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ASSESSMENT OF VERY HIGH-TEMPERATURE REACTORS IN PROCESS APPLICATIONS. APPENDIX III. ENGINEERING EVALUATION OF PROCESS HEAT APPLICATIONS FOR VERY-HIGH TEMPERATURE REACTORS

UNITED ENGINEERS AND CONSTRUCTORS, INC.
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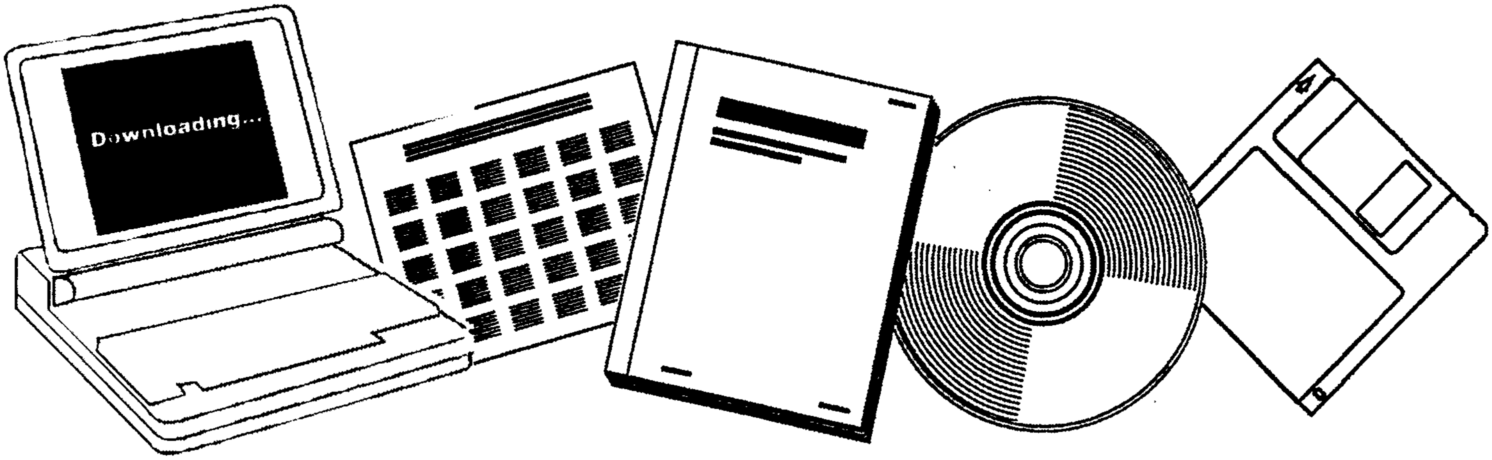

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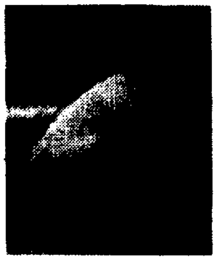

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Engineering Technology Division

ASSESSMENT OF VERY HIGH-TEMPERATURE REACTORS IN PROCESS APPLICATIONS

APPENDIX III — ENGINEERING EVALUATION OF PROCESS HEAT APPLICATIONS FOR VERY-HIGH TEMPERATURE REACTORS

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The complete report is composed of the following:

ORNL/TM-5242 — Assessment of Very High-Temperature Reactors in Process Applications

ORNL/TM-5409 — Evaluation of the Reactor System
(Appendix I)

ORNL/TM-5410 — Very High-Temperature Reactor Process Heat Application Studies
(Appendix II)

ORNL/TM-5411 — Engineering Evaluation of Process Heat Applications for Very High-Temperature Reactors
(Appendix III)

PREPARED FOR

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REVISION RECORD

Revision #1	Updated capital costs, prepared new sections on methanol and comparison of systems.	3/9/76
Revision #2	Revised capital cost of Westinghouse VHTR.	5/10/76

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FOREWORD

This one of a series of reports on nuclear process heat. The overall summary is *Assessment of Very High-Temperature Reactors in Process Applications* (ORNL/TM-5242). Details and background information are presented in Appendix I — *Evaluation of the Reactor System* (ORNL/TM-5409); Appendix II — *VHTR Process Application Studies* (ORNL/TM-5410): and Appendix III — *Engineering Evaluation of Process Heat Applications for VHTRs* (ORNL/TM-5411).

ABSTRACT

An engineering and economic evaluation is made of coal conversion processes that can be coupled to a very high-temperature nuclear reactor heat source. The basic system developed by General Atomic/Stone & Webster (GA/S&W) is similar to the H-coal process developed by Hydrocarbon Research, Inc., but is modified to accommodate a nuclear heat source and to produce synthetic natural gas (SNG), synthesis gas, and hydrogen in addition to synthetic crude liquids. The synthetic crude liquid production is analyzed by using the GA/S&W process coupled to either a nuclear- or fossil-heat source. Four other processes are included for comparison: (1) the Lurgi process for production of SNG, (2) the Koppers-Totzek process for production of either hydrogen or synthesis gas, (3) the Hygas process for production of SNG, and (4) the Westinghouse thermal-chemical water splitting process for production of hydrogen. The production of methanol and iron ore reduction are evaluated as two potential applications of synthesis gas from either the GA/S&W or Koppers-Totzek processes. The results indicate that the product costs for each of the gasification and liquefaction processes did not differ significantly, with the exception that the unproven Hygas process was cheaper and the Westinghouse process considerably more expensive than the others.

1. INTRODUCTION

1.1 Authorization

United Engineers & Constructors Inc. has evaluated coal liquefaction and gasification processes using nuclear and fossil fuels to provide process heat. This evaluation was supported by the United States Energy Research and Development Administration under contract number AT (11-1) - 2477.

The initial evaluation report was published in June 1975. As a result of recent studies by United Engineers and General Atomic, the June 1975 report has been revised and is herewith reissued.

Under the ERDA contract and a subcontract with Union Carbide Corporation Nuclear Division, United Engineers has prepared the following related studies:

- A Cost Comparison of Very High Temperature Nuclear Reactors for Process Heat Applications, March 1975.
- Evaluation of a Coal Liquefaction Process Using Either a Nuclear or Fossil Heat Source, February 1976.
- Evaluation of Pollution and Water Consumption by Selected Coal Conversion Processes, February 1976.

1.2 General Conclusions

Coal gasification and liquefaction processes and costs have been reviewed. Nuclear-heated processes based on 3000 MWt reactors were compared to the principal fossil-heated coal conversion processes of similar size. The published processes are technically feasible, and the costs are consistent with United Engineers' experience.

If a 15 percent fixed charge rate and $\$0.50/10^6$ Btu coal can be obtained, and costs published by proponents of these processes are valid:

- Coal can be converted to pipeline gas at a cost competitive with unregulated natural gas.
- Coal can be converted to heavy naphtha and heavy oils at a cost competitive with the present cost of imported oil.
- Synthesis gas ($CO + H_2$) which is obtained from coal can be converted to methanol which is significantly cheaper than the current market price of methanol. This application is the most promising under current economic conditions.
- Synthesis gas is also attractive for the reduction of iron ore.

The costs of coal conversion are very sensitive to the fixed charge rate and the coal costs, and may not be competitive for higher values.

At low coal costs, nuclear-heated processes are more expensive than fossil-heated. At high coal costs, nuclear processes become less expensive than coal.

The applicability of nuclear-heated coal conversion is limited because the size of the plant is too great for many mine-mouth operations. Much smaller fossil-heated plants can be built without a significant increase in product cost.

1.3 Scope of Evaluation

United Engineers reviewed coal conversion processes which can be combined with a very high temperature nuclear reactor (VHTR) heat source. The basic process was developed by General Atomic Company and Stone & Webster Engineering Company (GA/S&W) and produces synthetic pipeline gas. This design was modified by United Engineers to produce synthesis gas or hydrogen. General Atomic later issued a similar modification for hydrogen production.

The preliminary design of a coal liquefaction process was prepared by United Engineers. This process is similar to the H-Coal process developed by Hydrocarbon Research Inc. but can be heated either by a nuclear or fossil heat source. Because the processes are identical except for the heat source, comparison of the costs is more reliable than for other processes with different bases.

Four other processes were selected for comparison with the nuclear-heated GA/S&W processes:

- The Lurgi pipeline gas process is available and operating commercially.
- The Hygas process is in a promising stage of development for pipeline gas production.
- The Koppers-Totzek processes for synthesis gas and hydrogen are commercially available with many plants operating successfully.

- The thermo-chemical hydrogen process recently proposed by Westinghouse is an alternate source of hydrogen.

These processes, and the products derived from them, are shown in Table 1-1.

Two potential applications of synthesis gas were reviewed for synthesis gas produced by the GA/S&W and Koppers-Totzek processes:

- Methanol is a major product manufactured from synthesis gas and it appears to have a rapidly growing future market.
- Iron ore can be reduced with synthesis gas; this process was compared to conventional ore reduction methods.

1.4 Results and Specific Conclusions

1.4.1 Verification of Reported Data

Heat and material balances and thermochemical calculations were performed to verify product yields and energy requirements under the conditions cited in the literature. Reported costs were compared with United Engineers' data base which has been developed from many chemical plant projects. These calculations and comparisons confirmed the basic chemical, energy and cost parameters reported for the selected processes.

1.4.2 Limits of Comparison - Basis

The processes evaluated in this report can be compared on a general basis. The process parameters and costs are taken from descriptions published by the developers and proponents. The two coal liquefaction processes were designed by United Engineers using a common basis, and are thus directly comparable. Each gasification process is based on a different coal and a different site, and has a unique product mix. A detailed comparison of gasification processes is misleading because the data reported is not based on a consistent set of initial assumptions.

Direct, detailed comparison of the gasification processes would require extensive preliminary design to place all processes on the same basis. Comparison of these processes using a common basis would inevitably place some processes at a disadvantage because some processes operate on a wider variety of coals than others.

For coal gasification, the fossil-heated processes are based on coal with a heating value of 8600-8900 Btu/lb, while the nuclear-heated processes use 12,000 Btu/lb coal. Comparing processes at a specific coal price penalizes the fossil-heated processes which use cheaper coal. The difference between fossil and nuclear heating is thus greater than is immediately obvious.

1.4.3 Limits of Comparison - Products

The product mix of each process is shown in Table 1-2. Most of the gasification processes also produce oil, naphtha or other liquids. When the energy output of the plant is calculated, these liquids may constitute a significant fraction of the plant production.

For this evaluation, by-products such as naphtha are considered to be products if the published description of the process treats them as products. As a result, tar and tar oil produced by the Lurgi process are products in this report, even though their market value (per unit energy) is lower than pipeline gas. Char produced by several processes has a heating value of 7800 to 8300 Btu/lb, and typically contains about 35 percent ash and 5 percent sulfur. Although this is similar to some natural coals, and thus may have value, it was not considered as a product because the process designers treat it as a waste product.

The choice between products and waste products has a significant effect on the efficiency and average product costs which are reported.

The efficiencies and costs reported in Tables 1-2 and 1-3 are based on the products reported in the literature. If the costs of the primary products are desired, the probable value of each by-product must first be determined. The marketability of the by-products is strongly related to plant size and location, thus making comparison of processes very difficult.

1.4.4 Comparison of Product Costs

The average product cost of each of the gasification and liquefaction processes reviewed is presented in Table 1-3. This cost is computed by dividing the total annual operating cost by the total energy available in the products defined in Table 1-2. Two costs are presented; the low cost is based on coal at $\$0.50/10^6$ Btu and a 15 percent fixed charge rate; the higher cost is calculated from $\$1.75/10^6$ Btu coal and 25 percent fixed charge rate.

There is not a significant product cost difference between the processes, with two exceptions:

- For synthetic pipeline gas production, the unproved Hygas process appears to be considerably cheaper than competing processes, due mainly to more efficient initial conversion of coal to methane. Costs of further processing are thus much less than other processes. In addition, the Hygas process has a higher fraction of pipeline gas in the total product and is less sensitive to by-product costs.

- The Westinghouse Hydrogen process is much more expensive, both for capital and operating costs, and much less energy efficient. This process is intended for the longer term when coal is in short supply, and is not competitive in current or mid-term markets.

In Figure 1-1, the costs of the various products are compared with the current costs of oil and natural gas and the assumed cost of coal. All costs in this chart represent the low cost basis defined in Section 2, and are thus the most optimistic estimates. Three natural gas prices are given. The lowest, $\$0.41/10^6$ Btu, is the current price for Federal Power Commission regulated gas. The other prices, \$1.25 and \$2.32, represent the range of unregulated natural gas prices during 1975. These prices all represent long-term contracts. Synthetic natural gas produced from naphtha is used by several utilities, and has recently sold as high as $\$4.75/10^6$ Btu (Wall Street Journal, December 25, 1975).

All of the synthetic natural gas processes have an average product cost which is competitive with the current range of unregulated prices. However, these prices depend on cheap coal and a relatively low fixed charge rate. If these conditions are not available, synthetic natural gas is likely to be much less attractive.

Also, if cheap coal and a low fixed charge rate are available, coal liquefaction is competitive with imported oil at \$13.50 per barrel.

1.4.5 Comparison of Nuclear and Fossil Heating

The coal liquefaction plants were designed by United Engineers using the same design bases. The two plants thus provide the most reliable comparison of nuclear and fossil process heating. Figure 1-2 shows the product costs of these two plants as a function of the price of coal and the fixed charge rate.

Because the nuclear plant has a higher coal conversion efficiency, it produces coal liquids more cheaply than the fossil-heated plant when coal is expensive. However, when coal is in the range of \$0.50 to \$1.00/10^6 Btu, the products of the fossil-heated plant are cheaper.

There is not a significant difference in the energy efficiency of nuclear-and fossil-heated process. The energy efficiencies of the coal liquefaction processes, in which the same bases were used for design, are identical. However, the nuclear-heated processes convert a much higher fraction of the feed carbon into useable products, which is desirable when hydrocarbons are in short supply.

The cost estimate of the VHTR plant includes an intermediate helium loop which places the reformers away from the nuclear system. If this intermediate loop is not required by the regulatory authorities, the VHTR cost would be reduced by \$100 million, with a corresponding reduction of about 5 percent in the product cost.

1.4.6 Plant Size

All plants in this report were sized either to use all heat from a 3000 MWt VHTR or to produce the same quantity of primary product as the VHTR plant. All of the fossil-heated plants consist of 8 to 15 parallel trains of standard equipment modules. For the cost estimate, United Engineers assumed no economy of scale when multiple trains are used.

The product costs of fossil-heated plants are thus applicable to much smaller plants, whereas the nuclear-heated product costs can be expected to be higher for a 2000 MWt or 1000 MWt plant. The VHTR is based on HTGR technology, so the effect of cutting the size by more than 1/3 is unknown.

The coal liquefaction plant which is based on a 3000 MWt VHTR consumes as much coal as the largest strip mine in the United States. Only ten mines in this country are capable of supplying half or more of the demand of a 3000 MWt VHTR (Coal Age, February 1976). The market for a mine mouth plant of this size is therefore very limited. However, synthesis gas and pipeline gas plants use smaller quantities of coal and would be somewhat more widely applicable.

Because the VHTR consists of several coolant loops, it is suitable for hybred process heat-electric generation operation, with some loops used for process heat and steam generation and others for high-efficiency combination of direct drive (Brayton Cycle) turbines and a bottoming cycle. This combination would produce synthetic fuels at the costs reported in this report, but would use coal at a more suitable rate. This combined use of a

VHTR plant seems economically more feasible than a single-product plant and should be studied in more detail.

1.4.7 Methanol Production

Synthesis gas produced by the Koppers Totzek and GA/S&W processes can be used to produce methanol at costs which are cheaper than the current methanol market value ($0.40 per gallon) if moderate coal prices and fixed charge rates can be obtained. The GA/S&W gas requires more processing and produces a significant quantity of other products, while the Koppers-Totzek gas can be converted to methanol without by-products. The values assigned to the GA/S&W by-products determine the relative prices of the two processes. Table 1-4 shows the costs of producing methanol using synthesis gas from these processes.

The production of methanol appears to be the most attractive application of coal gasification under present economic conditions.

1.4.8 Iron Ore Reduction

The costs of reducing iron ore using synthesis gas from the GA/S&W and Koppers-Totzek processes are compared to standard ore reduction processes in Table 1-5. The coal conversion processes are more expensive than the direct reduction process, which is based on reforming natural gas. However, this estimate is based on a natural gas price of $\$0.79/10^6$ Btu. If natural gas costs more than $\$1.40/10^6$ Btu (less than the current unregulated price), then ore reduction using gasified coal is economically attractive.

1.4.9 Hydrogen Production

The Westinghouse hydrogen production process, using VHTR nuclear heat and water splitting, does not appear competitive with nearer term energy sources based on coal, or on hydrogen from coal using a VHTR.

TABLE 1-1

PROCESS APPLICATIONS STUDIED

Process	Pipeline Gas	Crude Liquids	H_2	Syn-Gas for Steel	Syn-Gas for Methanol
General Atomic Stone & Webster	X	X	X	X	X
Lurgi	X				
Hygas	X				
Koppers-Totzek			X	X	X
Coal Liquefaction/Fossil		X			
Coal Liquefaction/Nuclear		X			
Westinghouse			X		

TABLE 1-2

COAL CONVERSION PRODUCTS AND EFFICIENCY

Process	Products	Energy Output 10^9 Btu/hr	% Efficiency
GA/S&W Synthetic Pipeline Gas	Pipeline Gas	24.5	
	Light Aromatic Liquid	5.0	
	Total	29.5	67
Lurgi Synthetic Pipeline Gas	Pipeline Gas	24.4	
	Naptha	0.79	
	Tar	2.82	
	Tar Oil	1.54	
	Total	29.55	68
Hygas Synthetic Pipeline Gas	Pipeline Gas	26.50	
	Heavy Oil	2.35	
	Benzene	0.64	
	Total	29.49	66
GA/S&W Synthesis Gas	Synthesis Gas	15.54	
	Light Aromatic Liquid	2.76	
	Total	18.30	64
Koppers/Totzek Synthesis Gas	Synthesis Gas	15.48	55
GA/S&W Hydrogen	Hydrogen	15.56	
	Light Aromatic Liquid	2.76	
	Total	18.32	64
Koppers-Totzek Hydrogen	Hydrogen	16.97	55
Westinghouse Hydrogen	Hydrogen	4.8	45
Coal Liquefaction Fossil Heat	Heavy Naphtha	13.92	
	Mid-Distillate	12.42	
	Heavy Oil	10.30	
	Heavy Gas Oil	1.83	
	Total	38.47	65
Coal Liquefaction Nuclear Heat	Heavy Naphtha	16.88	
	Mid-Distillate	12.42	
	Heavy Oil	11.92	
	Heavy Gas Oil	1.83	
	Coker Gas	2.17	
	Total	45.22	65

Note: Data presented are not directly comparable. See Section 1.4 for discussion.

TABLE 1-3

PROCESS COST COMPARISONS

JULY 1974 DOLLARS

Coal Conversion Process	Primary Product	Capital Cost $ Million	Average Product Cost (2) $/10^6 Btu of Product (1)
GA/S&W	Pipeline gas	1,594.0	1.87 - 4.12
Lurgi	Pipeline gas	1,089.1	1.79 - 4.09
Hygas	Pipeline gas	784.8	1.47 - 3.68
GA/S&W	Syn-gas	1,363.9	2.22 - 4.63
Koppers-Totzek	Syn-gas	1,073.	2.34 - 5.47
GA/S&W	Hydrogen	1,466.4	2.33 - 4.95
Koppers-Totzek	Hydrogen	1,180.	2.34 - 5.45
Westinghouse	Hydrogen	1,178.	5.17 - 8.80
Heated/Fossil	Liquids	1,241.	1.70 - 4.04 (3)
Heated/Nuclear	Liquids	1,909.	1.80 - 4.06 (4)

NOTES:

(1) Includes all products shown in Table 1-2.

(2) For Low and High Cost Assumptions - See Section 2 Basis of Study.

(3) Equivalent to $10.47 - 24.90/barrel

(4) Equivalent to $11.10 - 25.03/barrel

(5) Data presented are not directly comparable. See Section 1.4 for discussion.

TABLE 1-4

METHANOL PRODUCTION COSTS

JULY 1974 DOLLARS

	GA/S&W	Koppers-Totzek
Plant Capacity, 10^9 Btu/day		
Methanol	195	293
Other Products	208	0
Total	403	293
Capital Cost, $ Million	1,537	1,296
Average Product Cost, $/$10^6$ Btu	2.72 to 5.47	3.41 to 7.47
Methanol Cost, $/Gallon (1)	0.24 to 0.61	0.22 to 0.48
Commercial Cost, $/Gallon	0.40	

Note: (1) Assuming $1.50/$10^6$ Btu for purge gas and $0.30/gallon for light aromatics.

TABLE 1-5

STEEL PRODUCTION ESTIMATED COSTS - SUMMARY

Case	Process	Plant Capacity Tons/Year (1)	Capital Cost $/Annual Ton (1)	Operating Costs (2) $/Ton (1)	Selling Price for 5-Year Recovery - $/Ton
1	Coke Ovens Blast Furnace Basic Oxygen Furnace	4,000,000	152.00	104.69	148.55
2	Conventional Reformer Direct Reduction Electric-Arc Furnace	1,060,000	93.60	104.44 (5)	131.44 (5)
3	3000 MWt General Atomic-Stone & Webster VHTR Synthesis Gas Feed Direct Reduction Electric-Arc Furnace	14,100,000	96.0	111.17-132.14 (3)	143.60-164.57 (3)
4	3000 MWt VHTR Equivalent Koppers-Totzek Synthesis Gas Feed Direct Reduction Electric-Arc Furnace	16,500,000	92.0	108.50-130.92 (4)	139.58-162.00 (4)

Notes: (1) Net Tons of Liquid Steel.

(2) Includes Direct Operating Expense, Depreciation and Interest during Construction.

(3) Based on Range of $1.78-3.78/$10^6$ Btu for reducing gas.

(4) Based on Range of $1.57-3.75/$10^6$ Btu for reducing gas.

(5) Based on $0.79/$10^6$ Btu for natural gas.

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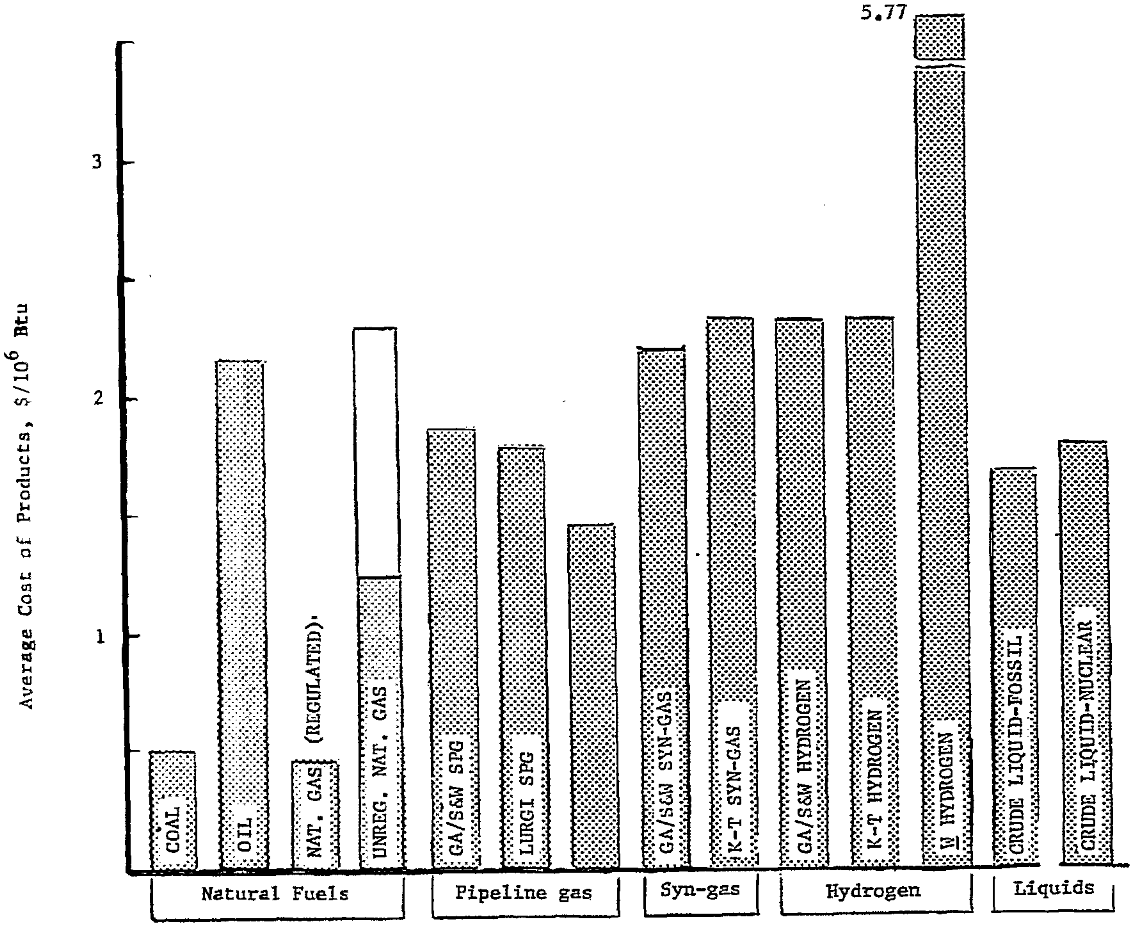

Fig. 1.1. Process Cost Comparison

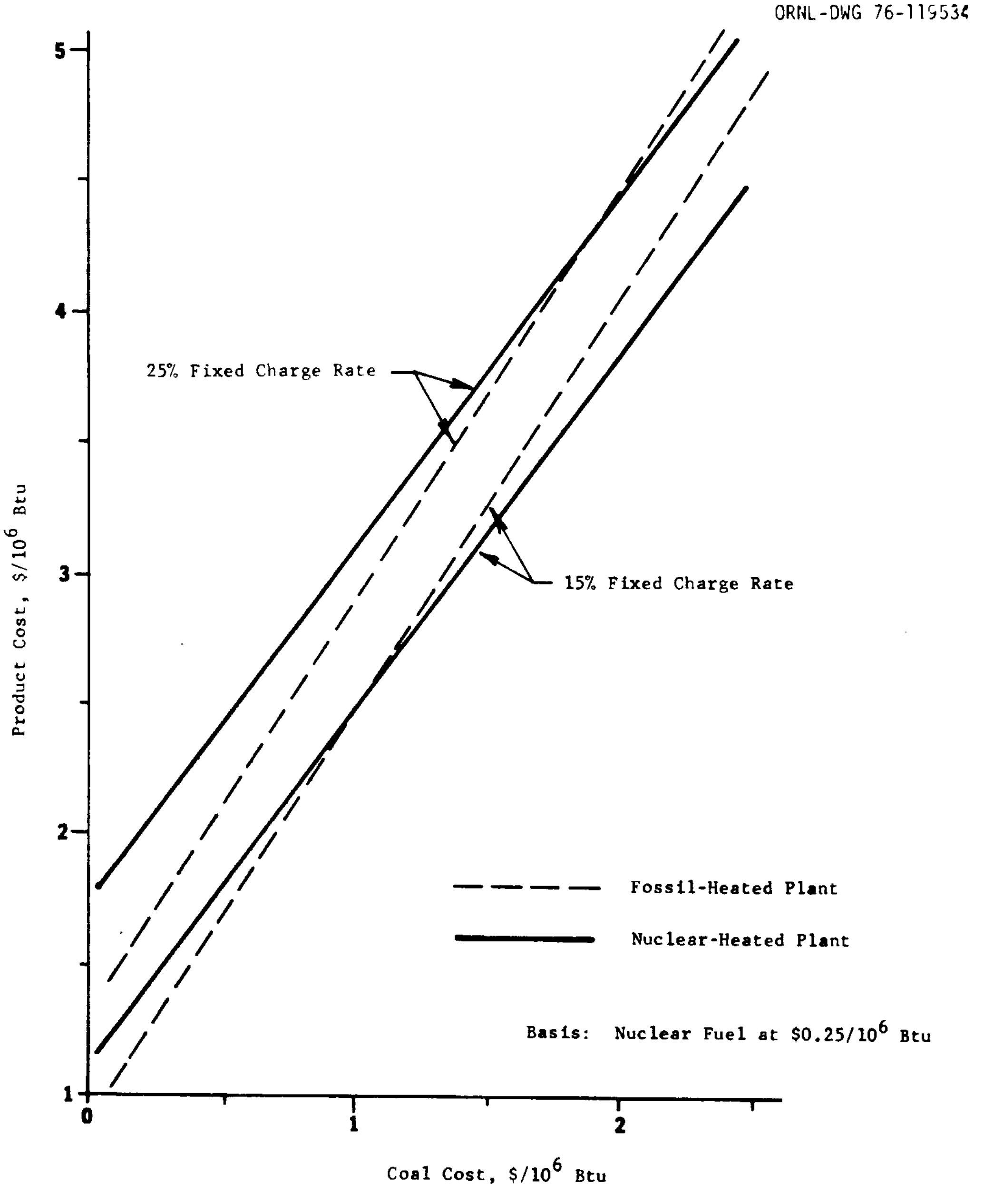

Fig. 1.2. Comparison of Nuclear and Fossil Heated Coal Liquefaction Costs

2. BASIS OF STUDY

2.1 Base Case - GA/S&W Synthetic Pipeline Gas Process

The GA/S&W Synthetic Pipeline Gas (SPG) process was made the basis for all comparisons. All of the heat from a 3000 MWt VHTR, equivalent to 10.2 x 10^9 Btu/hr, was utilized in a coal conversion process to produce a high Btu gas suitable for pipeline transmission and a minor amount of a liquid fuel. No coal was consumed as fuel. When modifications were made to the GA/S&W SPG base process for production of synthesis gas, hydrogen or crude liquids, these modified processes also utilized the full 3000 MWt heat from the VHTR. The amount of coal input and the point of product withdrawal was varied in accordance with chemical principles to obtain the various products.

