

CONVERSION OF FOSSIL FUELS TO UTILITY GAS

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Natural gas has nearly completely replaced the use of coal as a source of utility gas in the United States. As gas has shifted its status as a byproduct of petroleum, prices at the well in the Southwest, despite government regulation, have gone up sharply during the past fifteen years. Coal prices at the mine, however, have remained fairly constant during this period.

Natural gas is purchased by distribution companies from the transmission companies in the coal-producing area of West Virginia at about \$0.37 per MMBtu (million Btu). Coal in the same area selling at the mine for \$4.00-4.50/ton is equivalent in price to \$0.16-0.18/MMBtu. As the differential between coal and gas prices increases, as is likely, the conversion of coal to gas at the mine becomes increasingly attractive.

It has been demonstrated in a small pilot plant at the Institute of Gas Technology that the organic content of oil shale hydrogenates to methane even more readily than does coal. There are vast reserves of oil shale in Colorado and Utah; thus, a large gas-making potential in the form of shale exists in that area. On the other hand, it is doubtful that Eastern shales are rich enough to produce gas economically in the foreseeable future.

The location of the rich Western shales is far from the large Eastern population centers. This means that gas from shale would have to be cheaper by about \$0.20 to \$0.30 per MCF (thousand cubic feet) than gas from coal to compensate for transmission costs to justify early development of these reserves for gas-making purposes. Markets on the West Coast are closer, but natural gas is appreciably cheaper there than in the East.

The rich shales of Colorado and Utah presently seem better suited for production of liquids by relatively simple retorting at atmospheric pressure to recover crude shale oil. By conventional hydrogenation processes, a high quality gasoline can be made from the oil at prices close to present gasoline prices. Thus, the development of processes to make gasoline from oil shale may occur before gas from shale is a reality.

A variety of processes for the conversion of distillate and residual oils to gas have been developed and are being used widely for baseload gas elsewhere in the world. In this country, oil is used to produce gas for peaking purposes on the East Coast only. Distillate fuels, which are relatively simple to gasify, cost \$0.10/gallon, a raw material cost of about \$0.70/MMBtu. Crude oil, and even residual oil, can be converted to gas by hydrogenation⁷ or thermal cracking.⁵ However, even at the low price of \$2.00/bbl, about \$0.33/MMBtu, the raw material cost would be approximately twice that of coal.

Therefore, although the same general coal and shale gasification and hydrogenation techniques that will be discussed in this paper can be applied to oil, and

even though plant investment costs for gasification of oil would be lower than for coal or shale, the cost of oil is much too high at present for consideration of it as a feedstock for a baseload plant. Consequently, only those processes that are most promising for coal and oil shale will be considered in this paper.

UTILITY GAS FROM COAL

Coal Gasification

An excellent summary of past work on coal gasification and hydrogasification is given by C. G. von Fredersdorff and M. A. Elliott⁹ in the recent supplementary volume of "Chemistry of Coal Utilization." No attempt will be made to review that field further in this paper.

A study of the economics of coal gasification indicates that it is preferable to gasify coal under pressure when a heating gas containing methane is desired. The only pressurized gasification process being used at the present time is the Lurgi. The Scottish Gas Board is currently using it in their plant at Westfield,⁶ and it is also being used to make gas in Australia. The advantage of the Lurgi process is that due to the pressure of approximately 400 psig of the system, methane in appreciable quantities is obtained in the raw gas. Inasmuch as the methane-forming reaction is exothermic, it is possible to decrease the amount of oxygen fed into the gasifier. The Lurgi process requires a fixed bed, with coal being fed into the top of the gasifier through lock hoppers, and steam and oxygen into the bottom of the gasifier. Operation is nonslagging, requiring excess steam to maintain the temperature of the bottom of the gasification zone sufficiently low to avoid slagging of the ash.

The heating value of Lurgi gas after purification to remove carbon dioxide is in the range of 400 to 450 Btu/SCF. It is possible to make approximately 1000 Btu gas by employing catalytic methanation as an upgrading step to convert hydrogen and carbon monoxide to methane following Lurgi gasification. About 600 CF of oxygen is required to make 1000 CF of methane, including that methane which can be made by catalytic methanation of the hydrogen and carbon monoxide. In other gasification processes, where a suspension of coal is used with oxygen and steam at high temperatures, resulting in little or no methane in the product gas, about 1200 CF of oxygen is required per MCF of methane. Thus, the advantages of the Lurgi gas scheme for making a high-heating-value gas are obvious. A raw gas analysis from the Lurgi gasifier is:

CO ₂	30.5%
H ₂ S	1.0%
C _n H _m	0.6%
CO	16.5%
H ₂	42.0%
CH ₄	8.6%
N ₂	0.8%

Fig. 1 is a simple flowsheet of the major steps in the Lurgi gasification-catalytic scheme to produce gas having a heating value of approximately 1000 Btu/SCF. Using Eastern coals, it would be necessary to pretreat the coal to avoid agglomeration of the coal in the gasifier; conventional pretreatment consists of mild oxidation in the temperature range of approximately 600° to 900°F.

If it were necessary at the present time to convert coal to utility gas, we would have to select Lurgi gasification as the most advanced commercial process to accomplish this. However, there is under development a scheme using hydrogenation of coal that is economically more attractive.

