% of operating labor, maintenance labor, and supervision	266,000		0.2
General plant overhead at 50% of operating labor, superivison, maintenance, and supplies	946,000		0.7
Total Indirect Costs		1,212,000	
Fixed Costs:			
Depreciation at 5% of fixed investment	1,850,000		1.4
Property taxes and insurance at 3% of fixed investment	1,110,000		0.9
Total Fixed Costs		2,960,000	
TOTAL ANNUAL OPERATING COST		18,832,000	
ESTIMATED DRY SYNTHESIS GAS PRODUCTION COST			14.5

---

(1) Maintenance is 70% labor, 30% material

Table II-8

Estimated Annual Operating Cost
Ammonia from Bituminous Coal

Basis: 600 Tons/Day of Ammonia
90% Stream Efficiency
Fixed Investment = $13,470,000 (Sections 300 through 600)

Item	$/Year		$/Ton NH_3
Direct Costs:			
Dry Synthesis gas (2,357 MSCF/hr x 7,884 hr/yr x 14.5¢/MSCF)	2,694,000		13.70
Operating labor (120 man-hr/day x $2.80/man-hr x 365 days/yr)	123,000		0.60
Supervision at 15% of operating labor	18,400		0.10
Utilities (Table II-12)	1,488,000		7.60
Catalysts and chemicals (excerpted from Table II-13)	98,000		0.50
Maintenance at 4% of bare cost per year (1)	457,000		2.30
Supplies at 15% of maintenance	68,000		0.30
Total Direct Costs		4,946,400	
Indirect Costs:			
Payroll overhead at 20% of operating labor, maintenance labor and supervision	92,200		0.50
General plant overhead at 50% of operating labor, supervision, maintenance, and supplies	333,000		1.70
		425,200	
Fixed Costs:			
Depreciation at 5% of fixed investment	674,000		3.40
Property taxes and insurance at 3% of fixed investment	404,000		2.00
Total Fixed Costs		1,078,000	
TOTAL ANNUAL OPERATING COST		6,449,600	
TOTAL AMMONIA PRODUCTION COST			32.70

---

(1) Maintenance is 70% labor, 30% material.

Table II-9

Estimated Annual Operating Cost
Methanol from Bituminous Coal

Basis: 300 Tons/Day of Methanol
90% Stream Efficiency
Fixed Investment = $4,460,000 (Sections 700 through 1000)

Item	$/Year		¢/gal CH_3OH
Direct Costs:			
Dry Synthesis gas (1,502 MSCF/hr x 7,884 hr/yr x 14.5¢/MSCF)	1,717,000		5.8
Operating labor (168 man-hr/day x $2.80/man-hr x 365 days/yr)	171,700		0.6
Supervision at 15% of operating labor	25,800		0.1
Utilities (Table II-12)	371,000		1.2
Catalysts and chemicals (excerpted from Table II-13)	27,000		0.1
Maintenance at 4% of bare cost per year (1)	151,000		0.5
Supplies at 15% of maintenance	22,600		0.1
Total Direct Costs		2,486,100	
Indirect Costs:			
Payroll overhead at 20% of operating labor, maintenance labor, and supervision	60,600		0.2
General plant overhead at 50% of operating labor, supervision, maintenance and supplies	186,000		0.6
Total Indirect Costs		246,600	
Fixed Costs:			
Depreciation at 5% of fixed investment	223,000		0.8
Property taxes and insurace at 3% of fixed investment	134,000		0.4
Total Fixed Costs		357,000	
TOTAL ANNUAL OPERATING COST		3,089,700	
ESTIMATED METHANOL PRODUCTION COST			10.4

---

(1) Maintenance is 70% labor, 30% material

Table II-10

Estimated Annual Operating Cost Hydrogen from Bituminous Coal

Basis: 35,000,000 SCFD of 99.0% H_2
90% Stream Efficiency
Fixed Investment = $4,790,000 (Sections 1100 through 1400)

Item	$/Year		¢/MSCF H_2
Direct Costs:			
Dry Synthesis gas (1,754 MSCF/hr x 7,884 hr/yr x 14.5¢/MSCF)	2,000,000		17.4
Operating labor (96 man-hr/day x $2.80/man-hr x 365 days/yr)	98,100		0.8
Supervision at 15% of operating labor	14,700		0.1
Utilities (Table II-12)	308,000		2.7
Catalysts and chemicals (excerpted from Table II-13)	46,000		0.4
Maintenance at 4% of bare cost per year (1)	162,000		1.4
Supplies at 15% of maintenance	24,300		0.2
Total Direct Costs		2,653,100	
Indirect Costs:			
Payroll overhead at 20% of operating labor, maintenance labor, and supervision	45,200		0.4
General plant overhead at 50% of operating labor, supervision, maintenance and supplies	149,600		1.3
Total Indirect Costs		194,800	
Fixed Costs:			
Depreciation at 5% of fixed investment	239,500		2.1
Property taxes and insurance at 3% of fixed investment	144,000		1.3
Total Fixed Costs		383,500	
TOTAL ANNUAL OPERATING COST		3,231,400	
ESTIMATED HYDROGEN PRODUCTION COST			28.1

---

(1) Maintenance is 70% labor, 30% material

Table II-11

Estimated Annual Operating Cost
Gasoline from Bituminous Coal

Basis: 5,650 Bbl/Day of Gasoline
90% Stream Efficiency
Fixed Investment = $53,200,000 (Sections 1500 through 1700)