2.2 Comparable Coal-fueled Coal Conversion Processes

Various other processes have been either commercialized or are under development for the production of the several gaseous and liquid fuel products that could be produced from the GA/S&W process. For production of pipeline gas, the Lurgi process is commercial and the Hygas process has progressed to the small demonstration plant stage. The Koppers-Totzek process has been commercialized for the production of synthesis gas and hydrogen. For production of crude liquids the H-Coal process has been operated on a small pilot plant scale. These processes have been evaluated economically for comparison to the GA/S&W processes and are described in Section 4.

2.3 Sources of Cost Data

Costs were obtained from data published in the open literature by the manufacturers of the processes. Capital costs of plants and/or sections of plants were updated to July, 1974, by use of the GE Plant

Cost Index, published biweekly in the Economic Indicators section of Chemical Engineering (CE) magazine by McGraw Hill, New York, N.Y. This date is the base for the comparison of VHTR's and the associated process heat applications. Chemicals, catalysts and direct operating or maintenance labor costs were adjusted to July 1974 by use of the following indexes published in CE: industrial chemical wholesale prices, hourly earnings and productivity.

2.4 Cost Factors and Methods Used

The cost factors used are shown in Table 2-1. No by-product credits were included; it was assumed that disposal of these would occur at no cost.

The capital and production cost estimates reported herein were all adjusted to the same basis, to the extent allowable with the published reference costs. No contigencies were added by United Engineers.

The VHTR capital cost was estimated by ORNL at $800 million and included all direct, indirect and interest during construction charges. The operating and maintenance costs for the VHTR were taken at $9 million per year. Nuclear fuel costs include all expenses such as all materials, uranium enrichment, fuel fabrication, fuel reprocessing, credits for materials of value in spent fuel, and carrying charges in all parts of the fuel cycle.

Many of the basic cost references utilized in the preparation of this report did not disclose the detailed composition of the capital and production costs that were developed and published. Thus, contractor's overhead and profit, engineering and construction management costs, contingency, etc. were not disclosed.

Capacities of the comparable coal-fueled coal conversion processes, chosen from published sources, were adjusted to make the heating value in Btu/hr of the products equal to the products from the GA/S&W versions, even though the chemical composition was different. Where another product, methanol or steel, was to be made from the synthesis gas, an appropriately sized plant was matched to the synthesis gas plant to suit the feed quantity and composition.

2.5 Scaling Factors

Scaling of capital costs of plants or sections of plants was made by use of the ratio of capacities or throughputs to the 1.0 power instead of fractional powers that are normally used for single train plants. This was based on the assumption that the new capacity would be obtained by adding or deleting one or more of the multiple trains which made up the reference plant.

2.6 Types of Coal Used

The coals used in the various studies varied in composition, depending on the source. The effect of these differences, however, was minimized by comparing costs on a Btu basis for both coal fuels and products. The coals used in the Lurgi and Hygas plant studies were low sulfur as compared to the GA/S&W, Koppers-Totzek and H-Coal plants; however, no cost adjustments were made, even though this penalized the high sulfur coal plant costs somewhat due to the higher H_2 consumption required for sulfur hydrogenation and the larger sulfur recovery plant requirement.

2.7 Efficiency

The usually quoted coal conversion efficiency is the ratio between the energy in the useful products and the energy contained in the coal which is actually converted to product. The energy used by the process, in the form of steam and electricity, is not used in the calculation. The quantity of outside energy may significantly reduce the real efficiency of the process, so comparison of process efficiencies calculated in this manner can be misleading if the processes are not similar.

Some of the processes in this report utilized significant sources of energy other than from the coal feed. As a result, United Engineers has defined the efficiencies in this report as the ratio between the energy available in the useful products and the sum of the off-site energy utilized by the process and the energy in the raw materials (which equals the total energy input to the process). Where electricity is used, a generation efficiency of 37% was assumed. This efficiency is typical of modern coal-fired or nuclear gas-cooled power plants.

2.8 Product Pressure

Adjustments were made to all processes with gaseous products to produce gases at 1000 psia.

TABLE 2-1

COST ESTIMATE BASES

July 1974 Dollars

	Low Estimate	High Estimate	Unit
Coal	0.50 (12)	1.75 (42)	$/Million BTU ($/Ton at 12,000 Btu/lb.)
Nuclear Fuel	0.25	0.60	$/Million BTU
Steam	1.00		$/M lbs @ 600 psi
Raw Water	0.30		$/1000 gallons
Electric Power	0.012		$/KWh
Fixed Charge Rate	15	25	%/year
Nuclear Reactor Cost			
Capital	$ 800 million		
Operating	$ 9 million/year		
Interest During Construction	8%/year		
Standard Operating Year	330 days		
Chemical Plant Lifetime	20 years		

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3. NUCLEAR-HEATED COAL GASIFICATION PROCESSES

3.1 GA/S&W Pipeline Gas Plant

The basic conceptual design and cost estimates for this process were obtained from References 3.1-1, 3.1-3 and 3.1-5, which should be consulted for additional detailed information. A simplified block diagram for the process, Figure 3.1-1, shows the major process sections and their sequence together with the amounts of raw materials, fuel products and by-products. A light aromatics liquid fuel is obtained, in addition to the synthetic pipeline gas fuel. The efficiency, obtained from the ratio of heat available in the fuel products to the heat available in the coal plus that obtained from the nuclear reactor (VHTR), is 67 percent for the heating values shown in the diagram.

3.1.1 Description of Process

An Oklahoma coal as received from the mine is stored, reclaimed, pulverized, and partially dried, using some of the by-product waste CO_2, in the coal preparation section. The pulverized coal is slurried with hot recycle solvent and reacted with recycle H_2 in the coal solution section. The depolymerized coal in solution is then sent to a solution hydrocracking area. Here, a portion of the solvent is separated and returned to slurry preparation. The remaining liquid is reacted with H_2 at high temperature and pressure in the presence of a catalyst to form mainly distillable liquids. The low and intermediate boiling distillates are sent to the hydrogasification section, recycle solvent is returned to slurry preparation, and the heavy distillates and unconverted carbon and ash are sent to a fluid bed coker.

The fluid bed coker concentrates the unconverted carbon and ash into a refuse stream. Heat supplied by combustion of some of the carbon with air cracks the heavy distillates into solvent for recycle and/or distillate for processing in the hydrogasification section. The flue gas containing SO_2 is reacted with H_2S from a following section in a sulfur recovery plant.

In the hydrogasification section the distillates are reacted with all but a small fraction of the hydrogen in the feed to produce mostly gaseous hydrocarbons. The reaction products are purified by removal of H_2O, NH_3, CO_2 and H_2S and separated into a light aromatics liquid fuel stream and a synthetic pipeline gas stream. Approximately 50 percent of this product gas is sent to the reforming section for production of H_2 for recycle and the remainder withdrawn as pipeline gas product at approximately 1000 psi and $100^{o}F$.

The crude methane stream from hydrogasification is catalytically reformed with steam at $1400^{o}F$ and 300 psig (Reference 3.1-2). Heat for the reformer is supplied by heat exchange from helium in an intermediate loop, so that the reformer can be located outside of the containment (Reference 3.1-3). The steam for reforming is generated from makeup water in a boiler heated by steam in a closed loop. The condensate produced passes to a steam generator located in the reactor core where heat is provided from helium in a primary loop. Steam from the generator passes to a power turbine to produce process power before exhausting to the water boiler.

The hydrogen production section takes reformed effluent through a shift converter where the bulk of the CO is converted to H_2 and CO_2 by reaction with residual steam. This product passes to a CO_2 removal system where most of the remaining water is condensed out and recycled to the reformer section. The hydrogen leaving is approximately 88 mole percent purity and is compressed for recycle to the coal solubilizing sections.

3.1.2 Features and Limitations

Like all coal solubilizing processes, any grades of coal may be employed. No coal is consumed as fuel in the process, allowing an 87% conversion of coal to fuel products. However, this is advantageous only when coal heat becomes relatively more expensive and scarcer than nuclear fuel. The fuels produced are clean (low sulfur and ash content) and would meet all pollution requirements of EPA. By-products disposal present no unusual environmental problems, the requirements being similar to or less severe than those encountered in other coal conversion processes.

The GA/S&W pipeline gas process appears quite similar to the modified H-Coal liquefaction process which is described in Section 5 of this report. However, an additional hydrogasification step is included in the GA/S&W process to convert the majority of the crude liquids from the coal to paraffinic gases. This process is employed to some extent in petroleum processing, but has not been demonstrated on coal liquids. The lack of bench scale or pilot plant data for the entire GA/S&W process would present obstacles to its early commercialization, even where the economics appear promising.

3.1.3 Economic Evaluation

The capital cost estimate for the process is given in Table 3.1-1. The production cost estimate is listed in Table 3.1-2 and presented graphically in Figure 3.1-2.

The accuracy of the capital estimate for the process and offsite sections of the plant might be questioned since it was prepared from a preliminary or conceptual design only. However, it does not appear to be excessively out-of-line with the other plant costs studied in this report. With respect to utilities (very small consumption in the process), catalysts and chemicals and operating labor, significant errors in the estimate would not seriously affect the total production cost.

3.1.4 Major Pollution Outputs

About 20 percent of the coal remains as refuse (unconverted carbon and ash) and will be returned to the coal mining area for disposal. The ash residue will retain approximately 20 percent of the coal sulfur.

The majority of the sulfur in the coal is converted to H_2S. This, plus SO_2 in the coker flue gas, is reacted and recovered as sulfur in Claus and Stretford units. All effluent streams from the process will meet the sulfur limitations of the EPA.

Water generated by the process will contain NH_3, H_2S and some phenols. The foul water will be stripped of NH_3 and H_2S, the former being concentrated for sale and the latter being routed to the Claus unit for conversion to sulfur. The phenolic water will be treated bacteriologically to reduce the phenols to acceptable levels before adding it as makeup to the cooling towers.

3.1.5 References

3.1-1 Final Report Application Study of a Nuclear Coal Solution Gasification Process for Oklahoma Coal, Volumes 1 and 2, May 1972, General Atomic Company/Stone & Webster. (GA-A12068)

3.1-2 High-Temperature Nuclear Heat Source Study, September 27, 1974, General Atomic Company.

3.1-3 A Cost Comparison of Very High Temperature Nuclear Reactors for Process Heat Applications, March, 1975, United Engineers (UE&C-AEC-750311).

3.1-4 Personal Communication, J. Jones, Jr., Holified National Laboratory, June 17, 1975.

3.1-5 Studies of the Use of High Temperature Nuclear Heat from an HTGR for H_2 Production, General Atomic Report GA-A13391, September 30, 1975.

TABLE 3.1-1

GA/S&W SYNTHETIC PIPELINE GAS PROCESS
CAPITAL COST ESTIMATE
JULY, 1974 DOLLARS

Section	Capital Cost $ million
Coal Preparation	20.2
Coal Solution	72.9
Solution Hydrocracking	75.5
Fluid Bed Coking	27.0
Hydrogasification	78.5
Reforming	99.8
Hydrogen Production	172.4
Offsites (2)	134.4
Subtotal	680.7
Interest During Construction (8% for 4 years)	113.3
Total	794.2
VHTR System (Including Interest during Construction at 8% for 8 years) (1)	800.0
Total	1,594.0

Notes:

(1) From Reference 3.1-4

(2) Includes $1.4 million credit for compressors

TABLE 3.1-2

GA/S&W SYNTHETIC PIPELINE GAS PROCESS
PRODUCTION COST ESTIMATE
JULY, 1974 DOLLARS

	Product Cost $\$/10^6$ Btu
Coal	.58 to 2.02
Nuclear Fuel	.09 to .21
Utilities, Catalysts and Chemicals	.05
Direct Operating Cost, Process Plant	.09
Direct Operating Cost, VHTR	.04
Fixed Charges	1.02 to 1.71
Total Production Cost	1.87 to 4.12

Basis:

Plant Capacity

Gas	588×10^9	Btu/day
Liquid	120×10^9	Btu/day
Total	708×10^9	Btu/day

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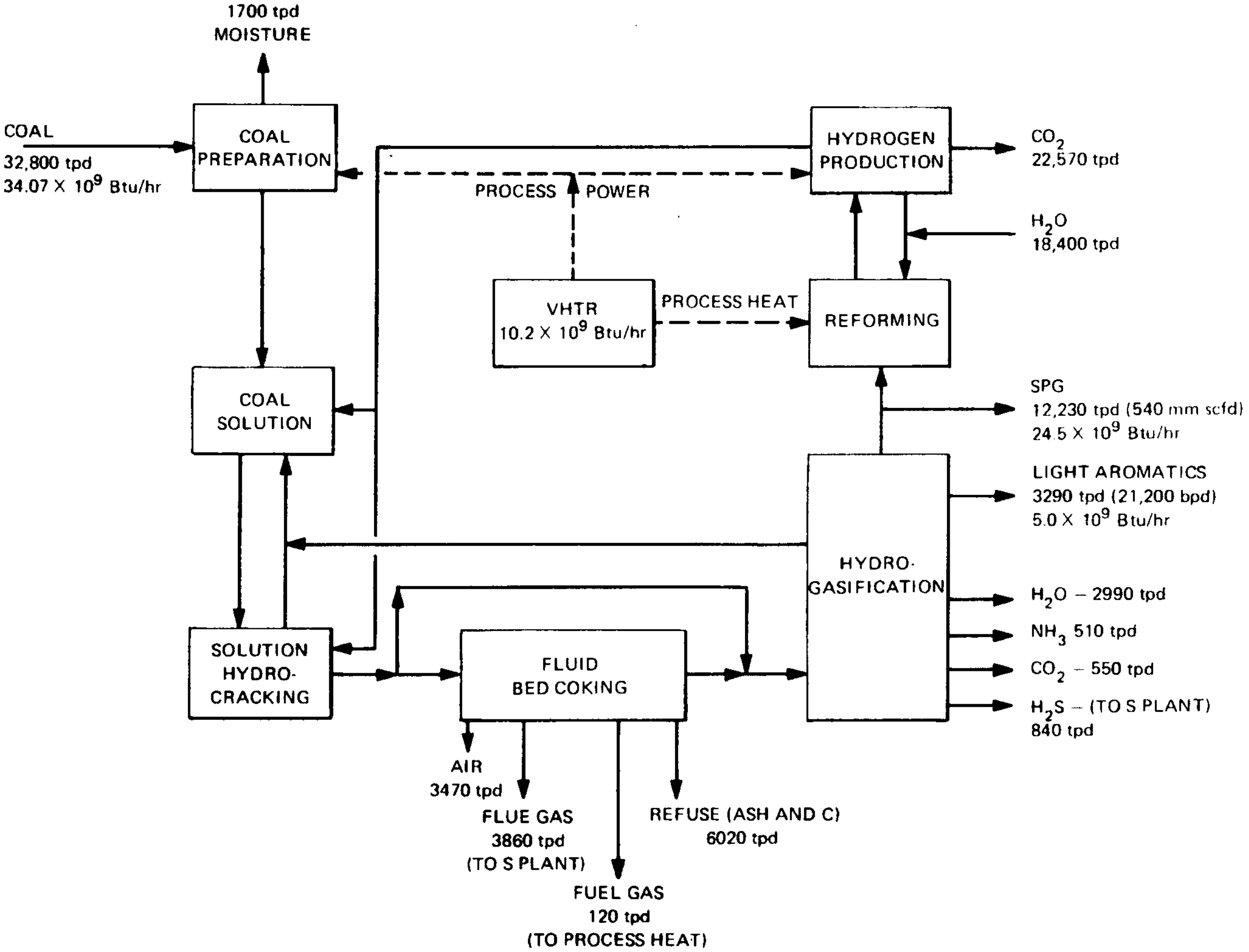

Fig. 3.1-1. Plant Block Diagram — GA/S&W Synthetic Pipeline Gas Process

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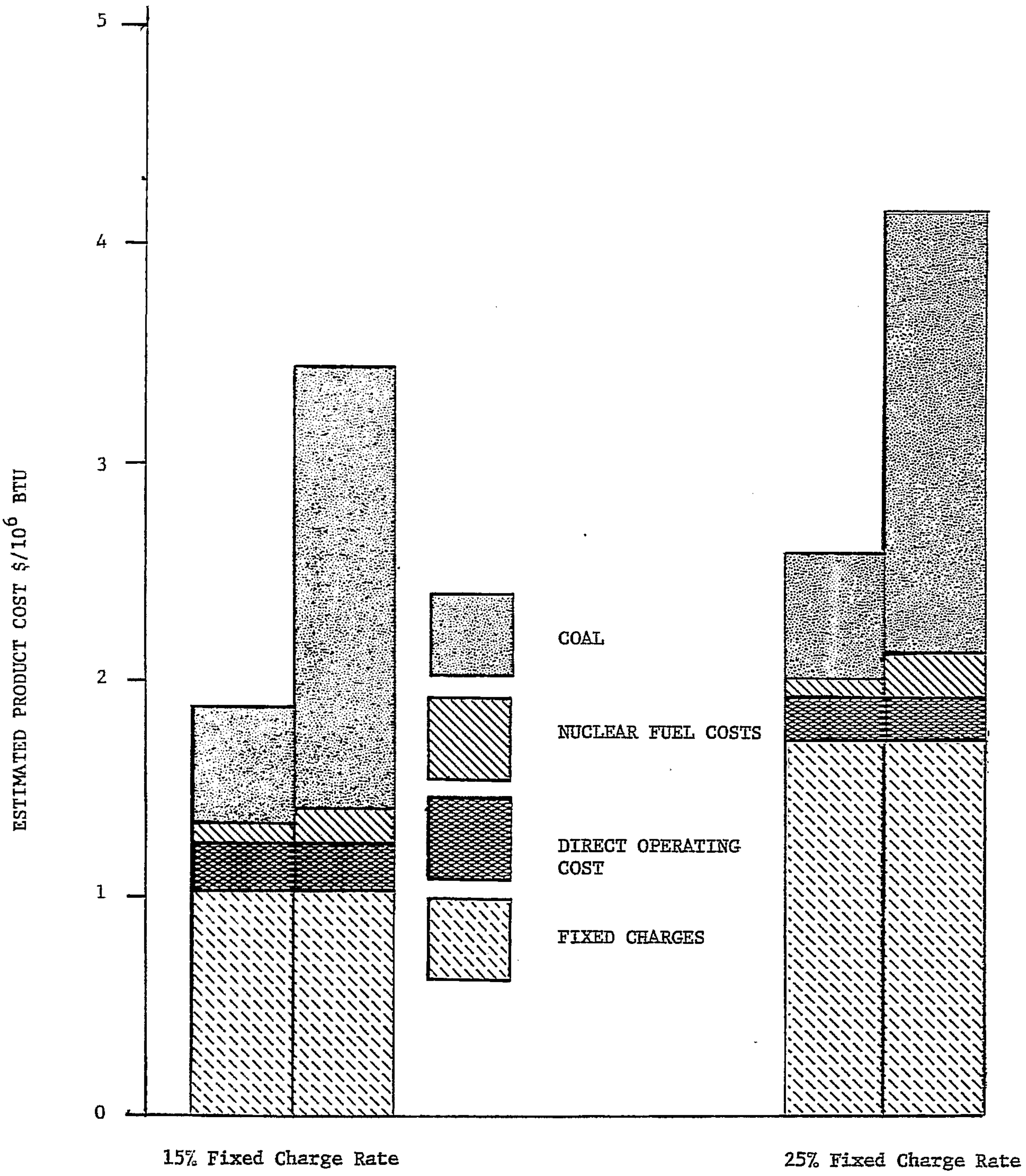

Fig. 3.1-2. Production Costs — GA/S&W Synthetic Pipeline Gas Process

3.2 GA/S&W Synthesis Gas Plant

3.2.1 Description of Process

A synthesis gas can be produced by changing the basic GA/S&W pipeline gas process described in Section 3.1 as follows:

1. Reduce the coal feed and hydrogen recycle by about 50 percent.

2. Send all of the product gas from the hydrogasification section to reforming.

3. Send approximately 50 percent of the crude synthesis gas from the reformer through shift conversion and CO_2 removal for compression to recycle.

4. The remaining crude synthesis gas can be withdrawn as product after condensing out most of the water vapor, or it can be sent to a CO_2 removal unit and then withdrawn as product, again after water condensation. A synthesis gas better suited for methanol production can be produced by adding approximately 33 percent of the CO_2 removed from the recycle hydrogen to the synthesis gas from the reformer. This latter method is the production method shown on the block diagram, Figure 3.2-1.

3.2.2 Features and Limitations

Approximately 50 percent of the base case light aromatics liquid fuel is also produced by the modified process. The fuel efficiency for this process is 64 percent. The product synthesis gas would be available for export at approximately 1000 psi and 100°F. The composition of the synthesis gases that can be produced are shown below.

