SECTION 6. EVALUATION OF SYNTHESIS GAS PROCESSES AS PROJECTED TO FULL-SCALE COMMERCIAL OPERATION

As a basis for making final recommendations regarding processes that merit further research and development to establish their full potential as gas generating systems, those processes selected in initial evaluation in Section 4 have been projected to full-scale commercial operation and an estimate made of the cost of the final product gas.

Processes for production of synthesis gas have been evaluated for use in a plant with a capacity of 250 MM scf per day of pipeline gas; those for production of fuel gas have been evaluated for use in a plant with a capacity of 100 MM Btu per hour.

One process for the production of a gas suitable for use in gas turbines has been evaluated on the basis of a single 3.7 meter ID Lurgi gasifier unit operating in combination with a gas turbine. And, finally, one other process has been evaluated for the gasification of char to produce a high-temperature high-pressure producer gas suitable for the steam-iron generation of hydrogen for subsequent use in the production of 250 MM scf per day of high-Btu pipeline gas by the hydrogasification of coal.

A. Basis for Evaluations

The basis for cost evaluations of synthesis gas processes, as agreed upon by BCR and OCR, is as follows:

a. A pipeline gas plant with a capacity of 250 MM scf per day of gas with a gross heating value of 928 Btu per scf will be used to evaluate the individual gasification systems.

b. Coal will be charged at $4.00 per ton as mined.

c. Fixed charges will be 15 percent per year of total fixed investment.

d. Labor cost will average $2.75 per hour for operating labor, plus 10 percent for supervision, plus 60 percent of labor and supervision for payroll overhead.

e. Repair and maintenance will be charged at an average yearly rate of 4.18 percent of total fixed investment. This includes maintenance labor and materials, plus overhead and supervision of labor. The 4.18 percent rate has been developed as a representative average for the gasification systems, process auxiliaries, and utilities evaluated in this report.

f. Make-up water will be charged at 10 cents per M gallons.

g. The plant load factor will be 95 percent; this is equivalent to 347 days operation per year at full capacity.

In addition, certain other basic assumptions have been made; they are as follows:

a. Pittsburgh seam coal without need for pretreatment to reduce or eliminate caking properties is usable in all the gasification systems which have been evaluated.

b. For those processes which have not been operated at 450 or 1050 psig, no changes in the fundamental mode of gasifier operation will be required by higher pressure operation. However, it is recognized that changes in pressure will change gasifier capacity, raw material requirements, and product composition, and that certain mechanical design modifications will have to be made to allow gasification systems to operate at these elevated pressures.

c. All processes will be credited with by-product sulfur production at a rate of $20 per ton of sulfur produced. Those fixed-bed processes producing phenols and ammonia have also been credited with 4 cents per pound for the raw phenols recovered from the waste, and with $24 per ton for the ammonium sulfate made by reacting the recovered ammonia with sulfuric acid produced, in turn, from part of the by-product sulfur.

d. Working capital will be borrowed, and 6 percent annual interest on this capital will be charged to the cost of the individual processes.

In accord with these assumptions, certain other procedures have been followed in making the process evaluations; they are:

a. Pittsburgh seam coal, a high volatile A bituminous coal, with the analysis shown in Table 6-1 has been used in the preparation of all material and heat balances unless otherwise indicated.

TABLE 6-1. ANALYSIS OF HIGH VOLATILE A BITUMINOUS PITTSBURGH SEAM COAL

	As Received	Dry, Ash-free
Proximate Analysis, Percent		
Moisture	1.2	
Volatile Matter	39.3	42.9
Fixed Carbon	52.4	57.1
Ash	7.1	
Calorific Value, Btu/lb	13,990	15,270
Ultimate Analysis, Percent		
Carbon	--	84.4
Hydrogen	--	5.7
Nitrogen	--	1.6
Oxygen (by difference)	--	5.6
Sulfur	--	2.7

b. Material and heat balances, based on Pittsburgh seam coal with 4.7 percent moisture as received, have been derived, wherever possible, from information from individual manufacturers who would supply the gasifiers.

c. The capacity of individual gasifiers has been based, wherever possible, on information received from the individual gasifier manufacturers.

d. The pipeline gas plants have been based on coal as the only source of energy and are designed to be self-supporting in all energy requirements.

e. The plant sites have been assumed to be at the mouth of the coal mine. No site acquisition costs have been included in these studies.

f. River water has been assumed to be available for the circulating cooling water system, the boiler feed-water preparation plant, and the drinking water preparation plant.

g. Waste treatment has been assumed to be negligible, except in the fixed-bed gasification processes, where phenolic waste treatment is required.

h. Atmospheric pollution has been avoided by producing elemental sulfur from the hydrogen sulfide formed in gasification; credit for the sulfur so produced has been taken. However, no sulfur dioxide removal has been attempted from the stack gases of the coal-fired preheaters, or the boiler plant.

i. In some processes, credit for excess char from gasification and excess fines from coal preparation has been assumed at coal Btu price equivalent.

j. The utility systems incorporated in the pipeline gas plants have been given a simplified treatment. Thus, no attempt to optimize utility conditions or to provide complete integration of utilities has been made, and it has been assumed that: (1) a cooling water system, providing 85 F water with a 30 F temperature rise, will be available; (2) steam will be generated at 600 psig, 750 F, for large drive turbines and for process use in the 450 psig gasification systems; (3) the super-pressure (1050 psig) processes will have steam generation at 1100 psig, 750 F, for the same purposes; (4) hot lime process softened water will be used for boilers up to 600 psig; and above this pressure, demineralized water will be used.

k. No allowance has been made for start-up expenses or for costs, if any, for "debottlenecking" the equipment to achieve 95 percent load factor.

l. No costs have been developed for access roads, railroads, dock facilities, and water lines that may be required outside of the pipeline gas plant site. Cost allowances have been made for normal on site auxiliaries such as change houses, guard houses, administration and laboratory buildings, roads, railroads, fences, sanitary and storm sewers, fire protection, etc.

m. The gasification systems operating at 450 psig have all been assumed to require 2.5 percent of the raw gas for use as lock hopper gas in pressurizing lock hoppers in the coal charging system. This lock hopper gas has been recovered and utilized at atmospheric pressure as fuel gas. The super-pressure systems have been provided with CO_2 compression to make CO_2 available at 1100 psig for coal charging; this eliminates fuel gas losses in the lock hoppers.

Those gasification systems using lump coal utilize the lock hopper gravity feed system developed by Lurgi for their pressure gasifiers. Those systems using fine coal utilize the lock hopper--pneumatic conveyor combination of Dr. C. Otto & Comp.

Data presented in these preliminary evaluations are approximate. Only those data have been developed which are deemed essential for the operating and capital cost structure of the processes. Investment costs have been estimated with emphasis on the relative accuracy of costs between processes, rather than the absolute accuracy of such costs.

Thus, no attempt was made to estimate any of the processes in sufficient detail to require the development of equipment lists. However, an approximate number of certain key process equipment items was obtained in order to make ratio estimates from unit costs. This preliminary equipment list for the synthesis gas processes is presented as Table 6-2.

The only written quotations obtained were from Lurgi and from the Dr. C. Otto & Comp. All prices for "cold box," steam boiler, compressor, turbine, shift reactor, condenser, acid gas removal system, etc., were obtained verbally from representative manufacturers. A summary of these quotations is presented as Appendix 6.1.

1. Lurgi Dry-ash Gasifier (Process 11): The Lurgi Dry-ash Gasifier (38) is an established process with over 25 years of commercial operation. Some 50 or more gas generators have been built and operated under pressure on lignite, brown coal, bituminous coal, and anthracite, in Germany, South Africa, the United Kingdom, Australia, Pakistan, and Korea.

Pittsburgh seam coal, with a high swelling and caking index, has been indicated to be suitable for use without pretreatment. A test with 100 tons of Pittsburgh seam coal was sponsored by the U.S. Bureau of Mines in a commercial Lurgi generator. The ash content of the material in the generator during the test was maintained in the range of 20 to 30 weight percent by addition of ash to the feed. However, future tests of longer duration could demonstrate that such ash addition is not necessary; even if some ash recycle were found to be necessary, it would have little influence on the overall cost of gasification. Therefore, the cost of addition of ash has been neglected for the purpose of this study.

Performance data expected for Pittsburgh seam coal were supplied by the Lurgi company. These data seem to be rather conservative in comparison with data obtained with a similar coal in commercial operation at Dorsten, Germany over a period of years. Minor adjustments in the data were necessary to obtain a precise material balance; these adjustments resulted in a slightly lower gasification efficiency than that calculated from Lurgi's own information.

Although availability of the Lurgi gasifier has been demonstrated to approach 95 percent in the SASOL, South Africa plant, three spare gasifiers have been allowed for 23 gasifiers in operation; this represents only 88.5 percent availability.

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

The size of gasifiers used in this analysis has been limited to the 3.7 meter (12.3 ft) size installed by Lurgi in South Africa, since this is the largest unit for which reliable costs are available. Development of the process on a commercial scale in the past few years has led to substantial increases in capacity and savings in steam, and on this basis, Lurgi supplied an estimate of the oxygen and steam requirements in a gasifier specifically designed for Pittsburgh seam coal. According to the Lurgi estimate such a gasifier would be 5 percent more expensive than the standard Lurgi unit because of the need for additional space required for the high swelling coal and the lower density ash.

A uniform coal of 1-1/4 x 1/8 inch size has been assumed as feed to the gasifier. The throughput rate for the Pittsburgh seam bituminous coal given by Lurgi is equivalent to 370 lb per hr per sq ft of grate area. This is a quite conservative rate, and compares well with the maximum rate of 400 lb per hr per sq ft obtained at Dorsten, Germany. This conservative gasifier capacity has been justified by Lurgi on the basis that demonstration tests will be required to prove the actual capacity of the gasifier for coal of such size and with such caking and swelling characteristics.

The Lurgi Dry-ash Gasifier produces byproducts, such as tar, oil, benzene, ammonia, and phenols in the raw gas. In the present evaluations, the low fuel value byproducts have been recovered and sold for credit. These include ammonia, phenols, and sulfur. The byproducts which can be used as fuel, such as tar, oil, and crude benzene, have been sent to the steam superheaters or the steam boilers for use as fuel in generating or superheating steam. A possible credit, after recovering and refining the benzene, is roughly estimated at 0.75 cent per M scf of pipeline gas. However, by-product sales of coal, tar, oil, and benzene have not been used in this report as a matter of conservatism. The dust carried over-heat from the Lurgi gasifier is recycled to the gasifier along with the heavy tar fraction.

An additional saving in the gas costs over the cost shown in this survey could be realized, if coal with a high ash content of 20 to 30 percent were available at a lower Btu unit price than the coal on which this evaluation is based. It should, therefore, produce gas with a lower cost since the Lurgi Dry-ash Gasifier shows relatively little cost increase for a 20 to 30 percent ash coal as compared to a 7 percent ash coal.

Also, if in the future it should be established that the high swelling and caking Pittsburgh seam coal requires ash recirculation to maintain an ash content of 20 percent or more in the Lurgi gasifier, a high ash coal could then be even more attractive because no ash recirculation would be required. Obviously this consideration would not apply for coals of lower free swelling index.

2. Lurgi Slagging Gasifier (Process 18, 19, 20): The Lurgi Slagging Gasifier is not a commercially established unit. However, experimental work has been performed in Lurgi's pilot plant facilities at Herten, Germany, and The Gas Council's 3.5 ft ID slagging gasifier at Solihull, England. Also, the U.S. Bureau of Mines has experimented with slagging gasification at elevated pressure at Morgantown, West Virginia, and is presently operating a 16-5/8 in. ID fixed-bed unit at Grand Forks, North Dakota.(39)

(39) See Processes 18, 19, 20, Table 3-1, and Appendix 3.5.