Item	$/year		¢/gal gasoline
Direct Costs:			
Dry Synthesis gas (10,840 MSCF/hr x 7,884 hr/yr x 14.5¢/MSCF)	12,390,000		15.9
Operating labor (1,080 man-hr/day x $2.80/man-hr x 365 days/yr)	1,104,000		1.4
Supervision @ 15% of operating labor	166,000		0.2
Utilities (Table II-12)	2,885,000		3.7
Catalysts and Chemicals (Excerpted from Table II-13)	684,000		0.9
Maintenance @ 4% of bare cost per year (1)	1,800,000		2.3
Supplies @ 15% of maintenance	270,000		0.3
Total Direct Costs		19,299,000	
Indirect Costs:			
Payroll overhead @ 20% of operating labor, maintenance labor, and supervision	506,000		0.6
General plant overhead @ 50% of operating labor, supervision, maintenance, and supplies	1,670,000		2.1
Total Indirect Costs		2,176,000	
Fixed Costs:			
Depreciation @ 5% of fixed investment	2,660,000		3.4
Property taxes and insurance @ 3% of fixed investment	1,600,000		2.1
Total Fixed Costs		4,260,000	
Co-product Credits (Table II-14)		(12,573,000)CR	(16.1)CR
TOTAL ANNUAL OPERATING COST		13,162,000	
ESTIMATED GASOLINE PRODUCTION COST			16.8

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(1) Maintenance is 70% labor, 30% material

Table II-12

Utilities Costs
Liquid Fuels and Chemicals from Bituminous Coal

Basis: Stream Efficiency = 90%
Unit Costs:

Steam:	¢/Mlb.		
1015 psia, 850° F	45	Electricity	0.8¢/kwh
500 psia, 700° F	39	Cooling Water	0.7¢/M gal
500 psia, saturated	33	Boiler Feed Water	20. ¢/M gal
190 psia, 390° F	27	Process Water	15. ¢/M gal
45 psia, saturated	20	Fuel	20. ¢/MM Btu

Plant Section	Synthesis Gas		Ammonia		Methanol		Hydrogen		Gasoline	
Product Rate	395,000,000 SCFD		600 T/D		300 T/D		35,000,000 SCFD		5,650 Bbl/Day	
Utilities Consumption:	Rate	$/year	Rate	$/year	Rate	$/year	Rate	$/year	Rate	$/year
Steam, M#/hr:										
1015 psia, 850° F	200	709,000	100	355,000	34.1	121,000	12.0	43,000	--	--
500 psia, 700° F	--	--	133	409,000	32.5	100,000	97.9	301,000	688	2,115,000
500 psia, sat'd.	925	2,405,000	7.7	20,000	--	--	--	--	--	--
45 psia, sat'd.	--	--	10	16,000	10	16,000	--	--	223	351,000
Electricity, kw	2,100	132,000	11,150	703,000	3,140	198,000	160	10,000	18,440	1,163,000
Cooling water, MGPM	41.4	137,000	42.6	141,000	15	50,000	12.2	40,000	106	350,000
Boiler feed water, M#/hr	795	151,000	50.8	10,000	31.6	6,000	55	11,000	319	60,000
Process water, M#/hr	56.1	8,000	40	6,000	31.5	4,000	--	--	60	9,000
Fuel, MM Btu/hr	35	55,000	--	--	--	--	--	--	55	87,000
Total Costs		3,597,000		1,660,000		495,000		405,000		4,135,000
Utilities Production:										
Steam, M#/hr:										
500 psia, sat'd.	742	1,930,000	31.4	82,000	--	--	--	--	78.6	204,000
190 psia, 390° F	--	--	--	--	--	--	--	--	24	51,000
45 psia, sat'd.	29.5	46,000	--	--	30.7	48,000	47.3	74,000	231	364,000
Boiler feed water heating, MM Btu/hr	--	--	--	--	--	--	14.5	23,000	400	631,000
Fuel, MM Btu/hr	769	1,213,000	57	90,000	48	76,000	--	--	--	--
Total Credits		3,189,000		172,000		124,000		97,000		1,250,000
NET UTILITIES COST		408,000		1,488,000		371,000		308,000		2,885,000

Table II-13

Catalysts and Chemicals Consumption
Liquid Fuels and Chemicals from Bituminous Coal

Basis: 90% Stream Efficiency

Section	Chemical	Quantity, lb./hr.	Unit Cost, cents/lb.	$/year
400	MEA	37	28.	82,000
	NaOH	52	3.8	16,000
800	Sponge iron	73	3	17,000
1000	Synthesis catalyst			10,000
1200	K_2CO_3	33	10	26,000
1400	MEA	3	28	7,000
	NaOH	42	3.8	13,000
1600	Sponge iron	440	3	104,000
1700	Makeup chemicals	460	15	540,000
	Iron ore for making synthesis catalyst	10(Tons/ Day)	12($/ton)	40,000
	TOTAL			855,000

Table II-14

Co-product Credits

Gasoline from Bituminous Coal

Basis: 5,650 Bbl/Day of Gasoline
90% Stream Efficiency

Co-product	Unit	Production Rate, units/hour	Unit Value, (1) cents	Credit, $/year
Ethylene	pound	6,600	4.5	2,340,000
Motor ethanol	gallon	855	45.	3,033,000
Propane LPG	gallon	985	7.6	590,000
Butane	gallon	610	8.8	423,000
Diesel oil	gallon	363	9.5	272,000
Acetic acid	pound	2,300	9.5	1,722,000
Waxy oil	gallon	225	5.0	89,000
Propanol	pound	1,350	11.	1,170,000
Propylene	pound	1,100	3.3	286,000
Acetone	pound	1,000	6.2	489,000
Propionic acid	pound	510	20.	804,000
1-Butanol	pound	440	13.	451,000
Methyl Ethyl Ketone	pound	330	12.	312,000
Acetaldehyde	pound	300	9.5	225,000
Isobutanol	pound	210	12.	199,000
Heavy alcohols	pound	142	15.	168,000
TOTAL CO-PRODUCT CREDITS				12,573,000

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(1) Credit taken at 95% of current selling price to allow for selling and distribution costs