Mole %

	From Reformer	With CO_2 Removed	With CO_2 Added
CH_4	9.5	10.95	9.0
H_2	70.5	80.7	66.5
CO	7.0	8.05	6.6
CO_2	12.9	.2	17.8
H_2O	.1	.1	.1

TABLE 3.2-1

GA/S&W SYNTHESIS GAS PROCESS
CAPITAL COST ESTIMATE
JULY, 1974 DOLLARS

Section	Capital Cost $ Million
Coal Preparation	10.9
Coal Solution	39.4
Solution Hydrocracking	40.8
Fluid Bed Coking	14.6
Hydrogasification	85.3
Reforming	108.5
Synthesis Gas & Hydrogen Production	102.3
Product Compression	8.3
Offsites	73.4
Subtotal	483.5
Interest During Construction (8% for 4 years)	80.4
Total	563.9
VHTR System	800.0
Total	1,363.9

TABLE 3.2-2

GA/S&W SYNTHESIS GAS PROCESS
PRODUCTION COST ESTIMATE
JULY, 1974 DOLLARS

	Product Cost $\$/10^6$ Btu
Coal	.50 to 1.77
Nuclear Fuel	.14 to .34
Utilities, Catalysts and Chemicals	.04
Direct Operating Cost, Process Plant	.07
Direct Operating Cost, VHTR	.06
Fixed Charges	1.41 to 2.35
Total Production Cost	2.22 to 4.63

Basis:

Plant Capacity

Gas	373×10^9 Btu/day
Liquid	66×10^9 Btu/day
Total	439×10^9 Btu/day

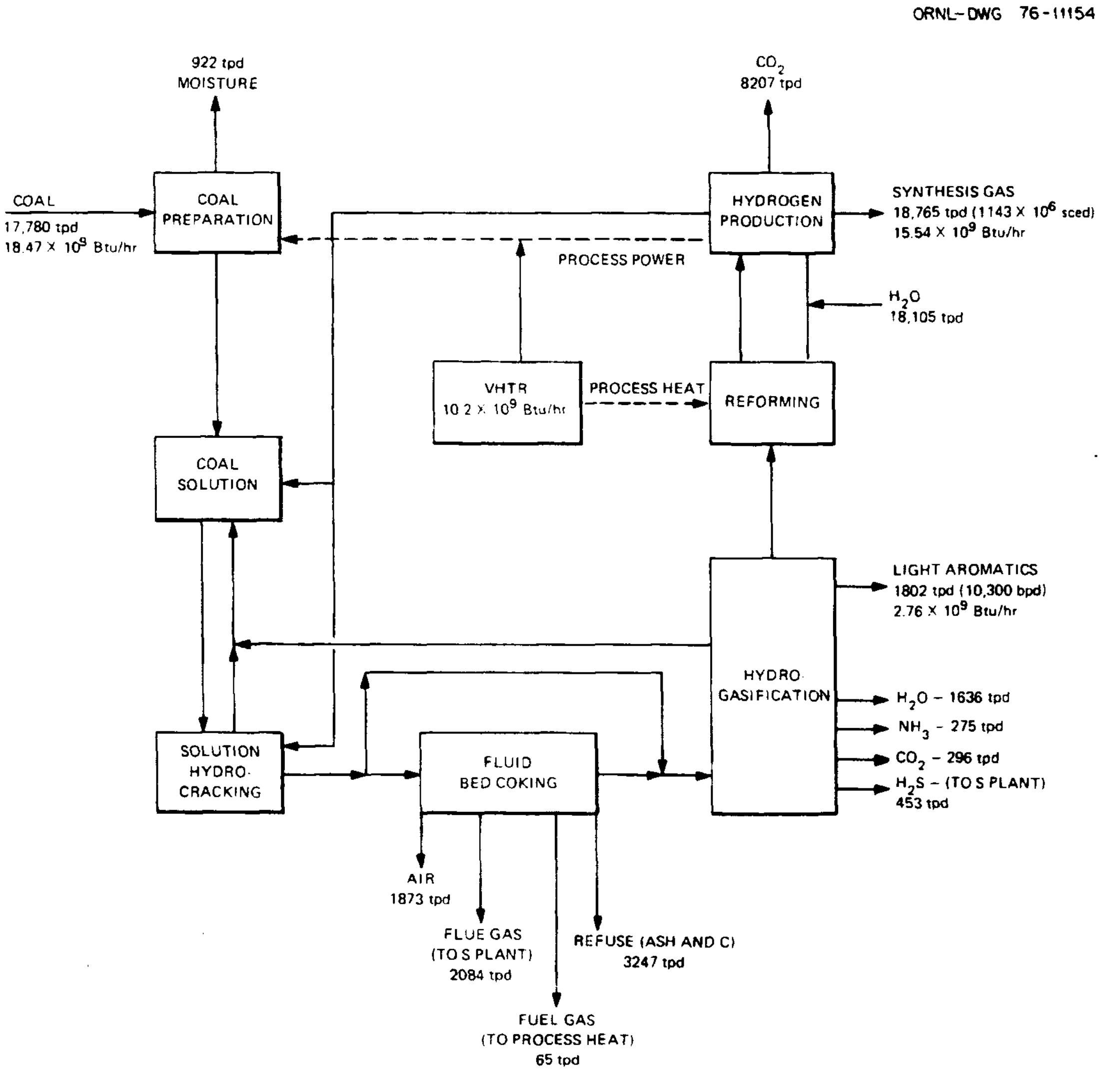

Fig. 3.2-1. Plant Block Diagram — GA/S&W Synthesis Gas Process

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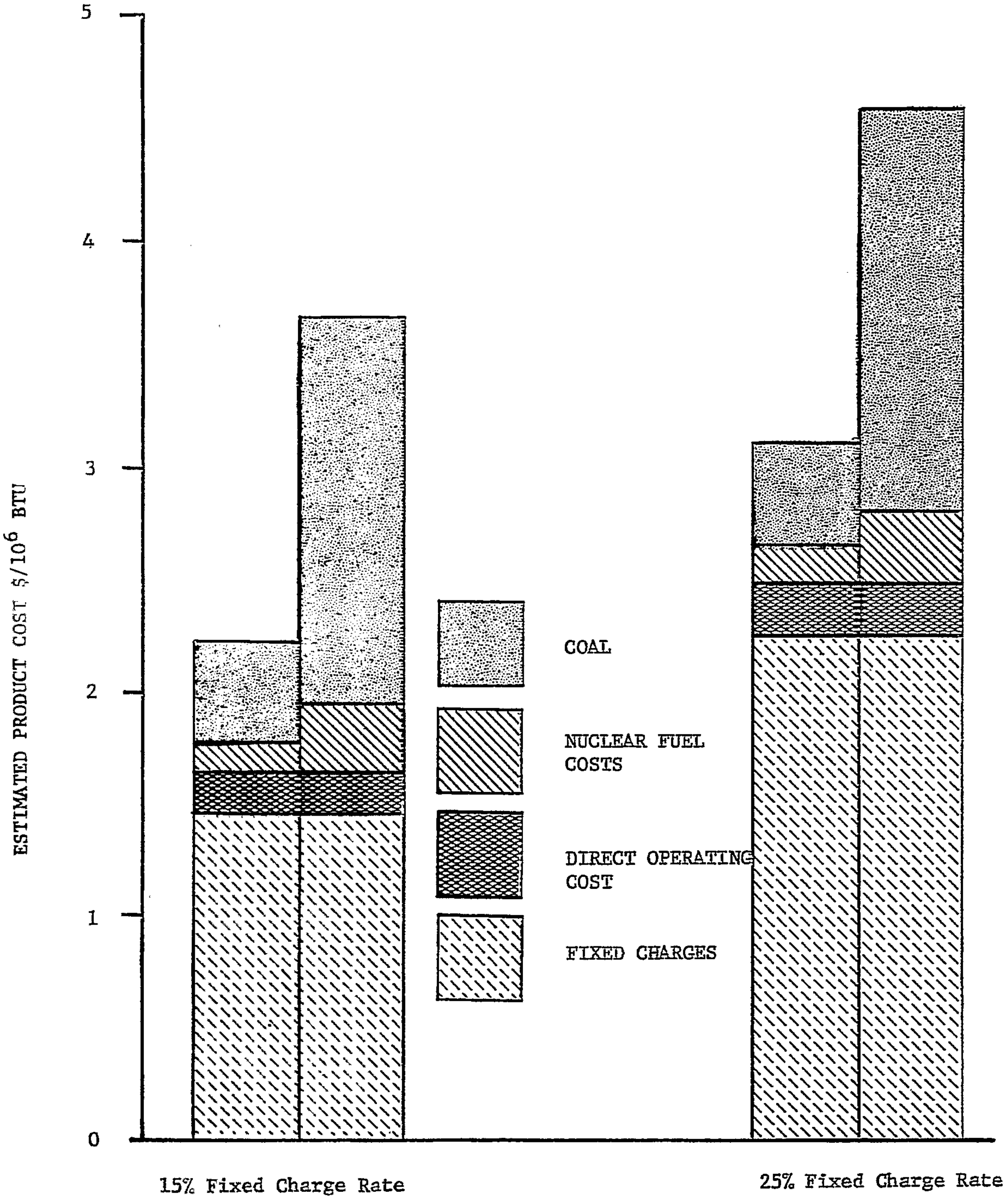

Fig. 3.2-2. Production Costs — GA/S&W Synthesis Gas Process

3.3 GA/S&W Hydrogen Plant

3.3.1 Description of Process

General Atomic has revised the process described in Section 3.1 to produce hydrogen gas. This hydrogen production process, and other potential hydrogen processes, is described in detail in Reference 3.3-1. The major differences between this plant and the pipeline gas plant described in Section 3.1 are:

- The coal feed is reduced by about 45 percent.
- All of the gas from the hydrogasification section is reformed and converted to crude hydrogen.
- Approximately 50 percent of the crude hydrogen is recycled to the coal solution stage.
- The remaining crude hydrogen is methanated to 87.9 mol % hydrogen and 11.9 mol % methane.
- The process energy efficiency is reduced to 62 percent.

The block diagram for the GA/S&W hydrogen process is shown in Figure 3.3-1.

3.3.2 Features and Limitations

This purity and composition of hydrogen is probably satisfactory for use as is in ammonia synthesis and various petroleum hydrogen treatment processes such as hydrodealkylation. The capital cost for the modified plant is shown in Table 3.3-1, and the production costs are presented in Table 3.3-2 and Figure 3.3-2.

Higher purity hydrogen, in the range of 97 to 99^{+} percent, should be readily attainable by passing the crude gas through a pressure swing adsorption (molecular sieve) unit. The separated methane could be credited as pipeline gas or recycled to the reformer. The additional purification cost would be in the range of 2 to 8 cents/10^6 Btu.

3.3.3 Economic Evaluation

The capital cost for the plant modified to produce higher purity hydrogen is $1,466.4 millions and is itemized in Table 3.3-1. The production costs are presented in Table 3.3-2 and Figure 3.3-2.

3.3.4 Major Pollution Outputs

The impacting pollutants will have approximately the same ratio to coal input as that described in 3.1.4 for the base case.

3.3.5 Reference

3.3-1 Studies of the Use of High Temperature Nuclear Heat from an HTGR for Hydrogen Production, General Atomic Report GA-A-13391, September 1975.

TABLE 3.3-1

GA/S&W HYDROGEN PRODUCTION PROCESS
CAPITAL COST ESTIMATE
JULY, 1974

Section	Capital Cost $ Million
Coal Preparation	10.9
Coal Solution	39.4
Solution Hydrocracking	40.8
Fluid Bed Coking	14.6
Hydrogasification	85.3
Reforming	108.5
Hydrogen Production	187.4
Product Compression	11.0
Offsites	73.4
Subtotal	571.3
Interest During Construction (8% for 4 Years)	95.1
Total	666.4
VHTR System	800.0
Total	1,466.4

TABLE 3.3-2

GA/S&W HYDROGEN PRODUCTION PROCESS
PRODUCTION COST ESTIMATE
JULY, 1974

	Product Cost $\$/10^6$ Btu		
Coal	.51	to	1.78
Nuclear Fuel	.15	to	.37
Utilities, Catalysts and Chemicals		.04	
Direct Operating Cost, Process Plant		.08	
Direct Operating Cost, VHTR		.07	
Fixed Charges	1.63	to	2.72
Total Production Cost	2.48		5.06

Basis:

Plant Capacity

Gas	343×10^9 Btu/day
Liquid	60×10^9 Btu/day
Total	403×10^9 Btu/day

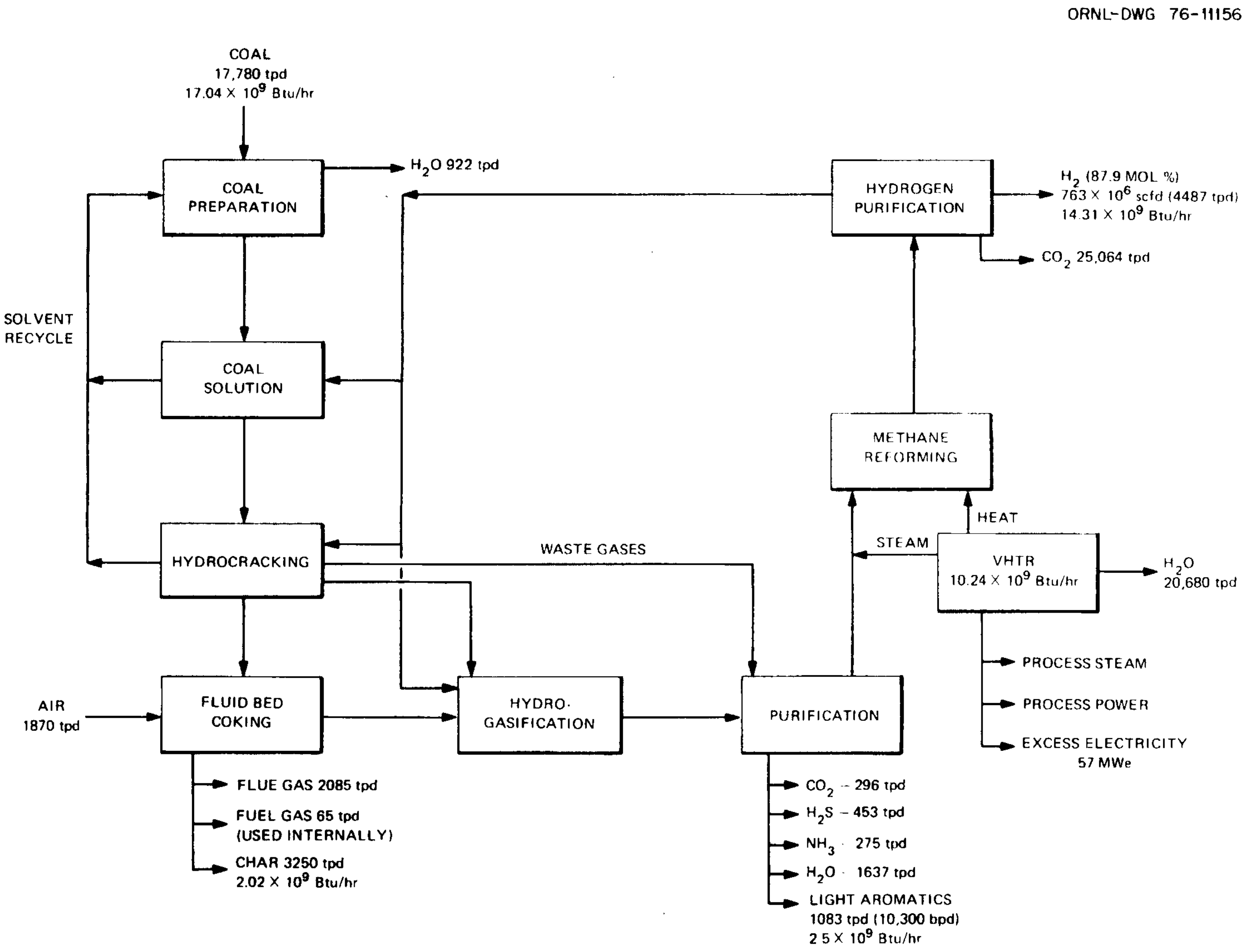

Fig. 3.3-1. Plant Block Diagram — GA/S&W Hydrogen Production Process

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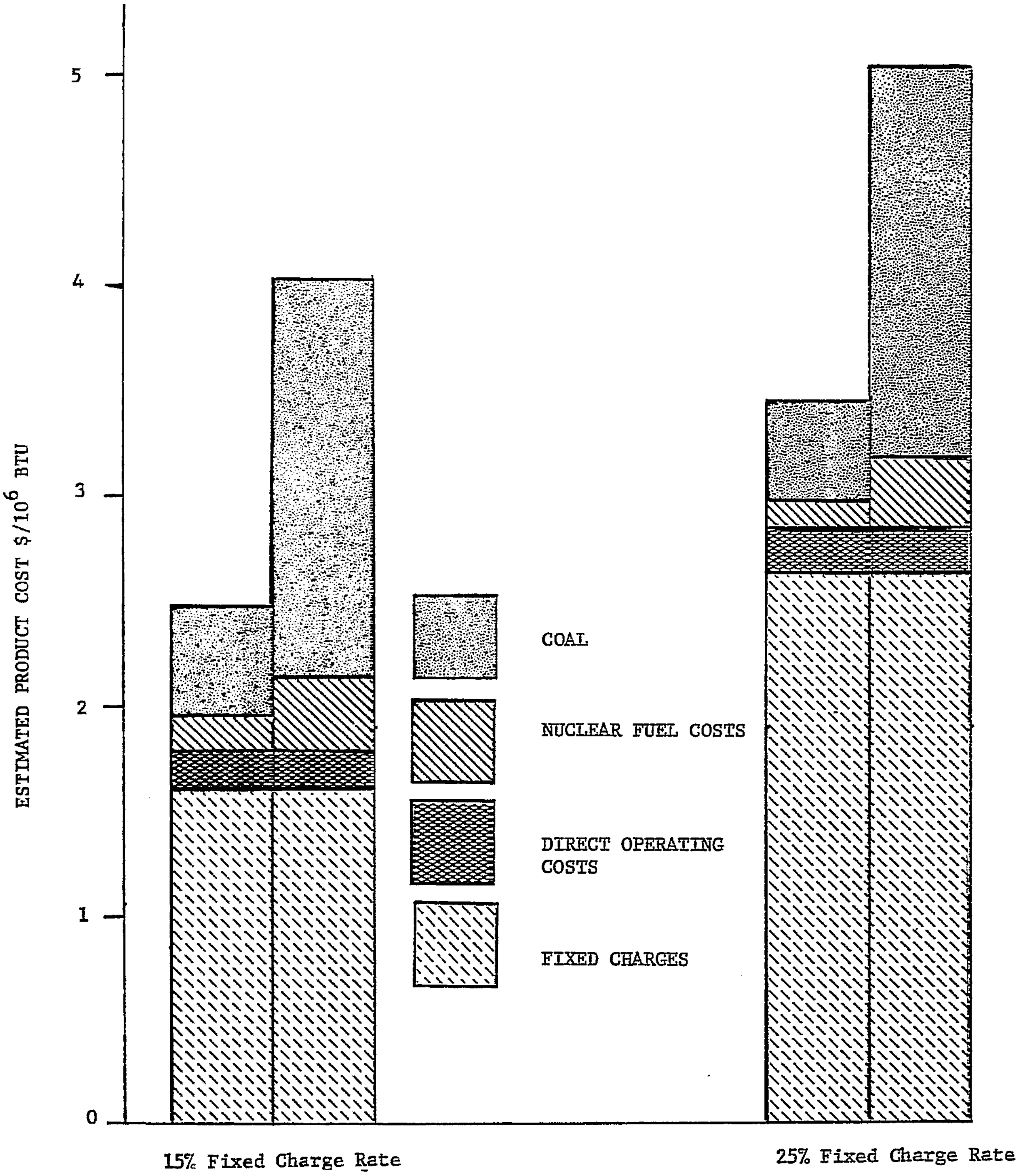

Fig. 3.3-2. Production Costs — GA/S&W Hydrogen Production Process

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4. COAL-FUELED COAL CONVERSION PROCESSES

Other processes have been commercialized or are under development for the production of the several gaseous and liquid fuel products. These are the Lurgi and Hygas synthetic pipeline gas processes and the Koppers-Totzek synthesis gas and hydrogen processes. These processes are described and evaluated economically in the following subsections.

4.1 Lurgi Synthetic Pipeline Gas Plant

4.1.1 Description of the Process

Figure 4.1-1 is a block diagram of the Lurgi Pipeline Gas Plant.

The Lurgi process utilizes a fixed bed gasifier operating at 350 to 450 psi and 1150 to 1400°F. Steam and oxygen are supplied to the bottom of the gasifier while sized coal enters from above by lock hoppers. The counter-current flow of coal and hot gases allows for efficient use of the heat released by the oxidation of coal at the base of the gasifier. This results in a smaller amount of oxygen required to heat the incoming coal, and thus a higher Btu raw gas product.

The raw gas typically has the following approximate analysis (Reference 4.1-1).

Component	Mole Percent
CO_2	28
CO	20
H_2	39
CH_4	11
Others	2

A portion of the raw gas is sent to crude gas shift conversion, where the H_2/CO ratio is adjusted. This stream is then mixed with the non-shifted raw gas stream. The final H_2/CO ratio is approximately 3.6, optimal for the methanation step.

Before the crude gases are fed to the low temperature purification process, they are cooled in waste heat boilers which generate steam at 60 psig. Further cooling is attained by the generation of low pressure steam, air coolers and finally cooling water.

The Lurgi Rectisol process, which utilizes low temperature methanol as an acid gas absorbent, is used for gas purification. In this step, most of the H_2S and CO_2 are removed. The purified gas stream (64 percent H_2 and 17 percent CO) is then passed through a methanation step where the final product is formed.

The temperature of the highly exothermic methanation reaction is controlled with a recycle stream which sets the concentration of reactants. Water is removed from product gases by compression and a conventional glycol system. The final product gas has a higher heating value of 954 Btu/Scf (dry), and is available at 1500 psi.

4.1.2 Features and Limitations

The overall energy efficiency of the Lurgi process is approximately 68 percent. Most of the non-recoverable heat is in the form of the coal burned to sustain the gasification step and for production of low Btu fuel for process power and steam generation.

One major disadvantage of the Lurgi gasifier is that without pretreatment, it can only process non-caking, or very weakly caking coals (Reference 4.1-2). This would eliminate its use with virtually all of the U.S. coals east of the Mississippi River (Reference 4.1-3). Another disadvantage is that the coal fines must be removed or briquetted before the feed coal enters the gasifier, which also adds to coal handling costs. Finally the fixed-bed gasifiers, as currently designed, are limited in capacity. For a large operation, many gasifiers would be required which would add significantly to piping complexity and cost.

4.1.3 Economic Evaluation

The capital costs for the Lurgi Process are presented in Table 4.1-1, which shows the investment required for process units, utilities, and off-site facilities. The total fixed investment, including interest during construction, is \$1089 million for a Lurgi pipeline gas plant producing 29.5 x 10^9 Btu/hr of fuel products.

Production costs are presented in Table 4.1-2 and Figure 4.1-2 as a function of both coal price and fixed charge rate. The minimum production cost is estimated to be \$1.79/10^6 Btu, corresponding to a coal price of \$0.50/10^6 Btu and fixed charge rate of 15 percent. The highest foreseeable production cost is \$4.09/10^6 Btu, which corresponds to a coal price of \$1.75/10^6 Btu and a fixed charge rate of 25 percent.

4.1.4 Major Pollution Outputs

The major pollution outputs from the Lurgi pipeline gas process will be:

1) Vent gas from the sulfur plant - Trace amounts of COS, CS_2, and H_2S will be present in the vent gases from the sulfur

recovery plant. However, these various gases are expected to comprise less than 100 ppm of the total vent stream; the major components will be CO_2 (82.5 Percent), N_2 (10.4 Percent) and H_2O (4.4 Percent).

2) Gas from the sulfur incinerator - This stream will be composed of mostly N_2 (45.0 Percent), CO_2 (42.0 Percent) and H_2O (11.0 Percent). Approximately 300 ppm of SO_2 will be included in the total gas flow from the incinerator.

3) Gas fired boilers - Flue gases from combustion will contain less than 0.16 lb SO_2, 0.20 lb NO_2, and 0.03 lb particulates/10^6 Btu fired.

4) Wet ash solids - Approximately 1.22×10^6 lb/hr of wet ash will require off-site disposal.

4.1.5 References

4.1-1 Second Supplement to Application of El Paso Natural Gas Co. for a Certificate of Public Convenience and Necessity, prepared by Stearns-Roger, Inc., FPC Pocket No. CP73-131, October 8, 1973.

4.1-2 Bodle, W. W. and K. C. Vyas, Clean Fuels from Coal, The Oil and Gas Journal, August 26, 1974.

4.1-3 Perry, Harry, Coal Conversion Technology, Chemical Engineering, July 22, 1974.

TABLE 4.1-1

LURGI PIPELINE GAS PROCESS
CAPITAL COST ESTIMATE
JULY, 1974 DOLLARS

	Capital Cost $ Million	
Process Units		
Gas Production	159.7	
Crude Gas Shift	19.0	
Gas Cooling, Purification & Refrigeration	120.7	
Methane Synthesis, Compression and Dehydration	60.0	
Gas Liquor Separation and Stripping	28.9	
Lock Gas Storage & Compression	4.8	
Sulfur Recovery	20.0	
Phenol Extraction	17.3	
Total Process Units		430.4
Utilities		
Fuel Gas Production, Cooling & Treating	67.4	
Air Compression, O_2 Production & Compression	123.3	
Steam and Power Generation	75.6	
Total Utilities		266.3
Offsites		
Water Treating, Cooling Water System and Miscellaneous Plant Utility Systems	45.1	
Ash and Raw Water Systems	50.9	
General Facilities	85.4	
Catalyst and Chemicals	9.9	
Coal Storage, Blending & Screening, Ash Handling	45.7	
Total Offsites		237.0
Total Capital Investment		933.7
Interest During Construction (8% for 4 years)		155.4
Total Fixed Investment		1,089.1

TABLE 4.1-2

LURGI PIPELINE GAS PROCESS
PRODUCTION COST ESTIMATE
JULY, 1974 DOLLARS

	Product Cost $\$/10^6$ Btu Product
Fuel and Raw Materials	0.74 to 2.57
Operation and Maintenance	0.35
Fixed Charges	0.70 to 1.17
Total Production Cost	1.79 to 4.09

Basis:

Plant Capacity

Gas	585×10^9	Btu/day
Liquid	124×10^9	Btu/day
Total	709×10^9	Btu/day

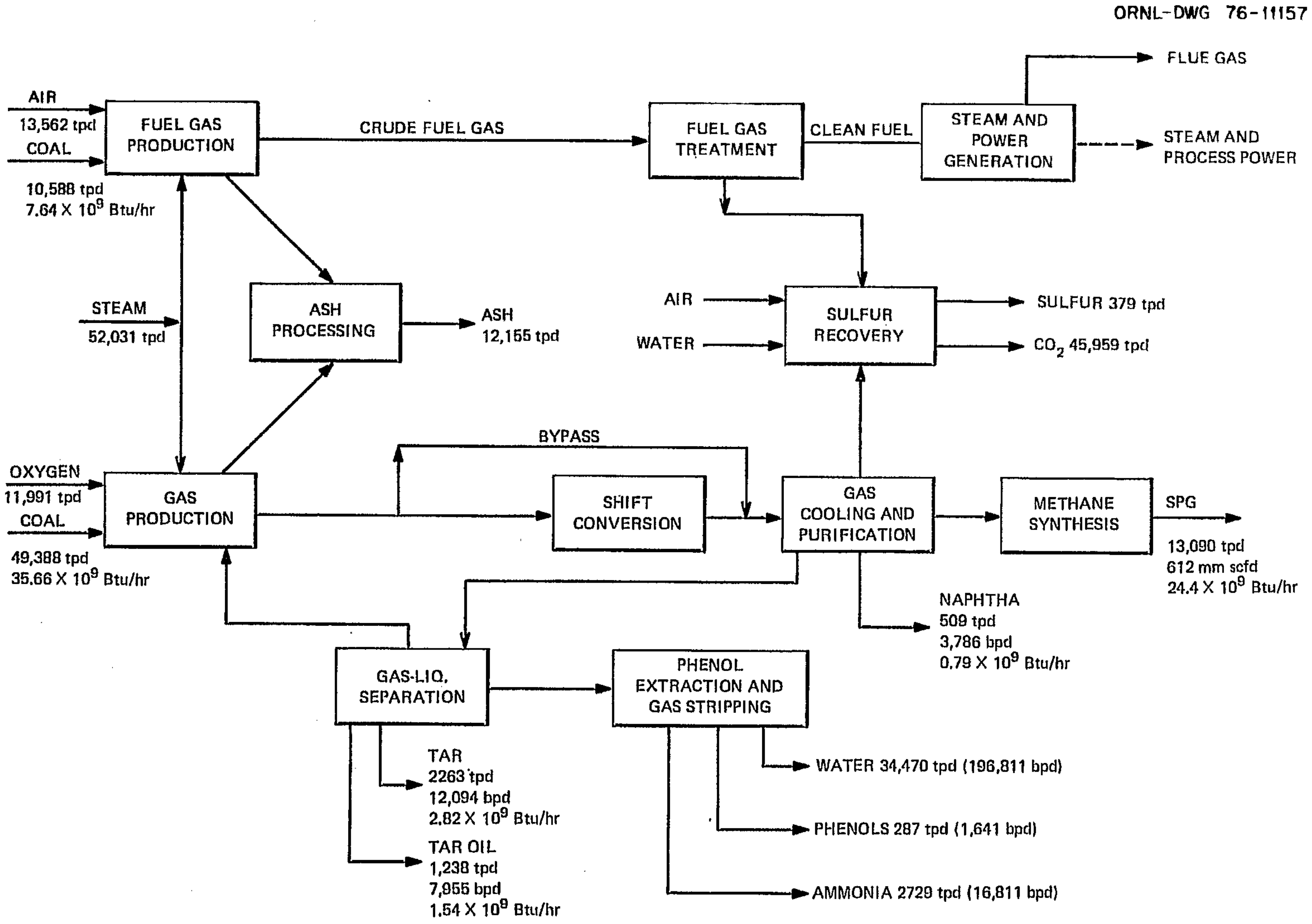

Fig. 4.1-1. Plant Block Diagram — Lurgi Pipeline Gas Process

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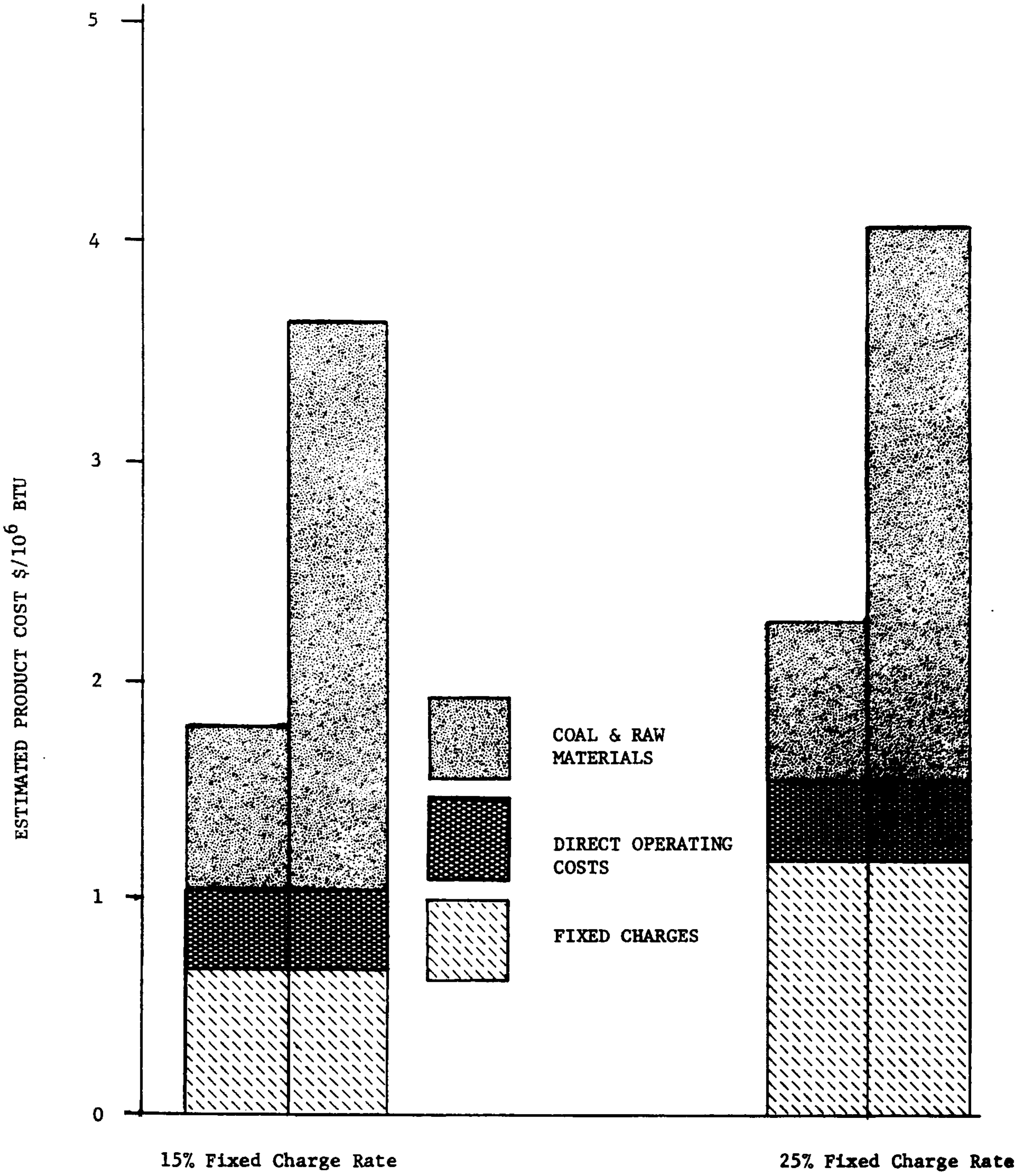

