SECTION 5. RESULTS OF INITIAL SCREENING OF PROPOSED GASIFICATION PROCESSES

The data and information collected on all processes were evaluated as a basis for the selection of processes for further consideration as projected for full-scale commercial operation.

The results of this initial screening are summarized in Table 5-1, together with the bases indicated on which the various processes were eliminated from further consideration and with references to text to where individual processes are discussed.

The gasification systems chosen for evaluation as projected to full-scale commercial operation include the following according to the type of product gas:

Synthesis Gas

Lurgi Dry-ash Gasifier (Process 11)
Lurgi Slagging Gasifier (Processes 18, 19, and 20)
Hydrocarbon Research Gasifier (Process 21)
Bamag-Winkler Atmospheric Gasifier (Process 7)
Rummel Single-shaft Pressurized Gasifier (Process 61)
Rummel Modified Single-shaft Pressurized Gasifier (Process 62)
Koppers-Totzek Pressurized Gasifier (Process 60)
Texaco Gasifier (Process 22)
Fixed-bed Super-pressure Gasifier (Process 56)
Fluidized-bed Super-pressure Gasifier (Process 57)
Two-stage Super-pressure Entrained Gasifier (Process 58)
Catalytic Steam Methanation Gasifier (Process 65)

Fuel Gas

Wellman-Galusha Gas Producer (Process 32)
IFE Two-stage Gas Producer (Process 33)

Gas Turbine Fuel

Airblown Lurgi Gas Producer (Process 37)

Steam-iron Reduction Gas

Two-stage Fluidized Super-pressure Gas Producer (Process 46)

Wherever possible, cost data for available commercial gasification processes were used as a basis for selecting gasification processes for further evaluation. Since the synthesis gas produced by gasification is ultimately to be converted to pipeline gas for use at 1000 psig, it was logical to consider the effect of pressure on gas cost. A comparison between the Lurgi process, the only commercial process operating at elevated pressure, and known atmospheric gasification processes made it apparent that operation at elevated pressure reduces investment and operating cost materially. Also, it was found that operation of the gasifiers at pipeline pressure i.e. 1050 psi, will lead to costs below those processes working at the present commercial level, 450 psi.

TABLE 5-1. SUMMARY OF RESULTS FROM INITIAL SCREENING OF PROPOSED GASIFICATION PROCESSES

I. COMMERCIAL SYNTHESIS GAS PROCESSES	Accepted for Evaluation as Projected to Commercial Scale	Excessive Fuel Costs	Not Amenable to Pressurization	Lump Fuels Only	Limited to Non-caking Fuels	Excessive Maintenance Cost	Excessive Equipment Cost	Excessive Operating Cost	Low Capacity	Remarks
Processes Using Oxygen and Coke										
1. UGI Converted*		x		x	x					The pressurized version of processes 1-6 using coal, the Lurgi process, is being evaluated.
2. Thyssen Galoczy		x		x	x					
3. Kerpely		x		x	x					
4. Leuna										
5. BASF-Leuna		x		x	x					
6. Wellman-Galusha		x		x	x					
Processes Using Oxygen and Coal										
7. Bamag-Winkler Atmospheric	x									
8. Koppers-Totzek	x									Pressurized version evaluated
9. B & W-DuPont										Suspension gasifier, see 8 and 10
10. Rummel Single-shaft	x									Pressurized version evaluated
11. Lurgi Dry-ash	x									
Processes Using Air and Coal										
12. Gas Integrale			x	x	x	x		x	x	
13. Pintsch Hillebrand			x	x	x	x	x	x	x	

*The cyclic water-gas process using air has become obsolete and many gasifiers have been converted to the use of oxygen.

TABLE 5-1. SUMMARY OF RESULTS FROM INITIAL SCREENING OF PROPOSED GASIFICATION PROCESSES (Continued)

II. PILOT-SCALE SYNTHESIS GAS PROCESSES	Accepted for Evaluation as Projected to Commercial Scale	Excessive Fuel Costs	Not Amenable to Pressurization	Lump Fuels Only	Limited to Non-caking Fuels	Excessive Maintenance Cost	Excessive Equipment Cost	Excessive Operating Cost	Low Capacity	Remarks
Processes Using Oxygen and Coal										
14. BASF-Flesch-Demag						x	x	x		Problems in building large units
15. Panindco										Suspension gasifier, see Processes 8 and 10
16. USBM Vortex										
17. Inland Steel										
18. Gas Council-Lurgi	x									Evaluations based also on data from Lurgi, Frankfurt
19. BCURA-Lurgi*	x									
20. USBM-Lurgi	x									
21. Hydrocarbon Research	x									
22. Texaco	x									
23. USBM Morgantown										Suspension gasifier, see Processes 8 and 10
24. Bianchi										
25. IGT Cyclonizer										
Processes Using Air and Coal										
26. ICI Moving Burden		x	x					x		
27. Heller Process			x				x	x	x	Material of construction not available
28. Rummel Double-shaft		x	x			x	x	x	x	Small units, low efficiency
29. USBM Annular Retort			x	x	x				x	Small, costly units
30. USBM Electrically Heated**			x	x	x				x	

*Uses Coke

**Uses no Air

TABLE 5-1. SUMMARY OF RESULTS FROM INITIAL SCREENING OF PROPOSED GASIFICATION PROCESSES (Continued)

III. COMMERCIAL FUEL GAS PROCESSES	Accepted for Evaluation as Projected to Commercial Scale	Excessive Fuel Costs	Not Amenable to Pressurization	Lump Fuels Only	Limited to Non-caking Fuels	Excessive Maintenance Cost	Excessive Equipment Cost	Excessive Operating Cost	Low Capacity	Remarks
Processes Using Air and Coal										
31. Power-Gas Mechanical										Fixed-bed producer, see Process 32
32. Wellman-Galusha	x									
33. IFE Two-stage	x									
34. Bamag-Winkler Atmospheric										Only large units economical; gas of low heating value
35. Ruhrgas Vortex*										Combination with power plants and gas turbines
36. LR Process*										
37. Lurgi Dry-ash	x			x						Combination with gas turbines
IV. PILOT-SCALE FUEL GAS PROCESSES										
Processes Using Air and Coal										
38. BCR-Kaiser				x				x		Operating problems inherent
39. BASF-Flesch-Demag			x			x	x	x		Problems in building large units
40. CEGB Marchwood*										Combination with gas turbines
41. Great Northern Railway										Suspension gasifier, see Process 35
42. Panindco										
43. B & W Cyclone										
44. FRS Cyclone										

*See Research Opportunities, Section IX.

TABLE 5-1. SUMMARY OF RESULTS FROM INITIAL SCREENING OF PROPOSED GASIFICATION PROCESSES (Continued)

V. CONCEPTUAL PROCESSES	Accepted for Evaluation as Projected to Commercial Scale	Excessive Fuel Costs	Not Amenable to Pressurization	Lump Fuels Only	Limited to Non-caking Fuels	Excessive Maintenance Cost	Excessive Equipment Cost	Excessive Operating Cost	Low Capacity	Remarks
Fuel or Producer Gas Using Air										
45. Bechtel Carbonizer*										Combination with gas turbine and power plant; see Process 40
46. Two-stage Fluidized Superpressure	x									Combination with steam-iron process
Synthesis Gas Using Air										
47. CO_2 Acceptor										Evaluation by others
48. Stookey			x	x				x	x	Similar to Process 12
49. Chemcoke*			x							Combination with power plants necessary
50. Nichols-Herreshoff			x				x			Small units
51. Cameron and Jones			x	x						
52. Standard Oil Fluidized-bed										Fluidized process, see Process 26
53. Mayland Pebble-bed			x			x		x		Operating difficulties
54. Jensen Electric		x						x		Feasible with low power cost

*See Research Opportunities, Section IX.