Figure II-4
Effect of Coal Cost, Cost of Nuclear Heat, and
Return on Investment on the Selling Price of Ammonia

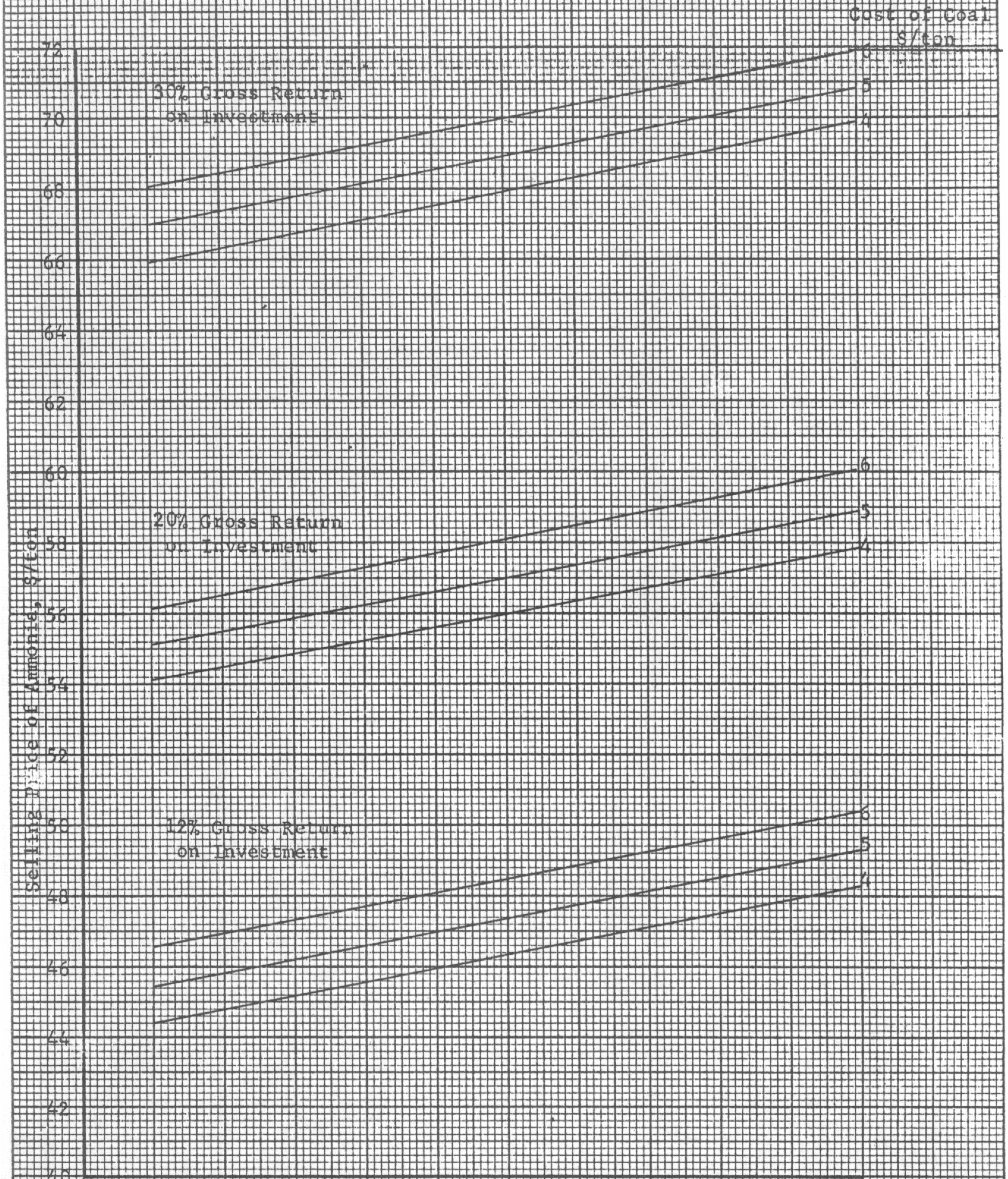


Figure II-5
Effect of Coal Cost, Cost of Nuclear Heat, and
Return on Investment on the Selling Price of Methanol