Hydrogenation of coal to form methane proceeds very rapidly above temperatures of 1400°F and pressures of about 1000 psig. When methane is the chief product, rather than the liquids that are obtained when the temperature is lower, the process is called hydrogasification. Early work on hydrogasification was done by F. J. Dent and associates of the Gas Council in England, and has been continued recently in this country by the U. S. Bureau of Mines and the Institute of Gas Technology. The reaction of hydrogen with carbon to produce methane is highly exothermic. Rather than attempting to control the temperature within the hydrogasification reactor by means of cooling coils, a major improvement in the technology is injecting steam along with the hydrogen. The heat from the exothermic methane-forming reaction can be utilized by the endothermic steam-carbon reaction. Thus, additional hydrogen and carbon monoxide are made which can subsequently be reacted catalytically to form additional methane. The effect of steam addition is to decrease the hydrogen requirement to about 70 percent. In addition, the reactor construction is greatly simplified by avoiding internal heat exchange surfaces.

The hydrogasification process can be operated with either fluidized beds or moving beds with countercurrent contact. It is believed that pretreatment of the coal to avoid agglomeration can be avoided by dropping the fine coal particles into a devolatilization zone at the top of the gasifier. Thus, there would be no loss of methane that would accompany pretreatment procedures.

A simplified flowsheet of the hydrogasification process is shown in Fig. 2. About 50 percent of the carbon is gasified in the hydrogasification reactor. The remaining 50 percent is used in the gasifier with oxygen and steam to make hydrogen for the hydrogasification operation. The gasification step is done at a lower pressure, about 400 psig, followed by a CO shift, gas purification to remove CO₂, and then compression to 1000 psig. Steam of about equal volume is added to the hydrogen for hydrogasification.

The crude gas from hydrogasification is subjected to CO shift to adjust the hydrogen/carbon monoxide ratio to about 3/1 for methanation purposes, gas purification to remove carbon dioxide and sulfur compounds, and finally, catalytic methanation with suitable iron or nickel catalysts. It is possible to reduce oxygen consumption to about 320 SCF/MCF of total methane made in this process.

Hydrogasification and the Steam-Iron Process

Inasmuch as the cost of the product gas is greatly affected by the oxygen cost, considerable thought has been given to development of processes that avoid the use of commercially pure oxygen. One such process would be the combination of hydrogasification with a modernized version of the steam-iron process to make hydrogen for the hydrogasification step. This system is being investigated by a group of three companies: the Consolidated Natural Gas System, Texas Eastern Transmission Corporation, and Consolidation Coal Company in a pilot plant of the Institute of Gas Technology. Preliminary work on the steam-iron process was done by the U. S. Bureau of Mines³ in their Bruceton Laboratories at 300 psig. The pressure in the Institute of Gas Technology pilot plant has been extended to 1000 psig with greater throughputs.

Hydrogasification would be carried out in the manner previously described with the exception that a stream of hydrogen and steam obtained from the steam-iron process is passed directly into the hydrogasification reactor. This scheme is shown in Fig. 3. Residual char from the hydrogasification step is sent without pressure reduction to a gas producer in which it is reacted with steam and air, rather than oxygen, to make producer gas. The producer gas reduces iron oxide, which is then reoxidized with steam in a separate vessel.

Because the steam and hydrogen can be made available at elevated pressures and temperatures, a considerable reduction of equipment is possible. Shifting of the carbon monoxide, scrubbing of the hydrogen stream to eliminate carbon dioxide, and subsequent compression are unnecessary. Injection and preheating of the steam for hydrogasification is avoided. The main advantage, however, is in the elimination of commercially pure oxygen. The spent producer gas still contains appreciable energy and can be expanded through a turbine to compress the air required for the gas producer. In addition, the spent producer gas may be burned to provide the steam which is required in the plant.

Gas Costs

Raw material costs and plant investment for gas made by these three schemes are summarized in Table 1. While data from pilot plants are by no means complete at this time, it is possible to make reasonable assumptions and to estimate the final gas costs. The plant size was taken as 90×10^9 Btu/day. Coal used in these estimates had a heating value of 12,500 Btu/lb. The pipeline gas composition was assumed to be the same in all cases, and had a heating value of 987 Btu/CF. By using a combination of iron and nickel methanation catalysts, it should be possible to produce a product gas that contains sufficient ethane and propane to yield a heating value close to 1000 Btu. It is possible to keep carbon monoxide below 0.1 percent. Plant costs, thermal efficiencies, and final gas costs are given. By use of the steam-hydrogen process as a source of hydrogen, plant costs for the hydrogasification plant can be reduced to about \$40 million from the \$65 million when hydrogen is made by oxygen-char gasification, and \$90 million for the Lurgi installation.

Fig. 4 shows graphically the gas costs under the three schemes, and the effects of coal and oxygen upon them. Final costs are complete and include con-

Table 1.-SUMMARY OF RAW MATERIAL REQUIREMENTS,
PLANT INVESTMENT, AND GAS PRICE
 90×10^9 Btu/Day Utility Gas From Coal

	Coal Cost: \$4.50/ton Oxygen Cost: \$7.00/ton (\$0.30 MCF)		
	Lurgi Gasification, Catalytic Methanation	Hydrogasification, H ₂ From Char, O ₂ , Steam	Hydrogasification, H ₂ From Steam-Iron
Coal consumption, tons/day	6,540	5,220	4,600
Oxygen consumption, tons/day	2,300	1,200	none
Utility gas analysis, %			
CH ₄	_____	91.0	_____
C ₂ H ₆	_____	3.0	_____
C ₃ H ₈	_____	0.4	_____
H ₂	_____	3.0	_____
CO	_____	0.1	_____
CO ₂	_____	0.5	_____
N ₂	_____	2.0	_____
Heating value, Btu/SCF	_____	987	_____
Plant investment, \$Million	90	65	40
Plant thermal efficiency, %	55	69	78
Price of gas, \$/MMBtu	0.95	0.70	0.52

ventional utility return on investment, and federal taxes, in accordance with the procedure recommended by the American Gas Association and summarized later in this paper. Plant life was taken at 20 years, and gas cost is averaged for the 20-year period. No credit was taken for byproducts. These estimates indicate that with a reasonable degree of success in improving coal gasification technology, it would be possible to decrease the cost of utility gas made from coal from \$0.95/MCF to \$0.52/MCF for a 90×10^9 Btu/day plant.