TABLE 5-1. SUMMARY OF RESULTS FROM INITIAL SCREENING OF PROPOSED GASIFICATION PROCESSES (Concluded)

V. CONCEPTUAL PROCESSES (Continued)	Accepted for Evaluation as Projected to Commercial Scale	Excessive Fuel Costs	Not Amenable to Pressurization	Lump Fuels Only	Limited to Non-caking Fuels	Excessive Maintenance Cost	Excessive Equipment Cost	Excessive Operating Cost	Low Capacity	Remarks
Synthesis Gas Using Oxygen										
55. Multi-stage Conveyor							x			
56. Fixed-bed Super-pressure	x									
57. Fluidized-bed Super-pressure	x									
58. Two-stage Super-pressure Entrained	x									
59. Mayland Pebble-bed			x			x		x		Operating difficulties
60. Koppers-Totzek Pressurized	x									
61. Rummel Single-shaft Pressurized	x									
62. Rummel Modified Single-shaft Pressurized	x									
63. Gas Council Fluidized-bed	x									Similar to Process 21
64. Maccormac-Rummel Double-shaft		x	x			x	x	x	x	Small units, low efficiency
65. Catalytic Steam Methanation	x									

The selection of synthesis gas processes for which yield and throughput data are available involved a straightforward estimate of cost, and a comparison with the Lurgi costs. In some cases, raw materials consumption alone was sufficient to determine whether or not a process merited further evaluation. In other cases, complete investment cost comparisons or considerations of the complexity of a process were necessary to show that a process did not merit further evaluation. In still other cases of pilot-plant or conceptual processes, similarities with either commercially developed processes or processes used in semi-commercial plants were so great that these conceptual or pilot-plant processes could be grouped with their commercial counterparts. In some cases, the requirement of a specific fuel limited the applicability of the process.

Similarly, comparisons were made for those processes producing fuel gases. Raw material, utility, and investment data for commercial systems were compared with various pilot-plant and conceptual processes.

Selection of processes for further study was possible, sometimes only on the basis of raw material requirements, and sometimes only on the basis of complete comparisons of gas costs.

As an aid in estimating cost data for auxiliary equipment and services required by the various gasification processes, a series of graphs was derived. They were used to determine directly the investment cost of oxygen, steam, and oxygen compression systems. They were also used to arrive at part of the plant steam balance, using curves for methanation steam production, shift steam requirements, and turbine steam rates for mechanical drive turbines with various steam inlet and back pressure conditions. These graphs are presented in Appendix 5.1, together with a brief discussion.

Discussions of individual gasification processes are given below with respect to their merit for evaluation as projected to full-scale commercial operation.

A. Review of Processes Eliminated

Certain processes were outside the scope of the present study. They included the CO_2 acceptor process, the molten salt bath process, and the Hydrogasification Process, as well as any adaptation of nuclear heat for the gasification of coal. Studies by others have already indicated that these processes may become economical if research and development on them proves successful.

Additional processes were eliminated as a group on the basis of method of heating or on the basis of process equipment design, while still others were eliminated on the basis of individual consideration.

1. Processes Eliminated on Basis of Method of Heating: The effects of the methods of heating have been included as a major factor in the evaluation of feasible commercial-scale processes. Methods of heating investigated included: (a) internal combustion, (b) internal circulation of gases, (c) internal circulation of solids, (d) internal electric heating, (e) cyclic operation, and (f) other.

a. Internal Combustion: The accepted method for supplying heat in coal gasification processes using air or oxygen to make producer or synthesis gas is by internal combustion. The processes which are being investigated in detail under the present contract use this principle.

b. Internal Circulation of Gases: The heat of reaction is supplied by internal circulation of gases in various proposed processes, including the Pintsch Hillebrand (Process 13) (7) and the Koppers-Cowper processes (8) which have been used for the gasification of brown coal. During the trip of the survey group to Europe, the two companies that have built such plants confirmed the high investment cost and small unit capacity which makes these processes obsolete.

The same applies to the Wintershall-Schmalfeldt process (9) which uses pulverized, suspended brown coal. A later development by Panindco (Process 15) (10) is based on this experience, but the principle was partially abaondoned--oxygen or air was introduced for the autothermal supply of heat.

All of these processes operate at atmospheric pressure. Operation at elevated pressure has been suggested (11), but so far, no experimental development has begun. The use of nuclear energy for this purpose is outside the scope of this contract.

c. Internal Circulation of Solids: Fluidized fuel beds form the basis of certain gasification processes and provide internal circulation of the solids. The Bamag-Winkler (Process 7) and the Hydrocarbon Research (Process 21) processes have been selected for evaluation in the present study. These processes use oxygen to supply the heat requirements and use the fluidized bed as a gas-solids contacting device rather than a specific means of supplying the heat of reaction.

Attempts to use circulating solids as a heat carrier material for the gasification of fine coal without oxygen have led to numerous patents. Two processes have been tested in pilot-size plants, and one process has been used for only the devolatilization, that is, partial gasification of coal.

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(7). "Ullmanns Encyklopaedie der technischen Chemie," Vol 10, 3rd Ed, Munich: Urban and Schwarzenberg, 1958. p 435.

(8) Anon., "Report on the petroleum and synthetic oil industry of Germany," BIOS, Overall Rept. No. 1, 25 (1947).

(9) Ibid, p 24.

(10) Foch, P., "The gasification of powdered fuels by the 'Panindco' process," Chim. Ind. 66, 639-47 (1951).

Foch, P. and Loison, R., "The gasification of pulverized coal by the Panindco process--Recent experiments," Intern. Conf. Complete Gasification Mined Coal, Liege, 1954. pp 224-34.

(11) Domann, F., "The circulating-gas process at atmospheric and higher pressures," Gas- Wasserfach 91 (13), 161-4 (1950).

d. Internal Electric Heating: For the carbonization of coal, internal electric heating has been proposed and experimentally used. A more recent proposal (12) could be suitable for the gasification of coal.

Electric heating, however, in spite of very favorable assumptions, has a higher sum of comparable cost elements than the Lurgi process. (See Appendix 5.2.) To attain the 32.5 cents per MM Btu raw material cost of the Lurgi process by electric heating, a power cost in the range of 0.29 cent to 0.36 cent per kwh is required.

A further study is required to establish whether or not the lower CO_2 content of the electrically generated gas could be a sufficient compensating factor to overcome the disadvantage of high raw material and operating cost.

Recently, electric resistance heating has been applied successfully to fluidized-bed reactors operated at 2300 C (4100 F).(13) The fluidized-bed reactor is expected to be suitable for the gasification of pulverized coal. Power consumption will be essentially as given for the fixed-bed reactor discussed above since there will be a similar gas exit temperature. The high temperature as well as the high heat transfer rates in the fluidized bed will not materially change the economic picture indicated for the electrically heated fixed-bed unit.

e. Cyclic Operation: The cyclic processes available for the gasification of coal are: (a) the water-gas process, and (b) the Flesch-Winkler process.

The water-gas process has been used for the production of synthesis gas using lump coke. The process is limited to operation at atmospheric pressure. Coal has been used instead of coke for the manufacture of gas but its use necessitates a great reduction in throughput. Manufacturers (Humphreys & Glasgow and Demag) as well as users (BASF) of equipment for this process state that equipment, fuel, and maintenance cost for this process are high in comparison with other processes. Therefore, no further study has been made, nor is it indicated.

Flesch-Winkler type processes are not limited to lump fuels, but are, however, limited to operation at atmospheric pressure. Test results by Demag in a pilot plant with a grate of about 10 sq ft have been satisfactory.

Operation of a commercial plant has shown--as discussion with Demag, the licensee, has revealed--that uniform flow in the much larger bed could not be maintained. This has led to overheating of the grate, nonuniform and unsatisfactory fuel gas composition, and higher coal consumption.

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(12) Jensen, O. J., "A new electric process for the carbonization of non-coking bituminous coal," J. Inst. Fuel 23 (129), 54-5 (1950).

(13) Anon., "Electrically heated fluidized-bed reactor," Chem. Eng. News 42 (45), 68-9 (1964).

Goldberger, W. M., Hanivay, J. E., Jr., and Langston, B. G., "The electrothermal fluidized bed," Chem. Eng. Progr. 61 (2), 63-7 (1965).

Discussions with BASF, who own some of the patents and sponsored earlier work done by Flesch, have indicated that problems of ash or slag agglomeration and removal exist, that the expected equipment and maintenance costs are high and that, therefore, development work on the process was terminated.

Since compression of the gas is needed, the gas cost structure will be similar to that of the ICI Moving Burden process (Process 26) and no further consideration of this process is suggested.

2. Processes Eliminated on Basis of Process Equipment Evaluation: In addition to systems using fixed, fluidized, or moving beds, various types of process equipment proposed for coal gasification have been considered and evaluated. They include:

Herreshoff furnace
Rotary kiln
Ball mill
Pug mill
Multi-stage conveyor system (Cochran)
Pebble heater (Phillips-Staber)
Multiple bed gasifier (Cameron and Jones)
Traveling grate coker

The first four items of equipment have one property in common; they are suitable for the contacting or mixing of different solids or liquids, whereby contact with the gas phase takes place primarily at the surface of the bed of solids, and the gases do not normally pass through the bed. Also, in this equipment, a gas velocity is maintained which minimizes dust carry-over. In the case of the Herreshoff and the rotary kiln, the solids are continuously turned over to attain contact of new parts of the solid with the gas phase.