Fig. 4.1-2. Production Costs — Lurgi Pipeline Gas Process

4.2 Hygas Pipeline Gas Plant

4.2.1 Description of the Process

This process, developed by the Institute of Gas Technology (IGT) of Chicago, Illinois under the joint sponsorship of ERDA (formerly Office of Coal Research) and the American Gas Association, is representative of the more advanced technologies being developed by several organizations for the conversion of coal into pipeline gas having a higher heating value of approximately 1000 Btu/scf. It is tailored to handle all types of coal, both caking and non-caking. The gasifier operates at high pressure, 1000 - 1500 psi, and has a significantly higher throughput than a comparable Lurgi unit. Also, the proportion of methane formed in the gasifier is much higher than in the Lurgi system, resulting in substantially smaller CO shift conversion and methanation units for the Hygas process. Therefore, the expectation is that the capital and operating costs for the Hygas and other advanced coal gasification processes will be significantly lower than those of the presently established commercial processes, represented primarily by Lurgi. Economic analyses (References 4.2-1 to 4.2-7) in general confirm this expectation and thus provide the justification for the extensive development effort now under way in this area.

The chemical reactions forming the basis of the Hygas process are similar to those occurring in other coal gasification schemes. Basically the Hygas process utilizes the reaction of hydrogen with coal to achieve as high as possible direct conversion of carbon to methane. The heat generated by this exothermic hydrogasification reaction has to be moderated, otherwise it would cause the temperature to rise uncontrollably and result in a reduction of the methane yield. To use this heat effectively and at the same time to control the reaction temperature, IGT adds steam to the hydrogen, thus causing exothermic hydrogenation and endothermic steam reformation to occur simultaneously. A plus for this arrangement is that the steam generates additional hydrogen needed by the process. On the negative side the carbon monoxide so formed has to be methanated at a later point in the process.

The hydrogen for the hydrogasification is provided by reacting char generated in the gasifier with steam and oxygen (to generate the required heat of reaction) in a separate gasifier. The char used in this reaction consists of less reactive coal carbon that could only with great difficulty be consumed in the hydrogasification step. Thus, with most coals a good balance is achieved between direct methane formation and hydrogen generation.

Overall, the Hygas process is almost thermally balanced. As a practical matter, however, significant amounts of energy are rejected to cooling water at low temperature and are therefore unrecoverable. The separation of oxygen from air and its compression to system working pressures requires considerable energy input. Therefore, the Hygas process requires substantial additional energy inputs. Only about 40 percent of the carbon ends up as fuel and thus most of the carbon acts as a reducing agent for the removal of hydrogen from water. This carbon ends up as carbon dioxide.

A block diagram of the Hygas plant is shown in Figure 4.2-1. This diagram and the plant description that follows are based on a number of published reports and articles (References 4.2-2, 4.2-3, 4.2-7 through 4.2-11 which at best disclose only minimum details of the Hygas system. The process itself is currently being evaluated in a 75 ton/day pilot plant at Chicago and the Institute of Gas Technology has taken the position that it will not release detailed information to others at this time. Therefore, the description in this report necessarily reflects this lack of availability of detailed process information. The plant has been sized so that the fuel output is the same as that of GA/SW process, namely 29.5×10^9 Btu/hr.

4.2.1.1 Coal Preparation and Pretreatment

Coal is ground to minus 8 mesh size and is fed to a pretreater if the coal is of the caking variety. The pretreatment consists of a mild surface oxidation consuming about 10 percent of the fuel value and produces a free flowing feedstock.

4.2.1.2 Slurry Preparation

Pulverized coal is mixed with an aromatic oil, a product of the gasification process, to form a heavy slurry (up to 45 percent solids) that is pumped to the hydrogasification reactor.

4.2.1.3 Slurry Vaporizer

The coal-oil slurry is sprayed onto a fluidized bed in the top of the hydrogasification reactor operating at 1000 - 1500 psi and a temperature of 600°F. The hot gases from the lower stages of the process flash the oil and dry the incoming pulverized coal, which is then piped down to the low temperature (first stage) hydrogasification reactor section. The vaporized oil is recovered in the quench unit for recycle to the process.

4.2.1.4 Low-Temperature Reactor

Dried coal from the slurry vaporizer enters an upward moving stream of hot gas where it is rapidly heated to 1200-1400°F.

It flows co-currently with the gas stream, the residence time, a matter of a few seconds, being adjusted so that about 20 percent of the coal is converted to methane. In a disengaging section the partially reacted coal drops through a connecting pipe to the high temperature (second stage) hydrogasification reactor section.

4.2.1.5 High-Temperature Reactor

The char from the low temperature reactor enters a fluidized bed operating at 1600-1800°F. The fluidizing gas, which is composed mainly of steam and hydrogen, converts an additional 25 percent of the coal carbon to synthesis gas and methane. The char at this point is considerably less reactive than in the low temperature stage so that the residence time is necessarily a matter of minutes. The hydrogasified char drops from the bottom of the fluidized bed into the steam-oxygen gasifier.

4.2.1.6 Steam-Oxygen Gasifier

The char from the high temperature gasification reactor section is contacted in a fluidized bed with steam and oxygen at 1800-1900°F to generate a hydrogen-rich gas. It is this hot gas that enters the high temperature gasifier and reacts with the coal char in the various stages of the gasification unit. A high ash char is discharged from the steam-oxygen gasifier via a water slurry. This char containing about 25 percent carbon has a heating value of about 4300 Btu/lb and might possibly be used as a utility fuel. In the present plant layout it is sent to ash disposal.

4.2.1.7 Oxygen Plant

Standard air separation plants are employed.

4.2.1.8 Quench

The gas leaving the top of the hydrogasification reactor passes through a quench system which removes fines, water-soluble trace components such as ammonia, and condenses product and slurry oil and excess steam. A portion of the oil is returned to the slurry preparation unit and the remainder is sent to the product oil recovery unit. The quench water is treated to remove pollutants and recover the dissolved ammonia.

4.2.1.9 Product Oil Recovery

The product oil stream is rectified into a benzene and heavy oil fraction.

4.2.1.10 Purification I

The gas leaving the quench unit is contacted with hot carbonate to remove carbon dioxide and hydrogen sulfide. The hot carbonate is regenerated by heating in a separate unit giving off a H_2S rich stream and a CO_2 rich stream. The CO_2 is vented after removal of H_2S by either incineration to SO_2 followed by scrubbing in a Wellman - Lord unit or by direct conversion to sulfur in a Stretford unit. The H_2S rich stream is sent to a Claus sulfur recovery unit.

4.2.1.11 CO Shift Converter

IGT has not investigated this part of the process but presumably will use the water-gas shift reaction to adjust the hydrogen to carbon monoxide ratio to slightly greater than 3 to 1.

4.2.1.12 Purification II

Again carbon dioxide and residual hydrogen sulfide are scrubbed from the shift converter gas by treatment with hot carbonate. Since almost total removal of sulfur compounds is required in order not to poison the methanation catalyst, the gas is further treated with zinc oxide and activated carbon.

4.2.1.13 Methanation and Dehydration

The purified gas passes through two fixed bed reactors in which the carbon monoxide and hydrogen are reacted over a nickel catalyst to form methane and steam. The initial temperature is about 550°F and rises to about 850°-900°F during the course of this highly exothermic reaction. Cooled product gas is partially recycle to control the reaction temperature. Additional cooling condenses steam from the product stream and the resulting dried gas is of pipeline quality ready for shipment.

4.2.1.14 Other Plant Auxiliaries

Supporting plant auxiliaries include a process steam generation plant equipped with particulate emission control and a Wellman-Lord SO_2 scrubbing unit, a Claus plant for the conversion of H_2S to sulfur, a Stretford unit for removal of residual sulfur pollutants, an ash disposal unit, coal storage and handling facilities, and miscellaneous other equipment such as for the treatment and handling of feedwater, cooling water, waste water and the distribution of electric power.

4.2.2 Features and Limitations

The energy recovery for the Hygas steam-oxygen process is shown in the following tabulation (Reference 6.2-2).

Input	10^6 Btu/hr	Percent Input
Coal to gasifier (2081.3 tons/hr x 2,000 x 8,806.2 Btu/lb)	36,657	82.5
Coal to utility (440.0 tons/hr x 2,000 x 8,806.2 Btu/lb)	7,750	17.5
	44,407	100.0

Output	MM Btu/hr	Percent Input
Gas (657.9 MMscfd x 966.5 Btu/scf x 1/24)	26,494	59.7
Benzene (35,500 lb/hr x 18,126.4 Btu/lb)	643.5	1.4
Heavy Oil (146,250 lb/hr x 16,096 Btu/lb)	2,354	5.3
Useful Fuel Output*	29,491.5	66.4

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* By-produces include ammonia (190.7 TPD), sulfur (173.4) TPD and char ash (6800 TPD) which nominally add 2,610 MM Btu/hr or 5.9 percent to the plant fuel output. Over 90 percent of this energy comes from the char.

The feedstock in this case is a Montana sub-bituminous coal which is believed to have the following characteristics:

Moisture	25.5 percent (wt)
Volatile Matter	28 percent (wt)
Fixed Carbon	38 percent (wt)
Ash	8.5 percent (wt)
Sulfur	0.75 percent (wt)
Btu/lb	8806
Ash Softening Temp, °F	2250

In pilot plant experience to date, which includes a 27 day continuous run and operation where no external heat has had to be added, some difficulties have been encountered with recovery of the slurry recycle oil. It is important to the process that the recovery rate be nearly 100 percent.

The Montana sub-bituminous coal feedstock used in the present study does not require pretreatment. Eastern and mid-western coals would require this treatment, adding significantly to both capital and operating costs.

4.2.3 Economic Evaluation

Capital costs for the Hygas Steam - Oxygen Process are reported in Table 4.2-1 for a plant equivalent in fuel output to that of the GA/SW plant. These amount to $784.8 million, exclusive of working capital and start-up expenses. The basis for this estimate is information recently published in Reference 4.2-2. Some comments are in order on several items contained in the cost estimate:

(1) The estimate contains no contingency factor. For an untried process of this type inclusion of a 15 percent contingency factor is usually recommended.

(2) The cost of the hydrogasifiers represents an area of considerable uncertainty. For example, C. F. Braun has calculated that three reactors, 22 feet in diameter and 250 feet tall would be required for a 250 billion Btu/day plant (Reference 4.2-12). Reactor weight would be 3000 - 4000 tons each.

Construction and installation costs for such huge, first of a kind vessels are difficult to estimate accurately.

(3) Based on UE&C's experience with SO_2 scrubbers the cost of Wellman-Lord unit has been checked out and found to be fairly accurate. It is fair to point out, however, that if other than a low sulfur (0.75 percent) Montana sub-bituminous coal is used as feedstock, then the cost of the Wellman-Lord unit could easily escalate by a factor of four or five or more.

Operating costs are detailed in Table 4.2-2. The effect of coal cost in the range $0.50 - 1.75/$10^6$ Btu on the product cost has been calculated. Similarly the effect of capital fixed charge costs at 15 percent and 25 percent is also shown. A graphical representation of the effect of these two parameters is shown in Figure 4.2-2. The fuel (pipeline gas and liquid hydrocarbon products together) cost is seen to vary between $1.47 and $3.68/$10^6$ Btu on this basis. These figures agree reasonably well with those published by IGT (Reference 4.2-2), since both were calculated from essentially the same data base.

4.2.4 Major Pollution Outputs

The major products of concern from the environmental viewpoint and their expected amounts are listed below:

Elemental Sulfur	173.4 TPD
Sulfur Dioxide	167.3 TPD
Ammonia	190.7 TPD

Hydrogen Cyanide	0 - 5 TPD
Phenols	25 - 170 TPD
Oil and Tars	Trace to 1000 TPD
Mercury	Less than 4 lb/day
Ash	7700 TPD

In the proposed plant operation, the sulfur dioxide will be scrubbed to the point where the SO_2 emissions will comply with federal standards ($1.2\ lb/10^6$ Btu). Elemental sulfur will be either sold or stockpiled and should not constitute a major pollution hazard in either case. Note, however, that if a high sulfur mid-western or eastern coal were to be used that the sulfur dioxide scrubbing and elemental sulfur disposal would be more expensive.

One of the principal waste streams will be the foul water from the quench system. This will contain ammonia, hydrogen cyanide, phenols, as well as small amounts of dissolved benzene, oils, and hydrogen sulfide. Little has been published on how these pollutants will be handled in the Hygas process, but the following procedures are believed to be applicable. The phenols, benzene and oils can be recovered by liquid extraction and either sold, used as fuel, or recycled to the gasifier. The ammonia and hydrogen sulfide can be stripped from the water using the Chevron Waste Water Treatment Process. The ammonia should represent a marketable item. By a modification of this process it may be possible to separate and recover the hydrogen cyanide, which then could be incinerated and disposed of. Finally, the residual water may have to undergo an activated sludge pond or active carbon treatment before it is suitable for re-use or disposal.

In general it has been proposed that the ash from the utility boiler combustion unit and from the steam - oxygen gasifier be handled by conventional power - plant disposal technology. It should be noted that the ash from the gasifier may be finer than normal power plant ash and, therefore, dustier. If recovery of fuel values from the gasifier ash and the pretreatment flue gas is attempted, then special equipment may be needed to trap the sulfur concentrated in these fuel sources.

The mercury in the coal in all likelihood will be trapped at several stages of the process, but will surely be removed by the activated carbon bed used to remove carbon dioxide and hydrogen sulfide from the product gas stream. Special provisions may be needed for its isolation during reactivation of the carbon.

4.2.5 References

4.2-1 H. M. Siegel and T. Kalina: The Oil and Gas Journal, Feb. 12, 1973, 87-94

4.2-2 K. C. Vyas and W. W. Bodle: The Oil and Gas Journal, March 24, 1975, 45-54

4.2-3 Process Design and Cost Estimate for a 258 Billion Btu/day Pipeline Gas Plant- Hydrogasification Using Synthesis Gas Generated by Electrothermal Gasification of Spent Char, Institute of Gas Technology Report No. OCR/R & D 22/Int. 3, January 1968

4.2-4 Electrothermal Hygas Process Escalated Costs, Institute of Gas Technology Report No. OCR/R & D 22/Int. 6, June 1971

4.2-5 U. S. Energy Outlook - Coal Availability, National Petroleum Council Report, 1973, 237-247

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4.2-7 J. P. Henry, Jr. and B. M. Louks: Chem. Tech., April 1971, 236-247

4.2-8 Symposium Papers - Clean Fuels from Coal, Institute of Gas Technology, Chicago, Illinois, Sept. 10 - 14, 1973, 219 - 240

4.2-9 Shaping Coal's Future through Technology 1974 - 1975, Office of Coal Research, 41-47

4.2-10 79th National Meeting - 8th Petrochemical and Refining Exposition of A.I.Ch.E., Houston, Texas, March 16 - 20, 1975, Institute of Gas Technology, "Progress in the Hygas Process" by F. C. Schora

4.2-11 Private Communication from C. L. Tsaros, Institute of Gas Technology, May 27, 1975

4.2-12 Evaluation of Coal Conversion Processes to Provide Clean Fuels Part II, February 1974, Electric Power Research Institute

TABLE 4.2-1

HYGAS PIPELINE GAS PROCESS
CAPITAL COST ESTIMATE
JULY, 1974 DOLLARS

	Capital Cost $ Million
Coal grinding	24.4
Slurry feed system	22.6
Hydrogasifiers	64.3
Steam - Oxygen gasifiers	21.1
Quench system	31.3
Benzene recovery	9.9
Purification I (hot carbonate)	19.6
CO shift conversion	17.5
Purification II (hot carbonate, active C, ZnO)	35.9
Methanation and drying	28.0
Oxygen manufacture	67.1
SO_2 removal, Wellman-Lord	26.2
Sulfur recovery, Claus plant and Stretford	5.6
Particulate emission control	6.9
Subtotal battery limit	380.4
Coal storage and handling	9.9
Waste water treatment	34.0
Char handling and ash disposal	17.8
Turbine generator	15.5
Process and turbine steam generation	76.3
In-plant electric power distribution	16.0
Cooling tower and off-site distribution	7.6
Boiler feedwater treating	16.5
Miscellaneous	3.6
General off-sites investment	34.0
Subtotal outside battery limits	231.2
Direct plant investment	611.6
Contractor's overhead and profit (10%)	61.2
Total plant investment	672.8
Interest during construction (8% for 4 years)	112.0
Total capital required, exclusive of working capital and start-up expenses	784.8

TABLE 4.2-2

HYGAS PIPELINE GAS PROCESS
PRODUCTION COST ESTIMATE
JULY, 1974 DOLLARS

	Product Cost $ per million BTU
Coal, feed and utility	0.76 to 2.63
Other raw materials and supplies	0.07
Utilities	0.01
Direct operating cost	0.13
Fixed charges	0.50 to 0.84
Total operating cost	1.47 to 3.68

Basis:

Fuel output	707.8×10^9 Btu/day

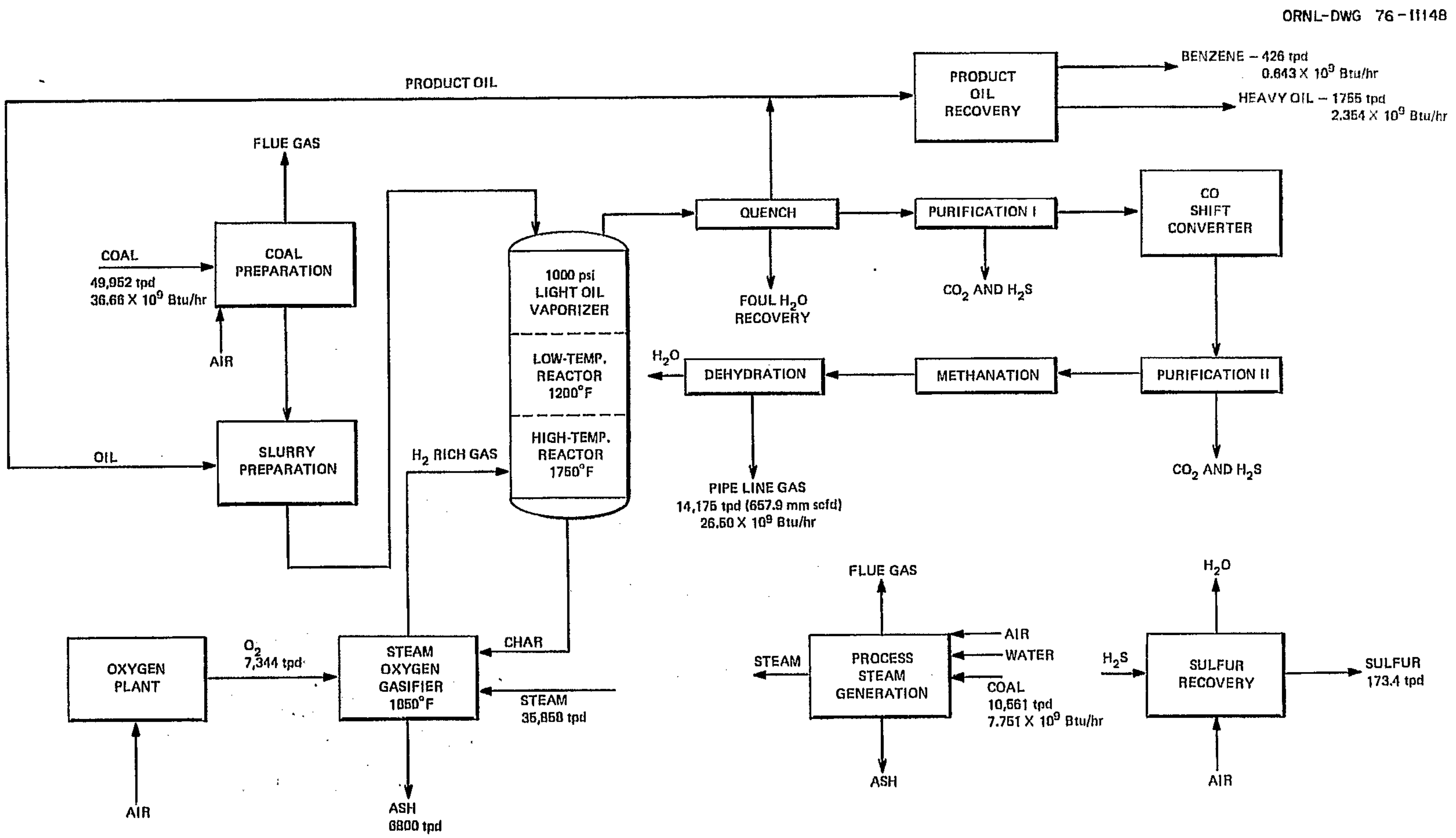

Fig. 4.2-1. Plant Block Diagram — Hygas Pipeline Gas Process

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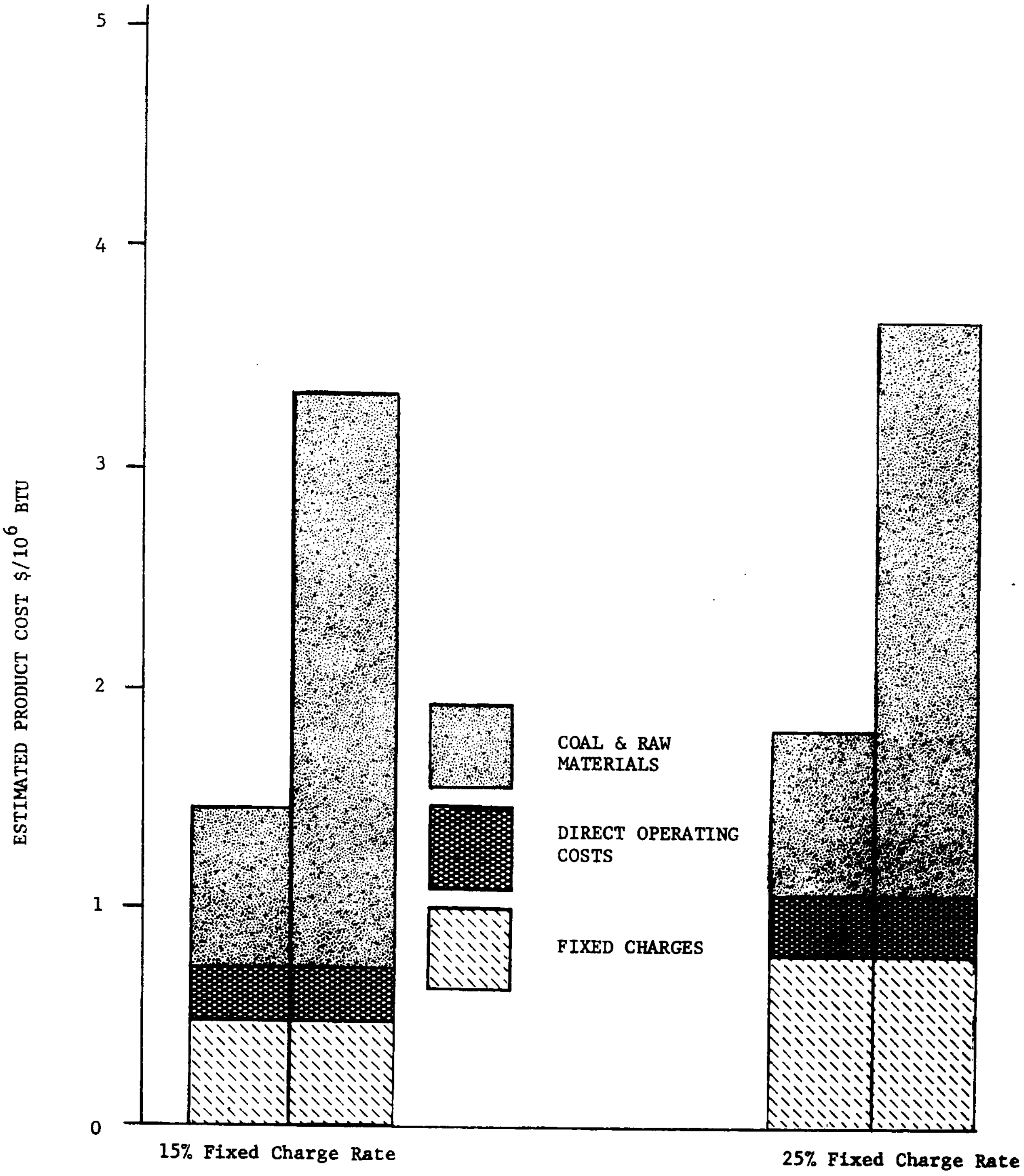

Fig. 4.2-2. Production Costs — Hygas Pipeline Gas Process

4.3 Koppers-Totzek Synthesis Gas Plant

4.3.1 Description of the Process

A block diagram of a synthesis gas process using Koppers-Totzek gasifiers is presented in Figure 4.3-1. The Koppers-Totzek process reacts pulverized coal, oxygen and steam at atmospheric pressure and up to 3300°F. The gasifier, as currently designed, is a refractory-lined horizontal vessel with either two or four opposing burners (Reference 4.3-1). A jacket around the gasifier is supplied to remove excess heat in the form of low pressure steam.