TABLE 6-2. PRELIMINARY EQUIPMENT LIST FOR PROJECTED COMMERCIAL-SCALE PIPELINE GAS PLANTS
(NUMBER OF PROCESS UNITS)

	Fixed-bed Processes		Fluidized-bed Processes		Entrained Processes				Super-pressure Processes					
Type of Unit	Process 11 Lurgi Dry-ash	Processes 18,19,20 Lurgi Slagging	Process 21 Hydrocarbon Research	Process 7 Bamag-Winkler Atmospheric	Process 61 Rummel Single-shaft Pressurized	Process 62 Rummel Modified Single-shaft Pressurized	Process 60 Koppers-Totzek Pressurized	Process 22 Texaco	Process 56 Fixed-bed	Process 57 Fluidized-bed	Process 58 Two-stage R1	Process 58 Two-stage R2	Process 58 Two-stage R3.4	Process 65 Catalytic Steam Methanation
Oxygen Cold Boxes (1,000 ton/day of 98% Purity)	7	9	7	11	12	11	16	13	6	6	8	7	6	3**
Gasifiers	26	14	15	7	8	6	4	9	16	8	5	5	5	6
Shift Reactors	2	5	4	7	7	5	7	5	2	3	6	5	5	2
Acid Gas Absorbers	10	12	10	13	14	11	16	15	4	4	5	4	4	3
Methane Synthesis Reactors	7	9	7	11	12	7	12	12	6	4	7	6	5	2
Coal Preparation Plants	1	1	2	3	2	2	3	3	1	2	2	2	2	2
Steam Boilers	2	2	2	2	2	2	2	1*	2	2	2	2	2	2

*Start-up Only

**800 ton/day

A

B

Volume I

The slagging gasifier is, at the top of the coal bed, similar to the dry-ash gasifier in physical performance. Therefore, the byproducts from the slagging gasifier would be much the same as those from the dry-ash gasifier, and the exit temperatures and the limitations on the swelling and caking characteristics of coal will also be similar for the two gasifiers. This means that the need for demonstration of feeding Pittsburgh seam coal directly to a slagging gasifier is the same as to a dry-ash gasifier.

Performance data expected for Pittsburgh seam coal in a slagging gasifier were supplied by the Lurgi company. These data appear to be quite conservative when compared to the data published on The Gas Council experiments with the slagging gasifier.

A material balance made from the data supplied by Lurgi showed a significant deficiency in hydrogen in the gasifier output. The hydrogen and methane contents shown in the product gas by Lurgi were lower than those obtained by the Bureau of Mines at Grand Forks, and substantially lower than those shown by The Gas Council. The hydrogen and methane contents of the product gas were therefore raised to be more nearly in agreement with The Gas Council and Bureau of Mines data, and to permit a precise material balance to be made. This resulted in a higher gasification efficiency of 84.9 percent, compared with that calculated from the Lurgi information, namely, 76.9 percent. However, the Lurgi data were quite conservative with respect to gasifier capacity, so that the overall effect of the revisions to the Lurgi data is to present what is believed to be a more realistic evaluation of the Lurgi slagging gasifier. The data used to evaluate the slagging gasifier are still somewhat conservative.

Lurgi states a requirement for the addition of limestone for fluxing, to allow the slag from Pittsburgh seam coal to flow freely. This requirement has been complied with, along with the added requirement by Lurgi that a substantial amount of the quenched slag be recirculated to minimize the need for limestone.

The capacity of slagging Lurgi gasifiers has been assumed to be double that of the Lurgi Dry-ash Gasifier. This is, as previously stated, a conservative assumption, since throughputs for fixed-bed slagging gasifiers have been shown to be more than twice those for dry-ash units. However, in view of the large diameter of the gasifier under consideration, that is, 3.7 meters (12.2 ft), it was not deemed realistic to assume a capacity of more than twice the capacity of the dry-ash unit. Problems with injection of steam and oxygen through tuyeres, and distribution of the gases in the slagging zone, lead Lurgi to believe it is unreasonable to apply a fourfold increase in capacity to a 3.7 meter gasifier, even though such an increase was actually experienced in the 3.5 ft ID experimental gasifier at Solihull.

The availability of the slagging gasifier has been assumed to be somewhat lower than that of the dry-ash gasifier, so that one spare unit is provided for each six on-stream slagging gasifiers.

A somehat lower cost for slagging gasification could be achieved if coal were available with slag characteristics that did not require fluxing. The elimination of limestone flux would save approximately one cent per M scf of pipeline gas.

3. Hydrocarbon Research Gasifier (Process 21): The Hydrocarbon Research Gasifier utilizes a fluidized fuel bed at 170 to 245 psig. It has been developed in a 26.5 in. ID, 40 ft high reactor by Hydrocarbon Research, Inc., (HRI) at Trenton, New Jersey.(40) The development work has all been done on a non-caking feedstock, namely anthracite. The batch method of charging coal into a fluidized bed would not lend itself to the feeding of caking material. HRI used dense phase pneumatic transport of feed material into the fluidized bed, transferring a large quantity very rapidly. Such a technique would almost certainly clog the fluidized bed if the feed material had high caking and swelling properties. However, a continuous feeder could be developed for transferring caking material into the fluidized bed and injecting it in such a way that the material would be blended into the bulk of non-caking char in the fluidized bed. On the basis that such a device is possible, HRI supplied the gasification characteristics for the reaction of Pittsburgh seam coal in a fluidized bed.

These characteristics are extrapolations of data obtained at three different concentrations of carbon in fluidized beds of an Ohio coal.

The feed to the reactor is pulverized dry coal which has been preheated in the drying system in an inert gas atmosphere. The fuel bed contains only 30 percent carbon. To effectively utilize such low carbon content in a fluid bed, HRI has developed internal redistribution devices which allow the use of a very high bed, to obtain the retention time necessary to react such material.

The expected performance data for Pittsburgh seam coal supplied by HRI show substantially more methane formation than one would expect from the normal methane formation equilibrium constant for their operating temperature of 1750 F. In a discussion of this increased methane formation, Squires (41) has pointed out, "There is great technical importance in determining (methane equilibrium) ratios for continuous feed of raw coals to a fluid-bed gasifier under pressure." From tentative estimates of increased methane formation, for raw bituminous coal feed, Squires has drafted a quasi-equilibrium curve showing that the increased "activity" of carbon for methane formation at the bed temperature of 1750 F used by HRI is 3.4 times that of beta graphite.

The arithmetically exact material balance, derived from the gasification parameters furnished by HRI for Pittsburgh seam coal is in good agreement with these estimates. With 30 percent carbon in the fluidized bed, the solid material leaving the fluidized bed per 1000 lb of coal amounts to 71 lb of ash plus 30 lb of carbon. Because of the uniformity of composition in a fluidized bed, it follows that to maintain the 30 percent carbon in the burden, the entrainment separation system must maintain the net carry-over from the fluidized system to no more than this 101 lb of solid material per 1000 lb of coal. Any net increase in the amount of material carried out of the fluidized bed would represent an increase in the amount of carbon in that output, and therefore, the percent carbon in the fluidized bed; the 71 lb of ash in the output cannot increase without an increase in coal feed rate, since otherwise this is all of the ash in

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

(41) Squires, A. M., "Steam-oxygen gasification of fine sizes of coal in a fluidised bed at elevated pressure," Trans. Inst. Chem. Engrs. 39 (1), 1-26 (1961).

the feed. This points out that a fluidized bed process requires a control of the amount of the solid material removed from the bed, so as to maintain a specific percent of carbon in the bed.

The heat balance has been based upon heating the coal, steam, and oxygen to an average of 1000 F before they enter the gasifier. Considering the increased capacity for gas production in the gasifier, compared to the Lurgi gasifiers, the heat loss obtained by difference in the heat balance is in line with that expected from such gasifiers.

HRI has assumed the utilization of a device for separating the solid particles entrained in the raw product gas, and recycling these solids into the fluidized bed. Such an entrainment separator is essential to good carbon utilization as discussed above, and the design of the separator must, of course, be demonstrated in order that operation of a fluidized bed containing only 30 percent carbon can be possible.

It again must be emphasized that demonstration of the HRI process would be required to determine the suitability of the caking coal feed, the high carbon activity for methane formation with only 30 percent carbon in the bed, and the entrainment separation to minimize carry-over of dust from the fluidized bed.

In the present evaluation, a 13 ft 6 in. OD gasifier has been chosen for the Hydrocarbon Research fluidized-bed unit in accord with the best judgment of HRI. The HRI data for the rate of gasification show that 13 of these gasifiers are necessary for a 250 MM scf per day pipeline gas plant. Two spare gasifiers have been designated for the 13 units on stream; this corresponds to approximately 90 percent gasifier availability.

4. <u>Bamag-Winkler Atmospheric Pressure Gasifier (Process 7)</u>: The Bamag-Winkler process is a well-established commercial entity, having been used since the 1930's at atmospheric pressure for gasifying brown coal.(42)

Tests have been made on the Winkler gasifier using bituminous coal, and the data resulting from those tests, and commercial operation on bituminous coal, were given to us by Pintsch-Bamag as a basis for our estimating the performance of pipeline gas production using a Bamag-Winkler gasifier.

The data from Pintsch-Bamag were for atmospheric operation of a fluidized bed. Information which would allow extrapolation of these data to a pressure operation such as Hydrocarbon Research used on a fluidized bed is not available. Bamag data used for the gasification were based on our Pittsburgh seam coal with an ash softening point of 2190 F. The use of the highly caking, highly swelling Pittsburgh seam coal will require the use of a grate in the gasifier, which increases investment cost somewhat.

The maximum gasifier size given by Bamag was 33 ft ID. This gasifier has been used as a basis for the present cost analysis of the Bamag-Winkler gasifier. The maximum capacity of 2500 normal cubic meters of raw gas per hour per square meter of gasifier area is given by Bamag, and on this basis, 6 of the 33 foot diameter units are required in operation in the 250 MM cu ft per day pipeline gas

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(42) See Process 7, Table 3-1, and Appendix 3.5.

plant. Assuming 90 percent availability, one gasifier spare would be ample for the six gasifiers in operation.

The coal feed to the Bamag gasifier will be of pulverized fuel size. A substantial portion of the carbon in the coal is carried over in the entrained solids in the raw gas, so that the carbon content in the bed is approximately 70 percent. With this high a carbon content in the fluidized bed, Bamag is able to get a rapid reaction without resorting to unusual bed heights as found in the Hydrocarbon Research reactor. The carry-over, containing 70 percent carbon, is a good fuel with a heating value of roughly 20 MM Btu per ton, so that a substantial part of the carbon fed to the gasifier is available as residual fuel for sale.

Since the Bamag-Winkler gasifier is operated at atmospheric pressure, it is necessary to compress the raw gas as soon as it has been cooled, so that the subsequent processing units such as the water-gas shift, CO_2 removal, and the methane synthesis can operate at a pressure where the number and cost of these process units is lower than it would be at atmospheric pressure.

If Pintsch-Bamag were to perform further tests, data could be obtained on the pressure operation of the Winkler gasifier. However, it is not expected that such an operation would yield results better than have been given by Hydrocarbon Research, so that no economic incentive is available to make these tests.

5. Rummel Single-shaft Pressurized Gasifier (Process 61): The Rummel Single-shaft Gasifier is a commercial suspension gasifier installed at Wesseling, Germany for operation on brown coal. The Wesseling gasifier has been operated experimentally at atmospheric pressure on a Ruhr coal, and the results published.(43)

Operation on Pittsburgh seam coal has been assumed similar to that on Ruhr coal, and operation at 450 psig has also been assumed to produce the same gas as at atmospheric pressure. Because of the nature of the gasification of coal in contact with slag in the Rummel gasifier, it is believed that the gas produced under pressure at the exit temperature of 2200 F would not vary significantly from the composition of the gas produced at atmospheric pressure.

The published data are based on preliminary experimental results from Rummel and have since been questioned by the Dr. C. Otto & Comp., which is responsible for the sale of the Rummel Single-shaft Gasifier. It was stated that the preliminary data published by Rummel show too high a steam decomposition; experimental work performed late in 1963 has shown that the steam decomposition is lower than the published figure, and that the actual steam decomposition would be about 40 percent. In any case, the published data on the Rummel Single-shaft Gasifier were used as a basis for the present economic evaluation of the system operated at 450 psig and, for the present purpose, it is designated as the Rummel Single-shaft Pressurized Gasifier.