UTILITY GAS FROM OIL SHALE

Comparison of Hydrogenation of Oil Shale and Shale Oil

Utility gas can be produced from oil shale by two routes: one is to hydrogenate the shale directly, and the other is to retort the material first and then hydrogenate the shale oil. Figs. 5 and 6 show process schemes for these two routes. In both processes sufficient shale is retorted to provide oil for making hydrogen and for boiler fuel requirements.

The same method for hydrogen manufacture is used. Synthesis gas is made by partial oxidation of shale oil, using 99 percent purity oxygen plus steam.^{4,8} Raw synthesis gas is scrubbed free of carbon and hydrogen sulfide prior to carbon monoxide shift, in which the carbon monoxide concentration is reduced from 46 to 1.3 percent. Following the shift reaction, the carbon dioxide is reduced to two percent of the process hydrogen stream by scrubbing with hot carbonate solution. Process hydrogen is compressed to hydrogasifier pressure, 1000 psig.

Hydrogasification of oil shale is carried out in a moving bed, with solids and gas downflow at 1000 psig, and at a temperature range of 1050° to 1350°F. Although most of the methane is produced in this step, the hydrogasifier effluent contains substantial amounts of carbon monoxide. This is catalytically shifted to adjust the H₂/CO ratio to a value suitable for methanation. Prior to the latter step, sulfur compounds, benzene, and ammonia are scrubbed from the gas.

The alternate to direct hydrogenation of shale is the hydrogasification of shale oil produced by retorting the shale. Work at IGT on high-pressure hydrogasification of petroleum oils⁷ showed that control of coke deposition from crude and residual oils would be necessary to permit continuous operation of a process not using some means of coke removal. A two-step process was developed. In the first step, the oil is catalytically hydrogenated at 3000 psig and 780°-790°F. Design of this step is based on work of the U. S. Bureau of Mines.^{1,2}

The hydrogasification step is based on work at IGT. Because of the small production of carbon oxides in the hydrogasification step, only final purification is necessary following hydrogasification of the oil.

An initial comparison of the economics of the two processes on the same cost basis showed no significant difference, with prices of 68 and 69 cents/MMBtu utility gas for oil shale and shale oil hydrogenation, respectively. Both processes were designed for hydrogen/shale or shale oil ratios of 100 percent of the stoichiometric. The oil shale hydrogenation design included a hydrogen-methane separation step.

Experimental work on hydrogasification of oil shale subsequent to this design indicated that successful operation could be carried out at hydrogen/shale ratios much less than stoichiometric. Utility gas plant designs based on these ratios showed investment savings from both the elimination of hydrogen-methane separation and the reduction of the size of hydrogen plants. Reduction of the hydrogen/shale ratio results in increased carbon formation, but this is discharged with the spent shale residue and causes no operating problem, such as the plugging of reactor tubes, that could occur in shale oil hydrogenation at drastically reduced

hydrogen/oil ratios. Since the direct hydrogenation of oil shale appeared to offer greater possibilities for utility gas cost reduction than shale oil hydrogenation, further economic studies were restricted to the former.

Reaction of Oil Shale with Synthesis Gas

One of the problems in hydrogasification of oil shale is the necessity of preheating the shale to about 1050°F to initiate the reaction. A solids downflow countercurrent solids-gas flow reactor is very advantageous for heat transfer, allowing the hot effluent gases to preheat the fresh shale at the reactor top, and the hot spent shale to preheat hydrogen at the bottom. However, since the shale must be preheated to 1050°F, while vaporization of hydrocarbons begins at around 700°F, a countercurrent flow will result in shale oil being carried out with the product gas. Recovery of this oil on spent shale, followed by hydrogenation in a second reactor, might be done, but that would complicate and increase the cost of the hydrogenation system. A solids-downflow cocurrent system prevents oil carry-over, but increases the heat transfer problem.

Experimental work in the pilot plant was carried out with cocurrent gas-solids downflow. Adapting such a system to a commercial installation would require that either the hydrogen stream be preheated sufficiently to bring the shale up to reaction temperature, or that a fluidized preheat section with internal heating tubes be installed in a section of the reactor. Bringing in sufficient heat with the hydrogen stream at the maximum temperature consistent with practical design requires a high hydrogen/shale ratio, which has been shown to be less desirable economically.

The use of raw synthesis gas from the oil partial oxidation reactors in place of hydrogen as the hydrogenating gas provides a way of preheating the shale as well as offering economic and process advantages. The heat-carrying capacity of the gas is increased by the carbon oxides and steam which accompany the hydrogen required for reaction. Computations show that a shale synthesis gas mixture temperature of 1050°F can be obtained with 2500°F synthesis gas from partial oxidation reactors if the shale is preheated to 500°F.