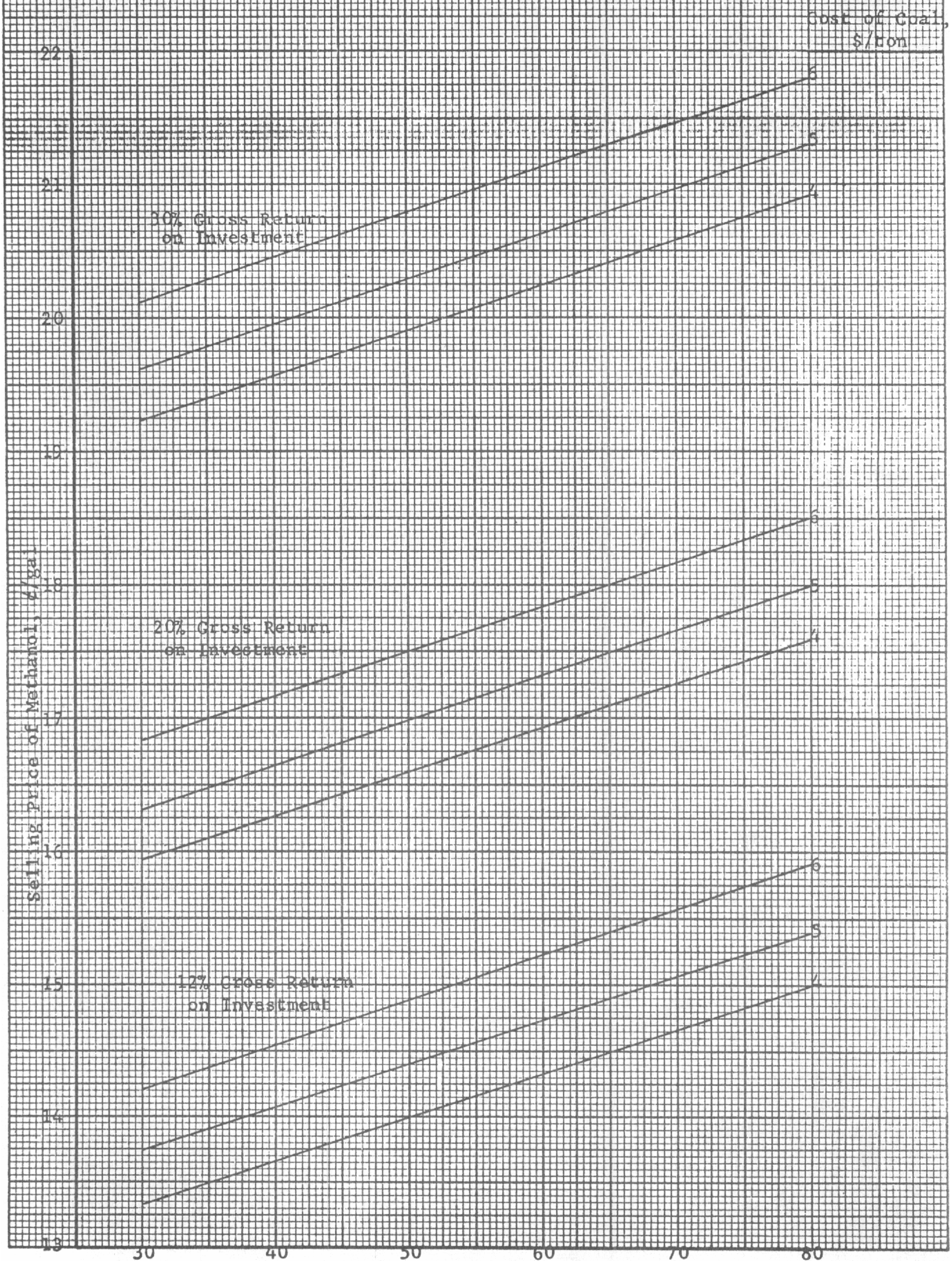


Figure II-6

Effect of Coal Cost, Cost of Nuclear Heat, and
Return on Investment on the Selling Price of Hydrogen

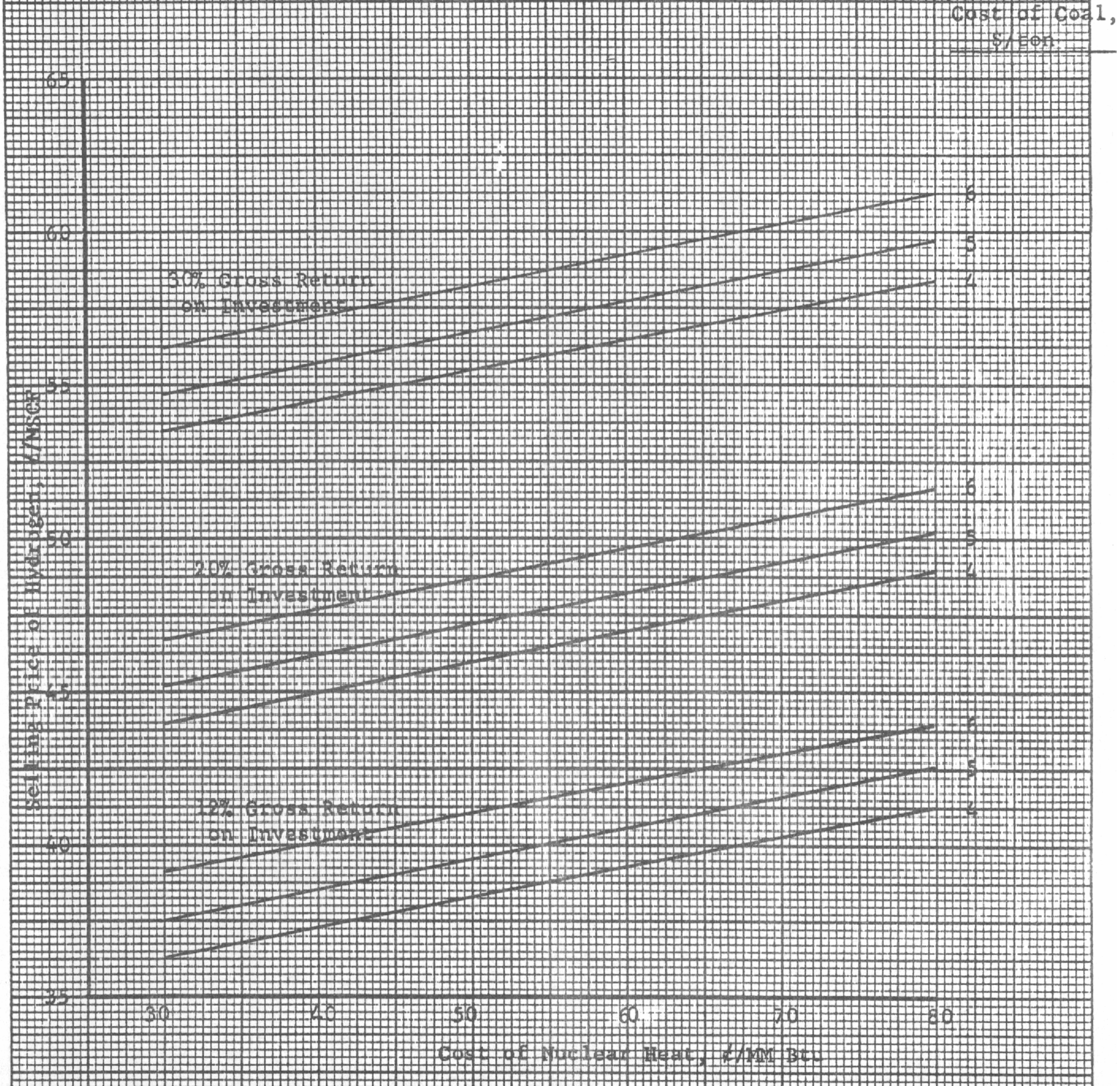


Figure II-7

Effect of Coal Cost, Cost of Nuclear Heat, and Return on Investment on the Selling Price of Gasoline