The results of the evaluation are such that even with the optimistic steam decomposition Rummel is purported to have found, the process is still

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(43) See Process 61, Table 3-1, and Appendix 3.5.

substantially more expensive for pipeline gas production than the fixed- or fluidized-bed gasifiers.

In the present evaluation, the gasifier size was chosen as 78 in. ID, the same as the Rummel Modified Single-shaft Gasifier quoted by the Otto company and described in the next chapter. It is assumed that the same raw gas production per square foot of gasifier area is found in the single-shaft gasifier as in the modified single-shaft gasifier. Since the modified gasifier produces a gas with considerably more methane, substantially less total gas volume is required than with the "unmodified" single-shaft unit; thus proportionately more of the single-shaft gasifiers are required. Seven operating gasifiers plus one spare are used, compared with five operating modified gasifiers with one spare, as quoted by the Otto company.

6. Rummel Modified Single-shaft Pressurized Gasifier (Process 62): The Rummel Modified Single-shaft Gasifier has not been operated commercially, but it was recommended by the Otto company, for the maximum production of methane from a suspension gasification system.(44) The modification involves feeding the pulverized coal into the top part of a suspension gasifier, where it is devolatilized by hot raw gas, and carried over to a separator. The devolatilized coal from the separator is then recirculated to the bottom of the gasifier and gasified in a slag bath with oxygen to produce the hot raw gas needed for devolatilizing the fresh coal in the upper part of the gasifier. This concept avoids the destruction of the methane formed during the devolatilization of coal; the fresh coal is not exposed to the extremely high temperature present in the slagging zone of the gasifier.

A similar concept for improving the efficiency of the gasification of coal in suspension was advanced by H. R. Hoy during the visit of the survey group to BCURA, Leatherhead, England.

Pittsburgh seam coal with its high caking and swelling indices would have to be fed quite carefully into the upper portion of the Rummel modified gasifier so as to avoid agglomerating and clumping, and thereby prevent the devolatilization which must occur in the upper stage. The ability to feed Pittsburgh seam coal into such a gasifier can be considered as proven by the operation of the U.S. Bureau of Mines pilot plant gasifiers. The ability of this gasifier to transfer heat to the coal and to devolatilize it rapidly, is subject to experimental demonstration.

Expected performance data for Pittsburgh seam coal as obtained from the Otto company were closely followed in the material balances that were made for the economic evaluation of the process operated at 450 psi, being designated for the present purpose as the Rummel Modified Single-shaft Pressurized Gasifier. The steam decomposition in the Rummel Modified Single-shaft Gasifier is only 40 percent, and the gasification efficiency is only 80 percent, according to the most recent data from the Otto company.

The low gasification efficiency was the primary reason for the slight improvement in economics indicated for the modified process as compared with the regular Rummel single-shaft process.

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(44) See Process 62, Table 3-2, and Appendix 3.5.

Since the modified gasifier produces a substantial amount of methane in the gasification step, it was expected to produce pipeline gas more cheaply than the regular single-shaft process. The Otto company was questioned on this point and it was indicated that steam does not enter into the reactions in the gas phase as much as the published data on the Rummel Single-shaft Gasifier indicate, and that the gasification efficiency was not as high as 88 percent. Also it was stated that the data on the Rummel modified single-shaft unit were not overly conservative, since they were the result of recent experiments with the regular Rummel Single-shaft Gasifier.

A gas exit temperature of 1650 F for the modified Rummel gasifier was indicated. At this temperature the water-gas shift reaction is still quite rapid. The equilibrium constant for the water-gas shift, as calculated from the obtained gas composition, is for a temperature of 3090 F; this is a far higher temperature than 1650 F. This indicates that it was not assumed that the water-gas shift reaction would be near equilibrium in these gasification calculations; we believe that it could be carried much closer to equilibrium. If the water-gas shift were to proceed to equilibrium at 1650 F, a substantially higher steam decomposition and hydrogen production would be realized, which in turn would allow more methane to form; the gasification economics would then be much better than the obtained data show.

The Otto company has supplied a complete documentation of the plant facilities required to produce 250 MM scf per day of pipeline gas, and have also supplied the overall cost of the gasification system for this size plant. These costs are on an erected in Germany basis and can be approximately converted to an erected in United States basis, allowing for ocean freight and duty, and higher labor costs.

The gasifier size of 78 in. ID is small compared to the sizes for the Lurgi gasifier which have been considered in the present evaluations. However, this size is obviously one that the Otto company has had experience in operating and we did not choose to deviate from it.

The quotation is for five gasifiers plus one spare to produce raw gas from 15,000 tons per day of coal. Actually the coal consumption for the Rummel Modified Single-shaft Pressurized Gasifier is such that only 4.3 gasifiers are needed in operation, but no credit has been taken for this and the gasification costs for six gasifiers have been used.

The coal feed size is not extremely critical and can be 1/8 inch to 0 size.

It must be emphasized that the data presented by the Otto company for the Rummel Modified Single-shaft Gasifier are conceptual in nature. A thorough analysis of the modified gasifier concept has been incorporated in the cost study of a two-stage gasifier operating at 1050 psig. A substantial improvement could be realized in the Rummel modified type two-stage gasifier if the upper or devolatilizing stage was a fluidized bed with separate steam injection; by this, the Squires "activity" for the hydrogen/carbon/methane reaction could be realized at 1650 F. Under these circumstances the methane formation would be such as to give a substantially more economic gasifier than the Otto company has shown in their expected performance data.

7. Koppers-Totzek Pressurized Gasifier (Process 60): The Heinrich Koppers GmbH in Essen has a well established commercial suspension gasification process using the Koppers-Totzek Gasifier at atmospheric pressure.(45) This gasification process has been installed in many plants for the utilization of various carbonaceous materials. Koppers has not had specific experience with Pittsburgh seam coal; but, because of the nature of the gasifier and its feed material, that is, a suspension of fine coal in the feed gas, Koppers expects no difficulty in gasifying such a coal.

The Pittsburgh seam coal for feeding a Koppers-Totzek Gasifier is a much higher grade material than is normally used in the gasifier. The Totzek gasifier can gasify efficiently almost any low grade fuel such as peat, lignite, or high ash materials.

Because of the high temperatures involved in the gasification reaction for the Koppers-Totzek Gasifier, there is practically no methane remaining in the raw gas. This indicates that the Koppers-Totzek Gasifier is primarily suitable for the production of carbon monoxide and hydrogen rather than the production of a gas which is 90 percent methane. Most of the installations of the Koppers-Totzek Gasifier have been for the production of synthesis gas for ammonia or methanol production.

Expected performance data for operation of the Koppers-Totzek Gasifier on Pittsburgh seam coal have been supplied by Heinrich Koppers GmbH in Essen. These data show a relatively high oxygen consumption and, as stated before, practically no methane formation. Thus, the Koppers-Totzek Gasifier becomes quite expensive as a source of pipeline gas. The data are for atmospheric operation but, because of the very high gasification temperatures, it is expected that operation under pressure would yield substantially the same gas. Koppers has begun investigations of feeding coal into a pressurized gasifier as the first step of a program of pressure gasification investigations.

In this study, it has been assumed that the standard size Koppers-Totzek Gasifier could be used at 450 psig and for present purposes it has been designated as the Koppers-Totzek Pressurized Gasifier. The gas produced would be essentially the same as that supplied by Koppers.(46) The capacity of the gasifier has been assumed to increase by ratio of the absolute pressure; thus, at 450 psig, each of the gasifiers could handle about 6000 tpd of coal. This is a very high capacity, and exceeds the present views of Koppers. However, for the production of methane under conditions of the present study, the Koppers-Totzek gasification system is expensive because of its consumption of oxygen and its lack of methane production. A more detailed gasifier cost study did not appear justified.

The method used as described above for calculating gasifier capacity shows that three Koppers-Totzek Gasifiers operating at 450 psig would be required for a 250 MM scfd plant; one spare has been added making a total of four.

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(45) See Process 60, Table 3-2, and Appendix 3.5.

(46) See Process 60, Table 3-2, and Appendix 3.5.

The coal feed to the Koppers-Totzek Gasifier is pulverized fuel and, as such, it requires preparation costs comparable to those for the fluidized-bed processes.

8. Texaco Gasifier (Process 22): The Texaco Gasifier is a pressurized unit which has been operated at Morgantown, West Virginia on a pilot plant scale using oxygen, and as a commercial scale using air.(47) More recently the process has been modified to eliminate the difficulties encountered in the Morgantown operations.

Texaco feels quite confident that the data they have quoted for the gasification of Pittsburgh seam coal is sufficiently representative to form a basis for economic analysis.

The Pittsburgh seam coal with its high swelling and softening properties will possibly be a problem in the Texaco Gasifier, because the coal is slurried with water and preheated before feeding to the gasification unit. During this preheat, the water is evaporated to form steam and the coal is entrained in the steam as a finely divided solid. Caution would have to be exercised to make sure that the preheat temperature would not be such that the coal would become plastic and agglomerate or stick to the tube walls.

Texaco has supplied a tabulation of expected performance data on Pittsburgh seam coal in the Texaco Gasifier using oxygen. The estimate is for production of 773 MM scf per day of hydrogen plus carbon monoxide; for the production of 250 MM scf per day of pipeline gas, approximately one billion cubic feet of hydrogen plus carbon monoxide is needed. Thus, the Texaco figures had to be prorated upward for the larger capacity that was required.

A material balance and a heat balance were based on the information from Texaco. The hot raw gas is quenched to 400 F as indicated by Texaco and a simple heat balance shows that the gas enters the quench at approximately 2175 F.

Texaco indicates seven gasifiers are necessary for achieving plant capacity. However, for 250 MM scf per day of gas, eight operating generators would be required, and a ninth generator has been added for a spare.

Discussions with Texaco about the preheat temperature on the coal and steam mixture and on the oxygen have revealed that, at whatever temperature these streams are heated, they must be held strictly constant during operation. For the coal/steam mixture, a temperature of 500 F has been assumed and for the oxygen going to the gasifier, a temperature of 750 F.

The Texaco Gasifier also requires pulverized-fuel sized coal feed so that it may be formed into a water slurry and pumped through the preheater to the gasifier. Because of the slurrying with water, no drying of the coal is assumed during the pulverization step.

As was stated for the Koppers-Totzek Gasifier, that also uses coal in suspension, the Texaco Gasifier is sufficiently expensive (because of the absence of methane formation and high oxygen consumption) that even without gasification

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

investment, the costs for the pipeline gas from the Texaco Gasifier are higher than the total cost, including gasification investment for processes using coal in fixed or fluidized beds. For this reason, only approximate figures were used for the Texaco Gasifier investment cost. No specific investment data were obtained from Texaco.

9. Fixed-bed Super-pressure Gasifier (Process 56) (48): The requirement for pipeline gas at 1000 psig suggests that a possible saving in investment for gasification and subsequent process equipment could be obtained, if the gas were generated at a sufficient pressure so that no gas compression would be required. A brief review of the economics of the various processes showed that a substantial potential saving in gas cost could also be realized, if the gasifiers were operated at the super pressure of 1050 psig; this is due to an increase in formation of methane.

Because the high pressure favors formation of methane in the raw gas from the gasifier, the gasification efficiency is improved, the oxygen requirements are less, and the capacity is substantially higher per gasifier, with lower heat losses and also with less heat in the form of sensible heat in the gas.

To realize these advantages, a fixed-bed super-pressure gasifier has been modeled after the Lurgi gasifier. The gas analysis for super-pressure operation was obtained by extrapolations of the gas analysis from a Lurgi unit at 30 atmospheres. Data by Danulat (49) on the effects of pressure on the formation of methane in the Lurgi gasifier were used as a basis for determining the amount of methane which would be formed at 1050 psig. The extrapolation is shown in Figure 6-1.