The use of raw synthesis gas directly in the hydrogasification reactors has economic advantages in addition to the above operating and cost advantages. The raw synthesis gas at 2500°F flows directly from the partial oxidation reactors to the hydrogasifiers. To avoid compression of the hot synthesis gas, it would be necessary to operate the synthesis gas generators at hydrogasifier pressure, 1000 psig. This raises reactor costs; however, synthesis gas cooling and scrubbing equipment, water-gas shift unit, contact tower and coolers, and hot carbonate scrubbing system, would all be eliminated from the hydrogen section in this scheme (Fig. 7).

At low equivalent hydrogen/shale ratios, with synthesis gas the heat-carrying capacity can be maintained by increasing the steam/oil ratio in the partial oxidation reactors, which increases the amount of hot gas for a given quantity of shale oil and oxygen.

Optimization of Hydrogen/Shale Ratio

A study of the economic effect of reducing the synthesis gas/shale ratio was made for equivalent hydrogen/shale ratios ranging from 61 to 0 percent of stoichiometric. Without external hydrogen, all hydrogen must be obtained from the oil shale, resulting in a low efficiency of carbon conversion. Shale preheat can be achieved either directly by a flue gas produced by combustion with oxygen, in the presence of steam, of enough oil to produce the required amount of flue gas, or indirectly by burning oil in air, and passing the hot combustion gas through heating tubes immersed in a fluid-bed shale preheat section. The latter appears to be a

cheaper method because of the elimination of the oxygen plant and synthesis gas generators, which more than compensates the added expense of indirect preheat.

In order to show the economic advantage of using synthesis gas instead of process hydrogen as a source of external hydrogen supply to the hydrogasifier, estimates of utility gas costs when using the latter were also made. These estimates covered the same hydrogen/shale ratio as in the synthesis gas cases, and are based on cost data derived from the latter.

Comparison of Costs

The effect of hydrogen/shale ratio on utility gas price for the two sets of estimates is shown in Fig. 8. The 20-year average price of gas represents capital and operating charges typical of utility financing for a 20-year plant life. Both sets of costs pass through a minimum at about one-third the stoichiometric hydrogen/shale ratio. Total plant investment and shale requirements also pass through a minimum at this point. The existence of a minimum results from the fact that as less external hydrogen is used, incremental decreases in hydrogen/shale ratio result in more than proportional increases in shale required for hydrogasification and all the attendant costs of increased solids usage. At some point they overbalance the savings in hydrogen supply facilities. At the minimum price for each method, the use of synthesis gas has a cost advantage of 3 cents/MMBtu utility gas.

Operation without external hydrogen is undesirable because of the low percentage of conversion of oil shale to gas.

Table 2 summarizes major process items for plant designs based on optimum hydrogen/shale ratios with hydrogen and with synthesis gas. Breakdown of

Table 2.-SUMMARY OF RAW MATERIAL REQUIREMENTS,
PLANT INVESTMENT, AND GAS PRICE
90X10⁹ Btu/Day Utility Gas From Oil Shale

Oil Shale Cost: \$0.72/ton
Oxygen Cost: \$7.00/ton (\$0.30/MCF)

	Reaction With Hot Synthesis Gas <u>32.5% Stoich H₂/Shale</u>	Reaction With Process Hydrogen <u>36% Stoich H₂/Shale</u>
Oil shale (40 gal/ton) consumption, tons/day	22,312	24,866
Oxygen consumption, tons/day	1,165	798
Utility gas analysis, %		
CH ₄	79.5	86.4
C ₂ H ₆	5.9	3.3
H ₂	10.0	4.3
CO	0.1	0.1
CO ₂	0.8	0.9
N ₂	3.7	5.0
Heating value, Btu/SCF	942	947
Plant investment, \$Million	57.1	60.2
Plant thermal efficiency, %	65	59
Price of gas, \$/MMBtu	0.56	0.59

plant shale requirements in tons/day for the two designs is:

	<u>Synthesis Gas</u>	<u>Hydrogen</u>
Hydrogasification	12,780	13,852
Synthesis Gas or Hydrogen	6,156	5,436
Retorting for Fuel	3,376	5,578

When process hydrogen is used to hydrogenate oil shale, more shale is required in the hydrogasifier and less is retorted for shale oil than when synthesis gas is used for hydrogenation. With synthesis gas less hydrogen is made by CO shift, making it necessary to gasify more oil. When the synthesis gas is cooled from 2800° to 1050° F during mixing with shale, only about 14 percent of the carbon monoxide is shifted by reaction with the water present. When hydrogen is produced as a separate stream, 96 percent of the carbon monoxide is catalytically shifted at a lower temperature and with a high steam/carbon monoxide ratio. With synthesis gas, more of the methane is produced by methanation of CO in the hydrogasifier effluent than when process hydrogen is used (20 percent over 13 percent). This requires less shale to be handled in the hydrogasifier system.

Elements of Utility Gas Price

Fig. 9 shows graphically the proportions of utility gas price represented by oil shale and oxygen, as raw materials, as separate items from the total of capital and operating costs for the rest of the plant. For the studies on which these cost estimates are based, a mined shale price of \$0.72/ton was used. At this price it represents about one-third of the utility gas price. Oxygen at \$7/ton (including excess capacity) is one-sixth and one-tenth of the gas price for the synthesis gas and hydrogen processes, respectively. These two elements represent about one-half the gas price.

Shale Mining Costs and Shale Richness

The mined shale price of \$0.72/ton was based on information supplied by Cameron and Jones, Inc., for a daily mining capacity of 25,000 tons. In order to allow for the possibility of variations in the cost of mined shale on gas price, Fig. 10 shows gas prices as a function of mined shale cost.