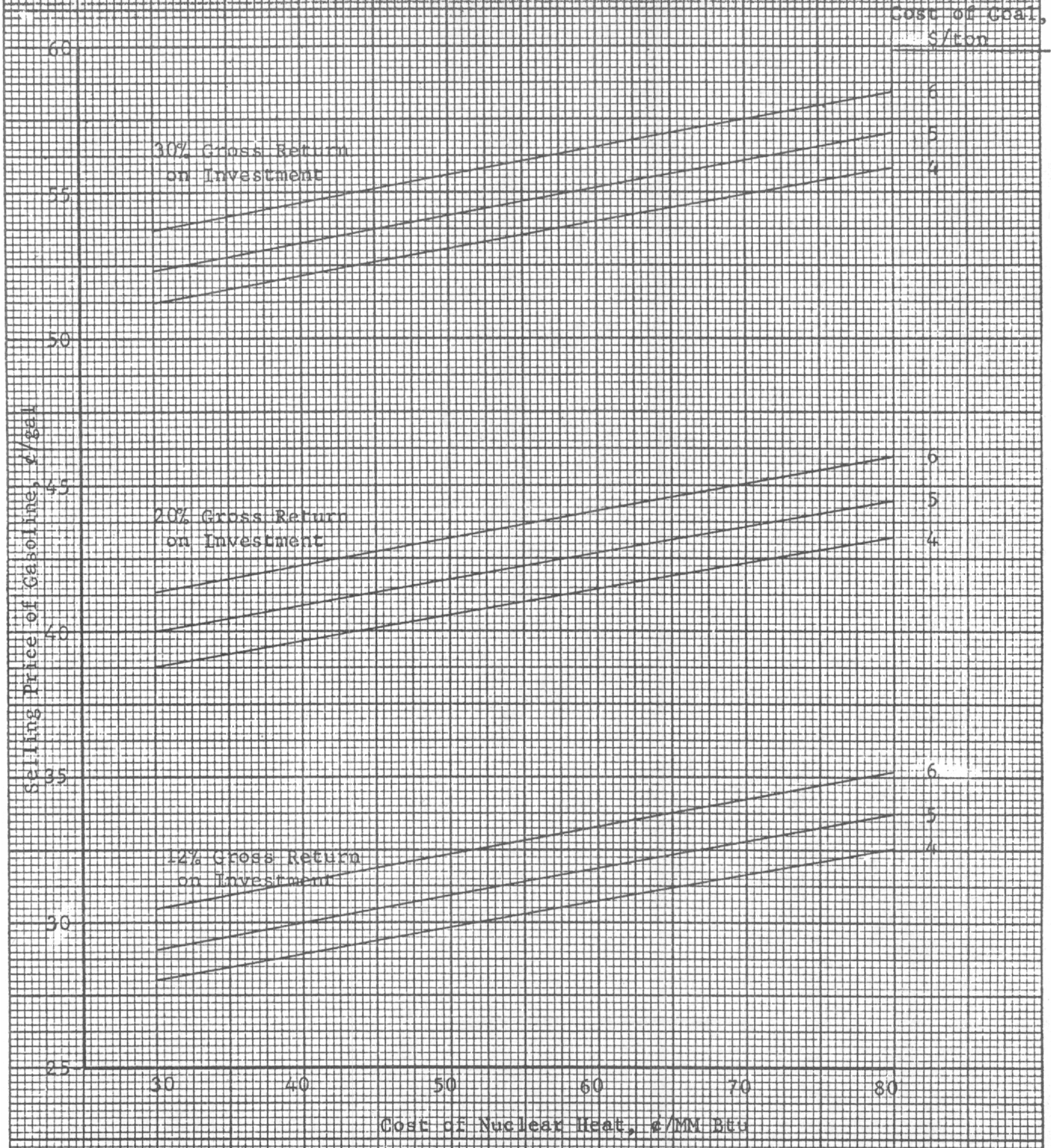
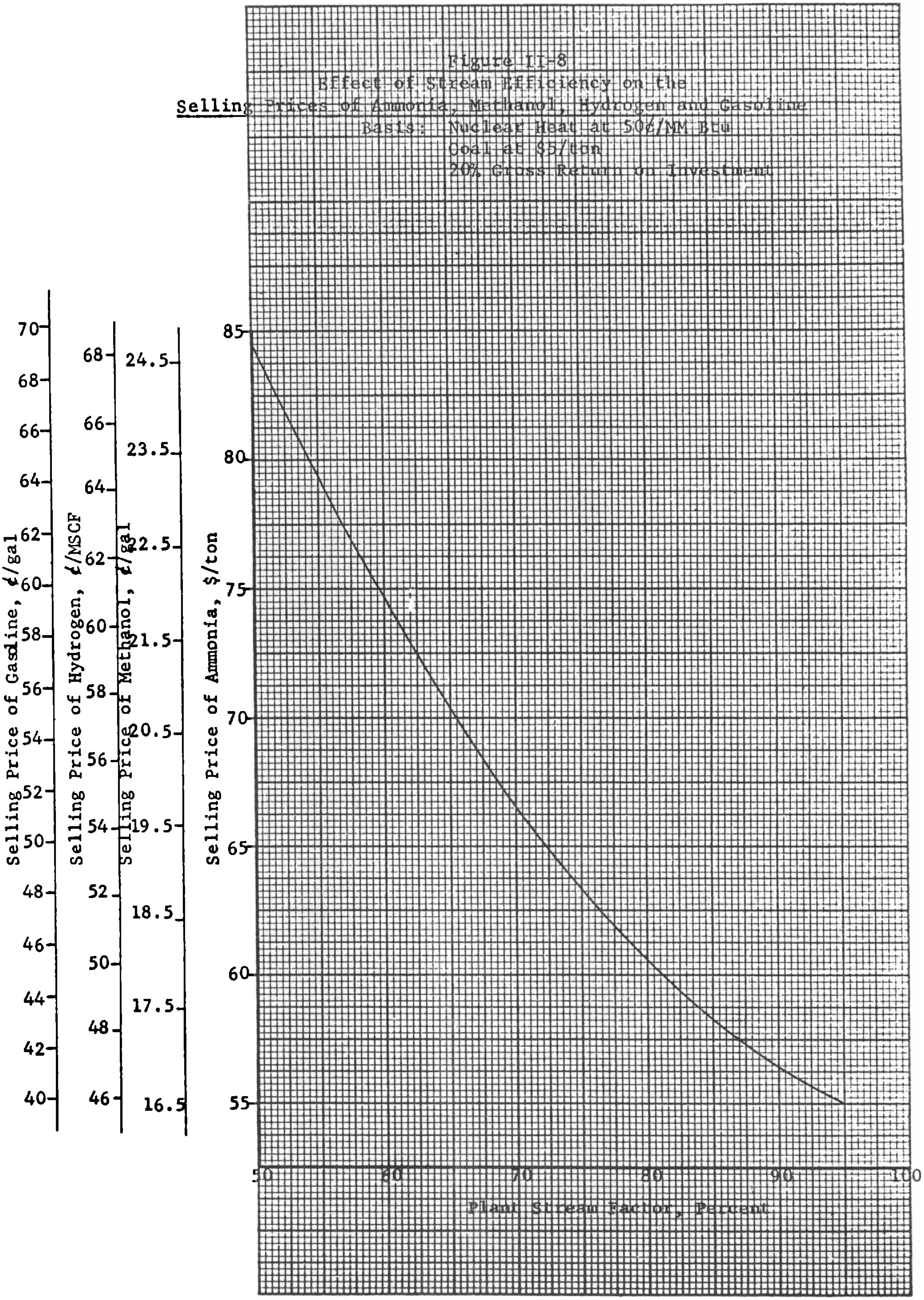