Since a substantial part of the methane in the raw gas from the Lurgi gasifier is formed by devolatilization of coal in the upper section of the fixed bed, an estimate of the amount of the volatile matter and its composition was made, and this portion of the raw gas was held to be independent of pressure. The remaining portion of the raw gas at 30 atmospheres was held to be subject to the Danulat equations for increased methane formation with pressure, so that the raw gas composition at 1050 psig was obtained by adding the assumed volatiles to the gas modified by the use of Danulat's data. By this method, it was determined that the amount of methane in the pipeline gas that is formed in the gasification step could be increased from 41 percent at 450 psig to approximately 47 percent at 1050 psig. This is quite a conservative increase for the fixed-bed gasification process and is well within the range of the data presented by Danulat.

The same qualifications for the use of Pittsburgh seam coal, with its high caking and swelling indices, in a fixed-bed gasifier are present for the super-pressure process as for the 450 psig process. No predictions can be made regarding the expected caking characteristics of the coal and its effect on gasifier operation at 1050 psig.

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(48) See Process 56, Table 3-2, and Appendix 3.5.

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

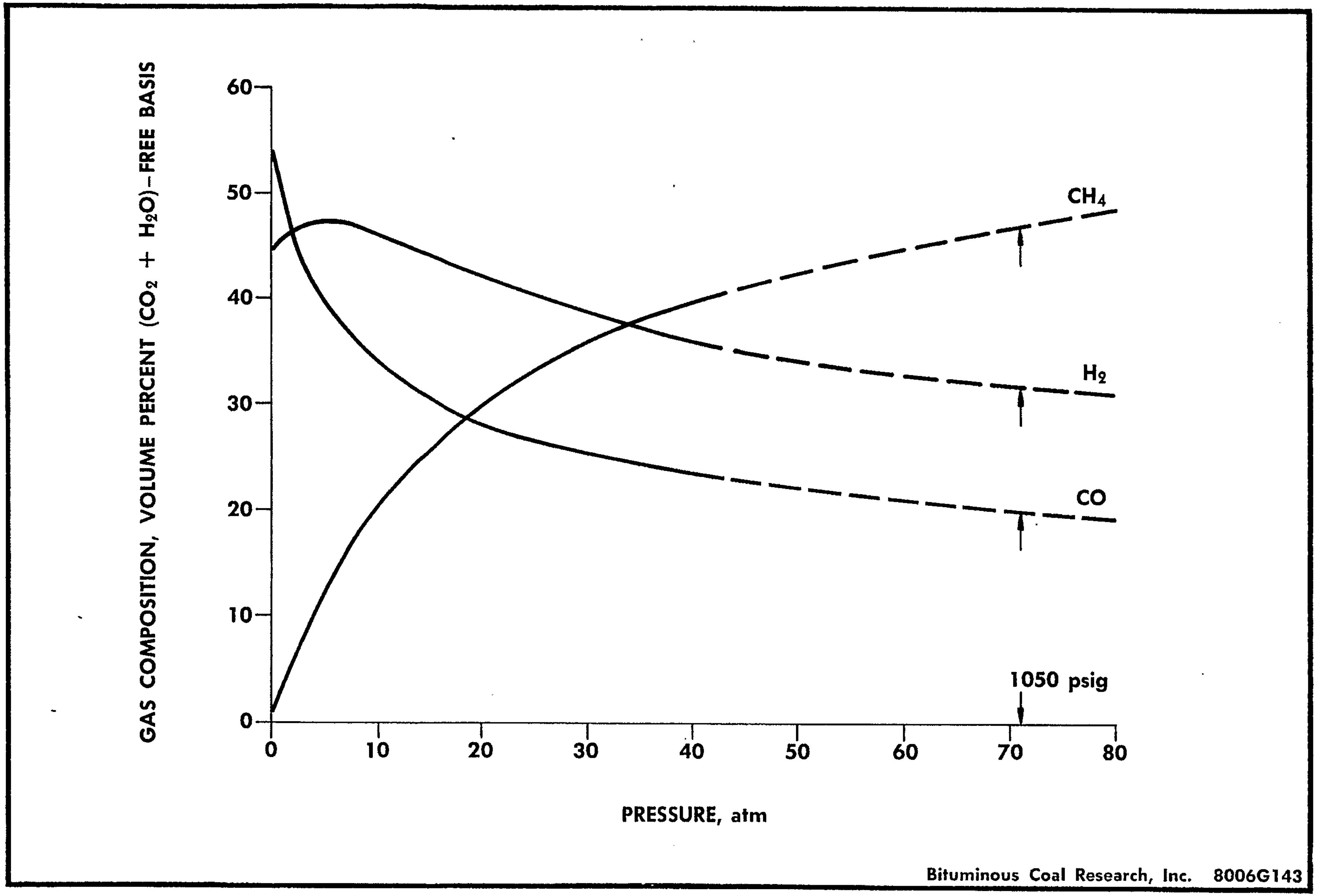

Figure 6-1 Effect of Pressure on Composition of Gases from Gasification of "Carbon" in Lurgi Gasifier as Based on Pure Gas Data of Danulat

When increasing gasifier capacity in proportion to the square root of the absolute pressure, the fixed-bed gasification at 1050 psig requires 14 gasifiers of the same external dimensions as the 450 psig units. To this, two spares have been added for a total of 16 gasifiers for operation at 1050 psig. The assumed capacity of the gasifiers at this pressure is 560 lb coal per hour per square foot of gasifier grate area. The limit of operability at 450 psig was judged by Lurgi to be 400 lb per hour per square foot. This is equivalent to 600 lb per hour per square foot at 1050 psig, so that the present design is well within this limit.

The same 1-1/4 x 1/8 inch size has been assumed for feed to the gasifiers as was used at 450 psig. The fines from the coal preparation plant will feed the boiler plant and excess fines will have to be sold.

At the 1050 psig operating pressure, the coal feeding into the fixed-bed gasifier would entail substantially greater loss of fuel gas through the lock hoppers than at 450 psig. Because of this, a carbon dioxide lock hopper pressurizing system has been included in the gasification unit so that as coal drops out of the lock hopper, carbon dioxide is automatically injected into it to maintain hopper pressure and to prevent loss of fuel gas from the gasifier. This carbon dioxide system eliminates the 2-1/2 percent gas losses assumed for lock hoppering for all of the 450 psig processes.

The fixed-bed gasifier, whether operated at 450 or 1050 psig, can handle substantially higher ash content fuels than the 7 percent ash content Pittsburgh seam coal assumed for this economic study. The higher ash content fuels presumably would be available at a lower Btu unit price and would produce pipeline gas at a more favorable cost than in the present study.

10. Fluidized-bed Super-pressure Gasifier (Process 57) (50): The same reasoning involved in assuming lower gas cost when operating a fixed-bed gasifier at 1050 psig is applicable to the fluidized-bed gasifier at similar high pressure. The high carbon activity of the Hydrocarbon Research Gasifier was used as the basis for estimating the characteristics of the fluidized-bed gasifier at 1050 psig.

Because of the uniformity of the material in a fluidized bed and the assumed method of feeding the caking coal continuously into a fluidized bed in such a way that a small part of caking material is mixed intimately with a large part of non-caking material in the bed, no distinction was made between methane formed by devolatilization, and methane formed by reaction, as had been done for the fixed-bed super-pressure gasifier.

The data that HRI furnished for gasifying Pittsburgh seam coal in their fluidized-bed unit at 1750 F, showed that the methane formation was 3.4 times as great as would have been found in equilibrium over beta graphite. This figure was based on quasi-equilibrium curve for methane in the carbon/hydrogen/methane reaction suggested by Squires for an average of bituminous and other coals. The data supplied by HRI on Ohio coal were calculated to have a similar activity of 3.9 at 1750 F. Using the same 3.4 carbon activity for methane formation at 1050 psig, as was used for the HRI data at 450 psig, it was possible to calculate a

(50) See Process 57, Table 3-2, and Appendix 3.5.

new gas composition from the gasifier based on only slightly increased gasification efficiency and also water-gas shift equilibrium at 1750 F. Because of the low bed temperature and the high activity for methane formation, it was found that the fluidized-bed process at 1050 psig would form approximately 58 percent of the methane required for the final pipeline gas in the gasifier, compared with only 43 percent preformed methane in the 450 psig Hydrocarbon Research Gasifier.

It is expected that the same limitations and qualifications on fluid-bed operation will apply at 1050 psig as applied at 450 psig; namely, the ability to feed the Pittsburgh seam coal directly into a fluidized bed, the rates and retention times for the reaction of carbon with the steam and oxygen to form gas with only 30 percent carbon content in the bed, and the ability of HRI to design the entrainment separators to limit excessive carry-over of material out of the fluidized bed.

The gasifier capacity in a fluidized bed is expected to increase in proportion to the square root of the increase in absolute pressure. In view of this and in light of the greater methane content which is expected at 1050 psig, the number of operating gasifiers required is considered to be 7 as compared with 13 that are required at 450 psig. One additional has been added as a spare, making a total of 8.

A gasifier with a 13-1/2 ft OD and a 40 ft bed height at 1050 psig would probably be fabricated in the field because of the heavy steel shell required for a unit of this size at such a pressure. However, it is entirely feasible to weld and stress relieve in the field sections of vessels of this type and size. This indicates that a cost reduction might be possible by using even larger vessels and fewer than 8 gasifiers. Such optimization is believed to be beyond the scope of this study.

Pulverized coal feed to the 1050 fluidized-bed process would require carbon dioxide pressurized feeders similar to those used for the fixed-bed super-pressure gasifiers. Preheat for coal, oxygen, and steam was maintained at an average temperature of 1000 F as was used for the 450 psig gasification system.

11. <u>Two-stage Super-pressure Entrained Gasifier R1 (Process 58) (51)</u>: The use of a two-stage gasifier may be more efficient and effective than other types of gasifiers. The economics of gasification for pipeline gas are shown to favor the operation of a gasifier at the pipeline pressure. This design involves a two-stage 1050 psig gasification process. The combination of simplicity, large capacity, and high efficiency for the suspension gasification of coal, added to the low oxygen consumption and favorable methane equilibrium at low temperatures for the fluidized bed, indicates that a combination two-stage suspension fluidized-bed gasification could be quite attractive.

In the lower, or first stage, of the proposed two-stage gasification, oxygen and steam will react with hot char removed from the raw gas, forming mostly carbon monoxide and hydrogen under slagging conditions. Into the hot stream produced from this first stage, fresh coal and steam will be injected. The easily gasified volatile matter will be converted into a gas with a high methane content in this second stage, and the highly active carbon in this zone will lead

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

to rapid reaction with hydrogen. Higher methane formation and lower gas exit temperature than in the single-stage gasifier lead to low oxygen consumption; this two-stage gasification scheme has been proposed for further experimental investigations.

Kinetic data available for the gasification of volatile matter of coal do not permit a reliable prediction of the size of the second stage, that is, residence time and carbon inventory required. It appears possible that gasification in suspension in this second stage will lead to sufficient carbon conversion and methane concentration. Experimental work is required to explore this question.

Based on the equilibrium conditions assumed to exist in a two-stage gasifier, gas compositions have been calculated for a unit using Pittsburgh seam bituminous coal. For the purpose of this evaluation, the equilibria calculated at the exit of the second stage have been calculated with beta graphite as the carbon. This would give the minimum methane content in the raw gas and would be the most conservative set of gasification parameters for the two-stage gasification.

Additional evaluations using carbon activities of 2 and 3.4 have also been made; they are presented under Two-stage Super-pressure Entrained Gasifier R2 and Two-stage Super-pressure Entrained Gasifier R3.4, respectively.

It must, of course, be demonstrated that the two-stage gasifier can operate on a direct feed of high caking, high swelling Pittsburgh seam coal. It must also be demonstrated that the heat content of the gas leaving the first stage is adequate to devolatilize the coal in the second stage, and that the residence time in the second stage is adequate to transfer the amount of heat required for the devolatilization.

To be quite conservative, a fluidized second stage has been assumed and the basic gasifier design modeled after the Rummel Modified Gasifier. Thus, the capacity of the gasifier is expected to increase in proportion to the square root of the absolute pressure from 450 psig to 1050 psig. On this basis, the number of gasifiers required for the super-pressure two-stage gasification is four in operation plus one spare, for a total of five.