Dried, pulverized coal is mixed with a stream of oxygen and steam and reacted in the gasifier in an entrained state. The carbon and volatile material are gasified while a portion of the ash and unreacted material are converted into molten slag and collected in the bottom of the vessel.

Gas leaving the gasifier can be direct water quenched to solidify and remove entrained slag particles, and then passed through a waste heat boiler capable of producing high pressure steam (up to 1500 psia). The raw gas composition typically has the following analysis. (Reference 4.3-2).

Component	Mole Percent
CO_2	7
CO	39
H_2	24
H_2O	28
Others	2

In order to produce a synthesis gas with a H_2/CO ratio of approximately two and a total H_2 plus CO content of over 90 percent, approximately one half of the raw synthesis gas is sent to a water shift conversion process, where CO and steam react to form CO_2 and H_2. This stream is then mixed with the unshifted raw gas stream. The resulting gas is compressed and purified by the Lurgi Rectisol Process, which removes both CO_2 and H_2S. The acid gases are sent to a sulfur recovery unit for final processing, where elemental sulfur is produced by the Stretford Process; the CO_2-rich stream is vented to the atmosphere. The resulting clean synthesis gas is approximately 65 percent hydrogen, 31 percent carbon monoxide and 4 percent carbon dioxide, and is suitable for both methanol production and iron ore reduction.

4.3.2 Features and Limitations

The Koppers-Totzek synthesis gas plant has an overall energy efficiency of 55 percent. This efficiency is based upon the assumption that all process steam and electricity requirements are produced on-site using coal as the energy source. It should be noted that a significant portion of the steam requirements of the plant can be supplied by the highly exothermic gasification and water gas shift reactions.

An important advantage of the Koppers-Totzek gasifier is that it can use all of the coal, including fines; separation is not necessary prior to the gasification step. Furthermore, any type or rank of coal is acceptable (Reference 4.3-3). As a result, a minimal amount of coal preparation is required for the Koppers-Totzek process when compared to the coal preparation requirements of the Lurgi process.

A typical Koppers-Totzek feed coal has the following analysis:

Component	Wt. Percent
Carbon	60.78
Hydrogen	4.23
Nitrogen	0.95
Sulfur	3.47
Oxygen	8.51
Ash	8.76
Moisture	13.30

The Koppers-Totzek gasification process has the disadvantage of operating at a low pressure, slightly above atmospheric. Large compressors are required to bring the product gas up to the discharge pressure of 1000 psi. As a result, a significant fraction of the total coal feed is burned in a utility boiler to provide steam for the high pressure turbines which drive the gas compressors. This requirement is reflected in the lower process efficiency compared to the other gasification processes which operate at elevated pressures.

The Koppers-Totzek gasification process also has the disadvantage of requiring more oxygen per ton of product gas than the Lurgi process. This requirement is due to the entrained mode of operation, which causes the crude gases to leave the gasifier at very high temperatures and increases oxygen consumption.

4.3.3 Major Pollution Outputs

The major pollution outputs from the process will be:

1) Vent gas from the sulfur plant - A gas stream consisting primarily of CO_2, N_2 and H_2O will contain a total of less than 100 ppm of COS, CS_2 and H_2S.
2) Gas from sulfur incinerator - Approximate 300 ppm of SO_2 will be included in this stream which is mostly N_2, CO_2 and H_2O.
3) Wet slag - Approximately 1006.9 TPD of wet slag will require disposal.
4) Emissions from the utility boiler will amount to 98 TPD of SO_2, 57 TPD of NO_x and 8 TPD of particulate.

4.3.4 Economic Evaluation

Capital costs for the Koppers-Totzek synthesis gas process are presented in Table 4.3-1. As shown, the investment includes process units, utilities and off-sites. The power requirements of the overall process were assumed to be produced on-site using a conventional coal fired steam electric generating facility.

The total fixed investment, capital investment plus interest during construction, for a Koppers-Totzek plant producing 15.5×10^9 Btu/hr of synthesis gas is approximately $1073 million. Production costs are presented in Figure 4.3-2 as a function of fixed charge rate and coal price. The high and low estimates are broken down into component costs and given in Table 4.3-2. The major production costs are for coal and those annual expenses expressed by the fixed charge rate (depreciation, return on investment, insurance, taxes, etc.). Other production costs, catalyst and chemicals and direct operating costs, account for only \$0.13/10^6 Btu of synthesis gas.

4.3.5 References

4.3-1 Bodle, W. W. and K. C. Vyas, Clean Fuels from Coal, The Oil and Gas Journal, August 26, 1974.

4.3-2 Information from the Koppers Company, Inc., Pittsburgh, Pennsylvania, April 16, 1974.

4.3-3 Perry, Harry, Coal Conversion Technology, Chemical Engineering, July 22, 1974.

4.3-4 Magee, E. M., Johnig, C. E. and Shaw, H., Evaluation of Pollution Control in Fossil Fuel Conversion Processes - Gasification, Section 1 - Koppers-Totzek Process, EPA-6501, 2-74-009a, January 1974.

TABLE 4.3-1

KOPPERS - TOTZEK SYNTHESIS GAS PROCESS
CAPITAL COST ESTIMATE
JULY, 1974 DOLLARS

Process Units	Capital Cost $ Million	
Coal Sizing	19	
Gas Production	278	
Crude Gas Shift	12	
Gas Cooling, Purification & Refrigeration	97	
Sulfur Recovery	47	
Product Compression	50	
Total Process Units		511
Utilities		
Air Compression and Oxygen Production	148	
Power and Steam Generation	101	
Total Utilities		249
Offsites		
Water Treating and Cooling Water System	45	
Slag Removal and Raw Water System	50	
General Facilities	65	
Total Offsite		160
Total Capital Investment		920
Interest During Construction (8% for 4 Years)		153
Total Fixed Investment		1,073

TABLE 4.3-2

KOPPERS-TOTZEK SYNTHESIS GAS PROCESS
PRODUCTION COST ESTIMATE
JULY, 1974 DOLLARS

	Product Cost $/10^6$ Btu		
	Low		High
Coal	0.90		3.16
Catalyst and Chemicals		0.03	
Direct Operating Cost		0.10	
Fixed Charges	1.31		2.18
Total Production Cost	2.34		5.47

Basis:

Fuel Output 371.4 x 10^9 Btu/day

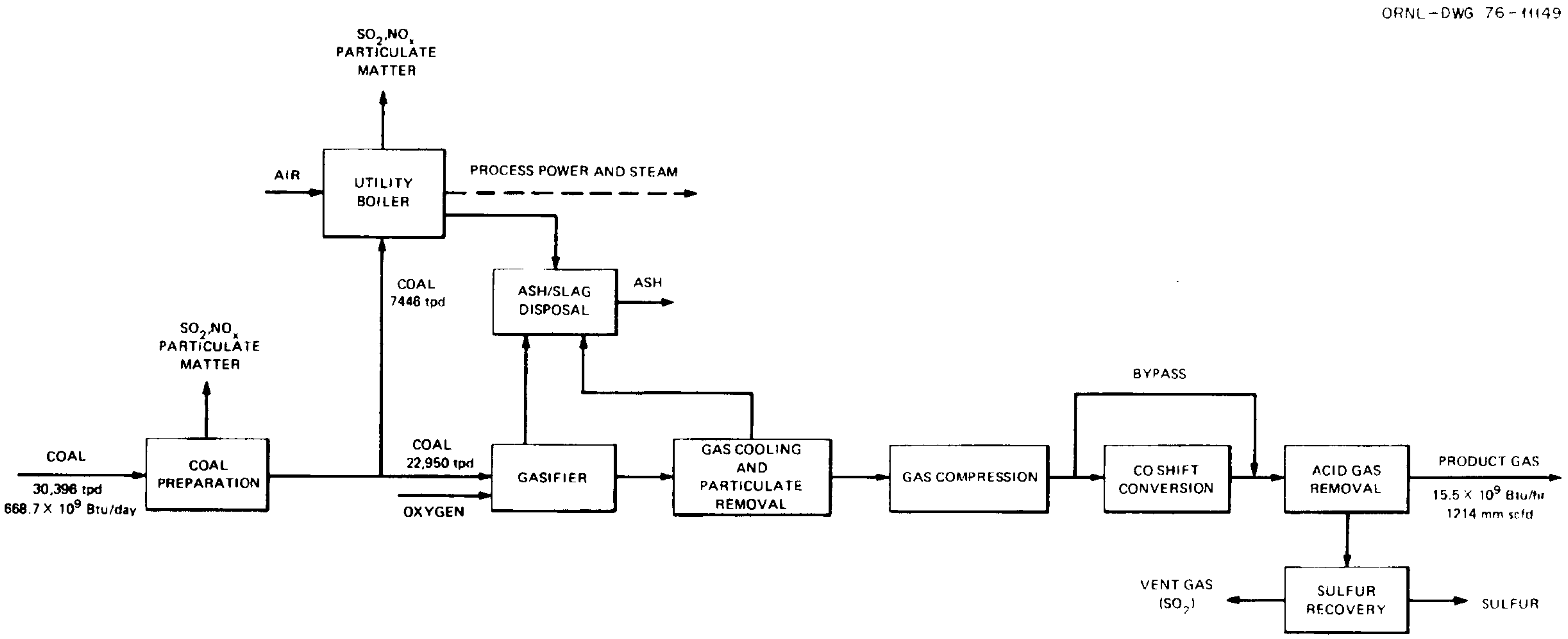

Fig. 4.3-1. Plant Block Diagram — Koppers-Totzek Synthesis Gas Process

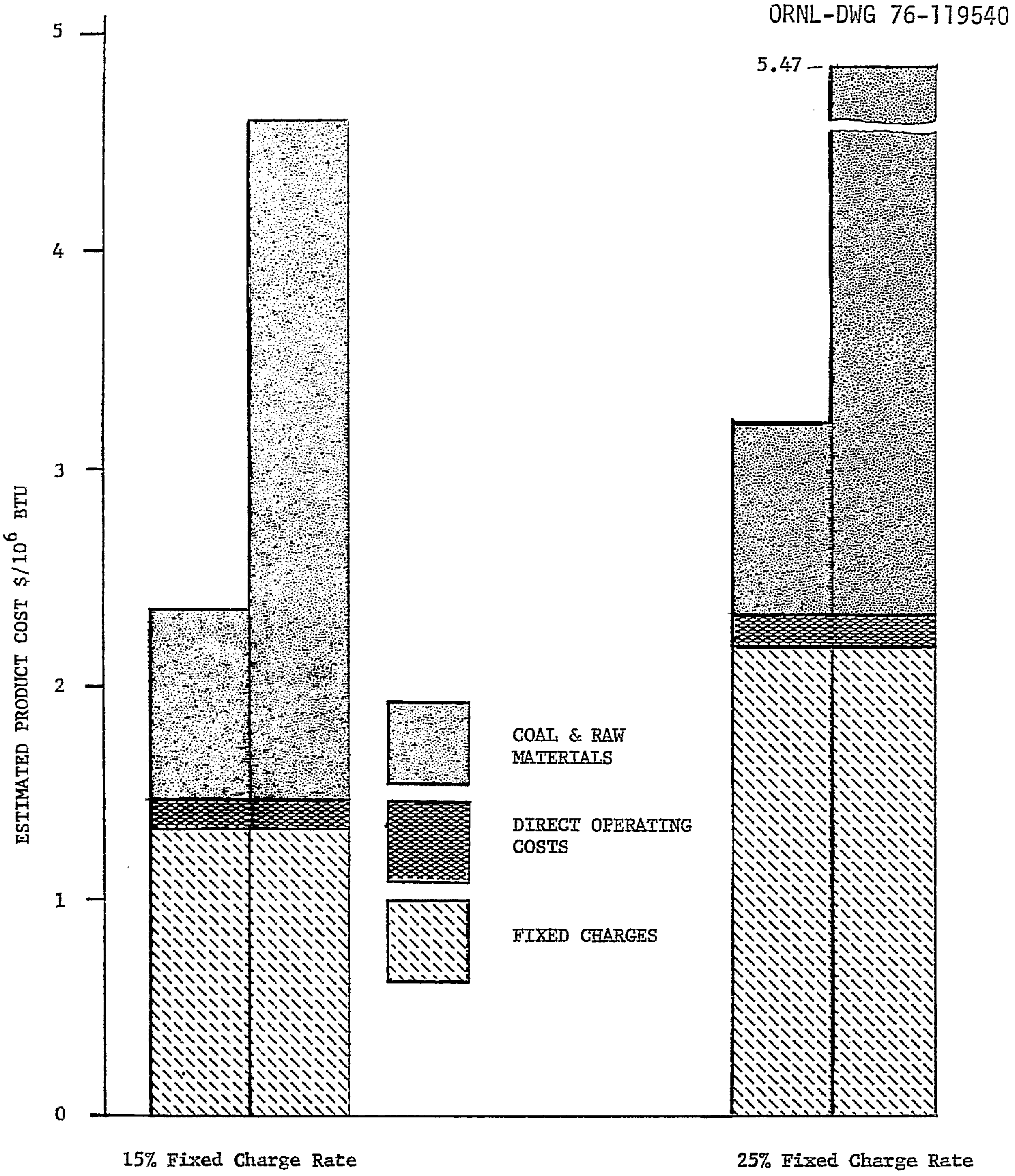

Fig. 4.3-2. Production Costs — Koppers-Totzek Synthesis Gas Process

4.4 Koppers-Totzek Hydrogen Plant

4.4.1 Description of the Process

Figure 4.4-1 is a block diagram of a hydrogen plant utilizing Koppers-Totzek coal gasifiers. The only significant difference between this plant and the previously described synthesis gas plant is that all of the crude gas is shifted; the same feed coal has been utilized. This results in a high concentration of hydrogen and carbon dioxide, with a very small percentage of carbon monoxide in the crude hydrogen stream.

This stream is then processed in a purification section where H_2S and CO_2 are removed by the Lurgi Rectisol Process, using low temperature methanol. The final product is approximately 97.5 percent H_2, with N_2, CO_2 and CO comprising the remaining portion. As in the synthesis gas plant, the H_2S is sent to a sulfur recovery plant, while the CO_2 stream is vented to the atmosphere.

4.4.2 Features and Limitations

Since the gross heating values of CO and H_2 are nearly equal on a Btu/lb mole basis, the Btu content of the gas stream remains relatively unchanged by converting a mole of CO to H_2 via the water gas shift reaction. As a result, the production costs expressed in $\$/10^6$ Btu are nearly equivalent for the Koppers-Totzek hydrogen and synthesis gas processes.

The hydrogen process has an overall energy efficiency of 55 percent.

As mentioned previously, the main advantage of the Koppers-Totzek gasifier is that any type or rank of coal can be used, including fines. The main drawback of the process is the low operating pressure of the gasification step which necessitates the use of large expensive compressors to bring the gas from near atmospheric pressure to 1000 psi.

4.4.3 Major Pollution Outputs

The major pollution outputs will be the same as those of the synthesis gas plant. In both cases, the single largest source of pollution is the utility boiler. Based on federal emission standards for fossil fueled steam generators, approximately 109 TPD of SO_2, 63 TPD of NO_x and 9 TPD of particulates will be produced.

More slag and ash will require offsite disposal (2925 TPD) than for the synthesis gas plant, because of the higher coal gasification rate required to produce the desired amount of hydrogem.

4.4.4 Economic Evaluation

Capital costs for the hydrogen plant are presented in Table 4.4-1. A total fixed investment of $1180 million is required to produce the desired amount of hydrogen, 16.965 x 10^9 Btu/hr. Production costs as a function of coal price and fixed charge rate are essentially equal to those for the synthesis gas plant and are shown in Figure 4.4-2. A breakdown of the production costs is summarized in Table 4.4-2.

TABLE 4.4-1

KOPPERS - TOTZEK HYDROGEN PROCESS
CAPITAL COST ESTIMATE
JULY, 1974 DOLLARS

Process Units	Capital Cost $ Million	
Coal Sizing	21	
Gas Production	304	
Crude Gas Shift	24	
Gas Cooling, Purification & Refrigeration	111	
Sulfur Recovery	47	
Gas Compressor	74	
Total Process Units		581
Utilities		
Air Compression and Oxygen Production	162	
Power Generation	109	
Total Utilities		271
Offsites		
Water Treating and Cooling Water System	45	
Slag Removal and Raw Water System	50	
General Facilities	65	
Total Offsite		160
Total Capital Investment		1,012
Interest During Construction (8% for 4 Years)		168
Total Fixed Investment		1,180

TABLE 4.4-2

KOPPERS-TOTZEK HYDROGEN PROCESS
PRODUCTION COST ESTIMATE
JULY, 1974 DOLLARS

	Product Cost $\$/10^6$ Btu		
	Low		High
Coal	0.88		3.09
Catalyst and Chemicals		0.04	
Direct Operating Cost		0.10	
Fixed Charge Rate	1.32		2.19
Total Production Cost	2.34		5.45

Basis:

Fuel Output 407.3×10^9 Btu/day

ORNL-DWG 76-11150

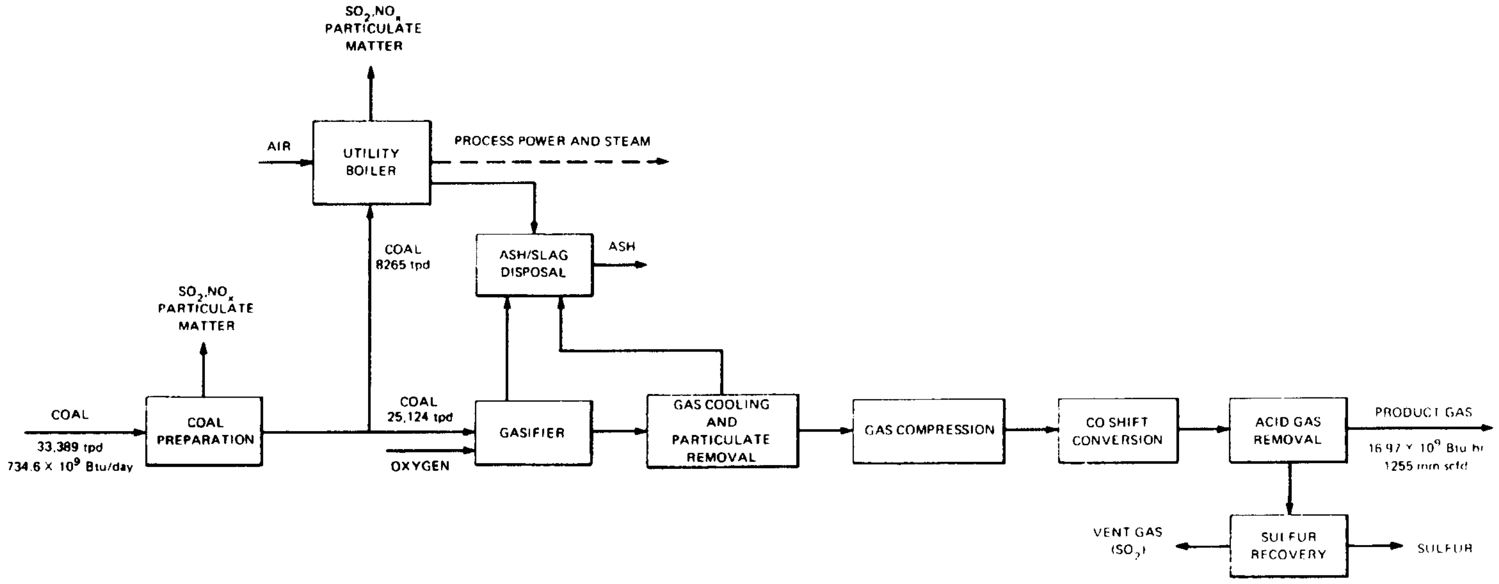

Fig. 4.4-1. Plant Block Diagram — Koppers-Totzek Hydrogen Process

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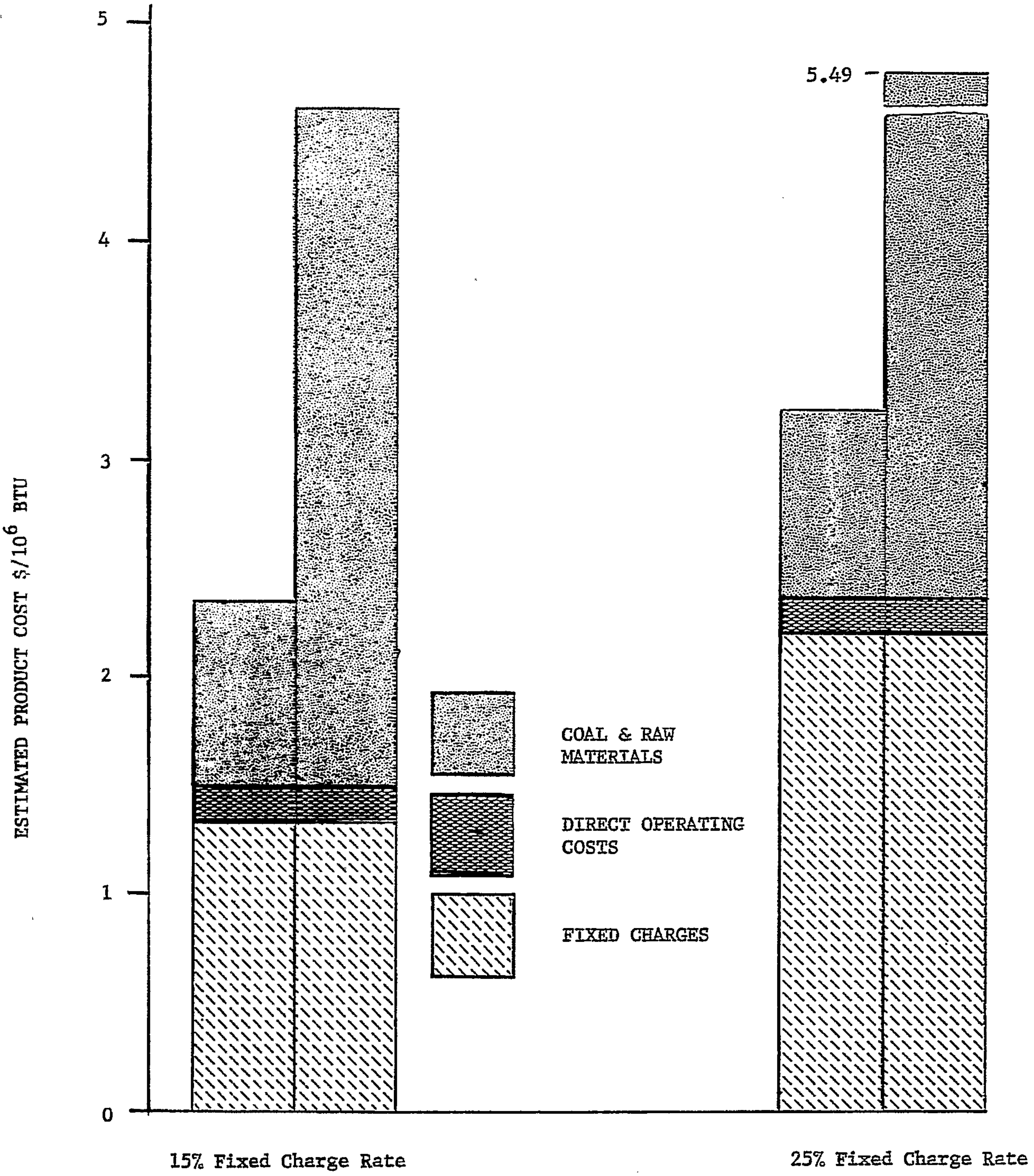

Fig. 4.4-2. Production Costs — Koppers-Totzek Hydrogen Process

5. COAL LIQUEFACTION

5.1 Description of the Process

Under a separate contract to Oak Ridge National Laboratory, United Engineers prepared the preliminary design of a coal liquefaction process which can be powered by either nuclear or fossil heat. This study is reported in detail in Reference 5-1 and is summarized here. Table 5-1 shows the operating characteristics of the plant.

The process designed by United Engineers is a modified and improved version of the H-Coal process (Reference 5-3). It produces heavy naphtha, heavy oils, and enough fuel gas to provide heat for the process. The plant has multiple trains of equipment in all sections except the nuclear-heated reformer. A block diagram of the plant is shown in Figure 5-1.

The process is designed to use 60,000 tons per day of Western Kentucky coal. The fossil-fueled plant produces 1.114×10^{12} Btu/day of liquids and char; the nuclear version produces 1.305×10^{12} Btu/day. The processes are energy self-sufficient, burning char to provide electricity which is required. The nuclear-heated plant produces some electricity using steam from the reactor.

5.1.1 Sections Common to Fossil-heated and Nuclear-heated Plants

The following sections of the process are common to both plants:

Section 100 - Coal preparation - Coal is reclaimed, pulverized and dried to 2 weight percent moisture using hot flue gas from fired heaters.

Section 200 - Coal hydrogenation - Coal is slurried with a heavy oil from the process, heated, and catalytically hydrogenated in an ebullated bed.

Section 300 - Product separation - Reaction products are separated by flashing, fractionation and absorption.

Section 400 - Fluid-bed coking - Oils are thermally cracked, producing lighter fractions. Heat is supplied by burning a portion of the char. A high-Btu fuel gas is produced and burned in the reformer or in fired heaters at various stages in the process.

Section 600 - Hydrogen production - Carbon monoxide and steam react to form hydrogen and carbon dioxide in a conventional shift converter.

Section 700 - Ammonia and sulfur recovery - Ammonia is recovered in its anhydrous form. Hydrogen sulfide is reacted with sulfur dioxide to form elemental sulfur.