Figure II-8
Effect of Stream Efficiency on the
Selling Prices of Ammonia, Methanol, Hydrogen and Gasoline
Basis: Nuclear Heat at 50¢/MM Btu
Coal at $5/ton
20% Gross Return on Investment



3. Design Considerations

a. Coal Pretreatment Process

The low-temperature carbonization pretreatment scheme is the same as that used in the Pipeline Gas Plant. The design bases for this operation have been discussed in detail in Section II-A-3-a.

b. Fluidized Bed Gasification

The gasification sequence used here is exactly the same as that in the Pipeline Gas Plant. The reader is referred to Section II-A-3-b for a thorough discussion of the design bases and assumptions for this process.

c. End-Product Synthesis

Design considerations for the processing schemes in the ammonia, methanol, hydrogen, and liquid fuels plants have already been discussed in Section I-B-3, of this report, and no further elaboration is necessary. Suffice it to say that all of these processes represent fields of well-developed technology and that no further development work is necessary to design and build these sections of the plant.

4. Optimization of Gasification Process Variables

The gasification process variables chosen for the design of this plant are exactly the same as those selected for use in the Pipeline Gas Plant. The optimization of these variables was exactly the same as was done for pipeline gas, details of which are presented in Section II-A-4 of this report.

Although there might be some economic incentive in gasifying at pressures higher than 450 psia in preparation for the ammonia and methanol syntheses, it should be a relatively small one on a plant-wide scale inasmuch as these two plants use a total of only about 22 percent of the raw synthesis gas. Also, a thorough evaluation of higher gasification pressures would have to include the effect of a high-pressure helium loop (required for balanced pressures in the gasifier tubes) on the cost of the nuclear reactor, which is beyond the scope of this work.

Providing a separate gasification step for each of the four plants might prove to be a more efficient means of producing the raw synthesis gas. For example, air could be added to the gasifier in the ammonia plant to provide the nitrogen required for synthesis. Likewise, carbon dioxide could be introduced into the gasifiers in the methanol and liquid fuels plants to reduce the effluent H_2/CO ratio, thus providing raw gases with compositions more favorable for these syntheses. It was agreed, however, that these variations in the designs would not be considered in the present study, although they could conceivably merit further investigation in the future.

5. Evaluation of Alternate Processing Schemes

The alternate pretreatment and gasification schemes are the same as those considered for pipeline gas in Section II-A-5. The conclusions reached there apply in the present situation, and no further discussion need be given.