12. Two-stage Super-pressure Entrained Gasifier R2 (Process 58) (52): The gasifier in this case is identical with the Two-stage Super-pressure Gasifier R1, except that in the calculation of equilibria for gases leaving the second stage, a carbon activity twice that of beta graphite has been assumed for the formation of methane from the carbon-hydrogen reaction. Also, the normal water-gas shift equilibrium has been assumed to be achieved at the 1700 F exit temperature from the second stage.

Use of a carbon activity of two results in a substantially higher methane content and a somewhat lower hydrogen and carbon dioxide content than for the beta graphite carbon activity assumed in the economic study of the gasifier R1. In that system, 35 percent of the methane required for the pipeline gas was formed in the gasifier; in the gasifier R2, 45 percent of the methane required for the pipeline gas is formed at 1000 psig in the gasifier. This is more than

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

the amount of methane formed in the Lurgi or Hydrocarbon Research gasification units at 450 psig.

This evaluation of the Two-stage Super-pressure Gasifier with a carbon activity of 2 was primarily made to obtain intermediate costs and process parameters between the carbon activity 1 and the carbon activity of 3.4 used by HRI in the data given for their fluidized-bed gasifier at 450 psig.

13. Two-stage Super-pressure Entrained Gasifier R3.4 (Process 58) (53): The gasifier in this case is identical with the Two-stage Super-pressure Gasifiers R1 and R2, except that the equilibria for gases leaving the second stage have been calculated assuming a carbon activity of 3.4 times that of beta graphite. As in the R1 and R2 cases, the normal water-gas shift equilibrium has been assumed to be achieved at the 1700 F exit temperature from the second stage.

The use of a carbon activity of 3.4 is based on the expected results of fluidized-bed gasification of Pittsburgh seam coal as used by Hydrocarbon Research, Inc., for 450 psig operation. The preformed methane in the gas from the super-pressure gasifier R3.4 is approximately 51.5 percent; this is still substantially below the figure of 58 percent formed in the fluidized-bed super-pressure process. However, the two-stage super-pressure process has the advantage of using a higher capacity gasification unit with simpler construction, and because of the slagging first stage, it has a better carbon utilization than the straight fluidized-bed reactor.

14. Catalytic Steam Methanation Gasifier (Process 65) (54): The following basis was used for an evaluation of the economics of a commercial pipeline gas plant using a catalytic steam methanation gasifier of conceptual design. Thus, not only the catalysis, but also the physical operation of such a gasifier remain to be demonstrated.

The two-stage gasifier is assumed to use a fluidized bed at 1050 psig and 1250 F. It is assumed that a catalyst will be available to convert 70 percent of the carbon in the feed during a retention time of 15 minutes. Pittsburgh seam bituminous coal and steam are fed into the middle portion of the fluidized bed, and recycle char and oxygen are fed into the lower slagging portion of the gasifier.

Results of equilibrium calculations on the fluidized gasifier show that approximately 79 percent of the methane required in the pipeline gas is formed in the gasifier. Such high methane formation leads to a very low oxygen requirement. The amount of carbon in the fluidized bed is approximately 50 percent of the total solids (including catalyst) in the bed. This carbon content was considered as that required to produce sufficient char to meet the fuel demand of the boilers and process heaters.

If five operating reactors are used to produce a sufficient amount of gas for a 250 MM scfd pipeline gas plant, the reactors would be 12 feet in diameter and 33 feet high for a carbon retention time of 15 minutes. There is no

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

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

certainty that a catalyst will be found to accomplish such a gasification reaction in 15 minutes, but for purposes of cost estimating, it has been assumed that 3 percent of the weight of the coal is added to the gasifier in the form of limestone, 1 percent of the weight of coal is added to the gasifier as iron ore, and 1/2 percent of the weight of the coal is added to the gasifier as soda ash. These three materials function as a catalyst mixture, and are carried out of the system as part of the by-product char.

B. Procedure for Evaluations

A discussion of the general procedure used in evaluating the synthesis gas processes for producing pipeline gas is given here; specific discussions of the procedure used for each of the processes is given under the individual processes.

For each process, a material balance and a heat balance were calculated for the gasifier alone to check the validity of the gasification data available, and to reduce the data from disparate sources to the common basis of 1000 lb of coal containing 1.2 percent moisture as gasifier feed. It soon became evident from the differences in gas composition that cost analysis of the gasification step alone could give misleading results. It was then necessary to include heat, energy, and equipment requirements for the total plant beginning with coal and ending with pipeline gas.

In most cases, the commercial data from the individual gasifier suppliers, and the gas analyses derived for the super-pressure processes, did not lead immediately to a precise material balance; errors of a few percent were found in carbon, hydrogen, oxygen consumption, or production. Each balance was adjusted to yield equal inputs and outputs for the various items in the balance.

Likewise, the individual heat balances were adjusted to equal input of heat to the gasifier and output of heat leaving the gasifier. The known elements of heat content leaving the gasifier, such as gas sensible heat content, gas heating value, ash sensible heat, etc., were totaled and subtracted from the total heat input to obtain the heat loss. In some cases this heat loss found by difference was inordinately small and was judged to be so because of possible inaccuracies in the method of deriving the heat and material balances. In no case does the heat loss amount to more than approximately 5 percent of the total heat input to the gasifier. All of the data used for the economic studies appear, therefore, to be well within normal industrial limits of accuracy.

After the heat and material balances were made, a simplified process scheme was drawn for each process showing the individual process steps involved in making pipeline gas from as mined coal. The generalized process scheme for all processes projected to full-scale commercial production of 250 MM scf per day of pipeline gas is shown in Figure 6-2.

These process schemes all involve similar process steps, but differ in detail. Each scheme shows the treatment of the raw gas leaving the gasifier, the generation of waste heat steam, and the removal of dust from the raw gas. The cooled, dust free gas then enters a shift converter, at a temperature of 750 F. In some cases, if the ratio of hydrogen to carbon monoxide is quite high, as in the Lurgi Dry-ash Gasifier, not all of the gas is sent to the shift converter; but part of it is allowed to by-pass the shift converter. Gas which is sent to the shift converter is reacted sufficiently to convert approximately 90 percent

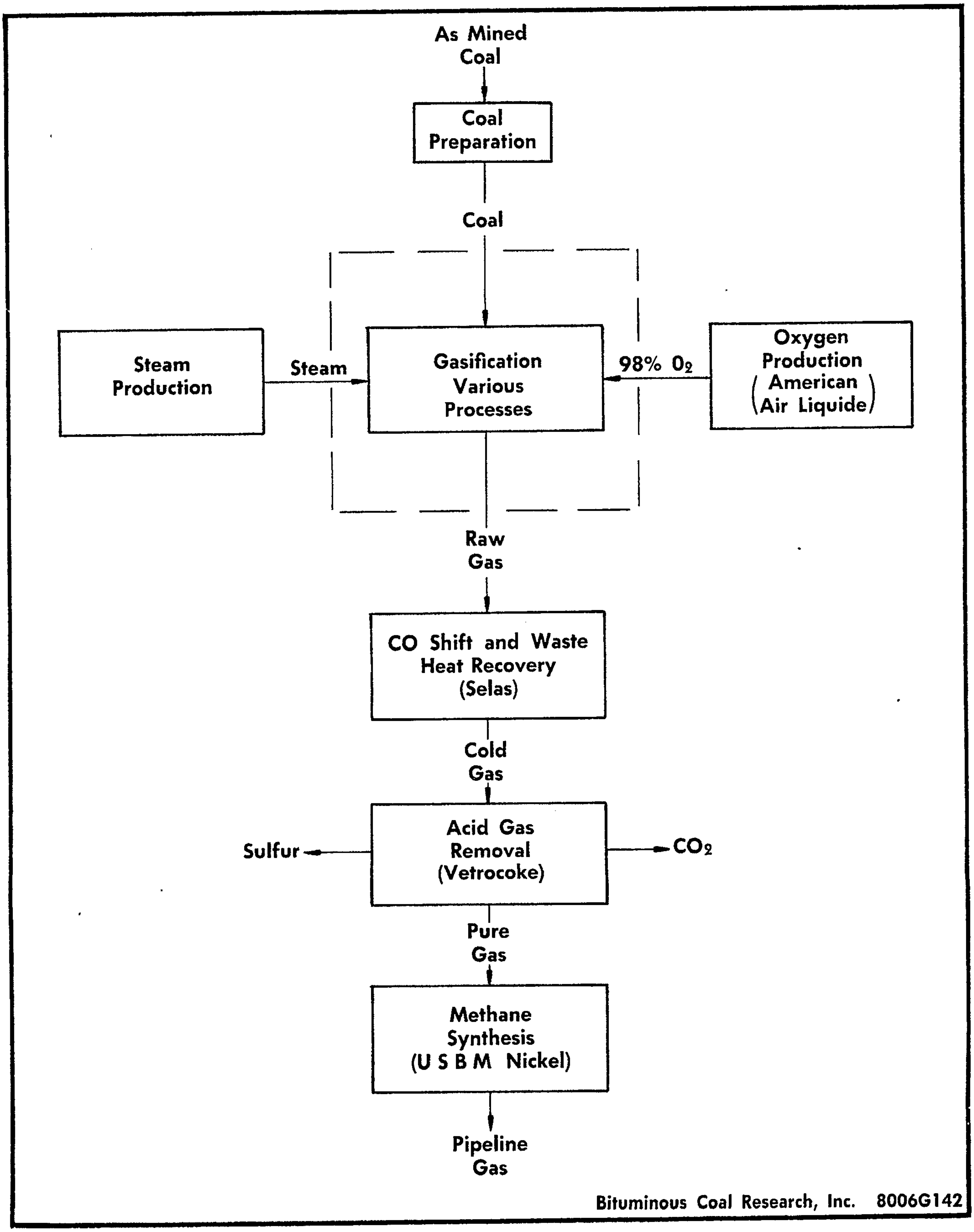

Figure 6-2 Block Diagram Showing Components of Overall Pipeline Gas Plant Used in Evaluation of Coal Gasification Processes

of the carbon monoxide to hydrogen. After the shift converter, a waste heat steam generator reduces the gas temperature to a point where cooling water can be used to further cool the gases down to a level suitable for acid gas removal.

The acid gas removal system is a two-stage Vetrocoke system in which the first stage removes hydrogen sulfide by a potassium arsenate/arsenite solution and the second stage removes carbon dioxide by an activated potassium carbonate solution.

The hydrogen sulfide is converted to sulfur directly in the regeneration system of the first-stage acid gas removal by blowing large volumes of air through the solution. Drive turbines for the airblowers require large amounts of steam, in addition to the 30 psig steam used for regenerating the carbon dioxide absorbent; this drive steam is shown separately in each process scheme.

Sulfur is obtained as a foam and is extracted and melted to produce molten sulfur for sale. The carbon dioxide from the second stage is vented to the atmosphere, except in those cases where some of it is compressed for pressurizing the coal feeding system. A final sulfur cleanup using activated carbon is necessary to reduce the sulfur content to 0.004 grain per C scf, which is the generally acknowledged amount tolerable in a methane synthesis reaction using Raney nickel catalyst. The pure gas then goes to a methanation unit, based on a Bureau of Mines process that uses a Raney nickel catalyst deposited on the outside walls of tubes cooled on the inside with dowtherm. The dowtherm in turn is used to generate steam. Dowtherm is used rather than water because at the nickel catalyst tube wall temperatures, the steam pressure necessary to provide proper cooling would be well over 2000 psig. After methanation, the gas is again cooled, first in waste heat boilers and then with cooling water. In the 450 psig processes, the cooled gas is compressed and dried before being sent to the pipeline. In the super-pressure processes the gas is sent directly through dryers to a pipeline at 1000 psig.