The cost estimates presented in this paper are based on 40 gal/ton Colorado oil shale. This is probably a higher quality raw material than would be available to plants manufacturing utility gas from oil shale, except through selective mining of wide areas.

The use of leaner shale increases the burden of unreactive rock that has to be mined, ground and sized, and processed in an oil shale conversion plant. Estimation of the effect of this on utility gas price is:

<u>Shale quality, gal/ton</u>	<u>Utility gas price, \$/MMBtu</u>
40	0.556
30	0.658
25	0.746

In summary, utility gas can be manufactured by the hydrogasification of oil shale at reasonable cost. The most important process variable influencing the cost of utility gas is the hydrogen/shale ratio, with the optimum value being about one-third the stoichiometric value. From both an operating and economic standpoint, the best way to supply hydrogen is by synthesis gas. Oil shale price and quantity exert a greater effect on gas costs than any of the individual process steps. The major problem in making utility gas from oil shale, in contrast to coal, is in solids handling. As oil shale richness decreases, the solids-handling problem becomes more important and might make oil hydrogasification more

attractive if it could be carried out at low hydrogen/oil ratios. This might be accomplished in a fluidized or moving coke bed as a means of removing carbon.

Correlation of Utility Gas Price, Investment, and Fossil Fuel Cost

The utility gas prices presented in this paper are 20-year average prices computed by an accounting procedure developed by the General Accounting Committee of the American Gas Association. This procedure is based on the financing of utility gas plants at 65 percent debt and 35 percent equity. Straight-line depreciation is assumed over a 20-year period and interest at 5 percent on the outstanding debt is charged. Seven percent return on undepreciated fixed investment is assumed. The 20-year effective average capital charge composed of federal income tax, debt, and net income amounts to about 5.8 percent. State and local taxes and insurance are taken at 3 percent and annual depreciation at 5 percent. This procedure has been programmed for computer operation.

Fig. 11 presents a generalized correlation of utility gas price versus total capital investment for a 90×10^9 Btu/day plant at various levels of fossil fuel cost. The latter parameter is the cost of the fossil fuel (coal, shale, or oil) in \$/MMBtu as fed to the plant, divided by the overall plant thermal efficiency of conversion to utility gas.

From a number of cost estimates of plants for making utility gas from coal and oil shale, relations between operating labor and daily material charges other than fossil fuel as percentages of equipment investment were derived for purposes of correlation. Operating labor and daily materials were taken as 2 and 0.5 percent, respectively, of total equipment, or bare cost. No byproduct credit is included. Capital investment is the bare cost plus contractor's overhead and profit, interest on fixed investment, and working capital.

Effect of Plant Size

Size of plant can have an appreciable effect on gas costs. A plant of only 90×10^9 Btu/day is not large enough to achieve the best economy. This is equivalent in product processed on a Btu basis to a petroleum refinery of only about 15,000 bbl/day capacity. The cost per unit of product is reduced by the petroleum industry by increasing the size of refineries to 50,000-100,000 bbl/day. It would be reasonable that coal gasification plants located at the mine would operate more economically if the size were increased to $300-400 \times 10^9$ Btu/day. Large pipelines readily transport $500-600 \times 10^9$ Btu/day.

Fig. 12 was prepared to show the effect of larger plants on the product gas cost. The effect of increasing the plant size on the unit cost, exclusive of fuel cost, was based on plant investment as a function of the 0.8 power of plant capacity. The raw material cost, coal or shale, was \$0.23/MMBtu. All other costs were then assumed to vary as the 0.8 power of plant size. This is a simplification, but serves to illustrate the savings possible with larger plants. Gas that costs \$0.60/MMBtu with a 90×10^9 Btu/day plant would cost about \$0.50 with a 400×10^9 Btu/day plant (Curve A), and \$0.50 gas from the smaller plant would cost \$0.43/MMBtu with the larger plant (Curve B). Gas made from hydrogasification with hydrogen from the steam-iron process could be reduced from \$0.52 to about \$0.445/MMBtu.

It is interesting to note that with the foregoing assumption of the 0.8 power, the investment cost of a coal-to-gas plant would be less than that of a 30-inch pipeline, 1,000 miles long, for equal daily capacities. This length of pipeline would be required to bring gas from Louisiana as far eastward as the coal fields of West Virginia.

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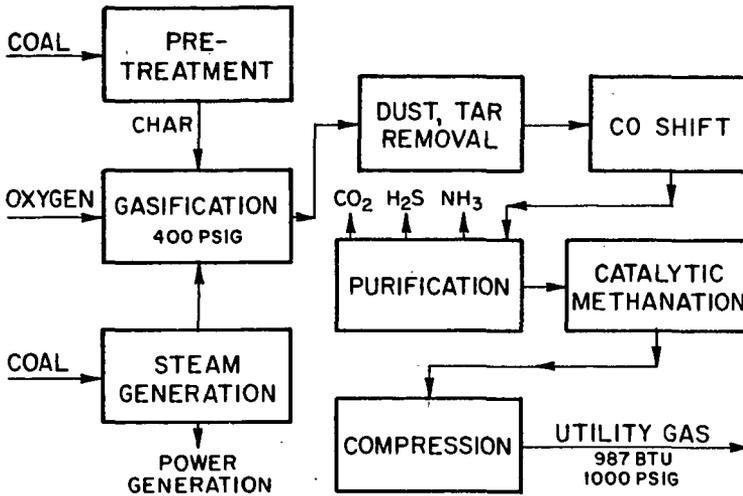