Herreshoff furnaces are presently employed, among other uses, to partially gasify coal, leaving activated carbon as a residual product of the gasification. Nichols Engineering, manufacturer of this equipment, have stated that a 22 foot, 3 inch diameter unit, the largest in this service, will gasify (burn-off) up to 1 pound of coal per hour per square foot of hearth area. The unit has 16 hearths, for a total usable area of 4,000 sq ft, which equals about 50 tons of coal gasified per day.

A comparable gasification unit, presently in commercial operation on bituminous coal to produce usable fuel gas, is the Wellman-Galusha gas producer. One of these producers, 10 feet in diameter, will gasify 85 tons per day. This producer would cost $145,000 erected, including a building. The Nichols-Herreshoff furnace would cost $200,000 erected, not including a building, for only 50 tons per day of gasification capacity. Nichols Engineering cannot give fuel efficiency or utility data for the Nichols-Herreshoff furnace gasifying bituminous coal. However, considering the physical construction of the two units under comparison, no advantages could be hoped for in the Nichols-Herreshoff furnace for gas capacity, efficiency, labor requirements, or maintenance. In addition, considerable problems are expected in designing and operating a Herreshoff type furnace at elevated pressure. Therefore, the overwhelming advantage in investment costs ($1,700 erected cost per ton of coal gasified for Wellman-Galusha versus more than $4,000 erected cost per ton of coal gasified for Nichols-Herreshoff) leads to the conclusion that furnaces of this type merit no further consideration as commercial gasifiers. It may also be mentioned that in modern, large installations for the roasting of pyrites, the Herreshoff furnace is being replaced by processes using fluidized beds.

Rotary kilns, as a rule, are used for the countercurrent heating of solids. To achieve high heating efficiency, kilns of great length are used, which leads to a long residence time of the gases. For a rotary kiln of 12 feet in diameter and 500 feet long, a throughput (14) of 500 tons per day of limestone is possible. From the fuel requirements for this service, a residence time of over 100 seconds is obtained for the flue gas. By comparison, gas residence time in gasification processes used at present are of the order of 1 second. Thus, the use of rotary kilns for the gasification of coal, especially at elevated pressure, appears uneconomical.

Ball mills and pug mills also appear uneconomical as contacting devices in coal gasification when they are considered in the same manner as the Herreshoff furnace.

The multi-stage conveyor system similarly leads to a gasifier of uneconomical size. However, this system incorporates a principle that may be used with advantage as one step in other gasification processes. The volatile material obtained in the uppermost stage is subjected to further treatment in the middle stage. Such treatment should convert tar into gas and thus eliminate the costly recovery of tar which is necessary in countercurrent fixed-bed gasifiers using coal.

The pebble heater is designed for the heating of gases to very high temperatures. High efficiency is obtained due to effective countercurrent heat exchange. These characteristics make it appear suitable for use in coal gasification. However, discussions with Babcock and Wilcox, Alliance, Ohio, the manufacturer of pebble heaters, and with Ruhrgas AG, Essen, Germany, have shown that this device is not suitable for the gasification of coal.

Babcock and Wilcox have stated that a pebble heater cannot be heated with coal because of coal dust distribution problems and spalling of pebbles due to interaction with coal ash. They see no possibility of designing a pebble heater for operation at elevated pressure.

Ruhrgas have stated that during the initial stages of the development of the LR Process for the devolatilization of coal in a fluidized bed of char, a moving bed of 3/8 irch alumina pebbles was used as the heat transfer medium. Ruhrgas confirms that for the heating of the pebble bed an ash free fuel, that is gas, is required. When gasifying bituminous coal, caking of the bed could not be avoided. The alumina pebble bed may possibly be suitable for the treating of anthracite, and preferably for more reactive fuels such as peat or lignite. In view of the problems encountered with the pebble bed, its use was abandoned and the process development concluded successfully using a bed of fluidized char as the heat carrier.

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(14) "Chemical Engineers' Handbook," 4th Ed, Perry, J. H.; ed., New York: McGraw-Hill, 1963. pp 20-4.

It appears significant that a similar development trend took place in the petroleum industry; development of the "continuous contact coking process," which used a pebble bed for the coking of residual oils was abandoned, while fluidized coking became a commercial process.

The multiple-bed gasifier as proposed by Cameron and Jones is limited to the use of lump coke or, more generally, non-caking lump fuels of high strength. The price of coke from conventional slot-type or beehive ovens makes the use of this gasifier unattractive. Also, the non-caking requirement eliminates the possibility of direct gasification of Pittsburgh seam bituminous coal in this type of gasifier.

Traveling grate stokers have been used for some time to produce coke suitable for chemical reduction purposes. In these, the volatile matter of the coal is partially burned with air to supply the heat required to obtain the coking temperature, and is thereby converted into a gas that is a mixture of producer gas and gaseous or volatile coal carbonization products. In existing plants this gas is either used for lime burning or for the generation of steam.

The plant of Chemcoke, Inc., now in operation at Columbia, Tennessee, was visited on the survey; the process was discussed together with unpublished results of laboratory experiments and data published by others.(15) Traveling grate stokers for this purpose were also discussed with the Riley Stoker Corporation in Worcester, Massachusetts.

Application of these processes was considered for: (a) coal feed preparation for fixed-bed gasification processes, such as water gas or the multiple bed process, and (b) production of fuel gas for local use by gasifying only the volatile matter of the coal and using the coke for other purposes.

Chemical coke of excellent quality is being obtained. Furthermore, coke of large size and strength can be obtained. The suitability of this coke for blast furnace use is being determined. This will also indicate whether or not this coke will be suitable for the use in water gas sets or the multiple bed process.

For an estimate of the cost of producing coke by this method, only the coking section of the plant will be considered on the following basis:

Coal throughput		325,000 tons/year
Coke production (56.5%)		183,500 tons/year
Cost of coking plant	$2,490,000	
Cost of steam plant	2,125,000	
Total	$4,615,000	

To obtain the cost of the coke, the credit for the gas must be established. For this, the cost of Wellman-Galusha gas from coal at $4 per ton or 15 cents per MM Btu will be used. Thus, for a coal with 27 MM Btu per ton, 52 percent,

(15) Grace, R. J. and Doherty, T. D., "Continuous coke production on a water cooled grate stoker," Presented at Annual Meeting, AIME, New York, 1956.

or 14 MM Btu, will be in the coke, and, assuming 90 percent gasification efficiency, 13 x 0.9 = 11.7 MM Btu per ton of coal will be in the gas. Now, according to Wellman Engineering, the cost of hot raw producer gas is 21.6 cents per MM Btu calculated as follows:

	Cents/MM Btu
Coal, 1.11 MM Btu (90% efficiency)	15.7
Operating cost, including 15% for investment	5.9
	21.6

The cost of coke on a yearly basis can be calculated together with its unit value as follows:

	Cost of Coke, Dollars per Year
Coal; 325,000 ton at $4/ton	1,300,000
Operating cost, including fixed cost assumed at 30 percent of coking plant cost	750,000
	2,050,000
Credit for gas at 21.6 cents per MM Btu, 325,000 x 11.7 x 0.216	(820,000)
Coke cost, (169,000 tons)	$1,230,000/year
Coke cost	7.28/ton
Coke cost	27.5 cents/MM Btu

This estimated value of 27.5 cents per MM Btu for coke corresponds to an 83 percent increase in the Btu cost of coke compared with the cost of the starting coal.

Conversely, it is evident that use of traveling grate stokers for the second purpose mentioned above would lead to a much higher gas price than that obtained from conventional producers if the coke were credited at the Btu price of coal.

Conversion of coal into coke by direct heating has been demonstrated in two types of circular ovens built by Salem-Brosius, Inc., Carnegie, Pennsylvania. The investment cost of an oven for the carbonization of non-coking Western coal was given as $200,000 for a unit producing 4 to 7 tons of coke per hour. The possibility of producing gas from coke made by these processes has not been explored. Therefore, evaluation has to await the time when such data become available and when these processes have shown the capability of producing coke of a size and quality suitable for use in units such as the Cameron and Jones multiple bed gasifier.