Based on the flow scheme in Figure 6-2, energy balances were made around each process so that the overall steam, fuel, and water requirements for each system could be evaluated. Process data for each process were summarized in tabular form showing the pertinent requirements for each process in the production of pipeline gas. As a final check on the data obtained in the material and energy balances, an overall material balance was made for the total coal, oxygen, boiler feed water makeup, etc., entering the system, and the pipeline gas, carbon dioxide, hydrogen sulfide, effluent water streams, and miscellaneous materials leaving the system.

Each of the pipeline gas plants has been charged with 20 thousand kilowatts of miscellaneous steam turbine drives to cover the approximate steam requirements for drive turbines other than those for the main turbines driving the oxygen compressors, the air compressors for the air separation plant, and the final gas compressors. In addition it has been assumed that 10 thousand kilowatts of power are required for lighting and control circuits and for miscellaneous small drives too small to use steam turbines.

The energy balances, material balances, and other process data are given in Appendix 6.2 for all processes.

1. Lurgi Dry-ash Gasifier (Process 11): The material balance for the Lurgi Dry-ash Gasifier was based on data specifically supplied by Lurgi for a fixed-bed gasifier designed for Pittsburgh seam coal. The amount of carbon in the ash, the composition of the tar, oil, phenol, and ammonia fractions, and the composition of the C_nH_m fractions were all obtained from Lurgi. It is to be noted that Lurgi generates jacket steam at gasifier pressure, and adds this jacket steam to the gasifier along with high pressure steam from an external source.

Fixed-bed processes can all be fed with coal containing the as mined moisture of 4.7 percent, so that for the material balance on the fixed-bed processes, the additional moisture in the coal above the basic 1.2 percent has been listed as 36.6 pounds of water per thousand pounds of coal. All of the sulfur in the coal is shown as being produced as hydrogen sulfide, even though in most of the processes under consideration a minor part of the sulfur is produced as organic compounds such as carbonyl sulfide.

Lurgi data showed approximately 2 percent more carbon in the output from the gasifier than in the input, so that the amount of gas produced had to be reduced to allow the carbon to balance. Once the carbon balance was established, the inlet steam was adjusted to allow a hydrogen balance, and then the amount of oxygen in the 98 percent oxygen inlet was adjusted to produce an oxygen balance. All of these adjustments were of the order of 2 percent or less of the total quantity of material being adjusted.

The heat balances were made using gross heating values for coal and for the combustible material in the product gases. The heat balances were made for a 100 F coal feed temperature, an oxygen temperature of approximately 250 F at the discharge of the compressor, and steam at 600 psig and 750 F. Jacket water is assumed to be at 225 F as it leaves the boiler feed water heater, and is charged as such in the heat input. Approximately 6 percent of the total hot raw gas heating value is for tar, oil, benzene, phenols, and ammonia. The hot gas leaving the gasifier at 1110 F is used as the outlet condition for the heat balance. The heat loss (by difference) of 2200 Btu per thousand pounds of coal, is low compared to an estimate by Lurgi for a normal heat loss of 600,000 Btu per thousand pounds of coal. However, this discrepancy amounts to only 3-1/2 percent of the total heat involved in the heat balance and is considered to be within the range of acceptable error for heat balances in industrial processes.

The simplified process scheme shows coal crushing and screening as the first step. The fines from coal crushing and screening are sent to the fired boilers, and the excess over the boiler requirement must be sold. It is estimated that this excess amount of coal fines could amount to several thousand tons per day depending on the specifications and performance of the coal crushing and screening equipment. However, no difficulty is anticipated in selling this fine coal at the price of $4 per ton.

Tar and oil from the gas cleaning system are burned in the fired boilers, as are benzene and fuel gas from the acid gas removal system. The lock hopper gas losses from the gasifier are burned to superheat the steam made in the methanation unit. All of the phenol containing effluents from the system are collected and sent to a Phenosolvan plant, and the recovered raw phenols are sold as a byproduct. The ammonia is stripped off the liquid effluents and sent to an ammonium sulfate plant where it is reacted with sulfuric acid made from a portion of the sulfur recovered in the Vetrocoke hydrogen sulfide removal unit. A biological oxidation plant is included to remove the few parts per million of phenol left in the water effluent after Phenosolvan plant treatment.

2. Lurgi Slagging Gasifier (Process 18, 19, 20): The material balance for the Lurgi Slagging Gasifier shows the addition of lime as 77 lb per M lb of coal. This lime, which would probably be added as limestone, is used for fluxing the ash so that the slag will flow freely at the operating temperature in the gasifier. Also, there is an ash recycle shown of 160 lb of slag per M lb of coal. This recycle is necessary to minimize the amount of lime added for fluxing.

When the data on operation of the slagging Lurgi were converted to a material balance, the gas quantity was adjusted to make a precise carbon balance. When a hydrogen balance was attempted, it became apparent that there was, according to the figures given us by Lurgi, an insufficient amount of hydrogen produced in the slagging gasifier.

The outlet hydrogen quantity was finally adjusted to give a precise hydrogen balance in agreement with data from The Gas Council in England and from the Bureau of Mines at Grand Forks. The oxygen balance then only required a slight reduction in the quantity of inlet oxygen.

The heat balance for the slagging gasifier was made on the same basis as that for the Lurgi Dry-ash Gasifier previously discussed, except that sensible heat is shown for the recycled slag. The heat loss shown for the Lurgi Slagging Gasifier is more nearly in line with the actual losses found by Lurgi. The heat loss in a slagging gasifier is of necessity higher than that for a dry-ash gasifier. Heat is lost to increased cooling water circulation required to keep the gasifier metal cool under slagging conditions; additional heat is lost by burning 2 percent of the raw gas under the slag taphole, to maintain a free flow of slag through the hole.

In the simplified process scheme, the coal fines from the slagging gasifier are used to fire the boiler with any excess being sold as was done in the Lurgi Dry-ash Gasifier. In the case of the Lurgi slagging pipeline gas plant, there is not sufficient low pressure by-product steam generated to provide the energy requirements for regenerating the carbon dioxide removal system solution; thus lock hopper gas, gas, benzene, and fuel gas are burned as a source of heat to reboil the Vetrocoke solutions. Since the Lurgi dry-ash system makes much more methane in the gasifier than the Lurgi Slagging Gasifier does, the carbon monoxide shift converter for the Lurgi slagging process is a larger unit, and has no gas bypass. Also, the relatively dry raw gas requires a substantial amount of additional steam for the shift conversion. Because of this, the steam and quench water requirements for the Lurgi slagging shift converter are substantially greater than those for the Lurgi dry-ash shift converter. The need for steam addition to the slagging gasifier raw gas before being shifted prevents the recovery of any substantial amount of waste heat steam from the raw gas after quenching.

3. Hydrocarbon Research Gasifier (Process 21): The material balance for the Hydrocarbon Research Gasifier was made directly from data furnished by Hydrocarbon Research, Inc. The results of the precise balance are in good agreement with the data supplied. The 101 lb of ash shown in the material balance is based on the assumed availability of an entrained dust separator for the process

as discussed above under "Basis for Evaluations."

The heat balance has been made on the basis of the coal, steam, and oxygen being heated to an average temperature of 1000 F before they enter the gasifier.

The simplified process scheme shows that a substantially more complex coal preparation is required for the Hydrocarbon Research Gasifier than for a Lurgi gasifier. This coal must be crushed to a much smaller size than in the case of Lurgi, and it must be dried and preheated in an inert gas atmosphere, before it is fed to the gasifier. However, the ability to handle fine material in the HRI gasifier eliminates the necessity for selling excess fines. As in the case with all of the 450 psig gasifiers, it was assumed that the coal lock hopper system requires the use of 2-1/2 percent of the raw product gas. This lock hopper gas is later used to superheat the high pressure steam from the methanation unit. Because the gas leaving the gasifier is assumed to contain no appreciable amounts of tar, oil, or phenols, the raw gas quench used in the fixed-bed process is not required in this process, and the high level heat in the raw gas can be used for superheating steam. Some of the raw gas can bypass the water-gas shift converters since the amount of carbon monoxide to be shifted is substantially lower than for those processes which make little methane in the gasification step. Also, there are no by-product recovery systems required for phenols or ammonia since there are no byproducts other than sulfur produced; the sulfur is made from hydrogen sulfide in the Vetrocoke acid gas removal system.

4. Bamag-Winkler Atmospheric Gasifier (Process 7): A material balance for the Bamag-Winkler Atmospheric Gasifier was made based on information obtained from Pintsch-Bamag. The ash composition shown was not specifically stated by Bamag; it was obtained by difference in the material balance, and is approximately that expected from the Winkler fluidized bed with its high entrainment of dust in the raw gas. The 67 percent carbon shown in the ash is, of necessity, the content of carbon in the fluidized bed also. The very small amount of methane shown in the gas from the Winkler atmospheric gasifier, compared with the large amount of methane in the gas from the Hydrocarbon Research Gasifier, shows the effect of the almost 20 times higher operating pressure on methane formation in a fluidized bed.

The heat balance also was made from Bamag data, but adapted to Pittsburgh seam coal with an ash softening point of 2190 F and using a gas exit temperature of 2100 F. The heat balance output shows a substantial part of the heat leaving as "ash combustibles heating value." This results from the 67 percent carbon in the ash. An ash produced with this carbon content is easily burnable, since it has a heating value of 20 MM Btu per ton. The remaining 1850 tons of char produced per day is assumed to be saleable at a Btu price equivalent to $4 per ton price for the as mined coal having 27 MM Btu per ton.

According to the simplified process scheme, the Bamag-Winkler Atmospheric Gasifier is operated at 10 psig, and the cooled, clean gas compressed to 450 psig.

The high temperature level of the gas leaving the gasifier allows steam to be superheated at two different temperature levels. The 600 psig saturated steam made in the methanation unit is superheated to 750 F, and the 30 psig steam at 350 F made from back pressure turbines in the synthesis gas compression system is superheated to 750 F for the gasifier.

After compression to 450 psig before the shift converters, the gas leaving the gasifier is processed in a manner similar to the other low methane synthesis gases.

5. Rummel Single-shaft Pressurized Gasifier (Process 61): The data used for the material balance for the Rummel Single-shaft Pressurized Gasifier are for operation of the Rummel gasifier on Ruhr coal, as discussed under "Basis for Evaluations." The data required very little modification for the use of Pittsburgh seam coal. The carbon, hydrogen, and oxygen balances required corrections of less than 2 percent. The material balance shows a high percentage of hydrogen in the raw gas, as a result of 75 percent steam decomposition. This value for steam decomposition is stated to be too high by Dr. Domann of Otto, as discussed under "Basis for Evaluations."

The heat balance shows that some water is fed to the jacket of the gasifier to produce 150 psig jacket steam. The heat losses, which are obtained by difference, are abnormally low for this process, as a result of the quite high gasification efficiency. The heat losses should be in the order of 200,000 Btu per thousand pounds of coal and would reduce the gasification efficiency by about 1 percent. Such a change is not of any significance for the purpose of this study.

The simplified process scheme shows that feed coal of 1/8 inch by 0 in size is required for the gasifier. There are no excess fines produced for sale, and no unusual features of the process flow diagram are to be noted. The high temperature level of the raw gas is used to superheat steam from the methanation unit; some of it is superheated to 1310 F and sent to the gasifier.

Since the gasifier produces a very small amount of methane, a substantial carbon monoxide shift conversion is required, even though the raw gas has an abnormally high hydrogen content. Part of the lock hopper gas and all of the fuel gas from acid gas removal is used as a fuel for heating reboilers to regenerate the circulating alkali solution in the Vetrocoke system.

6. Rummel Modified Single-shaft Pressurized Gasifier (Process 62): The material balance for the Rummel Modified Single-shaft Pressurized Gasifier was based on information from Otto as provided by Dr. Domann. The steam input to the gasifier was not specifically stated, but from the steam output in the gas and the hydrogen and oxygen content of the gas, it follows that steam decomposition in the gasifier is approximately 40 percent. From this, the steam in the inlet gas was derived by material balance.

The substantial amount of methane formed in the gasifier is the result of feeding coal into the upper part of the gasifier shaft in such a way that the volatile materials are not decomposed by the intense heat from the slagging section of the gasifier.