Appendix

Fuels and Chemicals from Bituminous Coal

Material Balances

Sections 100 and 150

Input:	Lb/Hr
Raw coal from Section 000	350,000
Air to compressors	190,200
Char quench water	27,000
Process water to wet dust collector	39,080
Char and steam from cyclones 204-G	29,920
	636,200

Output:	Lb/Hr
Char and steam to cyclones 204-G	306,920
Tar product to pump 203-J	32,995
Condensate from separator	20,000
Waste gas from wet dust collector	163,720
Slurry from wet dust collector	34,080
Make gas from tar recovery section	78,485
	636,200

Section 200

Input:	Lb/Hr
Char and steam from Section 150 to 204-G	306,920
Tar product from Section 150 to 203-J	32,995
Coal and char from settling pond to 202-L	4,080
Steam to 201-D	925,000
Residue quench water to 203-F	405,000
	1,673,995

Output:	Lb/Hr
Raw synthesis gas to ammonia plant	165,985
Raw synthesis gas to methanol plant	90,359
Raw synthesis gas to hydrogen plant	123,715
Raw synthesis gas to liquid fuels plant	778,926
Residue slurry from 203-F	485,090
Char and steam from 204-G to Section 150	29,920
	1,673,995

Section 300 (Ammonia Plant)

Input:	Lb/Hr
Raw synthesis gas from Section 200	165,985
Steam to shift converters	133,146
Boiler feed water to shift converters	18,432
	317,563

Output:	Lb/Hr
Synthesis gas to Section 400	111,169
Lock hopper pressurizing gas to 202-F	5,896
Condensate from separator	200,498
	317,563

Section 400 (Ammonia Plant)

Input:	Lb/Hr
Synthesis gas from Section 300	111,169
Nitrogen from Section 1700	44,430
	155,599

Output:	Lb/Hr
Synthesis gas to Section 500	55,887
Waste gas from CO_2 stripper	98,275
H_2S from caustic scrubber	54
COS from activated carbon drums	6
Condensate from separator	1,377
	155,599

Section 500 (Ammonia Plant)

Input:	Lb/Hr
Synthesis gas from Section 400	55,887

Output:	Lb/Hr
Synthesis gas to Section 600	55,693
Condensate from separator	194
	55,887

Section 600 (Ammonia Plant)

Input:	Lb/Hr
Synthesis gas from Section 500	55,693

Output:	Lb/Hr
Ammonia product	50,000
Purge gas to fuel	5,693
	55,693

Section 700 (Methanol Plant)

Input:	Lb/Hr
Raw synthesis gas from Section 200	105,946

Output:	Lb/Hr
Synthesis gas to Section 800	57,324
Condensate from separator	48,622
	105,946

Section 800 (Methanol Plant)

Input:	Lb/Hr
Synthesis gas from Section 700	57,324

Output:	Lb/Hr
Synthesis gas to Section 900	30,896
Waste gas from flash drum	26,223
H_2S from iron oxide drums	31
COS from activated carbon drums	174
	57,324

Section 900 (Methanol Plant)

Input:	Lb/Hr
Synthesis gas from Section 800	30,896

Output:	Lb/Hr
Synthesis gas to Section 1000	30,833
Condensate from separator	63
	30,896

Section 1000 (Methanol Plant)

Input:	Lb/Hr
Synthesis gas from Section 1000	30,833
Process water	1,515
	32,348

Output:	Lb/Hr
Methanol product	25,000
Purge gas to fuel	3,128
Waste gas	1,440
Condensate from separator	2,780
	32,348

Section 1100 (Hydrogen Plant)

Input:	Lb/Hr
Raw synthesis gas from Section 200	123,715
Steam to shift converter	34,920
Boiler feed water to shift converter	6,282
	164,917

Output:	Lb/Hr
Synthesis gas to Section 1200	90,616
Condensate from separator	74,301
	164,917

Section 1200 (Hydrogen Plant)

Input:	Lb/Hr
Synthesis gas from Section 1100	90,616

Output:	Lb/Hr
Synthesis gas to Section 1300	16,863
Waste gas from CO_2 stripper	73,753
	90,616

Section 1300 (Hydrogen Plant)

Input:	Lb/Hr
Synthesis gas from Section 1200	16,863
Steam to shift converter	63,018
	79,881

Output:	Lb/Hr
Synthesis gas to Section 1400	16,626
Condensate from separator	63,255
	79,881

Section 1400 (Hydrogen Plant)

Input:	Lb/Hr
Synthesis gas from Section 1300	16,626

Output:	Lb/Hr
Hydrogen product	8,532
Waste gas from CO_2 stripper	7,846
Condensate from separator	248
	16,626

Section 1500 (Liquid Fuels Plant)

Input:	Lb/Hr
Raw synthesis gas from Section 200	763,339

Output:	Lb/Hr
Synthesis gas to Section 1600	413,563
Condensate from separator	349,776
	763,339

Section 1600 (Liquid Fuels Plant)

Input:	Lb/Hr
Synthesis gas from Section 1500	413,563

Output:	Lb/Hr
Synthesis gas to Section 1700	267,052
Waste gas from flash drum	144,562
H_2S from iron oxide drums	683
COS from activated carbon drums	1,266
	413,563

Section 1700 (Liquid Fuels Plant)

Input:	Lb/Hr
Synthesis gas from Section 1600	267,052
Makeup chemicals	460
Steam	95,100
Makeup water	2,440
Air to oxygen plant	180,000
	545,052

Output:	Lb/Hr
Gasoline	58,200
Ethylene	6,600
Motor Ethanol	5,600
Propane LPG	4,200
Butane	3,000
Diesel Oil	2,500
Acetic Acid	2,300
Waxy Oil	1,700
Propanol	1,350
Propylene	1,100
Acetone	1,000
Propionic Acid	510
1-Butanol	440
MEK (methyl ethyl ketone)	330
Acetaldehyde	300
Isobutanol	210
Heavy Alcohols	142
High-purity Nitrogen to Section 400	44,430
Waste Nitrogen	93,870
Waste Gas	55,730
Condensate	261,000
Coke	540
	545,052

Overall Energy Balance

Datum Temperature = 60°F

Input:	MM Btu/Hr	Percent of Total
Coal to process	4,530	59.8
Coal to boilers	1,662	21.9
Heat transferred from helium	1,385	18.3
	7,577	100.0

Output:	MM Btu/Hr	Percent of Total
Ammonia product	442	5.8
Methanol product	240	3.2
Hydrogen product	470	6.2
Gasoline	1,175	15.5
Miscellaneous fuels and chemicals from Section 1700	520	6.9
Losses to cooling water	3,294	43.5
Electric power consumption	140	1.8
Sensible and latent heat of waste gases	388	5.1
Sensible heat of water purge streams	96	1.3
Losses from boiler plant (unburned fuel, sensible heat of flue gas, uncondensed water, convection loss, etc.)	350	4.6
Convection losses, miscellaneous, etc.	462	6.1
	7,577	100.0

Thermal Efficiency, (percent of total energy input appearing in combined product streams) = 37.6%

Utilities Summary

Steam

I. 1015 psia, 850°F

A. Generation

Steam is generated at 1015 psia, 850°F in two boilers each one capable of generating 875,000 pounds per hour. Fuel for the boilers includes purge gases from Sections 600 and 1000, residue from the gasifiers in Section 200, and pulverized coal.