In summary, the traveling grate stoker is suitable for the production of coke and gas. It will be economical for the purpose of fuel gas production, if the coke produced can be sold at a value that is considerably above the Btu cost of the starting coal. The further development in this field will be observed, since major improvements in coke quality and investment cost reductions could lead to a process of interest to the present program.

A tentative flow sheet for one possible integrated gas-power generating plant is given as Figure 5-1. The basic unit could be a traveling grate stoker, rotary oven, or a LR Process unit.

3. Processes Eliminated on Basis of Individual Process Evaluation: Various individual processes were eliminated from further consideration as projected to full-scale commercial operation in the present studies, only after they had been evaluated with regard to process raw material requirements, operating difficulties, or potential for development to large-scale operation at elevated pressures.

a. ICI Moving Burden Process (Process 26): The ICI Moving Burden Process (16) gasifies coal by means of steam in a fluidized bed. Heat to the bed is supplied by withdrawal of char (high ash residue) from the bed and recycling it via a heating zone. In the heating zone, the char is reheated by being partially burned with air. After separation from the flue gas, the hot char is returned to the fluidized gasification bed. Satisfactory operation of a 3.5 ft ID gasifier at atmospheric pressure has been reported. The process has a low gas production rate per cross-sectional area and a high dust emission because of attrition caused by the high char recycle rate.

Compression of the make gas to an elevated pressure such as that used in the Lurgi process is costly. Operation of this process at elevated pressure is possible only in theory. Discussions at the Stoke Orchard Laboratory of the National Coal Board, as well as with Ruhrgas in Essen, have confirmed that the amount of air needed as lift gas would increase with increases in operating pressure. This would, in turn, cause an unbalance in the process; that is, there would be a production of heat by char combustion greater than that required. Recycle of the products of combustion as lift gas does not appear practical in view of their dust content and temperature. Utilization of the excess flue gas at elevated pressure in gas turbines would require dust removal and would lead to the production of excess power.

Process data for a large-scale ICI Moving Burden plant are given in Table 5-2. For comparison purposes, the corresponding costs for the Lurgi process (17) are given in Table 5-3. The cost for producing 1 MM Btu of gas at 450 psi is 1.4 cents lower for the Lurgi process.

There are additional advantages of the Lurgi process, such as higher methane content of the gas (which means lower heating value loss in a smaller methanation plant), larger unit size, and simpler gasifier plant. Also, the char recycle rate in this process is stated to be 40-80 lb per pound of coal

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(16) See Process 26, Table 3-1, and Appendix 3.5.

(17) See Process 11, Table 3-1, and Appendix 3.5.

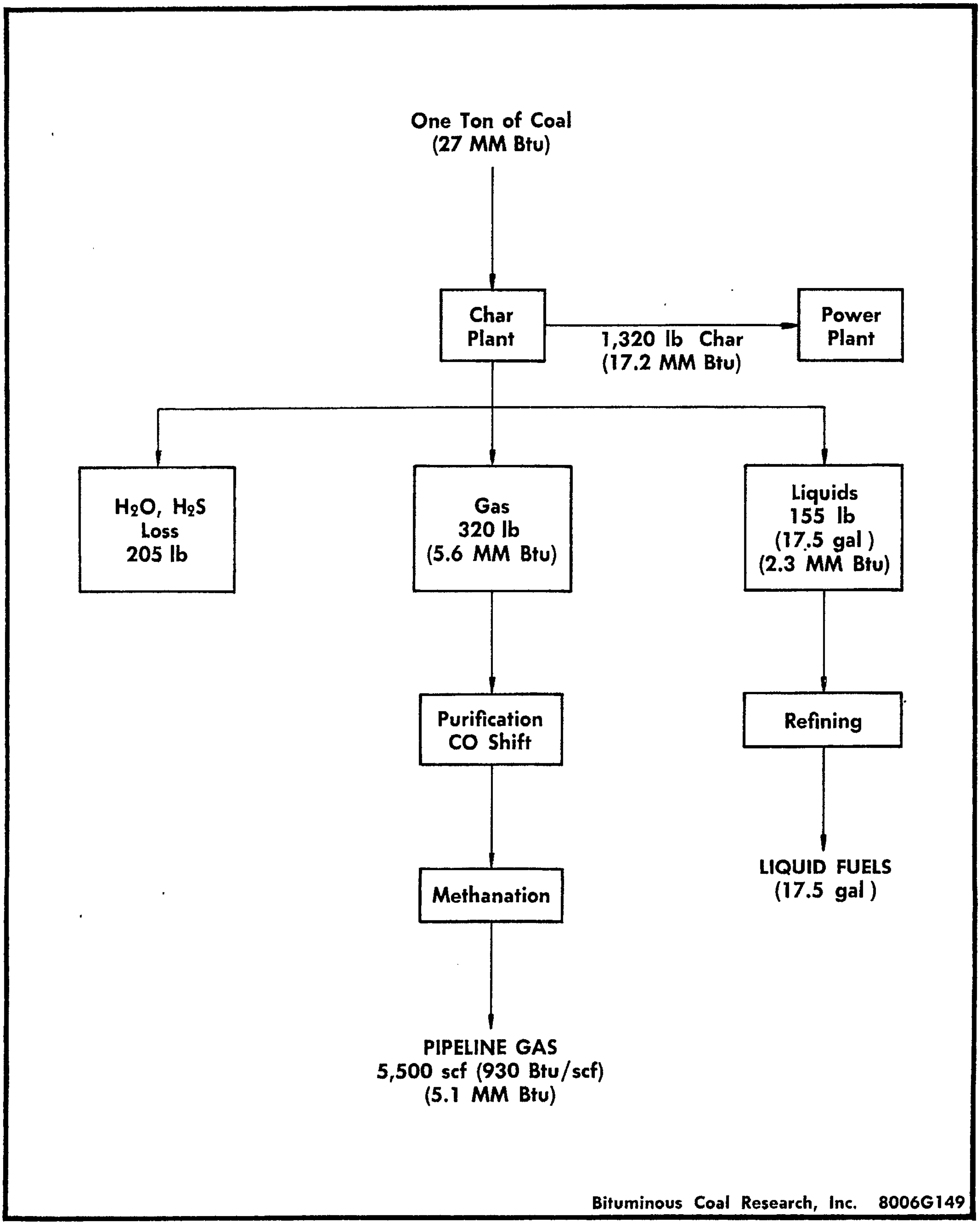

Figure 5-1 Simplified Process Scheme for Integrated Gas-power Generating System

TABLE 5-2. PROCESS DATA AND COSTS FOR
ICI MOVING BURDEN GASIFICATION PLANT

	kg/1,000 Ncu m $CO + H_2$	lb/M scf
Process Requirements		
Coal (6,300 kcal/kg) (11,340 Btu/lb)	880	10.6
Steam (15 psi)	42	0.5
Steam Credit (300 psi)*	2,600	31.1

Process Costs	Cents/MM Btu in Gas
Coal: 1.85 MM Btu at 15.4¢/MM Btu	28.5
Steam Credit: 400 lb equivalent to 0.48 MM Btu of coal at 15.4¢/MM Btu	7.4
Net Costs	21.1
Compression Costs	
Compression of 3,640 scf gas containing 12.5 percent CO_2 and 2 percent N_2 to 450 psi at 3.6¢/Mcf**	13.1
TOTAL COST OF COMPRESSED GAS	34.2

* The steam is generated in waste heat boilers using combustion, and therefore in this context, only the heat content can be credited.

** See cost data for auxiliary equipment and services for use in scoring gasification processes, Appendix 5.1.

TABLE 5-3. PROCESS REQUIREMENTS* AND COST FOR LURGI DRY-ASH GASIFICATION PLANT

Item	Unit Cost	Amount	Cents/MM Btu in Gas
Coal	15.4¢/MM Btu	1.15 MM Btu	17.8
Oxygen	$5/ton	39.5 lb	9.8
Steam	35¢/M lb	149 lb	5.2
TOTAL COST OF COMPRESSED GAS			32.8

* Heating value of tar-free raw gas 12,120,000 Btu per 100 lb coal.

feed at a temperature of about 1900 F. By comparison, the dolomite recycle rate in the CO_2 Acceptor process (18) is estimated at about 2.5 lb per pound of coal feed.