Section 1000- Cooling water system.

Section 1100 - Water treatment system.

Section 1200 - Residue handling system.

Section 1300 - Wastewater treatment system.

5.1.2 Fossil-Heated Plant

In the fossil version of the liquefaction plant, 52 x 10^9 Btu/day of high-Btu gas produced in the process are burned in a conventional steam-methane reformer. The steam-methane mixture is reformed in the presence of catalysts to form hydrogen and carbon monoxide.

The reformed gas then passes through a catalytic shift converter to form additional hydrogen and carbon dioxide.

5.1.3 Nuclear-Heated Plant

The intermediate loop helium coolant in a 3000 MWe VHTR heats the steam-methane reformer in the nuclear-heated plant. The reaction conditions and shift conversion are the same as the fossil version of the plant.

The intermediate loop helium is heated by exchange with hotter helium circulating in a primary loop from the VHTR core to the intermediate exchanger to a power steam generator and then back to the core. The power steam generators produce high pressure steam to drive the helium circulators and the high pressure power turbine. The exhaust from the power turbine is condensed by exchange with boiler feed water and returned to the power steam generator. The steam produced by exchange is fed to the reformer. Reference 5.2 provides a detailed description of the VHTR system.

5.2 Economic Evaluation

The capital cost of the coal liquefaction plants were estimated using United Engineers' coal and chemical processing experience and estimates prepared by others for similar plants. Costs for the process and off-site sections were based on estimates published by Hydrocarbon Research, Inc. (Reference 5-3), American Oil Company (Reference 5-4), the Bureau of Mines (Reference 5-5), and General Atomic Company (Reference 5-6 and 5-7). Applicable costs of equipment and/or sections were converted to July 1974 dollars using the Chemical Engineering (CE) plant cost index. Because the plant consists of multiple, identical equipment trains, the cost of the equipment was assumed to be directly proportional to the plant capacity. The escalated and extrapolated costs were then compared and adjusted as required for differences in design. The costs selected were based on agreement between two or more independent referenced estimates and/or the conservative judgement of United Engineers. Electricity generating costs are based on recent United Engineers experience.

The estimated capital cost includes equipment, structures, engineering, fee, 15% contingency, field indirects, and administration and overhead. The estimate does not include licensing fees or royalties, process development costs, environmental impact studies, land acquisition, and other items not normally included. These excluded costs have a negligible effect on the cost of the products of these plants.

The costs of catalysts and other chemicals were obtained from the referenced cost estimates, and escalated to July 1974 dollars using the CE industrial chemicals wholesale price index. These costs were adjusted for the size and stream factor of the plant.

Direct operating costs were also obtained from the referenced estimates. Operating and supervisory labor costs were adjusted for plant size, stream factor, the CE chemical products hourly earnings index, and the CE chemical products productivity index. Maintenance was assumed to be 4% of the capital cost for on-site facilities and 2% for off-sites.

The capital, operating, and fuel costs for these plants were calculated in dollars per million But's of products, and are presented in Table 5-3 and shown in Figure 5-2. Two cases, high and low, are shown for each plant. The material and fuel costs for these cases correspond to the high and low guidelines presented in Section 2.

Also shown in Table 5-3 are credits for ammonia, sulfur, and the residual char. The ammonia and sulfur are useful industrial-quality products if a customer is found for these quantities. The values of the ammonia and sulfur were obtained from the Bureau of Mines (Reference 5-5). The char has a heating value of 8235 Btu/lb., but the sulfur and ash contents are fairly high (4.3 and 36 weight percent). The credit of $3.00 per ton of char was assumed by United Engineers.

5.3 Major Pollution Outputs

The pollution which will be produced by the liquefaction plants is discussed in detail in References 5-1 and 5-8. As shown in Table 5-4, the major sources of air pollution are the utility boiler, the fired process heaters, and the reformers in the fossil-heated version. Air emissions were designed to meet all existing federal regulations except those for plants operating in specific air basins. The nuclear-heated plant, of course, shows much less air pollution than the fossil version.

Waste water from the progress will be treated to neutralize the water and remove oil. The treated water will be used as makeup to the cooling towers.

Solid wastes will be primarily the ash from the utility boiler and will be disposed using conventional techniques. If the residue from the liquefaction process cannot be sold or used economically as fuel, the volume of solid waste will be much larger.

The rate of water consumption for the fossil-heated plant is about 36 million gallons per day (mgd). The nuclear-heated plant uses approximately 32 mgd. The difference in use is due mostly to the smaller demand for fossil-generated electricity.

5.4 References

5-1 Evaluation of A Coal Liquefaction Process using Either Nuclear or Fossil Heat Source, February 1976, United Engineers & Constructors Inc. (UE&C-UCC-760223).

5-2 A Cost Comparison of Very High Temperature Reactors For Process Heat Applications, March 1975, United Engineers & Constructors Inc. (UE&C-AEC-750311).

5-3 Project H-Coal Report on Process Development, Office of Coal Research, R & D Report No. 26 - Final Report, Hydrocarbon Research, Inc., December, 1968.

5-4 Evaluation of Project H-Coal, Office of Coal Research, R & D Report No. 32 - Final Report, American Oil Company, December 1967.

5-5 H-Coal Process, Liquid Fuels from Wyodak Coal - 25,000 Barrels per Day Liquid Fuels Plant, an Economic Analysis, Report No. 75-23, Process Evaluation Group, Bureau of Mines, March 1975.

5-6 Final Report Application Study of a Nuclear Coal Solution Gasification Process for Oklahoma Coal, Volumes 1 and 2, May 1972, General Atomic Company/Stone & Webster (GA-A12068).

5-7 High-Temperature Nuclear Heat Source Study, September 27, 1974, General Atomic Company.

5-8 An Evaluation of Pollution and Water Consumption by Selected Coal Conversion Processes, February 1976, United Engineers & Constructors Inc. (UE&C-UCC-760227).

TABLE 5-1

COAL LIQUEFACTION PROCESS

PLANT OPERATING CHARACTERISTICS

	Fossil	Nuclear
CAPACITY		
Crude Liquids, 10^9 Btu/day	923	1,033
All Products, 10^9 Btu/day	1,114 (1)	1,305 (2)
EFFICIENCY		
Energy Efficiency excluding Char, %	65	65
Energy Efficiency including Char, %	78	78
Coal Use Efficiency, %	78 (1)	90 (2)
MATERIAL AND ENERGY REQUIREMENTS		
Nuclear Heat, 10^9 Btu/day	--	244.8
Coal (7.5% moisture), Tons/day	60,000	60,000
10^9 Btu/day	1,431	1,431
Water, Tons/day	150,000	150,000
Catalysts and Chemicals, $/year	21.2×10^6	21.2×10^6
PRODUCTS		
Heavy Naphtha (18,200 Btu/lb.), 10^9 Btu/day	334	405
Mid-distillate (17,430 Btu/lb.), 10^9 Btu/day	298	298
Heavy Oil (17,250 Btu/lb.), 10^9 Btu/day	247	286
Heavy Gas Oil (17,010 Btu/lb.), 10^9 Btu/day	44	44
Coker Gas (1,080 Btu/scf), 10^9 Btu/day	--	52
BY-PRODUCTS		
Ammonia, Tons/day	441	441
Sulfur (Elemental), Tons/day	1,200	1,200
Char (8,235 Btu/lb.), Tons/day	11,518	13,280
POLLUTANTS, Tons/day		
Sulfur Dioxide	113	28
Nitrogen Oxides	45	12
Particulates	2.6	1.1
Solid Waste	1,128	494

Notes: (1) Includes char.
(2) Includes char and coker gas.

TABLE 5-2

COAL LIQUEFACTION PROCESS

CAPITAL COST ESTIMATE

JULY 1974 DOLLARS

Section		Capital Cost, Million $	
		Fossil	Nuclear
100	Coal Preparation	80	80
200	Coal Hydrogenation	320	320
300	Liquid Products/Solvent Recovery		
400	Fluid Coking	60	60
500	Reforming & Shift Conversion	320	235
600	Hydrogen Production		
700	Ammonia & Sulfur Recovery	9	9
1000, 1100, 1200, 1300,	Misc. Off-sites	178	178
1400	Electricity Generation (By Char Combustion)	64	36
	Tankage	33	33
	Subtotal	1,064	951
	Interest During Construction (8% for 4 years)	177	158
	VHTR System(1) (Includes IDC at 8% for 8 years)	---	800
	Total	1,241	1,909

---

(1) See Section 2, Basis of Study

TABLE 5-3

COAL LIQUEFACTION PROCESS

PRODUCT COST ESTIMATE

JULY 1974 DOLLARS

	Product Cost, \$/Million Btu (1)			
	Fossil Heat (2)		Nuclear Heat	
	Low	High	Low	High
Coal	.78	2.71	.66	2.31
Nuclear Fuel	-	-	.06	.14
Raw Water	.01	.01	.01	.01
Catalysts & Chemicals	.12	.12	.10	.10
Direct Operating Cost				
Process Plant	.15	.15	.12	.12
Electricity Generation	.03	.03	.02	.02
HTGR	-	-	.03	.03
Fixed Charges	.61	1.02	.80	1.33
Total Product Cost	\$ 1.70 to \$4.04		\$1.80 to \$4.06	
Potential Credits				
Ammonia	.03		.02	
Sulfur	.03		.03	
Char	.04		.04	
Potential Product Cost	\$1.60 to \$3.94		\$1.71 to \$3.97	

---

(1) \$/million BTU of gas and liquids, excluding char.

(2) For low and high cost assumptions, see Section 2, Basis of Study.

TABLE 5-4

COAL LIQUEFACTION PROCESS

POLLUTION SUMMARY

	Tons/Day				
Fossil Fuel Reformer	Utility Boiler	Reformer	Fired Heaters	Sulfur Recovery	TOTAL
SO_2	30.1	28.1	53.3	6.0	117.5
NO_x (as NO_2)	18.1	20.9	5.9	-	44.9
Particulates	2.6	-	-	-	2.6
Solid Waste	1128	-	-	-	1128
Nuclear Reformer					
SO_2	13.6	-	8.6	6.0	28.2
NO_x (as NO_2)	7.9	-	3.9	-	11.8
Particulates	1.1	-	-	-	1.1
Solid Waste	494	-	-	-	494

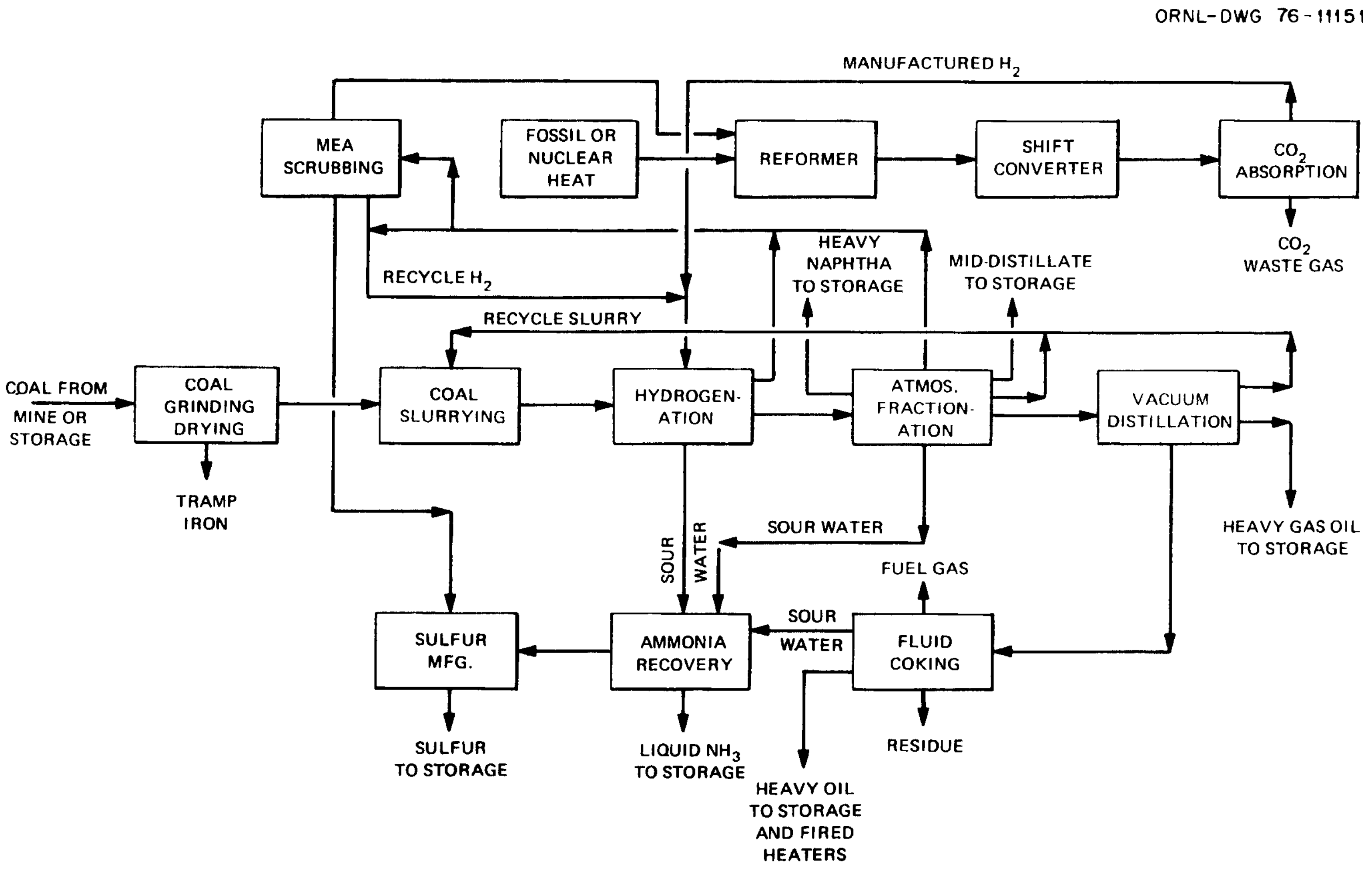

Fig. 5.1. Coal Liquefaction Process — Block Flow Diagram

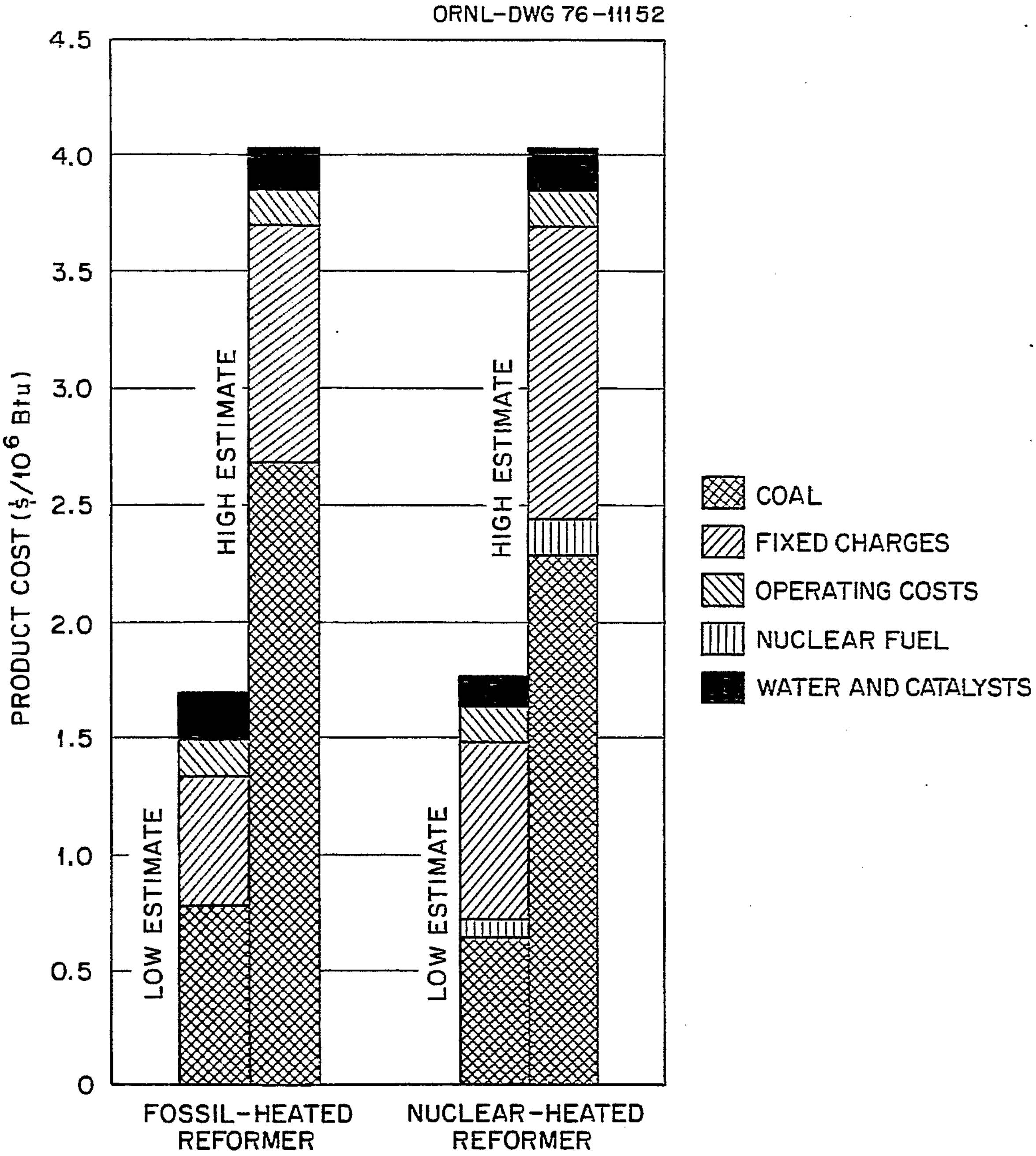

Fig. 5.2. Coal Liquefaction Process — Product Costs

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6. APPLICATION OF SYNTHESIS GAS TO METHANOL PRODUCTION

6.1 Methanol Process Requirements

Since methanol is an appropriate alternate to hydrocarbon liquid fuels, United Engineers has reviewed the suitability of synthesis gases from two processes as feed for the Imperial Chemical Industries, Low-Pressure/Low Temperature (ICI LP/LT) Methanol Process. The following discussion from ICI Company Literature (Reference 6-1) is pertinent.

> "It is well known that for correct overall stoichiometry of the methanol synthesis reaction, based on steam/hydrocarbon reforming, the feedstock to the reformer should have an empirical formula close to CH_2:
>
> $$CH_2 + H_2O \rightarrow CO + 2H_2 \rightarrow CH_3OH$$
>
> With gaseous feedstocks, particularly natural gas, the H/C ratio is too high for production of a perfectly stoichiometric synthesis gas. Conventionally, a gas stream containing carbon (in most cases CO_2) was added to the system. Methanol manufacturers have in the past gone to some length to locate their production facilities adjacent to ammonia plants to gain access to a supply of CO_2. Where this proved impracticable, reformer flue gases were scrubbed with amine solution and the recovered CO_2 recycled to the reformed gas. This not only involves high initial capital investment but considerable increases in steam and cooling water consumption. It is also necessary to maintain close control to ensure that the CO_2 produced is of a purity suitable for the synthesis make-up gas, since MEA solution degrades continously due to the oxygen content of flue gases. Corrosion problems are also inherent in this type of scrubbing system.
>
> An alternative Low Carbon Concept system, pioneered by Davy Powergas for the LP/LT process, is to pass the reformed gas directly into the synthesis loop without CO_2 addition, the excess hydrogen being purged from the loop. This purge gas flow is used in the reforming furnace as fuel and hence the overall thermal efficiency of the plant is maintained at a very high level.
>
> The loss of hydrogen as a process reactant means that the capacity of the reforming plant must be greater than if CO_2 were added into the make-up gas and also that the compression duty on the synthesis gas machine must be increased. However, the extra costs involved are less than those of the alternative flue gas CO_2 recovery system."

Another reference (6-2) states:

> "In conclusion, it may be stated broadly that make-up gas mixtures containing either a high content of nonreactants or a preponderance of one or the other reactant are unsatisfactory for efficient low-cost methanol production, since the excess reactant or inert gas must be purged off in synthesis. Purge of these constituents cannot be effected relatively, with the consequence that reactant gases and methanol are lost together with the undesired component(2). The value of the gases lost by purge is often less significant than the loss of energy in the form of compression, since the latter, in many plants is the most costly operation."

Also:

> "The nature of the reaction between hydrogen and carbon dioxide, with or without the presence of carbon monoxide, is such that the crude methanol produced from these reactants will at best be more dilute than the product obtained from synthesis involving carbon monoxide only. Refining, although following similar principles, has to be more elaborate but there is no difference in the primary end-product."

Thus, any process which produces synthesis gas for the ICI LP/LT Methanol Process will sustain an economic penalty dependent on the extent to which the gas deviates from the stoichiometric ideal of CH_2 and to its content of nonreactant gases unless auxiliary processes are added to use the excess gases or they are combusted in an efficient heat recovery system.

6.2 Synthesis Gas Suitability

Methanol can be formed by either of two reactions:

$$CO + 2\ H_2 \rightarrow CH_3OH$$
$$CO_2 + 3\ H_2 \rightarrow CH_3OH + H_2O$$

For converting carbon monoxide to methanol, the stoichiometric ideal ratio of hydrogen to carbon monoxide is thus 2:1. When carbon is supplied as carbon dioxide, the ideal ratio of hydrogen to carbon dioxide is 3:1. Because the $CO_2 + H_2$ reaction produces water which is soluble in methanol, and must be subsequently removed, the first reaction is the more desirable.

Synthesis gas produced in many industrial processes is not stoichiometrically ideal for methanol synthesis. The methanol manufacturer makes the composition of these gases stoichiometric by either withdrawing or adding quantities of hydrogen, carbon monoxide, or carbon dioxide as discussed above.

Table 6-1 shows the compositions of the synthesis gases which are produced by the processes described in Sections 3.2 and 4.3 of this report, and a gas having the ideal composition.

Synthesis gas from the Koppers-Totzek process is appropriate as feed to the ICI LP/LT system because both processes have been commercially proven and are available for producing synthesis gas and methanol. The Koppers-Totzek process has the advantage of producing a synthesis gas which is free of liquid products. Its synthesis gas is also free of methane, which, if present, would have to be removed either by purging or other separation before methanol synthesis. A partially offsetting disadvantage of the Koppers-Totzek process is that, since it operates at atmospheric pressure, its synthesis gas product must be compressed to 1000 psi before entering the ICI LP/LT process. The costs of this compression have been factored into cost and efficiency calculations.

The synthesis gas as produced by the GA/S&W reformer is too rich in hydrogen for methanol production. For every 100 moles of gas, it would be necessary to remove 17.8 moles of hydrogen and 9.5 moles of methane in order to obtain a stoichiometric gas. The methanol produced would contain 0.65 moles of water for every mole of methanol. This water must be removed after the reaction.

The carbon dioxide can be removed from the GA/S&W gas after it comes out of the reformer but the resulting ratio of hydrogen to carbon monoxide would be about 10:1, a more unfavorable ratio. The large excess of hydrogen and considerable methane must be removed.

Carbon dioxide can also be added until the stoichiometric composition is reached, also shown in Table 6-1. This gas converts to methanol with maximum efficiency, but the methanol contains a high proportion of water.

When excess hydrogen and accompanying gases are purged for removal purposes, they can be used to provide heat to the gasification process and the heating valve can be recovered. However, if a VHTR provides most of the process heat, the excess hydrogen is available for other applications. Ammonia plants are often constructed together with methanol plants to utilize excess hydrogen.

6.3 Costs of Methanol from Selected Synthesis Gases

Costs were developed for the production of methanol by the ICI LP/LT process from the GA/S&W synthesis gas, Section 3.2,with CO_2 added and the Koppers-Totzek synthesis gas, Section 4.3 of this report. The cost data presented in Reference 6-3 were used to obtain the capital and production costs for the ICI process portion of the total plant. The synthesis gas costs were obtained from Section 3.2 and 4.3.

The GA/S&W synthesis gas with CO_2 added from the hydrogen recycle treatment has the correct theorectical ratio of carbon (from CO and CO_2) to hydrogen for methanol production, but also has a significant amount of nonreactant methane. A high yield of methanol is obtained by recycling gas, after condensation of methanol, to the methanol converter. However, in order to maintain the methane content at a reasonable level it is necessary to purge approximately half of the gas from the condensation step. This purge gas of 570 Btu/lb and available at 1000 psi is considered to be a fuel-by-product. The conversion of useful carbon to methanol is approximately 86% and 10,000 tons per day can be produced.

The Koppers-Totzek synthesis gas from Section 4.3 also has the approximately correct theoretical ratio of carbon to hydrogen for methanol production, but it does not contain any methane of significant amount. Very little purging from the synthesis loop should be required and the purge gas is assumed to have no significant fuel value. At 96% conversion, approximately 15,000 TPD can be produced.

The lower total product cost of the GA/S&W process can be misleading because the methanol constitutes less than 50% of the total products. All of the Koppers-Totzek product is methanol. If the purge gas and light aromatic liquids are priced at approximate current market values ($\$1.50/10^6$ Btu and \$0.30/gallon, respectively), then the low basis cost for methanol is about the same for the two processes. Because the Koppers-Totzek synthesis gas requires less processing and produces only methanol, it is more attractive. However, a significant increase in the value of the GA/S&W by-products would make methanol from the GA/S&W synthesis gas much less expensive.

On February 9, 1976 the Chemical Marketing Report quoted a price of $0.40 per gallon for synthetic methanol in large quantities at the producing plant. Both these processes are thus very attractive when compared to the current price for methanol.

6.4 References

6-1 Davy Powergas, Inc. Methanol, Publication No. 3-19274.
P.O. Box 36444, Houston, Tx 77036

6-2 Kirk-Othmer, "Methanol" Encyclopedia of Chemical Technology
2nd. Ed. Interscience Publishers N.Y. CR 1963 V. XIII pg. 384.

6-3 Vyas and Bodle, "Coal and Oil - Shale Conversion Looks Better",
Oil and Gas Journal, March 25, 1975 pg. 45.