About 10 percent excess capacity is available during normal operation.

Section	Normal Generation, Lb/Hr
3000	1,552,104

B. Consumption

Section	Normal Consumption, Lb/Hr
100 and 150	17,800
200 - Turbine drives on helium circulators, 201-J	182,000
500	100,400
900	34,100
1200	8,520
1400	3,500
3000	1,205,784
	1,552,104

II. 500 psia, 700°F

A. Generation

Steam at 500 psia, 700°F is generated in Section 3000 from 1015 psia, 850°F steam as exhaust steam from one of the turbogenerators.

Section	Normal Generation, Lb/Hr
3000	951,584

B. Consumption

Section	Normal Consumption, Lb/Hr
300	133,146
1000	32,500
1100	34,920
1300	63,018
1700	688,000
	951,584

III. 500 psia, saturated

A. Generation

Steam at 500 psia, 467°F is generated in Section 3000 in a boiler capable of producing 160,000 pounds per hour. The remainder of this steam required for the process is generated by waste heat throughout the plant.

Section	Normal Generation, Lb/Hr
200 - Waste Heat Boilers, 204-C	742,000
300	31,420
1500	78,650
3000	140,630
	992,700

B. Consumption

Section	Normal Consumption, Lb/Hr
200 - Gasifiers, 201-D	925,000
400	7,700
3000	60,000
	992,700

IV. 190 psia, 390°F

A. Generation

Section	Normal Generation, Lb/Hr
1700	24,000

B. Consumption

Section	Normal Consumption, Lb/Hr
3000	24,000

V. 45 psia, saturated

A. Generation

Section	Normal Generation, Lb/Hr
150	29,500
700	30,650
1300	47,340
1500	230,510
	338,000

B. Consumption

Section	Normal Consumption, Lb/Hr
400	10,000
800	10,000
1600	30,000
1700	193,000
3000	95,000
	338,000

Power

A. Generation

Electric power is produced at 13,800 volts by five 9,000-kw turbogenerators with steam turbine drives. About 10 percent excess capacity is available during normal operation. An electric substation is provided to reduce the voltage to 4160, 440 and 110 volts.

Section	Normal Generation, kw
3000	41,000

B. Consumption

Consumption	Normal Consumption, kw
000	400
100	1,500
200	200
300	40
400	50
600	11,060
700	20
800	2,000
1000	1,120
1100	20
1200	50
1300	20
1400	70
1500	40
1600	5,000
1700	13,400
3000	6,010
	41,000

C. Summary for Section 200

Item	Normal Consumption, kw
202-J, Water Pumps	40
203-J, Tar Pumps	50
201-L, Coal Distributors	100
202-L, Coal Elevator	10
	200

Cooling Water

A. Generation

Cooling water is available from the nearby Monongahela River. During the warm summer months, however, when the river is normally low, it may prove inadequate as a heat sink. Accordingly, the plant cooling system is based on tower cooling of recirculated water, with the necessary makeup water obtained from the river.

Cooling water is available throughout the plant at 40 psig and a maximum temperature of 85°F.

Section	GPM
3000	278,212

B. Consumption

Section	GPM
150	6,130
200	35,310
300	3,660
400	14,505
500	21,615
600	2,840
700	3,080
900	7,232
1000	4,700
1200	7,770
1300	2,540
1400	1,890
1500	22,440
1700	84,000
3000	60,500
	278,212

C. Summary for Section 200

Item	GPM
201-J, Surface condensers on steam turbines	34,500
203-F, Residue quench	810
	35,310

Boiler Feed Water

A. Generation

Boiler feed water at 250°F is produced by the following sequence:

(1) River water is filtered and treated with chemicals to remove dissolved solids.

(2) Further purification is effected by passing the treated water over anion- and cation-exchange resins.

(3) Purified makeup water is mixed with returning condensate and heated to about 200°F in condensate heaters in Section 1700.

(4) Low-pressure steam is used for deaeration.

(5) Deaerated water is heated to about 250°F in boiler feed water heaters in Sections 1100 and 1700.

Section	Normal Generation, Lb/Hr
1100	403,000
1700	2,593,114
	2,996,114

B. Consumption

Section	Normal Consumption, Lb/Hr
150	29,500
200 - Waste Heat Boilers, 204-C	765,000
300	50,832
700	31,600
1100	6,282
1300	48,800
1700	319,100
3000	1,745,000
	2,996,114

Process Water

A. Generation

Process water at 90°F is produced by filtering river water and treating it with chemicals to precipitate dissolved solids.

Section	Normal Generation, Lb/Hr
3000	187,645

B. Consumption

Section	Normal Consumption, Lb/Hr
150	56,080
300	40,000
700	30,000
1000	1,565
1500	60,000
	187,645

Fuels

A. Generation

Section	Item	MM Btu/Hr
000	Raw coal to boilers	1,662
150	Net make gas	158
600	Purge gas	57
1000	Purge gas	48
3000	Char residue from settling pond	611
		2,536

B. Consumption

Section	Item	MM Btu/Hr
100	Dryer Furnaces	35
1700	-	55
3000	Boilers	2,446
		2,536

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