On the basis of all these facts, the moving burden process is being eliminated from further consideration.

b. LR Process (Process 36): The LR Process is pilot-plant process using solid heat carriers.(19) At first, countercurrent operation in a pebble bed was studied. Later, the char itself was used as a heat carrier in a fluidized bed. Use of the pebble bed was limited to partial gasification of non-caking fuels and required gas as fuel for the heating of the pebble bed. Higher rank coals caused caking of the pebble bed. Non-caking bituminous coal could only be partially gasified, that is, essentially only devolatilized due to temperature limitations caused by reaction of the pebbles with the coal ash.

The fluidized-bed process using recycle char as the heat carrier was developed by pilot-plant operation. Two commercial LR Process units were erected in Yugoslavia by Lurgi. According to cost data based on German conditions for a two-unit LR Process plant, 25.5 percent of the heating value of the coal is converted into a gas of 460 Btu per scf with the remainder available as char for use as power plant fuel. (See Tables 5-3 and 5-4.) Each of the two units would produce 2600 MM Btu per day of gas; this is about the size of a fuel gas plant or about 1 percent of the size of the high Btu pipeline gas plant visualized in the present process evaluations. The net cost of LR gas

(18) See Process 47, Table 3-2, and Appendix 3.5.

(19) See Process 36, Table 3.1, and Appendix 3.5.

of 57.5 cents per MM Btu is higher than the cost of producer gas. Thus, the LR Process is not competitive for the production of local fuel gas unless a very high premium could be attached to its lower nitrogen content, or the char utilized at a premium price.

TABLE 5-4. PROCESS DATA AND COST FOR TWO-UNIT LR PROCESS PLANT

Coal Consumption, tpd	800
Gas Production, MM scfd	11.2
MM Btu/day	5,200
Gas Heating Value, Btu/scf	460
Boiler Fuel Production, MM Btu/day	15,000
Plant Investment	DM 14.5 million ($3,625,000)

For the conversion of LR gas into high Btu pipeline gas, compression, purification, and methanation would be required. For an LR Process plant producing 250,000 MM Btu per day of gas, a power plant of 3,500,000 kw capacity would be required to utilize the char. In view of this and the high fuel gas cost, the LR Process is not being included among the processes deserving further consideration in the present process evaluations. (See Table 5-5.)

c. BCR-Kaiser Gas Producer (Process 38): From 1949 to 1954, Bituminous Coal Research, Inc., supported a program for the development of a dual flow producer at Battelle Memorial Institute. Lump coal fed to the top of the gasifier; after being devolatilized, it descended the gasifier shaft together with air, the gasifying medium. In the downdraft zone, liquid and carbonization products were converted into gas, which together with the producer gas, was withdrawn near the center of the gasifier shaft. The coke descended further in the gasifier shaft and was gasified in a countercurrent, that is upflow, stream of air and steam.

TABLE 5-5. GAS PRODUCTION COSTS FOR
LR PROCESS PLANT OPERATING 7,200 HR/YR

Charges	Cents per MM Btu
Coal ($4 per ton)	62.0
Utilities, Materials, etc.	7.0
Labor (German rates)	8.5
Capital Connected Costs at 15 Percent	35.0
Total Charges	112.5
Credits	
Char and Boiler Gas (coal equivalent)	35.4
Filter Cake (wet char)	7.6
Tar and Oils (8 cents per gal)	12.0
Total Credits	55.0
NET COST OF GAS	57.5

The process was tested in a pilot-plant producer (20) with a coal devolatilizing grate area of 4.5 sq ft and a downdraft section of 3 sq ft cross section. The updraft section contained a Wellman-Galusha type rotary grate.

The manufacture of a tar-free gas was possible. The lack of control of the size consist and size distribution of the char led to irregularities in the downdraft section of the gasifier. This did not permit the design throughput being achieved.

In view of these problems, the BCR-Kaiser producer is considered not to be competitive with commercially-available gasifiers for either synthesis or producer gas production. One principle of this gasifier, namely, the demonstration that complete gasification of liquid coal carbonization products in the presence of carbon but the absence of oxygen is possible, is of value to the development of the two-stage super-pressure gasifier.

(20) See Process 38, Table 3-1, and Appendix 3.5.

d. Great Northern Railway Gas Producer (Process 41): The Great Northern Railway gasification process was developed, at first, to produce a reducing gas for the beneficiation of iron ore.(21) As described in U.S. Patent 3,110,578, this is an entrained coal gasification process using induced turbulence in the reaction zone to improve contact between fuel and gasification medium. The patent limits the amount of air to be used for the gasification process below that theoretically required for complete combustion and describes and claims operation at substantially atmospheric pressure. The use of preheated air leads to a 25 percent increase in efficiency and an increase in the $CO + H_2$ content in the gas from about 22 to 26 percent at the optimum air/fuel ratio.

A comparison is given in Table 5-6 of the analyses of producer gas made from lignite, subbituminous coal, and bituminous coal, by this process and of producer gas from bituminous coal obtained in the Ruhrgas vortex gasifier.(22)

TABLE 5-6. COMPARISON OF COMPOSITION OF GAS FROM GREAT NORTHERN RAILWAY AND RUHRGAS VORTEX GAS PRODUCER

Gas Composition, %	Great Northern Railway Lignite	Great Northern Railway Subbituminous Coal	Great Northern Railway Bituminous Coal	Ruhrgas Vortex Bituminous Coal
Carbon Dioxide	14.7	12.9	9.5	5.1
Carbon Monoxide	11.8	12.8	11.5	22.8
Hydrogen	8.2	8.2	3.7	8.0
Methane	0.1	0.2	0.0	0.0
Nitrogen	65.2	65.9	75.3	64.1
Gross Heating Value, Btu/scf	65	69	50	98

(21) See Process 41, Table 3-1, and Appendix 3.5.

(22) See Process 35, Table 3-1, and Appendix 3.5.

The better efficiency of the Ruhrgas process as indicated by the higher heating value of the gas is no doubt, at least, to a great extent, due to the high air preheat temperature of 1300 F and the use of a much larger gasifier with lower heat losses. This points out the difficulty in comparing data obtained in equipment differing in size. However, it can be concluded that the Great Northern Railway process does not have any apparent advantages over other commercial scale entrained coal gasification processes.

e. Inland Steel Processes (Process 17): The Inland Steel Co. has proposed both a two-stage and a single-stage atmospheric pressure suspended steam-oxygen-coal gasification process for the production of synthesis gas.(23)

A high percentage of carbon conversion is claimed for both process variations but the specific oxygen consumption is higher in each than it is for a commercial unit such as the Koppers-Totzek gasifier. The specific oxygen consumptions are quoted 35.2 to 37.9 lb/M scf ($CO + H_2$) for the single stage and 37.6 to 39.3 lb/M scf ($CO + H_2$) for the two-stage system. These are significantly higher than the specific oxygen consumption of 31.4 lb/M scf ($CO + H_2$) of the commercial Koppers-Totzek unit.

Based on the cost of the gas produced, the atmospheric pressure Koppers-Totzek process is not considered well-suited for producing pipeline gas and estimates for pressurized operation do not improve the costs substantially. It was, therefore, concluded that a commercialized Inland Steel gasifier would probably not be significantly better than a Koppers-Totzek; hence, it was not selected for detailed evaluation.

U.S. Patents 3,002,736 and 2,989,297 assigned to Inland Steel are concerned with a combination of iron melting and coal gasification processes. Since the metallurgical part of the total process will determine the amount and quality of the gas, and the timing of gas production, this is not considered an independent coal gasification process for evaluation on the present program.

f. Stookey Gas Producer (Process 48): From information supplied by Thermal Processes, Inc., the Stookey Gas Producer is conceptual in design derived from or connected with the experience gained in oil gasification procedures for water-gas production.(24) The gas producer contains two fuel beds which can be used in a similar manner as the two fuel beds of the BASF-Flesch-Demag type processes, although fluidization is not mentioned by Stookey. In view of this, the area of application, as well as the economy of the process, is considered to be identical with the BASF-Flesch-Demag processes.

The Flesch process was discussed during the survey trip with Demag, the inventor and builder. Demag has built a small commercial plant for the production of fuel gas. This plant, in contrast to the pilot plant, had operating problems caused mainly by uneven gas distribution. The process was said not to be adaptable for pressure operation, and thus is ruled out for the manufacture of synthesis gas suitable for conversion into pipeline gas. The

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(23) See Process 17, Table 3-1, and Appendix 3.5.