The heat balance shows the effect of the solids (char) recycle from the raw gas dust separator. The temperature of this recycle of char is 730 F after the waste heat has been recovered from the raw gas/char mixture. The char is sent to the bottom portion of the gasifier, where it is gasified in contact with the slag bath. The heat loss obtained by difference in the heat balance is a reasonable one. This indicates that the thermal values assigned to the other streams in and out of the gasifier are reasonable.

The simplified process scheme for this process is not unusual in any respect other than the hot char recycle mentioned above. The steam generated in the gasifier jacket is used to drive small miscellaneous drives as part of the 20,000 kilowatt total of such drives.

7. Koppers-Totzek Pressurized Gasifier (Process 60): The material balance for the Koppers-Totzek Pressurized Gasifier is based on data obtained from the Heinrich Koppers GmbH of Essen for atmospheric gasification of Pittsburgh seam bituminous coal. At the very high temperatures of gasification, practically no methane is formed, so that the Koppers gasifier is primarily a producer of a carbon monoxide and hydrogen synthesis gas. The precise material balance was obtained by only minor modifications of the data supplied by Koppers.

The heat content of the slag in the output of the material balance has all been shown as "ash sensible heat," even though some of the heat is in entrained particles, and could be listed as "entrained solid sensible heat." The heat loss found by difference is somewhat higher than should normally be expected; this is probably due to an inaccuracy of 1 or 2 percent in the "hot raw gas heating value." However, a few percent inaccuracy in the heat balance makes little difference in the final pipeline gas cost.

The simplified process scheme shows that the gasifier feed is pulverized fuel. The gasifier requires a feed at least as small as pulverized fuel, and possibly smaller. The Koppers gasifier is commonly known as a dust gasifier. The flow diagram is a conventional one for processes containing practically no methane in the raw gas from the gasifier. Because of the very high heat content of the gas from the gasifier, a substantial amount of waste heat steam is formed in this process. The high temperature level of the raw gas is also used to superheat the 600 psig saturated steam from the methanation unit.

8. Texaco Gasifier (Process 22): The material balance for the Texaco Gasifier was based on information given by Texaco. Coal is fed in a slurry with water and the water in the slurry vaporized in a preheater. A high steam-coal ratio thus is used in this process. A precise material balance was made from the data with almost no changes required. The data supplied by Texaco were for entrained gasification, which produces no phenols, tars, or oils, and which was specifically based on Pittsburgh seam coal by extrapolation of actual experimental data for a Japanese coal.

The heat balance is made with the hot exit gas as the reference point for the output gases. The quenched gas according to Texaco is available at 450 psig at 400 F. A heat balance on the quench gave a gas temperature of approximately 2175 F for the hot raw gas entering the quench zone. This is the temperature used for the output streams in the gasifier heat balance.

In the simplified process scheme some of the gas is bypassed around the hot gas shift. Because of the high water vapor content of the quenched gas, and the relatively high hydrogen to carbon monoxide ratio in this gas, approximately 30 percent of the gas can be bypassed around the shift converters. Also, because of the large amount of steam produced in methanation, plus the fact that the Texaco process produces its own steam for gasification by vaporizing the water from the coal/water slurry fed to the gasifier, no normal steam production is required from the steam boiler in this process. The only boiler required is that for start-up service.

The heat loss calculated by difference in the heat balance is a reasonable figure and attests to the accuracy of the thermal values of the input and output streams. All of the water used to slurry the coal fed to the gasifier is vaporized, and all of this steam enters the gasifier at 500 F. Also, the excess of 30 psig steam produced in the "waste heat out" of the shift converters is "compressed" in a steam ejector with 600 psig steam to make 150 psig drive steam for small drives. It has been assumed that all the heat for oxygen preheating, coal water slurry preheating, and steam superheating is obtained from coal; these items account for approximately 1300 tons per day of coal in addition to the coal required for gasification.

9. Fixed-bed Super-pressure Gasifier (Process 56): The assumptions for the material balance for the Fixed-bed Super-pressure Gasifier have been described in the "Basis for Evaluations." This is a fixed-bed process; and the formation of tars, oils, phenols, and ammonia must be accounted for in the material balance. Also, some of the steam shown as required by gasification in the material balance is made in the jacket of the gasifier; 190 lb of boiler feed water per thousand pounds of coal is required. This amount of jacket steam has been determined by prorating it according to the jacket area, which is less than that for the conventional fixed-bed gasifier, due to the increased gasifier capacity gained by higher pressure operation.

The heat balance is made with the heating values of the tar, oils, benzenes, and ammonia shown as part of the raw gas heating value. Once again, the coal and oxygen are assumed to be unpreheated, the coal being fed at ambient temperature and the oxygen being fed at oxygen compressor discharge temperature. The 1100 psig steam necessary for the gasification is assumed to be fed at 750 F. The heat loss of 29,000 Btu obtained by difference is much too small for a commercial fixed-bed unit at 1050 psig. This indicates that the gasification efficiency assumed for this process is somewhat high, but it is still well within the tolerable limits of accuracy considering the data available.

The simplified process scheme for the super-pressure gasifier differs in several important respects from the process schemes for the 450 psig gasifiers. Pipeline gas compression is not required, since the gas is produced at pipeline gas pressure. Also, the gas losses associated with lock hopper feeding of coal into the gasifier at 1050 psig would be so high as to be economically prohibitive for this feeding method. Therefore, carbon dioxide from the acid gas removal system is compressed to 1100 psig and is used as a pressurizing gas for the coal feed system. Thus, there are no lock hopper gas losses for the super-pressure processes. Also, it has been assumed that the fuel gas losses in acid gas removal will be practically eliminated by a partial flash of the solvent to release fuel gases which would be absorbed in the solvent, and by a recompression and recycling of them into the feed gas. Steam for gasification is required at 1100 psig; this has been used as the pressure for generating steam in the fired boilers for gasification as well as for the drive turbines for the oxygen plant and the carbon dioxide compressors. In all other respects, the process scheme is quite similar to the scheme presented for the Lurgi Dry-ash Gasifier at 450 psig.

10. Fluidized-bed Super-pressure Gasifier (Process 57): The material balance for the 1050 psig fluidized-bed gasifier has been obtained by assumptions as discussed under "Basis for Evaluations." The same operating temperature, that is 1750 F, has been assumed for operations at 1050 psig. Also, the same carbon "activity" has been assumed at 1050 psig as was used at 450 psig by Hydrocarbon

Research. These assumptions, plus a slight increase in gasification efficiency because of higher gasifier capacity and lower heat loss, and application of the water-gas shift equilibrium, allowed the calculation of a gas composition for a fluidized-bed system at 1050 psig. The amount of carbon in the fluidized bed was maintained at 30 percent, as has been done for processes operating at 450 psig.

The heat balance shows the entering coal, oxygen, and steam as being superheated to an average temperature of 1000 F, and the raw gas leaving the gasifier as being at 1750 F. The gasification efficiency assumed is reasonable, since the heat losses by difference are about what would be expected for a commercial unit of the capacity considered here.

The simplified process scheme for the Fluidized-bed Super-pressure Gasifier is quite similar to the process scheme for the 450 psig Hydrocarbon Research Gasifier, except that pipeline gas compression is not necessary and compressed carbon dioxide has been provided as the gas for feeding coal to the gasifier. Again, the process scheme is simpler than that for the fixed-bed gasifier because no byproducts such as tar, oils, or phenols are formed.

<u>11. Two-stage Super-pressure Entrained Gasifier R1 (Process 58)</u>: A material balance for the Two-stage Super-pressure Gasifier R1 has been derived from calculations described under "Basis for Evaluations" of the processes. The calculation for this material balance is based on a carbon activity equal to beta graphite (i.e., a carbon activity of one) in the carbon/hydrogen/methane equilibrium. Because of the two-stage nature of the process, the devolatilized char from the upper stage is gasified in the lower stage, and ash with essentially no carbon content is produced as shown in the material balance. The methane shown in the raw gas is obtained from devolatilization of the coal in a second stage and from reaction between the char so produced and the hydrogen in the gas from the first stage.

The heat balance shows the coal entering the gasifier at a temperature of 210 F, as it would leave a coal drying system. The 1100 psig steam entering the first stage is assumed to be at 800 F, and the steam entering the second stage to be at 932 F. Raw gas and char leave the gasifier at 1700 F, and the separated char is returned to the first stage of the gasifier at 1110 F. Some heat loss is accounted for in the temperature drop in the char in going from the raw gas to the entrance of the first stage. The remaining heat loss by a difference is very close to what would probably be a reasonable heat loss in a commercial unit.

The simplified process scheme for the Two-stage Super-pressure Gasifier R1 generally follows the line of the previously discussed fluidized-bed super-pressure gasifier, except that an external solid char recycle to the gasifier is required, and different conditions of preheat are required for these streams fed to the gasifier.

<u>12. Two-stage Super-pressure Entrained Gasifier R2 (Process 58)</u>: Heat and material balances and a simplified process scheme for this gasifier are in every respect similar to that for the Two-stage Super-pressure Gasifier R1 except that the calculations for the raw gas methane content were made based on a carbon activity of two with respect to beta graphite as described under "Basis for Evaluations."

13. Two-stage Super-pressure Entrained Gasifier R3.4 (Process 58): The material balance, heat balance, and simplified process scheme for this gasifier are also similar in every respect to the gasifier for the Two-stage Super-pressure Gasifier R1, except that a carbon activity of 3.4 has been used, as described under "Basis for Evaluations" of the processes.

14. Catalytic Steam Methanation Gasifier (Process 65): Material and heat balances have been derived for the gasifier described under "Basis for Evaluations." For the purpose of economic evaluations, an amount of catalyst equal to 45 lb per 1000 lb of coal was assumed. This quantity of catalyst has been used to arrive at an assumed catalyst cost, and has not been included as one of the chemical reactants shown in the material balance. The heat balance was made for the conditions shown in the simplified process scheme. It is assumed that the coal-catalyst mixture is preheated to 615 F.

The procedure for evaluating the catalytic steam methanation gasification system as shown in the simplified process scheme for pipeline gas production is as follows. Run-of-mine coal is crushed and ground to 1/32 inch x 0 size, and catalyst is added to it before preheating. A coal-catalyst mixture is preheated to 615 F and injected into the gasifier using hot carbon dioxide as a carrier gas. Hot recycle char at 1100 F and oxygen at 1150 F are injected into the gasifier along with the coal and catalyst. The raw gas and char leaving the gasifier enter a char separation system, and the clean gas leaving the char separation system enters a superheater and waste heat boiler that cools it to 700 F. During this cooling, the gas gives up sufficient heat to superheat all of the steam required for gasification from a temperature from 750 F to 1150 F. In addition to this superheating, the raw gas waste heat is sufficient to generate approximately 6,000,000 lb per day of 1150 F steam from 225 F boiler feed water. The 700 F gas enters a carbon monoxide shift conversion unit; 45 percent bypasses the converter and the other 55 percent is shifted to produce a 3 to 1 hydrogen to carbon monoxide ratio in the final recombined gas.

A waste heat boiler and boiler feed water preheating system following the carbon monoxide shift converter cools the combined gas from approximately 775 F to nearly ambient temperature. This gas then enters a dual Vetrocoke acid gas removal system, where it is first cooled with cooling water and then scrubbed with a potassium arsenate solution for H_2S removal. Following H_2S removal, the gas is contacted with an activated potassium carbonate solution for carbon dioxide removal. Some of the carbon dioxide so removed is recompressed to approximately 1200 psig and is used for coal feeding in the gasification system. The H_2S removed in the Vetrocoke system is converted to sulfur in air regeneration towers, and this sulfur is extracted and either sold as a byproduct or is used to produce sulfuric acid. This, in turn, is reacted with the ammonia recovered to make ammonium sulfate. The gas leaving the acid gas removal systems enters a methane synthesis unit where sufficient additional methane is synthesized to form a gas with a gross heating value of 928 Btu per standard cubic foot. Finally, the gas leaving the methane synthesis is cooled by waste heat boilers and cooling water, and desiccant beds used to give a 40 F dew point at 1000 psig.