Fig. 1.-UTILITY GAS BY LURGI GASIFICATION FOLLOWED BY METHANATION

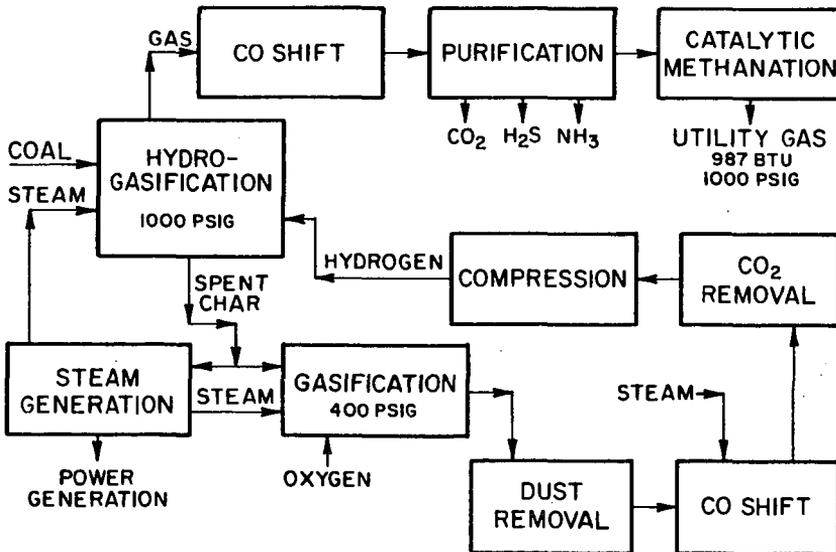


Fig. 2.-UTILITY GAS BY HYDROGASIFICATION, WITH H₂ FROM GASIFICATION OF CHAR, O₂, AND STEAM

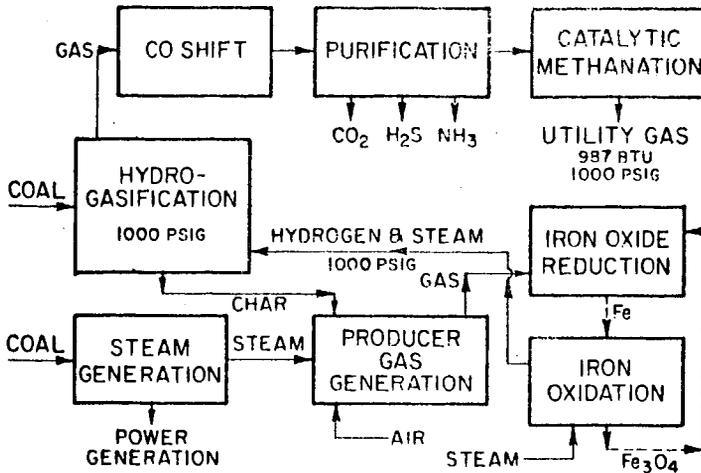


Fig. 3.-UTILITY GAS BY HYDROGASIFICATION, WITH H₂ FROM THE STEAM-IRON PROCESS

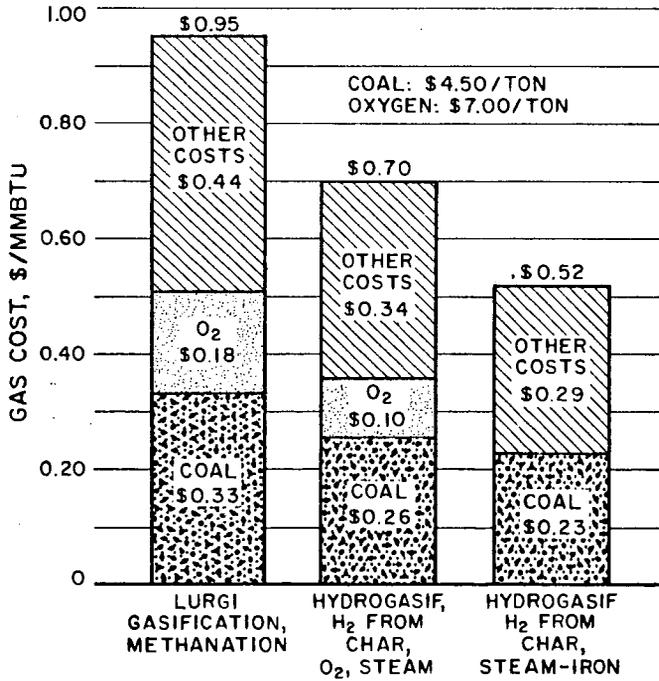


Fig. 4.-ELEMENTS OF UTILITY GAS COST IN GASIFICATION OF COAL (90×10^9 Btu/day)