(24) See Process 48, Table 3-2, and Appendix 3.5.

costs given for the manufacture of fuel gas for local use were higher than those for a standard producer. In view of this, the Stookey conceptual producer does not warrant further consideration at this time.

g. Cameron and Jones Gas Producer (Process 51): The Cameron and Jones process is one that continuously produces water gas by use of moving beds of lump fuel that are alternately blown with steam and air.(25)

The process requires lump fuel of good mechanical strength and is limited to operation at or near atmospheric pressure. Therefore, it does not appear suitable as a candidate process for pipeline gas manufacture. The study of the process is incomplete, but will be continued to determine whether there are areas of application of the process which appear promising.

B. Processes Accepted for Further Evaluation

Processes were finally selected for evaluation as projected to full-scale commercial operation primarily on the basis of information available on pilot-scale or commercial-scale operations. The following commercial synthesis gas processes were selected for full-scale evaluation as used at present or by using data extrapolated by manufacturers to operation at 450 psi:

Lurgi Dry-ash Gasifier (Process 11)
Lurgi Slagging Gasifier (Processes 18, 19, 20)
Hydrocarbon Research Gasifier (Process 21)
Bamag-Winkler Atmospheric Gasifier (Process 7)
Rummel Single-shaft Pressurized Gasifier (Process 61)
Rummel Modified Single-shaft Pressurized Gasifier (Process 62)
Koppers-Totzek Pressurized Gasifier (Process 60)
Texaco Gasifier (Process 22)

The Winkler process for which large atmospheric pressure units can be built was evaluated on that basis to obtain a cost comparison between operation at atmospheric and elevated pressure.

In view of the advantages of operation at 450 psi, a further increase in operating pressure to 1050 psi, the pressure needed for pipeline transportation, was expected to offer a further cost reduction. Therefore, data for operation at 1050 psi were extrapolated for the following processes:

Fixed-bed Super-pressure Gasifier (Process 56)
Fluidized-bed Super-pressure Gasifier (Process 57)

For the manufacture of fuel gas for local use, the following two available commercial processes were selected for full-scale evaluation:

Wellman-Galusha Gas Producer (Process 32)
IFE Two-stage Gas Producer (Process 33)

In addition, certain gasification systems were selected on the basis of studies made on the project, namely:

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(25) See Process 51, Table 3-2, and Appendix 3.5.

Two-stage Super-pressure Entrained Gasifier (Process 58)
Catalytic Steam Methanation Gasifier (Process 65)
Airblown Lurgi Gas Producer (Process 37)
Two-stage Fluidized Super-pressure Gas Producer (Process 46)

Further data for these processes follow.

1. Two-stage Super-pressure Entrained Gasifier (Process 58): For the gasification of coal with oxygen to produce synthesis gas, three contacting principles are known. They are a fixed fuel bed, a fluidized fuel bed, and an entrained or suspended fuel bed.

Table 5-7 gives coal and oxygen consumption data for these types of processes.

In addition to the differences in process requirements, these processes have other typical advantages and disadvantages. The main advantages of gasification processes using coal in suspension are that they can handle any type of coal, use simple equipment, and can accommodate large units for operation at high pressure. The main disadvantage is their high oxygen consumption. However, a substantial reduction in oxygen consumption should result from proposed modifications and improvements in the entrained gasification of coal. This appears possible by converting the conventional one-stage gasification process into a two-stage process with separate zones for reaction with oxygen and with steam.

TABLE 5-7. OXYGEN AND COAL CONSUMPTION FOR PRINCIPAL TYPES OF GASIFICATION PROCESSES

Process Requirement/MM Btu in Cold Synthesis Gas	Fixed-Bed	Fluidized-Bed	Entrained
Process Coal, MM Btu	1.1-1.3	1.35-1.6	1.2-1.5
Oxygen, lb	41-45	44-80	68-98

In the first stage of such a two-stage gasifier, coal or char removed from the product gas would be gasified with oxygen under slagging conditions. Fresh coal and steam would then be injected into the hot-gas stream leaving this first stage and entering the second stage. Here, in the second stage, the coal would be partially gasified with steam; the char residue would be carried out of the second stage in the product gas but would be separated from the gas and recycled back into the first stage for complete gasification.(26)

In support of such a gasification scheme, the following can be derived from data published for the gasification of anthracite and bituminous coal in a pilot plant at atmospheric pressure.(27) The oxygen consumption was 120-140 lb per MM Btu in the synthesis gas from anthracite as compared to 94-107 lb when using high volatile A bituminous coal. In bituminous coal with about 38 percent volatile matter, about 47 percent of the heating value resides in the volatile matter. Using the oxygen consumption figures for anthracite for the gasification of the char from this bituminous coal in the first stage and assuming complete gasification of its volatile matter in the second stage by means of the hot gases from the first stage, an oxygen consumption of 55-65 lb per MM Btu in the total gas is obtained. This is about 60 percent of the oxygen consumption of 94-107 lb given for the gasification of this bituminous coal in the same one-stage pilot plant. By using this 60 percent figure to convert the oxygen consumption for entrained gasification in Table 5-7 to that for a two-stage entrained gasification plant, an oxygen consumption of 41-59 lb per MM Btu in the synthesis gas is obtained.

If this reduction in oxygen consumption can be verified experimentally, the proposed two-stage entrained process will have a considerable investment and operating cost advantage over fixed- or fluidized-bed processes.

The proposed two-stage procedure would avoid the partial combustion of the readily gasifiable volatile matter in coal which occurs in conventional one-stage suspension gasification. It would permit complete utilization of the volatile matter for the production of gas, rather than for supplying heat by reaction with oxygen. It would also make possible a high concentration of coal in the second stage. This, together with the use of a high pressure of 1000 psi, would permit a high rate of conversion of carbon into carbon monoxide and hydrogen by reaction with steam. In addition, the quenching effect of coal and steam in the second stage would lead to a comparatively low gas exit temperature; at the high pressure this would, in turn, permit a higher concentration of methane. Also, since methane formation can be considered an exothermic reaction, a further reduction in the oxygen consumption would be possible.

The increase in methane concentration when using elevated pressure is a well-established fact shown by many commercial Lurgi plants. That the increase in methane concentration and decrease in oxygen consumption continues at pressures above those used commercially has been concluded by the inventors of the Lurgi process.(28) Figure 5-2 shows their predicted data. The Lurgi process pressure of 25-30 atm was originally selected since it produced from Central German brown coal a gas of about 500 Btu per scf and thus conforming to

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(26) See Process 58, Table 3-2, and Appendix 3.5.

(27) Strimbeck, G. R. et al, "Gasification of pulverized coal at atmospheric pressure," U.S. Bur. Mines, Rept. Invest. 5559 (1960).

(28) Danulat, F., "Reaction between gas and fuel in pressure gasification," Gas- Wasserfach 85, 557-62 (1942).