6-4 Davy Powergas Inc., Preliminary Data for LCF/Ammonia Facility,
P.O. Box 2436 Lakeland, Florida

6-5 Pettman, M. J. and Humphreys, G. C. "Improve Designs to Save Energy" Hydrocarbon Processing, Jan. 1975 pg. 77

6-6 Royal, M. J. and Mimms, N.W. "Why LP Methanol Costs Less" Hydrocarbon Processing, March 1969 pg. 147

6-7 Imperial Chemical Industries, Ltd. "Methanol (ICI Low Pressure) Process", Hydrocarbon Processing, Nov. 1975 pg. 147

6-8 Ellwood, Peter "Methanol Process Makes Production Possible in Small Plants", Chemical Engineering, Feb. 12, 1968 pgs. 104-6

6-9 Richard Savage, Davy Powergas Inc., Personal Communication

TABLE 6-1

COMPOSITIONS OF VARIOUS SYNTHESIS GASES

(Mole %)

Component	Ideal Composition	Koppers-Totzek(1)	General Atomic/Stone & Webster(2) From Reformer	 CO_2 Removed	 CO_2 Added
CH_4	-	-	9.5	10.95	9.
H_2	66.7	65.	70.5	80.7	66.5
CO	33.3	31.	7.0	8.05	6.6
CO_2	-	4.	12.9	0.2	17.8
H_2O	-	-	0.1	0.1	0.1

Notes:

(1) Section 4.3

(2) Section 3.2

TABLE 6-2

METHANOL PRODUCTION FROM SYNTHESIS GAS USING THE ICI METHANOL PROCESS

JULY 1974 DOLLARS

	GA/S&W	Koppers-Totzek
Plant Capacity, 10^9 Btu/day		
Methanol	195	293
Purge Gas (Fuel)	142	-
Light Aromatics	66	-
Total Output	403	293
Capital Cost, $ Million		
Synthesis Gas Plant	1,363.9	1,073
ICI Methanol Plant	173	223
Total Capital Cost	1,536.9	1,296
Average Product Cost, $/$10^6$ Btu Product		
Synthesis Gas	2.42 to 5.04	2.91 to 6.74
Methanol Conversion	.30 to .43	.50 to .73
Total Product Cost	2.72 to 5.47	3.41 to 7.47
Methanol Cost, Assuming values of $1.50/$10^6$ Btu for Purge Gas and $0.30/gal for light aromatics		
$/$10^6$ Btu	3.74 to 9.43	3.41 to 7.47
$/Gallon	.24 to .61	.22 to .48

7. APPLICATION TO STEEL PRODUCTION

Most of the data used herein for steel making costs by various routes havebeen obtained from the several reports of the American Iron and Steel Institute (AISI) Committee on the use of nuclear energy in steel making (References 7-1 to 7-5).

7.1 Nuclear Energy Applications in Steel Making

The most viable concept for applying nuclear energy to steel making combines two well-known processes: direct reduction in a shaft furnace and refining in an electric furnace.

Although direct reduction of iron ore is a fairly new process, it is well developed and in commercial use in various parts of the world where low-cost natural gas is available. In this process, iron ore is reduced in the solid condition by a synthesis gas ($CO + H_2$) derived from steam reforming of natural gas to a product known as sponge iron. The reaction requires high temperatures and heat. Nuclear energy could be used to provide the heat needed to produce the reducing as for the direct reduction of iron ore and the electricity needed to refine the resulting sponge iron to steel in an electric-arc furnace.

Production of steel by electric-arc furnaces is a long-established commercial technology. Electric-furnace capacity in the United States today is about 30 million tons a year. Almost all of that tonnage is made with scrap as the only ferrous charge, although a number of plants presently use at least some sponge iron in their charges. Sponge iron could be used for a large portion of that charge if the cost of the sponge were competitive with scrap, and provided the gangue content were low enough. Electric-arc refining uses about 650 KWH/ton of steel.

The impurities, or gangue constituents, in iron ore (mainly silica and alumina) are not removed in direct-reduction processes as they are when ore is reduced to liquid iron in the blast furnace. Instead, these impurities remain in the sponge iron and increase the amount of slag that must be melted in subsequent refining in the electric-arc furnace. Since this imposes a substantial cost penalty, low-gangue ores are favored for direct-reduction processes.

Processes for direct reduction differ in certain details, but almost all of them use as a reductant a gas mixture of carbon monoxide and hydrogen at temperatures in the range of 1500-1800^{o}F. For the efficient reduction of iron ore, the $CO + H_2$ content of the reducing gas should be above 90 percent.

Iron ore can also be reduced to sponge iron with hydrogen alone, but there are drawbacks. First, the iron tends to sinter or stick together during hydrogen reduction, and second, there is a strong tendency for the reduced iron product to reoxidize. Both of these problems must be overcome before hydrogen reduction can be utilized commercially on a large scale.

In a steel making system involving direct reduction and refining in an electric-arc furnace, nuclear energy can be used to:

a) Provide high-temperature heat for the production of a gas suitable for the reduction of iron ore to iron. The reducing gas, after removal of sulfur and other impurities, is at moderate temperature and pressure and can be transported a reasonable distance from the nuclear heat source.

b) Produce electricity for the operation of electric-arc furnaces to refine the sponge iron. The electricity for this step may be generated in any kind of a nuclear power plant; thus, direct reduction does not necessarily have to be close-coupled with steel making.

Although reforming, direct reduction, and electric-arc furnace refining are all well-developed processes, combining a VHTR with the reforming process will require extensive experimentation, development, and engineering. A system coupling a VHTR to a coal conversion plant for the production of a synthesis or reducing gas has been studied and the economics reported in Section 3.2 of this report. Furthermore, although this system conserves the use of natural hydrocarbons, it still requires reformed gas made from a carbon source, the future domestic availability and/or cost of which is uncertain.

7.2 Economics of Nuclear Steel Making

At present, the vast majority of steel produced from ore is made by conventional coke oven/blast furnace/basic-oxygen furnace refining (CO/BF/BOF) with much smaller amounts produced by conventional reforming of natural gas/direct reduction/electric-arc-furnace refining. Accordingly, economic assessment of nuclear steel making systems must be compared to these conventional systems.

Systems for which economic comparisons were made are:

Case 1 - Conventional coke oven/blast furnace/basic-oxygen-furnace refining (Table 7-2).

Case 2 - Conventional reforming of natural gas/direct reduction/electric-arc-furnace refining (Table 7-3).

Case 3 - General Atomic-Stone & Webster Synthesis gas feed/direct reduction/electric-arc furnace refining (Table 7-4).

Case 4 - Koppers-Totzek Synthesis gas feed/direct reduction/electric-arc furnace refining (Table 7-5).

Figures 7-1 and 7-2 present flow sheets for the conventional coke oven/blast furnace/basic-oxygen-furnace system (Case 1) and the conventional reformer/direct-reduction system (Case 2). Figure 7-3 is the flow sheet for the direct-reduction/electric-arc furnace process fueled by gas reformed by heat from a VHTR (Case 3). The fourth case, shown in Figure 7-4, is a similar process using gas produced by the Koppers-Totzek process as reported in Section 4.3 of this report.

Metallic balances are shown in Tables 7-6 and 7-7.

7.2.1 Basic Assumptions for the Individual Cases

Case 1 - Blast-Furnace - BOF Route

Fuel credit only taken for the by-products

Blast-furnace burden was 100 percent oxide pellets

A blast-furnace product rate of 6 tons/day/square foot

A 1,000 pound coke rate with tar injection was assumed for a modern furnace.

Plant production was 4 million annual net tons of liquid steel.

Two blast furnaces were required and the size was based on the above criteria.

Case 2 - Natural Gas Reformer - Shaft Reduction, Electric-Arc Furnace Steel Making Plant.

High-temperature (1600 to 1700°F) natural gas-steam reformers with low steam-to-methane ratios (near stoichiometric) were used to produce a "one-step" reducing gas (92 percent $H_2 + CO$).

Reduction was accomplished in a shaft-type, continuously-fed reduction furnace, utilizing the bottom section of the shaft for cooling.

The shaft reducer utilized oxide pellets and produced a 95 percent reduced product.

The steel making plant utilized two modern high-powered electric furnaces with continuous charging practice.

The steel making plant was sized to produce 1,060,000 annual net tons of liquid steel, based on a 27 percent of metallics scrap charge.

Cases 3 & 4 - Reformer, Shaft Reduction, Electric-Arc Furnace

The plant capacity was based on the reducing gas from a 3000 MWt VHTR-coal gasification (which is an economically sized reactor) process (Case 3), and the reducing gas from a Koppers-Totzek process of comparable heating value output (Case 4).

The synthesis gas to Case 3 was passed through a secondary reformer to reduce the methane content to less than 1 percent. Top gas, after removal of H_2O and CO_2, was recycled through a reheat section of the secondary reformer. Top gas was also used to fire the reformer-reheater.

The synthesis gas to Case 4 was passed through a heater to raise the temperature to that required by the shaft furnace. Top gas, after removal of H_2O and CO_2, was recycled through the heater, which was fired by top gas.

In both cases the reducing gas was utilized in a shaft-type, continuously-fed reduction furnace.

The reduction plant was designed to produce 95 percent reduced pellets from a feed of oxide pellets.

The reduction plant utilized as many 16 foot diameter shaft reducers as were required to produce the specified capacity.

The steel making plant utilized modern high-powered electric furnaces with continuous charging practice.

7.2.2 Capital and Production Costs

Table 7-1 summarizes the capacities and estimated capital costs for each process system. The costs for Cases 1 and 2 were developed by the AISI Committee, reviewed and updated by UE&C. Costs for Cases 3 and 4 were adjustments made by UE&C to the Case 2 cost to account for the slight differences in these three cases.

Each case is considered to be a totally new facility and includes the following off-site facilities: raw materials handling facilities, electric power distribution, water distribution, tracks and roadways, service, waste treatment (not including spent nuclear fuel), administrative buildings, laboratories, warehouses, and shop, fire, and safety facilities. The off-site facilities are estimated to be 35 percent of the battery limit plant costs in all cases.

The plant is assumed to be located at a Middletown, USA site. Costs for transportation of raw materials are not included.

All processes used purchased iron oxide pellets, with an analysis of 66 percent Fe, 3 percent Si02, and low phosphorus. Pellets were assumed to be delivered.

Although a product mix was not selected, it was assumed that the revert scrap is equivalent to 27 percent of the metallic charge to the BO furnace or 27 percent of the metallic charge to the electric furnaces.

In all cases, the evaluation included costs only to the production of liquid steels. It is assumed that the costs from this point to finished product are the same for all routes. For this study, the liquid steel will have 0.05-0.10 percent carbon and 0.16-0.20 percent sulfur.

Raw materials delivered to USA ports were estimated as follows:

Blast Furnace Flux: \$6/NT
Burnt Lime: \$25/NT
Electrodes: \$0.48/lb
Oxide Pellets: \$0.355/Gross Ton Fe Unit
Fluorspar: \$75/NT

An average labor rate of $9.60/manhour including fringe benefits was used for all cases. All other labor-associated items, such as supervision and plant overhead were estimated to be 65 percent of direct-labor costs. Repair and maintenance, including labor and materials, was estimated to be 4 percent of investment.

Direct and variable costs supplied by the AISI Committee were reviewed by United Engineers and adjusted to July, 1974 rates (see Tables 7-2 through 7-5).

The economics of the alternate processes is done by a return-on-investment analysis which incorporates both capital and operating costs. This was done by calculating a hypothetical selling price for liquid steel on the basis of obtaining a five-year capital recovery.

7.3 Pollution

Use of nuclear energy in basic steel making will reduce both air and water pollution, because of the substitution of nuclear heat for burning of coal.

The coke oven-blast furnace-basic oxygen process is well known as a major air and water pollution control problem in the steel industry. A modern installation can be constructed to meet present standards, but as standards become tighter, meeting them will become progressively more difficult and expensive.

Control of emissions (mainly dust) in the closed direct reduction process does not appear to present serious problems. The dust emissions from arc-furnace can be satisfactorily controlled with the use of exhaust systems and baghouse type air filters.

The most difficult emission and control problem in arc-furnaces is due to varying contaminants in the scrap charge (oils, etc.). The substitution of a large percentage of sponge iron for scrap would greatly reduce this problem.

7.4 References

7-1 Status of Armco's Reduction Process, American Iron and Steel Institute (AISI), May 22, 1974.

7-2 Integrated Steel Making SIDBEC-DOSCO Using Direct Reduction and Electric Furnaces, AISI, May 22, 1974.

7-3 Utilization of Direct Reduction Iron in Electric Steel Making, AISI, May 22, 1974.

7-4 The Use of Nuclear Energy in Steel Making - Prospects and Problems, AISI, May 23, 1974.

7-5 Private Communications from the AISI General Research Committee on the Use of Nuclear Energy in Steel Making.

TABLE 7-1

ESTIMATED CAPITAL COSTS

Case	Process	Steel Plant Capacity Tons Liquid Steel/Year	Capital Cost $/Annual Ton
1	Coke Ovens Blast Furnace Basic Oxygen Furnace	4,000,000	152.00
2	Conventional Reformer Direct Reduction Electric Arc-Furnace	1,060,000	93.60
3	3000 MWt General Atomic-Stone & Webster VHTR Synthesis Gas Feed Direct Reduction Electric-Arc Furnace	14,100,000	96.0
4	3000 MWt VHTR Equivalent Koppers-Totzek Synthesis Gas Feed Direct Reduction Electric Arc-Furnace	16,500,000	92.0

TABLE 7-2

CASE 1

COKE OVEN - BLAST FURNACE - BOF OPERATING COSTS

	Unit	Quantity	Unit Price $/Unit	Cost $ Per NT Liquid Steel
Ferrous Materials				
Oxide Pellets	NT Fe	0.792	31.70	25.11
Alloys	NT	0.01	290.00	2.90
Revert Scrap	NT Fe	0.273	100.00	27.30
Subtotal				55.31
Dust & Sludge (Credit)	NT Fe	0.027	17.50	(0.47)
Net Ferrous Materials				54.84
Fuel				
Coal	NT (Dry)	0.585	40.00	23.40
Net Excess Fuel (Credit)	MM Btu	3.5	1.48	(5.18)
Electric Power - Aux.	KWH	45.0	0.012	0.54
Net Fuel				18.76
Other Materials				
Oxygen	CCF	18.0	0.05	0.90
Flux - Raw Stone	NT	0.21	6.00	1.26
Burnt Lime	NT	0.08	25.00	2.00
Spar	NT	0.004	75.00	0.30
Refractories - BOP	$/NT Liq. Steel	1.0	0.71	.71
Total Other Materials				5.17
Labor Direct & Indirect Incl. Supervision	Manhours	0.3	9.60	2.88
Maintenance Labor & Mat'l.		4.0% x $152/Annual Ton		6.08
Miscellaneous Expenses		65% of Labor Cost		1.87
Direct Operating Expense				89.60
Depreciation		5.0% x $152/Annual Ton		7.60
DOE + DEP				97.20
Average Interest @ 8% of Unrecovered Capital Over 5 years				7.49
Profit Before Federal Income Tax				43.86
SELLING PRICE FOR 5-YEAR RECOVERY				148.55

TABLE 7-3

CASE 2

NATURAL GAS REFORMER - DIRECT REDUCTION - ELECTRIC ARC-FURNACE OPERATING COSTS

	Unit	Quantity	Unit Price $/Unit	Cost $ per NT Liquid Steel
Ferrous Materials				
Oxide Pellets	NT Fe	0.802	31.70	24.52
Alloys	NT	0.01	290.00	2.90
Revert Scrap	NT Fe	0.273	100.00	27.30
Subtotal				55.62
Dust & Sludge (Credit)	NT Fe	0.027	17.50	(0.47)
Net Ferrous Materials				55.15
Fuel				
Natural Gas	MM Btu	11.0	0.79	8.69
Electric Power - Melting	KWH	625	0.012	7.50
- Aux.	KWH	53	0.012	.64
Net Fuel Cost				16.83
Other Materials				
Oxygen	CCF	1.50	0.07	0.10
Flux - Burnt Lime	NT	0.125	25.00	3.12
Spar	NT	0.004	75.00	0.30
Electrodes	Lb	12.00	0.48	5.76
Refractories - Elec. Furnace	$/NT Liq. Steel	1.00	2.63	2.63
Carbon Additions	Lb	12.00	0.023	2.77
Total Other Materials				14.68
Labor Direct and Indirect Including Supervision	Manhours	0.3	9.60	2.88
Maintenance Labor & Materials		4.0% x 93.6/Annual Ton		3.74
Miscellaneous Expenses		65% of Labor Cost		1.87
Direct Operating Expense				95.15
Depreciation (20 years)		5.0% x 93.6/Annual Ton		4.68
DOE + DEP				99.83
Average Interest @ 8% of Unrecovered Capital over 5 years				4.61
Profit Before Federal Income Tax				27.00
SELLING PRICE FOR 5 YEAR RECOVERY				131.44

TABLE 7-4

CASE 3

REDUCING GAS FROM GENERAL ATOMIC-STONE & WEBSTER VHTR AT COST GAS REHEATING-DIRECT REDUCTION-ELECTRIC-ARC FURNACE OPERATING COSTS

	Unit	Quantity	Unit Price $/Unit	Cost $ per NT Liquid Steel
Ferrous Materials				
Oxide Pellets	NT Fe	0.802	31.70	25.42
Alloys	NT	0.010	290.00	2.90
Revert Scrap	NT Fe	0.273	100.00	27.30
Subtotal				55.62
Dust & Sludge (Credit)	NT Fe	0.027	17.50	(0.47)
Net Ferrous Materials				55.15
Fuel				
Reducing Gas	MM Btu	8.7	2.22-4.63	19.33-40.30
Electric Power - Melting	KWH	625	0.012	7.50
- Aux.	KWH	93	0.012	1.12
Net Fuel Cost				27.95-48.92
Other Materials				
Oxygen	CCF	1.50	0.07	0.10
Flux - Burnt Lime	NT	0.125	25.00	3.12
- Spar	NT	0.004	75.00	0.30
Electrodes	Lb	12.00	0.48	5.76
Refractory-Mtl. & Labor	$/NT	1.00	2.63	2.63
Coke	Lb	12.00	0.023	2.77
Total Other Materials				14.68
Labor Direct & Indirect Incl. Supervision	Manhours	0.300	9.60	2.88
Maintenance (Reduction & Elec. Furnaces) - Labor & Mat'ls.		4.0% x 96.0/Annual Ton		3.84
Miscellaneous Expenses		65% of Labor Costs		1.87
Direct Operating Expense				106.37-127.34
Depreciation		5.0% x 96.0/Annual Ton		4.80
DOE+DEP				111.17-132.14
Average Interest @ 8% of Unrecovered Capital Over 5 Years				4.73
Profit Before Federal Income Tax				27.70
SELLING PRICE FOR 5 YEAR RECOVERY				143.60-164.57

TABLE 7-5

CASE 4

REDUCING GAS FROM KOPPERS-TOTZEK PROCESS AT COST, GAS REHEATING-DIRECT REDUCTION-ELECTRIC-ARC FURNACE OPERATING COSTS

	Unit	Quantity	Unit Price $/Unit	Cost $ per NT Liquid Steel
Ferrous Materials				
Oxide Pellets	NT Fe	0.802	31.70	25.42
Alloys	NT	0.010	290.00	2.90
Revert Scrap	NT Fe	0.273	100.00	27.30
Subtotal				55.62
Dust & Sludge (Credit)	NT Fe	0.027	17.50	(0.47)
Net Ferrous Materials				55.15
Fuel				
Reducing Gas	MM Btu	7.4	2.30 - 5.33	17.02 - 39.44
Electric Power - Melting	KWH	625	0.012	7.50
- Aux.	KWH	93	0.012	1.12
Net Fuel Cost				25.64 - 48.06
Other Materials				
Oxygen	CCF	1.50	0.07	0.10
Flux - Burnt Lime	NT	0.125	25.00	3.12
- Spar	NT	0.004	75.00	0.30
Electrodes	Lb	12.00	0.48	5.76
Refractory-Mtl. & Labor	$/NT	1.00	2.63	2.63
Coke	Lb	12.00	0.023	2.77
Total Other Materials				14.68
Labor Direct & Indirect Including Supervision	Manhours	0.300	9.60	2.88
Maintenance (Reduction & Elec. Furnaces) Labor & Materials		4.0% x 92.0/Annual Ton		3.68
Miscellaneous Expenses		65% of Labor Costs		1.87
Direct Operating Expense				103.90 - 126.32
Depreciation		5.0% x 92.0/Annual Ton		4.60
DOE + DEP				108.50 - 130.92
Average Interest @ 8% of Unrecovered Capital Over 5 Years				4.53
Profit Before Federal Income Tax				26.55
SELLING PRICE FOR 5 YEAR RECOVERY				139.58 - 162.00

TABLE 7-6

METALLIC BALANCE

COKE OVEN - BLAST FURNACE - BOF

CASE 1

Item Charge			Item Yield		
	Net Ton	Net Ton Iron		Net Ton Iron	Remarks
Blast Furnace					
Oxide Pellets	1.170	0.772	Hot Metal (+C.I.)	0.772	93% Iron
BOF Slag	0.100	0.020	Dust and Sludge	0.012	For Credit
			Unaccounted Loss	0.008	
TOTAL		0.792		0.792	
BOF					
Hot Metal	0.830	0.772	Liquid Steel	1.000	Basis
Scrap	0.278	0.278	BOF Slag	0.020	
			Dust and Sludge	0.015	For Credit
			Unaccounted Loss	0.010	
TOTAL	1.103	1.045		1.045	

TABLE 7-7

METALLIC BALANCE

DIRECT REDUCER - ELECTRIC-ARC FURNACES

CASES 2, 3 & 4

Item Charge	Net Ton	Net Ton Iron	Item Yield	Net Ton Iron	Remarks
Shaft Reducer					
Oxide Pellets	1.220	0.802	Reduced Pellets	0.782	92 % Iron
			Dust and Sludge	0.012	For Credit
			Unaccounted Loss	0.008	
TOTAL	1.220	0.802		0.802	
Electric-Arc Furnace					
Reduced Pellets	0.850	0.782	Liquid Steel	1.000	
	0.273	0.273	Electric Furnace Slag	0.030	Dispose
			Dust and Sludge	0.015	For Credit
			Unaccounted Loss	0.010	
TOTAL	1.123	1.055		1.055	

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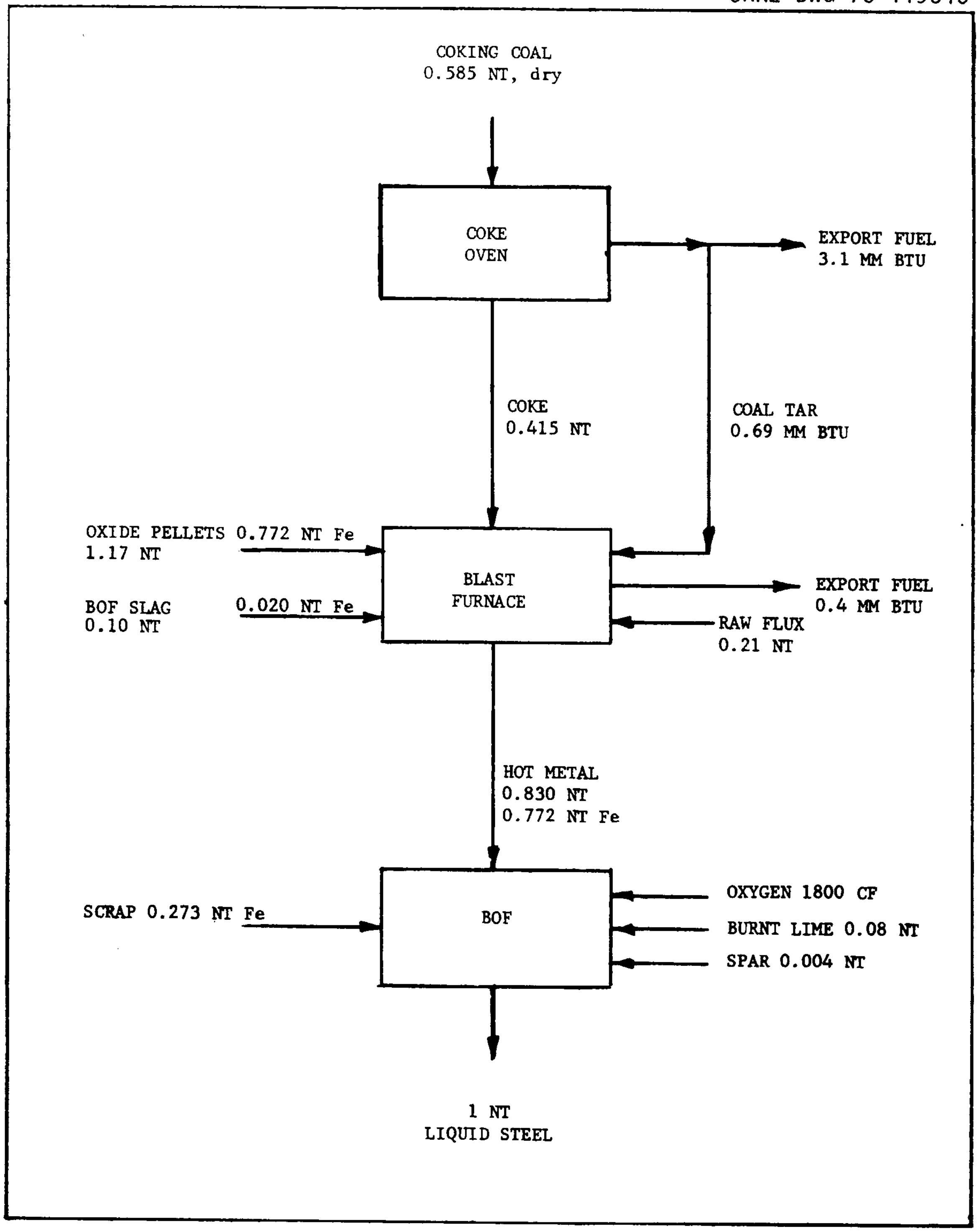

Fig. 7.1. Case 1 Flow Sheet — Coke Oven/Blast Furnace/BOF

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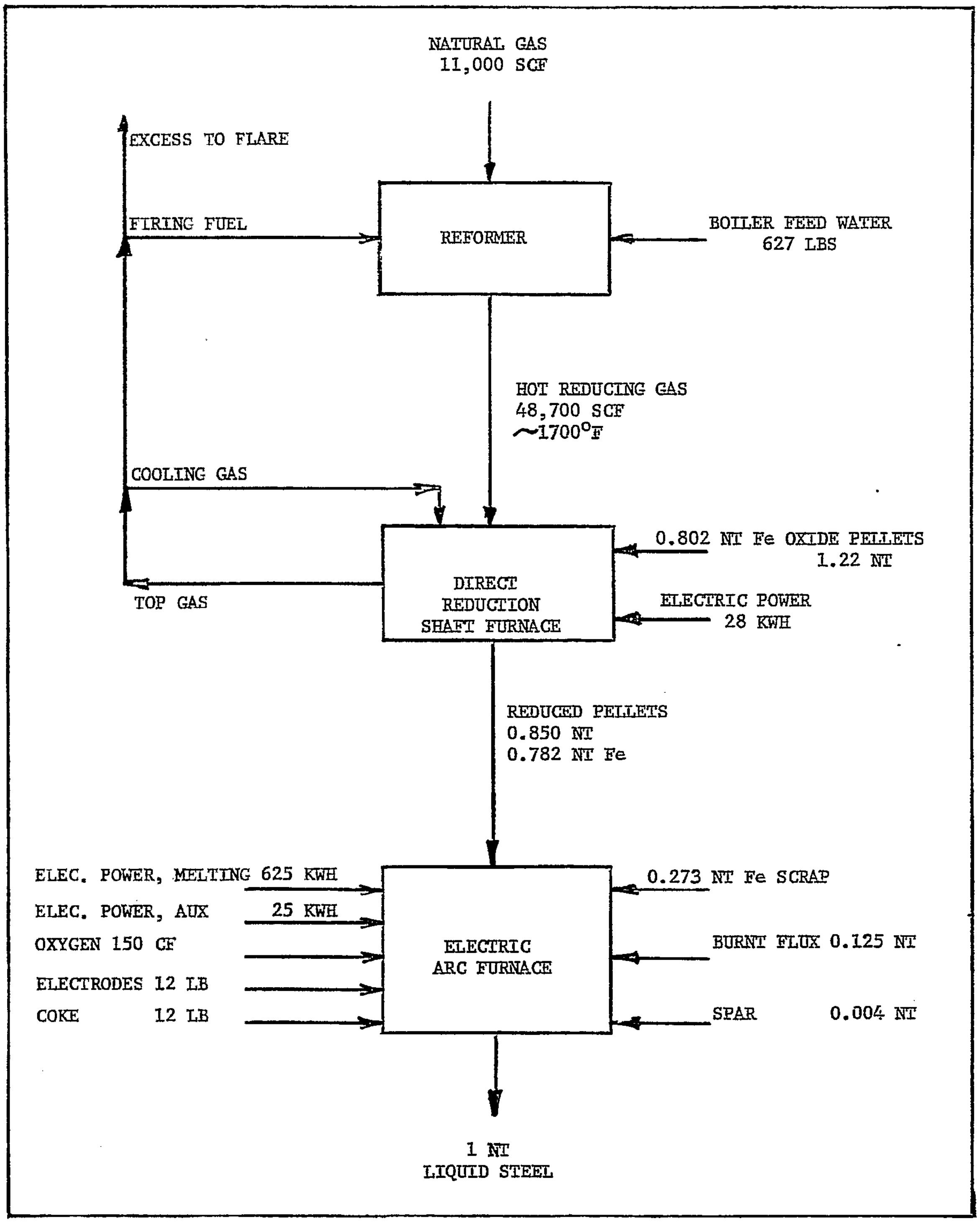