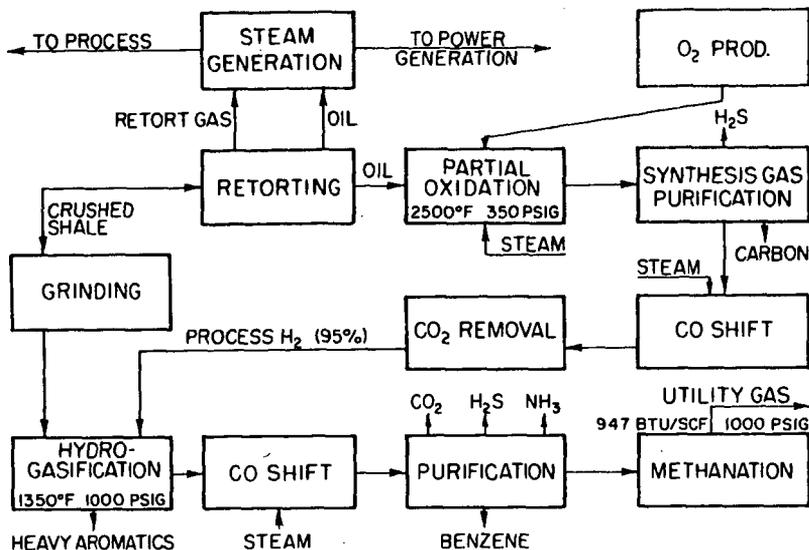


Fig. 5.-UTILITY GAS BY HYDROGASIFICATION OF OIL SHALE, REACTION WITH HYDROGEN

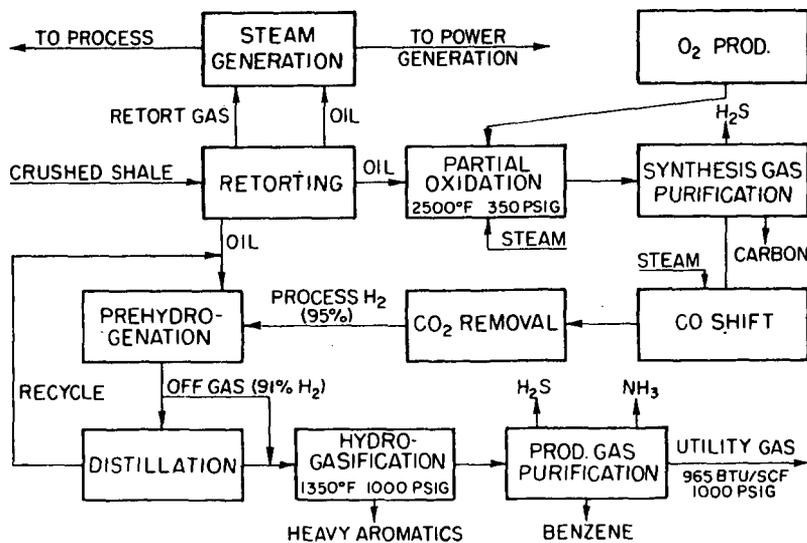


Fig. 6.-UTILITY GAS BY HYDROGASIFICATION OF SHALE OIL

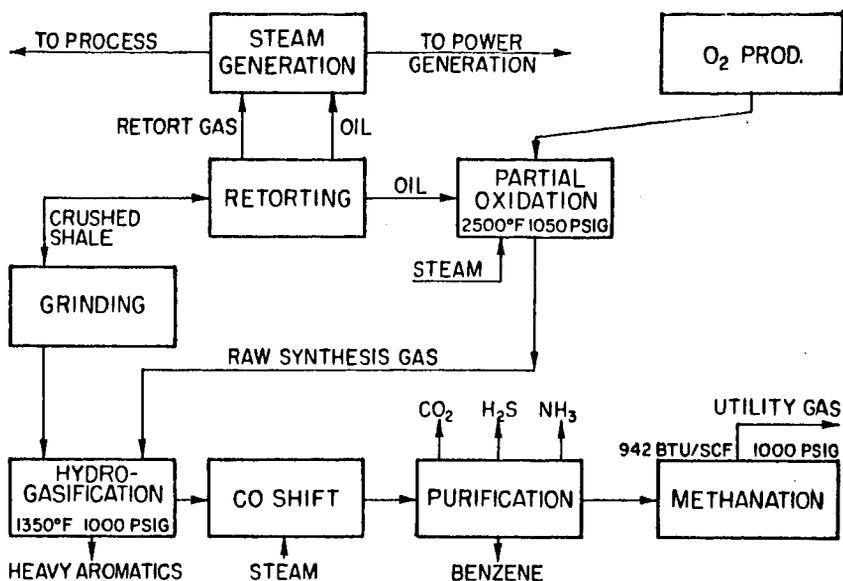


Fig. 7.-UTILITY GAS BY HYDROGASIFICATION OF OIL SHALE, REACTION WITH RAW SYNTHESIS GAS

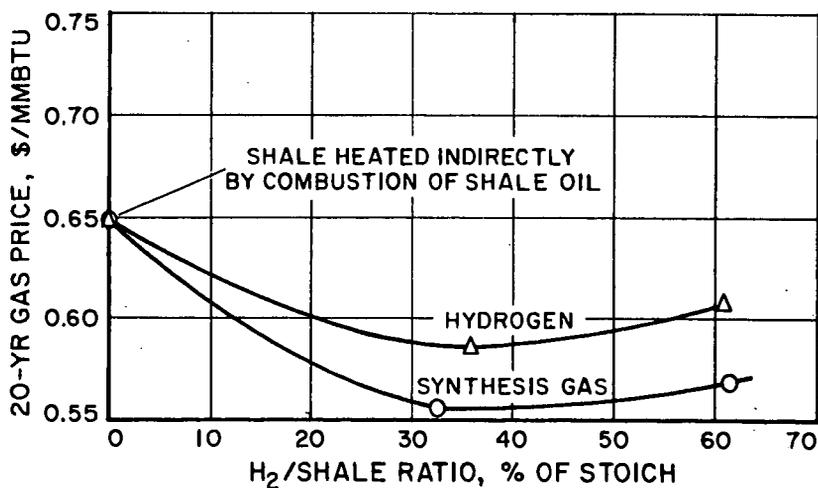


Fig. 8.-OPTIMIZATION OF HYDROGEN/SHALE RATIO, HYDROGASIFICATION WITH SYNTHESIS GAS AND PROCESS HYDROGEN

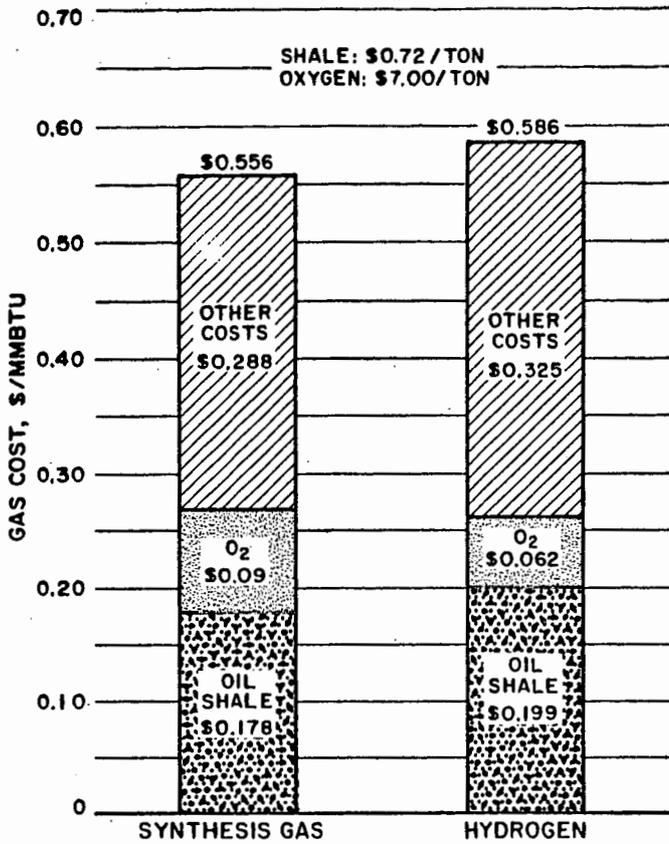