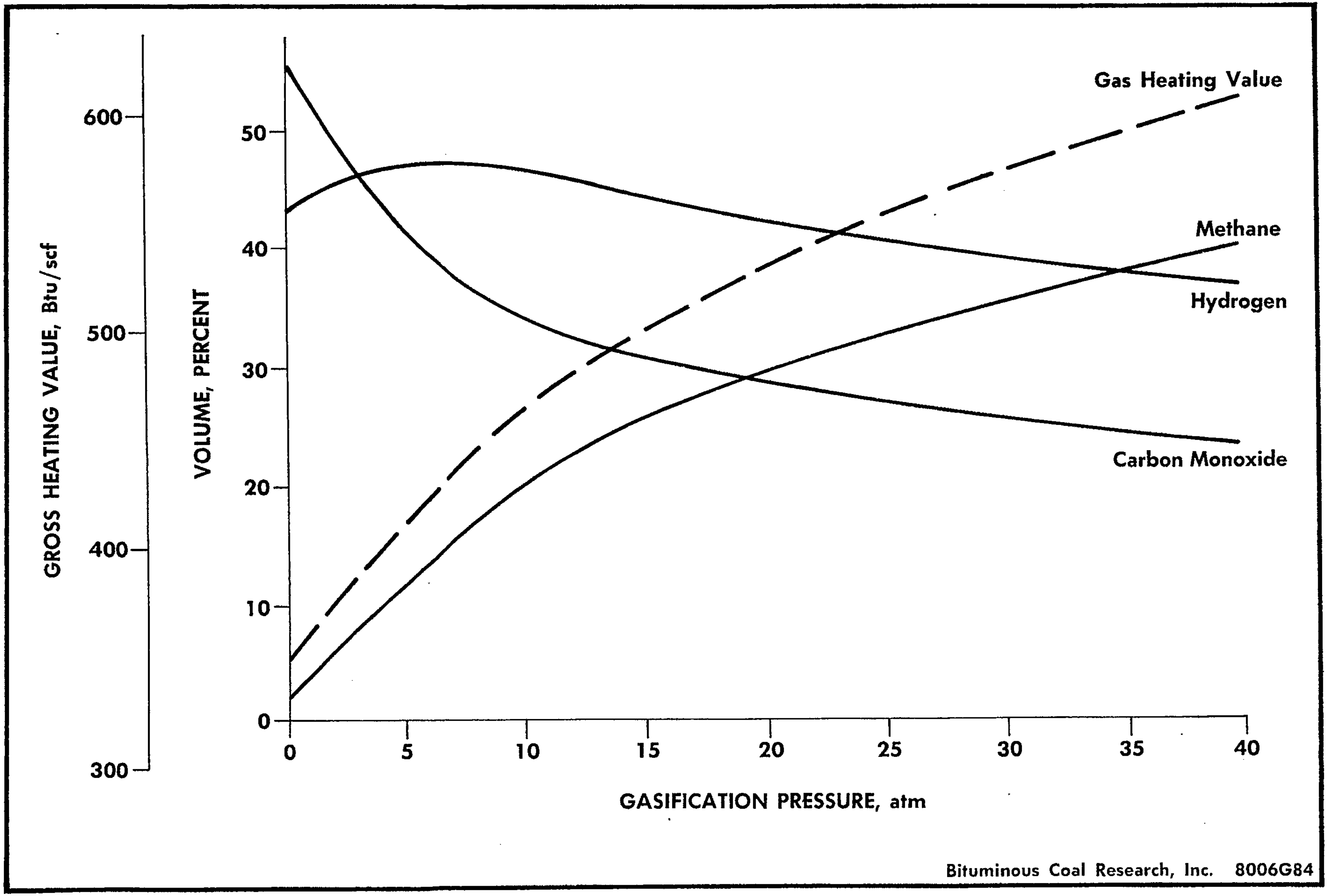

Figure 5-2 Theoretical Dependence of Pure Gas Composition and Gross Heating Value upon Pressure

town gas specification. This pressure also was used in later plants. At this time a higher pressure appears more economical for the production of pipeline gas in the 1000 Btu range.

Thermal annealing has been offered by Ergun (29) as an explanation for the decrease in the rate of combustion of carbon filaments above 1200 C (2192 F). In conventional entrained gasifiers using oxygen, such temperatures are exceeded in the initial reaction zone where slagging occurs; the coke or char leaving this zone can, therefore, be expected to have lesser reactivity in comparison to that produced from coal in the proposed two-stage gasification process. In this process, fresh coal and steam are injected into the gases issuing from the first stage and are mixed with them before excessive temperatures that may cause annealing are reached.

An additional advantage of this two-stage gasification process is that difficulties, due to vaporization of slag and the sticking of slag as the temperature of high slag viscosity is approached, will be avoided by the rapid cooling of the slag below its melting point by the injection of coal and steam into the second gasification zone.

Data for the first stage are available from the literature on gasification of char and anthracite with oxygen. However, data are not available for the yield and composition of gas to be produced by heating coal to the temperatures and pressures existing in the second stage. However, certain assumptions have been made concerning the reactions taking place in the second stage and data have been calculated. (See Appendix 5.3.)

For the kinetics of the reaction of the volatile matter, little information is available. The reaction has been investigated (30) in the presence of a large excess of hydrogen at a pressure of 1500 psi and higher. Very high reaction rates compared with the carbon-steam reaction were found. The reaction rate of the volatile matter of bituminous coal was much higher than the reaction rate of the fixed carbon of the coal or of char. This supports the assumption that rapid gasification of volatile matter occurs in the second stage.

For the amount of methane formed in the second stage, various assumptions can be made. As shown in Appendix 5.3., an estimate can be made using the concept of carbon "activity." However, experimental investigations for different residence times, various pressures, and coals of different rank are needed.

From the overall material balance for the system, incorporating all the indicated improvements, a major reduction in cost of gas is indicated and further evaluation of the two-stage gasifier extrapolated to commercial scale is desirable. For these evaluations, carbon activities of 1, 2, and 3.4 are recommended.

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(29) Ergun, S., "Kinetics of the reactions of carbon dioxide and steam with carbon," U.S. Bur. Mines, Bull. 598 (1962).

(30) Feldkirchner, H. L. and Linden, H. R., "Reactivity of coals in high-pressure gasification with hydrogen and steam," Ind. Eng. Chem., Process Design Develop. 2 (2), 153-162 (1963).

2. Catalytic Steam Methanation Gasifier (Process 65) (31): Present coal gasification processes for the commercial production of synthesis gas, or pipeline gas, require oxygen to supply the heat needed at high temperatures for the highly endothermic reaction between carbon and steam leading to carbon monoxide and hydrogen.

The gasification reaction,

$$2C + 2H_2O = CH_4 + CO_2,$$

is only slightly endothermic; thus its utilization without the use of oxygen for the generation of heat should be possible. The equilibrium constants for this reaction indicate that favorable yields should be obtainable at lower temperatures and higher pressures than used in present coal gasification processes. Favorable yields are indicated in the range of 1100 to 1300 F, at pressures between 1000 and 3000 psi or more. Higher pressure and lower temperature favor methane formation. (See Appendix 5.4.)

However, satisfactory reaction rates in the indicated temperature range will only be achieved if effective catalysts for the reaction can be found. The literature indicates that a number of catalysts appear suitable for this reaction; the alkali carbonates are mentioned as being the most active.

In view of the potential simplication in process design which catalytic steam-methanation might afford, evaluation of this conceptual process as projected to full-scale commercial operation is considered justified.

3. Two-stage Fluidized Super-pressure Gas Producer (Process 46): The steam-iron process for the production of hydrogen has been improved by the use of moving beds of iron oxide rather than the fixed beds of the old established cyclic process.(32) An estimate for the cost of hydrogen from this process as developed by the U.S. Bureau of Mines has been published.(33) For use as a hydrogen source for the coal hydrogasification process, further reduction in

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(31) See Process 65, Table 3-2, and Appendix 3.5.

(32) Feldkirchner, H. L. and Huebler, J., "Reaction of coal with steam-hydrogen mixtures at high temperatures and pressures," Am. Chem. Soc., Fuel Div. Preprints 8 (1), 160-83 (1964).

Gasior, S. T. et al, "Production of synthesis gas and hydrogen by the steam-iron process: Pilot plant study of fluidized and free-falling beds," U.S. Bur. Mines, Rept. Invest. 5911 (1961).

Wen, C. Y., "A kinetic study of coal hydrogasification - The rapid initial reaction," Am. Chem. Soc., Fuel Div. Preprints 8 (1), 147-59 (1964).

(33) Katell, S. and Faber, J. H., "What hydrogen from coal costs," Hydrocarbon Process. Petrol. Refiner 43 (3), 143-6 (1964).