Fig. 7.2. Case 2 Flow Sheet — Natural Gas Reformer/Direct Reduction/EAF

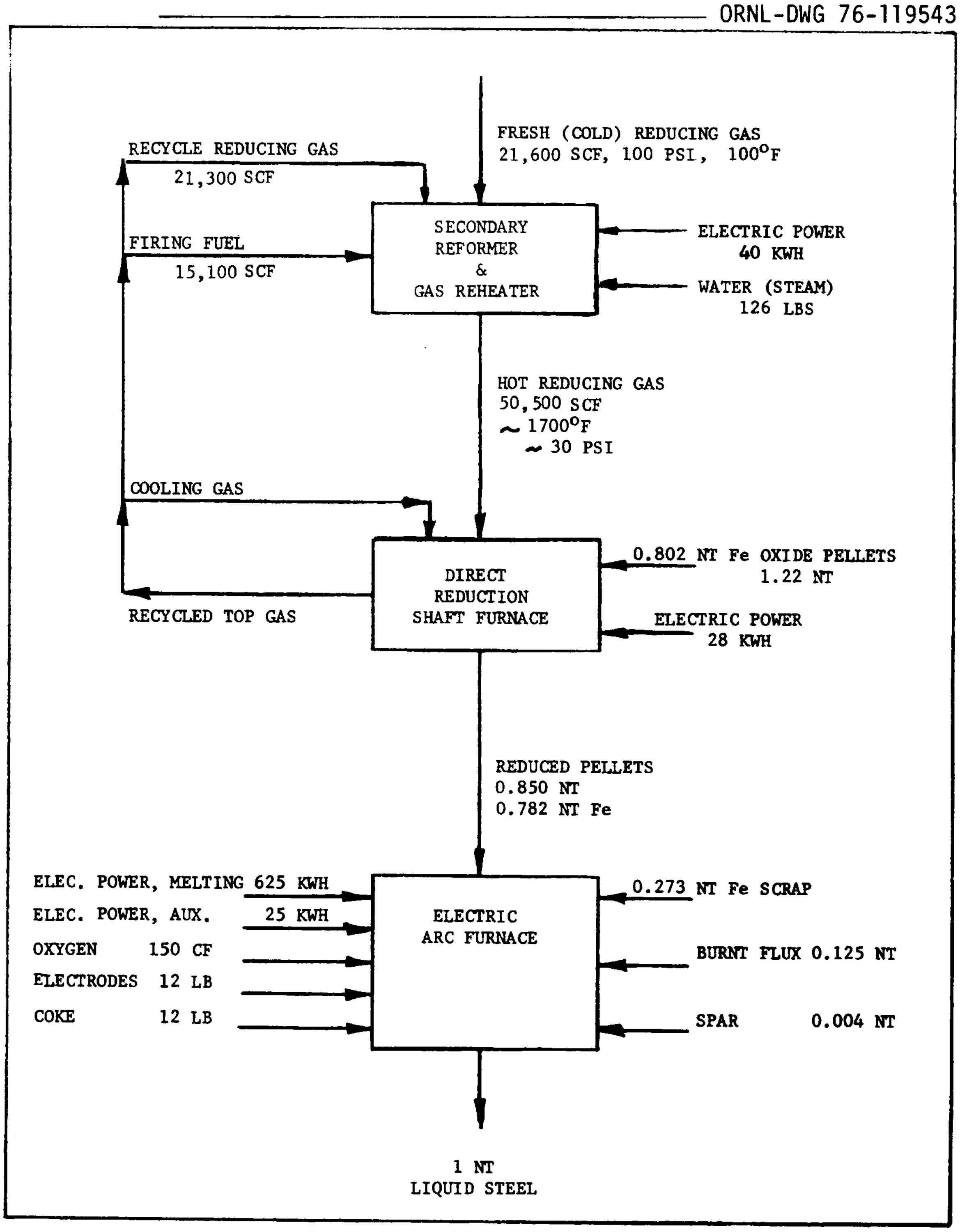

Fig. 7.3. Case 3 Flow Sheet — General Atomic VHTR/Reformer/EAF

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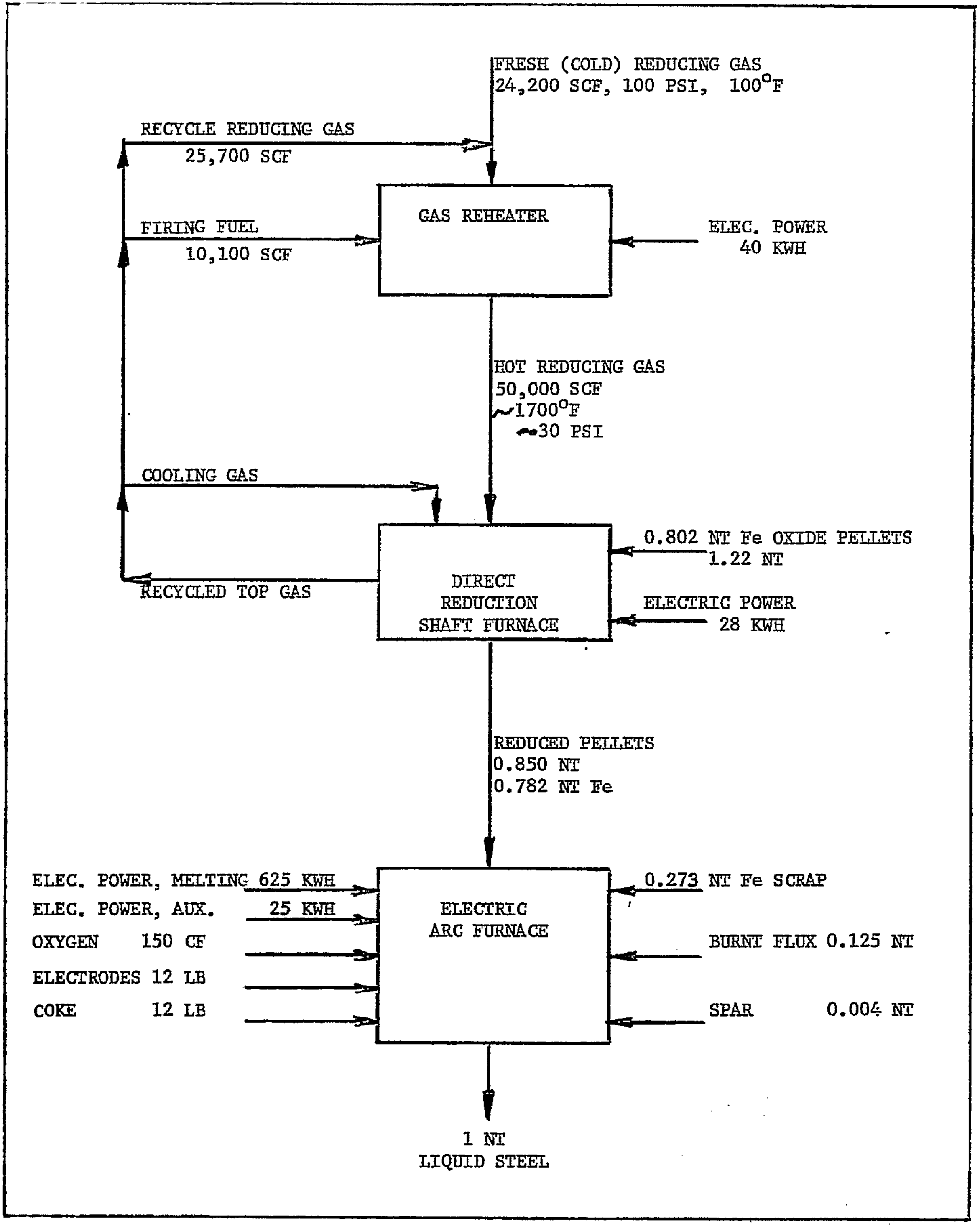

Fig. 7.4. Case 4 Flow Sheet — Koppers-Totzek/Reformer/EAF

8. WESTINGHOUSE HYDROGEN PRODUCTION PROCESS

8.1 Description of the Process

This process represents a novel approach by Westinghouse Astronuclear Laboratory (WANL) to solve the problem of splitting watter economically into hydrogen and oxygen through the application of nuclear heat. Since water is too stable to be decomposed by the direct application of heat, the usual approach is to react water (steam) with a carbon-containing reducing agent such as methane. In this way the chemical energy of the carbon-containing species is used to strip hydrogen away from the oxygen and the carbon ends up as waste carbon dioxide. As long as carbon and/or hydrocarbons remain cheap and plentiful, this approach is the preferred one. However, when this situation no longer exists, then the energy to split water must come from some other source such as that suggested by Westinghouse.

The Westinghouse process is based on the combination of two energy forms, both ultimately derived from a very high temperature gas cooled nuclear reactor. The first form of energy is electric power in which water is decomposed electrolytically to hydrogen and an in situ oxidizing agent. The decomposition voltage is reduced significantly through the use of sulfur dioxide as an anodic depolarizer. The second energy form is high temperature process heat by which the sulfuric acid generated by the oxidation of the SO_2 at the anode is thermally decomposed to sulfur dioxide, water and oxygen. The SO_2 is recycled to the electrolytic cell. The reactions occurring in the various process steps are:

SO_2 solution (anode compartment) $SO_2 + x\ H_2O \rightarrow SO_2 . x\ H_2O$

Anode reaction $SO_2 .\ x\ H_2O + 2H_2O - 2e \rightarrow H_2SO_4 .\ x\ H_2O + 2H^+$

Cathode reaction $2H^+ + 2e \rightarrow H_2$

Thermal decomposition $H_2SO_4 .\ x\ H_2O \rightarrow SO_2 + x\ H_2O + H_2O + 1/2\ O_2$

Overall reaction $H_2O \rightarrow H_2 + 1/2\ O_2$

By reducing the voltage and hence the power consumption in the electrolytic cell and substituting heat energy for it, the inefficiency associated with electric power generation is minimized, and hence the overall cycle efficiency is increased.

Westinghouse has confirmed in the laboratory that SO_2 functions effectively as an anodic depolarizer in the electrolytic decomposition of water. They also have obtained some laboratory data on the thermal decomposition of sulfuric acid. These experimental data plus supporting information from the literature from the chemical basis for this process.

The process is shown in a block diagram, Figure 8-1 and is described in detail below.

8.1.1 Electrolysis

The electrolyzers operate at approximately 0.7 volts compared to about 2 volts required for normal water electrolysis. The electrolyte is 70 to 80% H_2SO_4 and SO_2 is introduced around the anode. Recycle and fresh feed water are also added to the system. Care must be taken to prevent SO_2

from reaching the cathode where it would be reduced to sulfur that might cause plugging. To maintain the system in balance, 80% H_2SO_4 is continuously removed from the anode compartment.

8.1.2 Sulfuric Acid Vaporization

Vaporization of the 80% H_2SO_4 represents one of the most critical steps in the Westinghouse process. For every pound of H_2 product, 49 pounds of H_2SO_4 have to be circulated and decomposed. Massive equipment must be involved to handle this very large process stream. Additionally, the proposed process conditions, 50 psig and 460-750°F, are those where the 80% H_2SO_4 is a highly aggressive oxidizing agent and cannot be contained by conventional materials of construction. Studies to date suggest that silicon-containing materials, e.g. Duriron, might hold up in this environment. The basic process equipment for acid vaporization is a Duriron plate heat exchanger design. Mechanical chemical and heat transfer aspects of the design need to be demonstrated in practice.

8.1.3 Sulfuric Acid Decomposition

The H_2SO_4 vapor, in the form of sulfur trioxide (SO_3) and water, is decomposed to SO_2, H_2O and O_2 at 1600°F by the absorption of heat from the helium coolant stream from the nuclear reactor. Inasmuch as the H_2SO_4 is now above its dew point, the corrosion problems are much more manageable in this particular part of the process.

8.1.4 Partial Condensation

In a series of partial condensers, undecomposed H_2SO_4 and H_2O are separately condensed. The H_2SO_4 is returned to the vaporizer and the water to the electrolyzer. The condensers must be of corrosion-resistant materials.

8.1.5 Sulfur Dioxide/Oxygen Separation

The SO_2 and O_2 left after removal of the H_2SO_4 and H_2O are cooled and separated by conventional compression liquefaction - fractionation procedures. Current plans call for venting of the oxygen. The SO_2 is returned to the anode compartment of the electrolyzer.

8.1.6 Heat Source

A 3220 MWt VHTR nuclear plant provides the necessary heat and electrical energy via helium gas flowing through an intermediate heat exchanger.

8.2 Major Pollution Outputs

Barring unforeseen process releases, the Westinghouse process is virtually non-polluting since it basically involves the decomposition of water into hydrogen and oxygen. The oxygen may contain 1-2 ppm SO_2 which, while not troublesome from the point of view of venting to the atmosphere, might have to be removed if other uses of the gas are contemplated.

8.3 Process Thermal Efficiency

Thermal efficiency, as presented in Table 8-1 is defined as the higher heating value of the product gas divided by the heat input to the total plant complex. Since the plant is self-sufficient from an energy viewpoint, i.e., no net sale or purchase of power or heat is required for operation of the VHTR and hydrogen plant, the heat input is the full thermal rating of the VHTR. As in the other hydrogen generation processes, the work of compression of the product gas to 1000 psia is included in the calculation of efficiency.

8.4 Costs

Under subcontract from Westinghouse, UE&C conceptually designed the W H_2 Hydrogen Process plant (Reference 8-1) and developed a preliminary capital cost estimate associated with the hydrogen production process (excluding the electrolyzers, the H_2SO_4 decomposition reactor, and the VHTR unit). Preliminary results of this study are now available and together with supplemental data provided by Westinghouse and Oak Ridge National Laboratory are reported in Table 8-2. The total plant investment amounts to \$1,178 million, including \$453 million for the hydrogen plant and \$725 million for the VHTR. The VHTR cost is the \$800 million assumed in Section 2.4, with the approximate reformer cost of \$75 million removed. Process heat exchangers are included in the Westinghouse process plant estimate. Operating costs are shown in Table 8-3. The H_2 product cost varies from \$5.77 to \$ 9.76 per million Btu, as the nuclear fuel cost increases from \$0.25 to \$0.60 per million Btu and the fixed capital charge rate from 15% to 25%. These results are shown graphically in Figure 8-2. The predominant contribution to the product cost is the large capital expenditure associated with the 3220 MWt nuclear power source.

The total capital cost of the VHTR (\$725 million including IDC) was estimated by Oak Ridge National Laboratory as the cost of a typical VHTR.

The ORNL estimate was obtained from a detailed study by UE&C (Reference 8-2) which estimated the VHTR cost using recent HTGR experience. On the basis of a conceptual VHTR design, Westinghouse estimates this cost as about $447 million. If the Westinghouse estimate is substantiated when more details of the design are developed, the cost of the H_2 product could be considerably less than estimated here.

8.5 References

8-1 Studies of the Use of Heat From High Temperature Nuclear Sources for Hydrogen Production Processes, Astronuclear Laboratory, Westinghouse Electric Corporation for National Aeronautics and Space Administration, NASA CR-134918, January, 1976.

8-2 A Cost Comparison of Very High Temperature Nuclear Reactor for Process Heat Applications, March, 1975, United Engineers & Constructors Inc. (UE&C-AEC-750311)

TABLE 8-1

WESTINGHOUSE HYDROGEN PROCESS

OVERALL PROCESS EFFICIENCY

Heat Input	
VHTR Thermal Output	11.0×10^9 Btu/hr
Heat Output	
Product Gas	15.8×10^6 SCF/hr
Heating Value	325 Btu/SCF
Total Heat Output	5.14×10^9 Btu/hr

Overall Efficiency

$$\frac{5.14 \times 10^9 \text{ Btu/hr}}{11.0 \times 10^9 \text{ Btu/hr}} = 46.7\%$$

TABLE 8-2

WESTINGHOUSE HYDROGEN PROCESS
CAPITAL COST ESTIMATE
JULY, 1974 DOLLARS

DIRECT COSTS H_2 PROCESS COMPONENTS	CAPITAL COST $\$/10^6$
Electrolyzer power supply	18
Electrolyzers	113
Sulfuric acid decomposition and partial condensation	83
SO_2/O_2 separation	29
Turbine-generator	31
H_2 product compressors	5
Subtotal battery limit	278
Cooling and water intake	6
Makeup and feedwater	7
Waste-water treatment	0.3
Electrical auxiliary power	17
General off-sites investment	6
Subtotal outside battery limits	37
Land and land rights	0.2
Special materials	0.4
Contingency @ 15%	47.
Subtotal Direct Costs	363
INDIRECT COSTS	
Construction Facilities, Equipment & Services	9
Engineering Services	15
Other Costs	4
INTEREST During Construction 8 % for 4 years	63
Subtotal Indirect Costs	91
TOTAL CAPITAL COSTS H_2 PRODUCTION PROCESS (1)	453
TOTAL CAPITAL COSTS VHTR (2)	725
TOTAL PLANT INVESTMENT	$ 1,178

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(1) From Reference 8-1

(2) From Section 2.3

TABLE 8-3

WESTINGHOUSE HYDROGEN PROCESS
PRODUCTION COST ESTIMATE
JULY 1974 DOLLARS

	Product Cost $ Per Million Btu
Nuclear Fuel (3)	0.55 to 1.28
Operating and Maintenance Costs	
VHTR (1)	.22
Production Plant (2)	.06
Fixed Charges (1)	4.34 to 7.24
Total Production Cost	5.17 to 8.80

Plant Capacity	
Scf/Year	1.25×10^{11}
Btu/Year	4.07×10^{13}
Stream Factor	330 Days/Year

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(1) Based on Section 2.3
(2) Estimated from Reference 8-1
(3) Reference 8-1 Adjusted for Section 2.3 Cost Range

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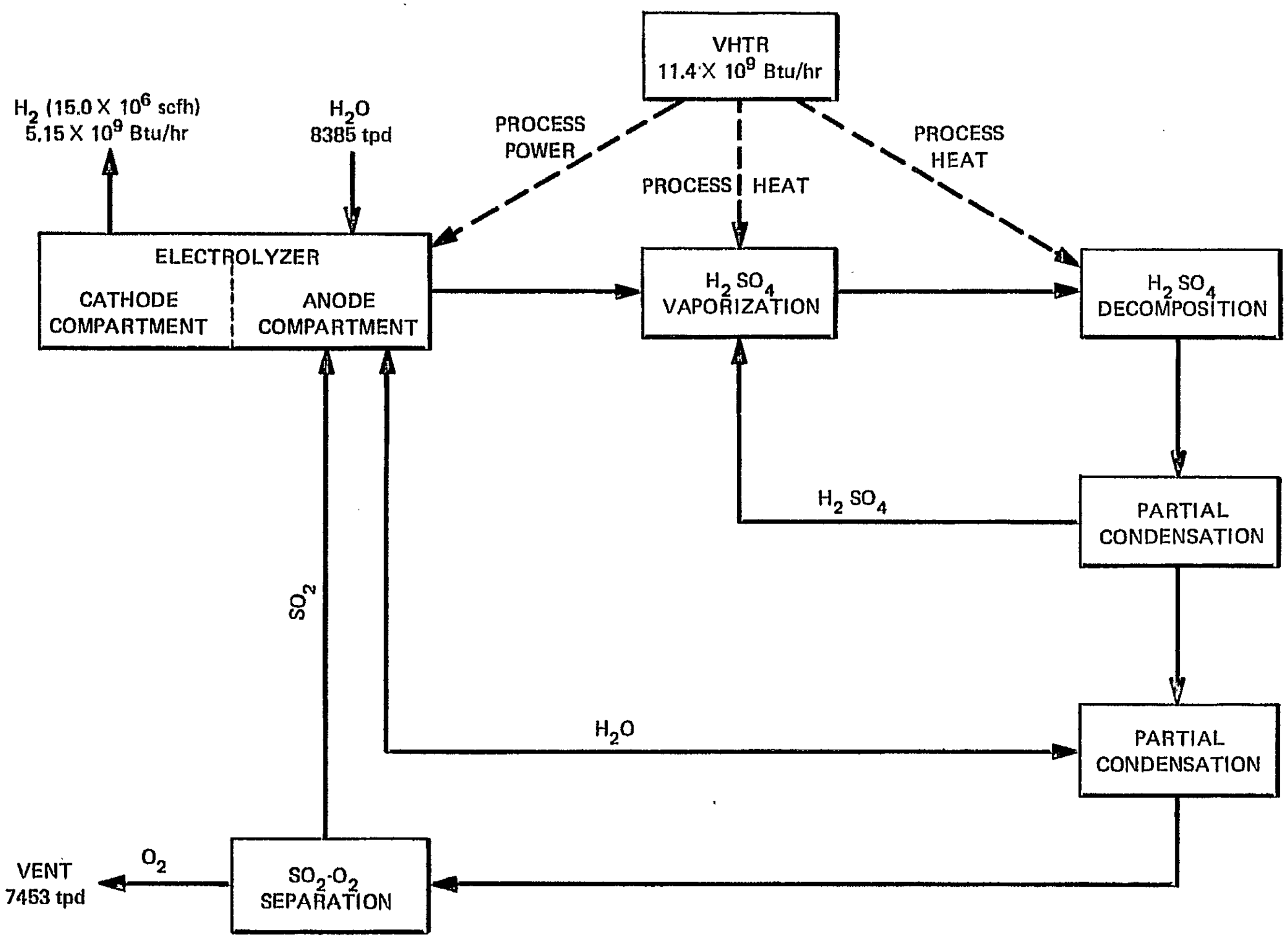

Fig. 8.1. Plant Block Diagram — Westinghouse Hydrogen Process

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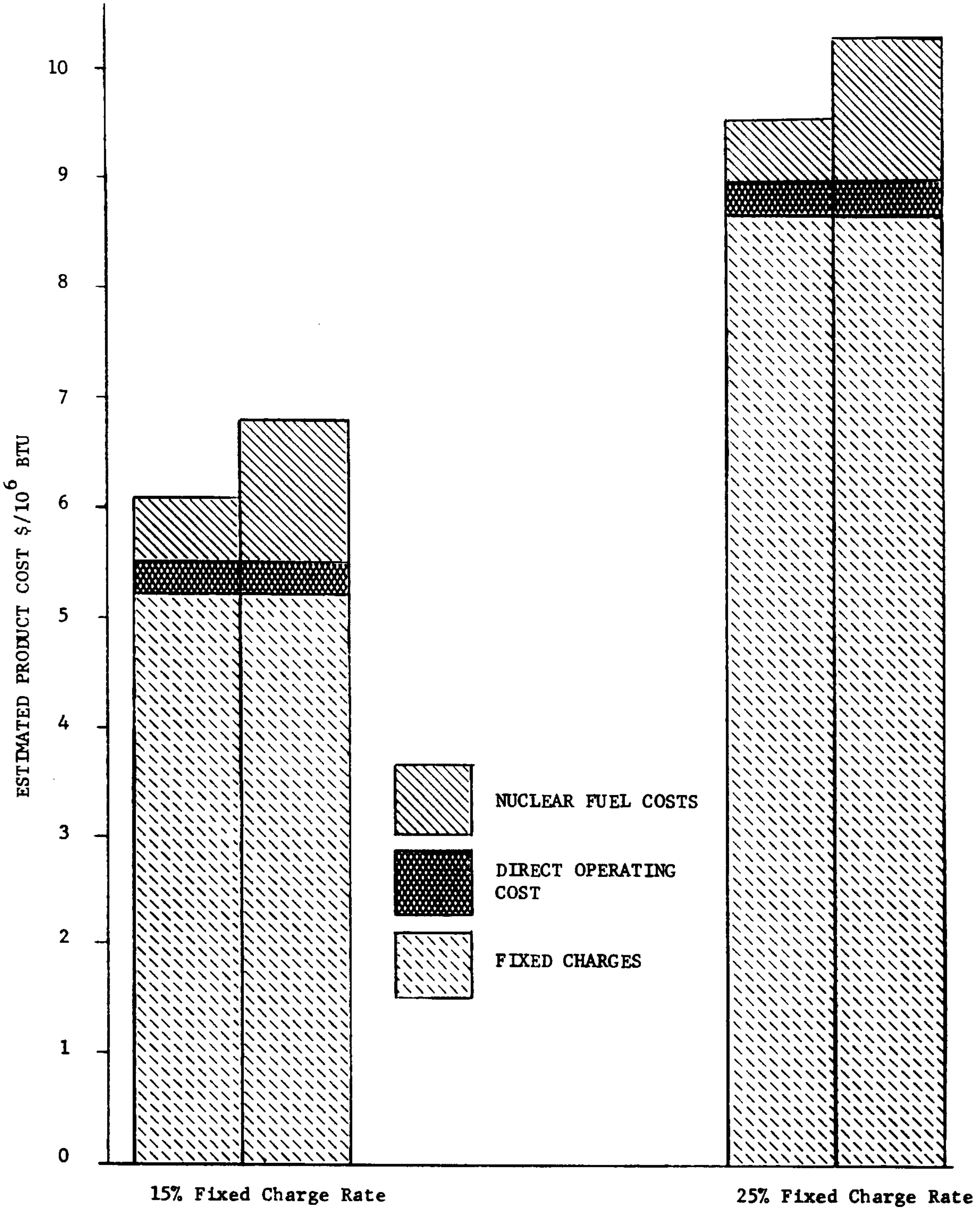

Fig. 8.2. Production Cost — Westinghouse Hydrogen Process

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