Fig. 9.-ELEMENTS OF PIPELINE GAS COSTS, HYDROGASIFICATION OF OIL SHALE BY REACTION WITH PROCESS HYDROGEN AND SYNTHESIS GAS AT ECONOMIC OPTIMUM H₂/SHALE RATIOS

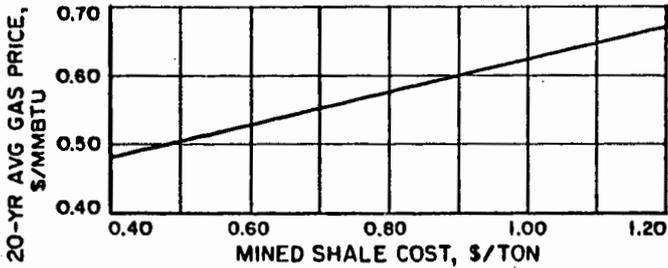


Fig. 10.-EFFECT OF MINED SHALE COST ON THE PRICE OF UTILITY GAS

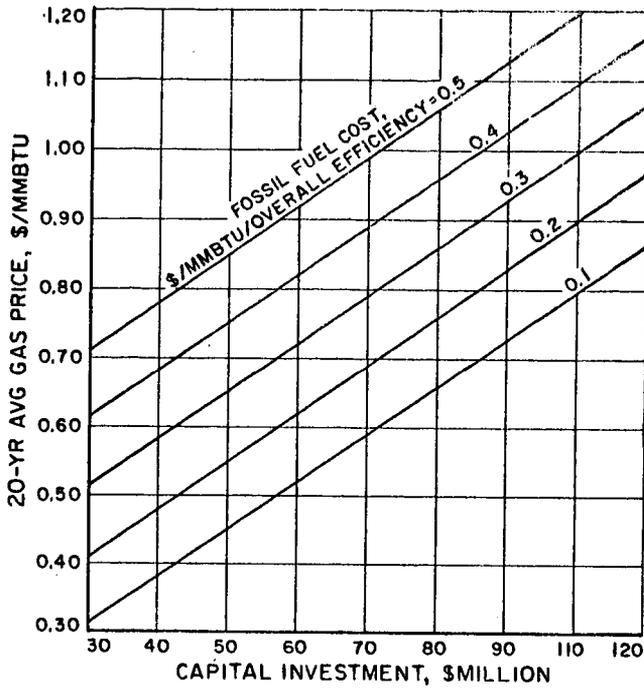


Fig. 11.-CORRELATION OF UTILITY GAS PRICE WITH INVESTMENT AND FOSSIL FUEL ENERGY COST (90×10⁹ Btu/Day Plant)

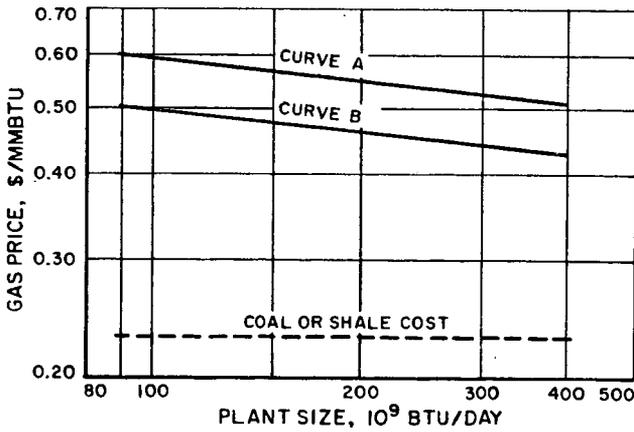


Fig. 12.-EFFECT OF PLANT SIZE ON PRICE OF PIPELINE GAS FROM COAL