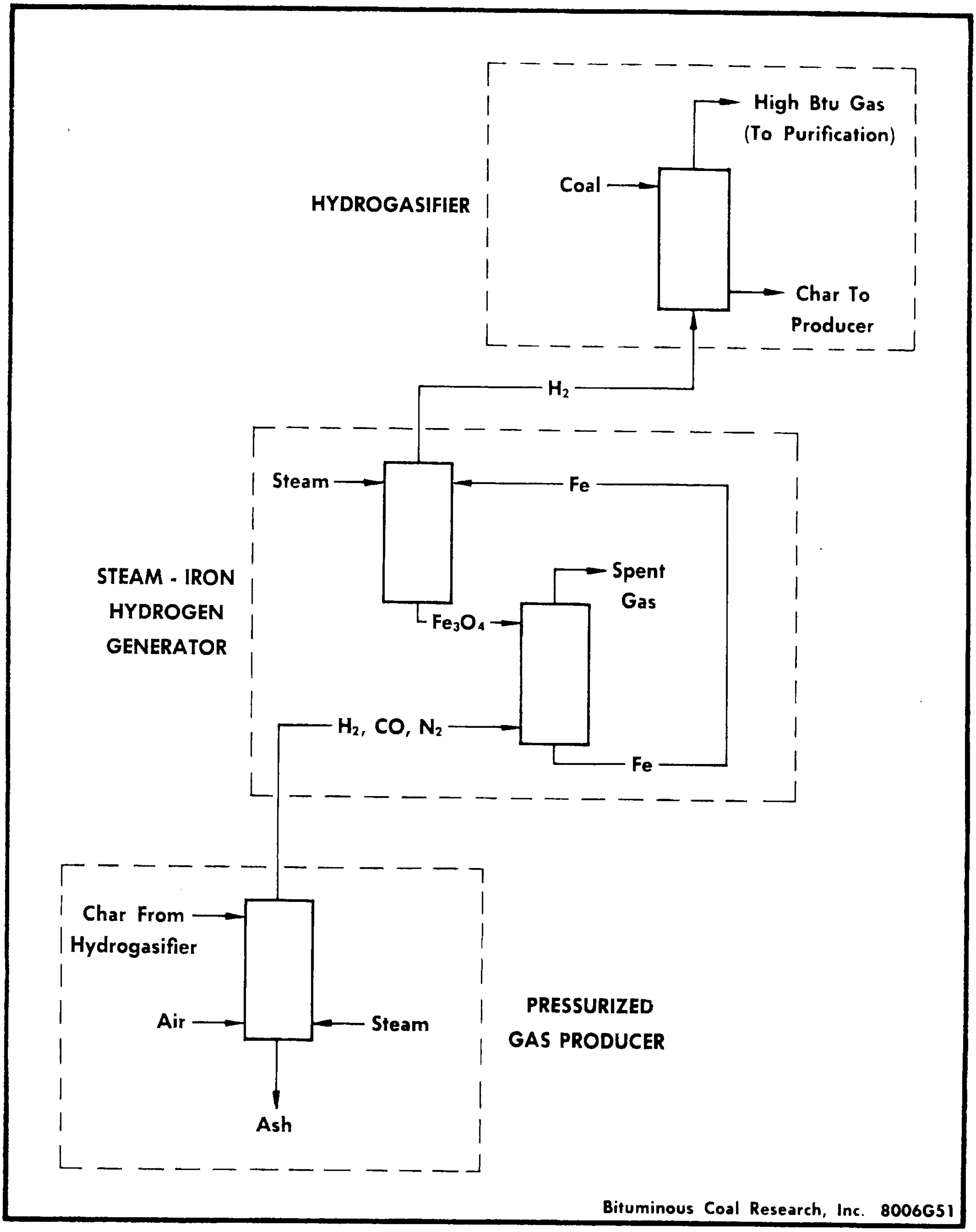

Figure 5-3 High Btu Gas Plant Based on Pressurized Gas Producer

cost of the combined processes (Figure 5-3) can be achieved by the following (34):

(a) Operation of the steam-iron process at hydrogasification pressure,

(b) Manufacture of producer gas for use as reducing gas without cooling at the same pressure, and

(c) Use of the char from hydrogasification as raw material for manufacture of the producer gas.

It is further known that for the reduction of the iron oxide in the steam iron process, a high ratio of ($CO + H_2$) to ($CO_2 + H_2O$) is required. Conversely, the reduction of iron oxide ceases when this ratio drops below a certain level.

A high ($CO + H_2$) to ($CO_2 + H_2O$) ratio in the producer gas can be achieved by cooling and scrubbing of the gas to remove CO_2. However, this is a costly step requiring high temperature heat exchangers if the gas must be reheated. To avoid it, a gasification process must be selected that leads to a gas with a low CO_2 and H_2O content. Studies have indicated that it will be possible to produce such a gas from hydrogasification char in a two-stage process operating at 1500 psi and similar in principle to the two-stage super-pressure synthesis gas process.(35)

The first stage is the slagging gasification of recycle char from the second stage in a gasification system similar to that used by Ruhrgas in their atmospheric pressure vortex gasifier using highly preheated air. Data for this step of the process are available (36) although extrapolation to high pressure is necessary. The hot gases from this stage supply the heat required for the gasification of hydrogasification char in the second stage with superheated steam.

In the proposed synthesis gas super-pressure process, only the volatile matter of the coal is gasified in the first stage and it is indicated that entrained gasification of fresh coal will be satisfactory. In the producer gas production, char with a low volatile matter content must be used and, therefore, a greater reaction between steam and fixed carbon is reached. Extrapolation of

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(34) Benson, H. E., "Process and cost considerations in making substitute natural gas from coal," Presented at Am. Gas Assoc. Operating Section, Transmission Conference, <u>1963</u>. CEP-63-10.

(35) See Process 46, Table 3-2, and Appendix 3.5.

(36) Nistler, F., "The Ruhrgas vortex gas producers," Coke Gas <u>19</u> (2), 54-7 (1957).

available steam-carbon reaction rate data at atmospheric pressure (37) to operation at a total pressure of 1500 psi indicates that this is possible at high temperatures in a fluidized bed. These data indicate that at 2000 F and 70 percent steam decomposition, about 5 pounds of carbon will be gasified per pound carbon present in the reactor in the form of coke. Operation at 1500 psi may increase this and further increase may be obtained with char of high reactivity. If entrained gasification were used with 20 seconds residence time, only about 1 pound carbon is present at any time in the gasifier for 50 pounds carbon to be gasified per hour. (See material and heat balances in Appendix 5.5.) It is thus evident that a fluidized bed with a carbon inventory about 10 times greater than that present in entrained gasification will be needed to reach sufficient carbon conversion and steam decomposition.

Experiments will be needed to obtain design data needed for the second stage, just as was the case for the two-stage super-pressure entrained gasifier for the production of synthesis gas.

A fluidized bed temperature of about 1800 to 2000 F and possibly higher is indicated not only to obtain a high steam decomposition and ($CO + H_2$) to ($CO_2 + H_2O$) ratio, but also to obtain a low methane concentration. Methane is not utilized for the iron oxide reduction. For the calculation of the methane content of the gas in the second stage, a carbon "activity" of 1, that is, that of graphite, has been used in view of the high reaction temperature. If experiments should show that the carbon has a higher activity resulting in a higher methane content, the yield of utilizable producer gas would decrease. Use of still higher temperature in the second stage would reduce the methane concentration, which is favorable; however, at the same time, the overall gasification efficiency would fall off. Experimental data must be awaited to obtain the information needed for an optimization of the process.

Production of a gas suitable for use in the steam-iron process thus appears promising and evaluation of the process as projected for full scale commercial operation is indicated.

Gasification data consistent with thermodynamic equilibrium and heat balance data have been calculated and are also presented in Appendix 5.5. They were used to obtain the amount of char which would be consumed in the producer gas section of a 250 MM scf per day pipeline gas plant, on the assumption that 1.0 mole of hydrogen is needed per mole CH_4 formed and that the producer gas

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(37) Dotson, J. M., Holden, J. H., and Koehler, W. A., "Rate of steam-carbon reaction by a falling particle method," Ind. Eng. Chem. 49 (1), 148-54 (1957).

Ergun, S., "Kinetics of the reactions of carbon dioxide and steam with coke," U.S. Bur. Mines, Bull. 598 (1962).

May, W. G., Mueller, R. H., and Sweetser, S. B., "Carbon-steam reaction kinetics from pilot plant data," Ind. Eng. Chem. 50 (9), 1289-96 (1958).

Squires, A. M., "Steam-oxygen gasification of fine sizes of coal in a fluidized bed at elevated pressure," Trans. Instn. Chem. Engrs. 39, 3-22 (1961).

will reduce iron oxide until its ($CO + H_2$) to ($CO_2 + H_2O$) ratio drops to 0.4. A reduction of the char consumption appears feasible by reducing part of the spent producer gas in the second stage of the gasifier using it in place of steam. The improvement possible by this measure will be evaluated later.

4. Air Blown Lurgi Gas Producer (Process 37): During the start-up operations of the SASOL Lurgi gasifiers, air was used.(38) Data obtained during these start-up operations indicate that the gas so produced under pressure would be a suitable fuel for the gas turbine operation. The low exit temperature for the product gas indicates that it should be possible to free the gas of dust without the need for complex equipment.

Compression of gas from a conventional air blown gas producer for use in gas turbine adds too much to the cost of cold clean fuel gas for it to be competitive with natural gas. However, a hot raw gas produced in the air blown Lurgi at a temperature of about 1100 F would be much more competitive.

Heat and material balances, together with summary evaluations, are given in Appendix 5.6. A 3.7 meter (12 ft 2 in.) ID Lurgi unit operating at 10 atm should have a capacity of 1.23 MM scfh of hot raw gas with 158 Btu/scf corresponding to a hot efficiency of 91 percent.

Thus, an evaluation of the air blown Lurgi operating at about 150 psig and as projected to full scale commercial operation appears desirable.

(38) See Process 37, Table 3-1, and Appendix 3.5.