TABLE 6-3. OVERALL MATERIAL BALANCE SUMMARY FOR PROJECTED COMMERCIAL-SCALE PIPELINE GAS PLANTS

	Fixed-bed Processes		Fluidized-bed Processes		Entrained Processes				Super-pressure Processes					
	Process 11	Processes 18,19,20	Process 21	Process 7	Process 61	Process 62	Process 60	Process 22	Process 56	Process 57	Process 58	Process 58	Process 58	Process 65
	Lurgi Dry-ash	Lurgi Slagging	Hydrocarbon Research	Bamag-Winkler Atmospheric	Rummel Single-shaft Pressurized	Rummel Modified Single-shaft Pressurized	Koppers-Totzek Pressurized	Texaco	Fixed-bed	Fluidized-bed	Two-stage R1	Two-stage R2	Two-stage R3.4	Catalytic Steam Methanation
INPUT TON/DAY														
Coal (4.7% Moisture)	11,780	12,900	11,230	16,280	13,060	12,650	15,250	14,300	10,880	10,500	11,200	10,650	10,350	10,460
Oxygen (98%)	5,700	8,050	5,870	10,000	10,700	9,880	15,100	11,680	4,440	4,650	7,200	5,910	5,020	2,080
Boiler Feed Water Makeup	31,700	23,450	20,200	25,700	23,500	23,450	27,750	19,220	28,200	17,400	18,000	15,600	15,000	14,900
Lime	---	960	---	---	---	---	---	---	---	---	---	---	---	470*
TOTAL INPUT	49,180	45,360	37,300	51,980	47,260	45,980	58,100	45,200	43,520	32,550	36,400	32,160	30,370	27,910
OUTPUT TON/DAY														
Pipeline Gas	5,440	5,440	5,440	5,440	5,440	5,440	5,440	5,440	5,440	5,440	5,440	5,440	5,440	5,440
Carbon Dioxide	16,050	18,500	15,600	19,100	21,700	20,400	26,100	22,450	14,570	14,450	17,600	16,050	15,100	11,350
Hydrogen Sulfide	300	330	290	190	340	320	390	370	280	290	290	280	270	270
Process Effluent Water	24,470	17,380	13,740	21,900	17,460	17,760	23,250	14,330	21,720	11,350	12,300	9,660	8,850	8,150
Ammonia	140	140	---	---	---	---	---	---	130	---	---	---	---	40
Gas Losses, Tar and Phenols	1,930	1,720	1,130	650	1,400	1,170	1,510	680	600	---	---	---	---	---
Ash, Slag or Char	850	1,850	1,100	4,700	920	890	1,410	1,930	780	1,020	770	730	710	2,660
TOTAL OUTPUT	49,180	45,360	37,300	51,980	47,260	45,980	58,100	45,200	43,520	32,550	36,400	32,160	30,370	27,910

A

*Catalysts

B

C. Results and Discussion

The results of the evaluations of synthesis gas processes for pipeline gas production are presented in a set of tabular summary sheets and graphs. The following discussions concern these summaries and graphs.

1. Overall Material Balance Summary: The overall material balance summary (Table 6-3) compares the material balances for process material in and out of each projected commercial-scale pipeline gas plant. The input for each plant is coal as mined with 4.7 percent moisture. This coal input figure is only the "process" coal going to the gasifier, and does not include the coal which is required by fired boilers, superheaters, or preheaters. The moisture in this coal is removed in the grinding and drying stage for all processes except the fixed-bed processes and the Texaco process. The 4.7 percent moisture which is removed during grinding and drying is accounted for as part of the process effluent water stream in the output.

The oxygen of 98 volume percent purity includes both nitrogen and some argon as impurities, which eventually leave in the pipeline gas. Boiler feed water shown in the input is used to make up the requirements for the boiler blowdown plus process steam used in gasification and elsewhere, such as in by-product recovery. A substantial part of the boiler feed water makeup used to produce process steam is decomposed in the gasifier, and leaves in the form of hydrogen, methane, carbon monoxide, and carbon dioxide in the output streams.

Among the output streams it has been assumed that all of the processes produce the same composition of pipeline gas, although this assumption is somewhat of an oversimplification. For example, the fixed-bed processes would produce pipeline gas with more ethylene than the processes which would decompose the ethylene before it appeared in the raw gas from gasification. The weight of pipeline gas shown is for 250 MM scfd of gas with a composition of 90 percent methane, the remainder being a mixture of unreacted hydrogen, carbon dioxide, nitrogen, some small amount of carbon monoxide, and a small amount of ethylene. This gas has 928 Btu gross heating value per standard cubic foot.

The carbon dioxide in the output streams contains the oxygen that has been added as 98 percent oxygen, plus the oxygen from water decomposed in the gasifier. The hydrogen sulfide and organic sulfur compounds in the gas leaving the gasifier originate from the sulfur content of the coal. These are completely converted to hydrogen sulfide in the hot gas shift. The process effluent water in the output is boiler blowdown plus condensate from the synthesis gas. This effluent could possibly be reused in the boiler feed water preparation unit, but in this simplified economic survey, it has been assumed that the process effluent water is discarded and the boiler feed water makeup is prepared from fresh river water. In the cases of the fixed-bed processes, some process water effluent streams contain phenols. They must be removed and are shown separately in the output tabulation. The ammonia output is shown as ammonia since the overall material balance was not made to include a by-product chemical plant which will convert ammonia to ammonium sulfate.

The figures shown under gas losses, tar, and phenols include oil and benzene as part of tar. This fraction is burned in the process auxiliaries such as the steam boiler, preheaters, and superheaters. The gas losses included in this item are for the lock hopper gas loss for gasification and the fuel gas losses entailed in the acid gas removal. The Bamag-Winkler gas losses are appreciably lower because the Bamag-Winkler unit operates at near atmospheric pressure and requires no lock hopper gas. Also, the super-pressure processes are designed to eliminate lock hopper gas losses by the use of a CO_2 purge, and eliminate fuel gas losses in acid gas removal by partial letdown and recompression of such gases. The final figure in the output column is for ash or, in the case of the slagging units, slag. The Bamag-Winkler Atmospheric gasification and the Catalytic Steam Methanation show figures substantially different from the others in this category. Bamag-Winkler produces ash which is a usable fuel containing 67 percent carbon. Of the 4700 tons of this fuel shown being produced per day by the Bamag-Winkler process, all but 1850 tons are used in firing the boiler for the process. The 1850 tons then is a product to be sold at the Btu value of the coal input to the process. Catalytic Steam Methanation produces 2660 tons per day of 50 percent carbon ash, all of which is used to fire boilers and heaters.

2. Process Data Summary: The data obtained from the material and energy balances discussed under "Procedure for Evaluations" are presented in tabular form as Table 6-4. This summary shows the total coal requirement; it is the sum of the 4.7 percent moisture coal to the gasifier and the coal to the fired boilers, preheaters, and superheaters. Requirements for 98 percent purity oxygen and boiler feed water makeup, as discussed for the "Overall Material Balance Summary" are also shown.

Cooling water makeup to replace windage losses and blowdown from the cooling tower basins is shown as 5 percent of the total cooling water circulation. The cooling water circulation for all coolers and condensers has been tabulated, based on a 30 F temperature rise. Steam produced in fired boilers is shown, as is the total steam production including waste heat boilers. The total dry raw synthesis gas is shown, as is the amount of this gas that goes to the carbon monoxide shift. The amount of methane which must be formed in the methane synthesis unit, and the total amount of CO_2 to be removed from the gas are tabulated. Elemental sulfur production is shown together with those processes which also have ammonia recovery; the amount of ammonium sulfate formed by the combination of this ammonia with sulfuric acid made from some of the sulfur is also shown. The amount of raw phenols produced by the fixed-bed processes and the amount of excess char produced by the Winkler atmospheric process are shown. Finally, the overall gasification efficiency from total coal to the gross heating value in the pipeline gas is given, based on 27 MM Btu per ton of as mined coal.

3. Labor Summary: The "Labor Summary" in Table 6-5 shows the total number of operators and assistants required for each process unit, as well as the number of laboratory technicians, guards, cleanup men, and foremen required for the plants. Labor requirement totals are used for the direct operating labor, with 10 percent added for supervision and 60 percent added for payroll overhead to arrive at operating costs. The labor requirements have been estimated assuming a modern, thoroughly automated plant and are only applicable if maximum use is made of instrumentation to maintain normal operation without manual assistance. It is entirely possible that a thorough study of the labor requirements for a well laid out plant could reduce the number of operators. However, no layouts were deemed necessary for arriving at the tabulated approximate operating labor figure.

4. Investment Summary: An estimate of approximate investment costs was made for each of the process units identified on the simplified process scheme. This is used for a relative evaluation of economics. Table 6-6 presents a summary of all these costs. Primary emphasis has been placed on evaluating the processes on a comparable cost basis. Thus the investment costs have been derived with more concern for accuracy between processes; that is, the use of the same basis for all processes. The costs presented herein are conservative. An estimate made in the manner of this one should not attempt to show the minimum costs for the equipment being estimated, since many small factors, which tend to be overlooked in budget estimates, will be adequately covered in cost because of the conservative estimating procedure.

a. Methanation: The first investment cost item is the methanation unit. The costs shown for methanation include a gas to gas heat exchanger for heating pure gas to near the 660 F methanation temperature, and the methanation unit itself. This latter unit consists of a dowtherm cooled tubular catalytic unit with Raney nickel catalyst deposited on the outside of tubes which are in contact with gas on the shell side of the methanation unit, and in contact with boiling dowtherm on the inside of the tubes. Dowtherm from the methanation unit passes to a waste heat boiler and is used to generate 600 psig or 1100 psig steam, depending on whether the process is a normal pressure, elevated pressure, or a super-pressure one. Also included in the methanation investment is the cost of a waste heat boiler for generating low pressure steam at approximately 30 psig, using some of the product gas heat, and further, the cost of the cooler for cooling the gas with cooling water to a temperature suitable for entering the pipeline gas compressors or the final dryer.

b. Pipeline Gas and CO_2 Compression: The costs for the pipeline gas compressors, their steam turbine drives and condensers, and compressed gas dryers, all installed, are shown next. For those super-pressure processes which operate at the pipeline gas pressure, costs shown in this item are for the carbon dioxide compressors and their turbines and condensers as required to furnish 1100 psig carbon dioxide for the coal charging facilities for the gasifier, plus the cost of final gas drying to a 40 F dew point at 1000 psig.

c. Shift Conversion: The next item is the investment for the carbon monoxide shift and its associated exchangers, waste heat boilers, and coolers. The cost of the carbon monoxide converters was obtained from the Selas Corporation of America, and the costs of the associated waste heat boilers, heat exchangers, and coolers were estimated based on large heat exchanger costs available from standard cost estimating methods in the literature.

d. Acid Gas Removal: Acid gas removal system costs have been estimated using the Vetrocoke processes for H_2S and CO_2 removal. Studies were made of these processes by the Chemical Plants Division of Blaw-Knox several years ago, using data obtained directly from Dr. Giammarco of Vetrocoke. These processes have been further developed since Blaw-Knox made its evaluations. It is to be expected that a careful evaluation of a combination purification step for a plant of this size would lead to savings in investment figures compared with those used here. The acid gas removal investment includes the cost of final cleanup of the acid gas to remove residual traces of hydrogen sulfide, using activated charcoal.

TABLE 6-4. PROCESS DATA SUMMARY FOR PROJECTED COMMERCIAL-SCALE PIPELINE GAS PLANTS

		Fixed-bed Processes		Fluidized-bed Processes		Entrained Processes				Super-pressure Processes					
Item	Units	Process 11 Lurgi Dry-ash	Processes 18,19,20 Lurgi Slagging	Process 21 Hydrocarbon Research	Process 7 Bamag-Winkler Atmospheric	Process 61 Rummel Single-shaft Pressurized	Process 62 Rummel Modified Single-shaft Pressurized	Process 60 Koppers-Totzek Pressurized	Process 22 Texaco	Process 56 Fixed-bed	Process 57 Fluidized-bed	Process 58 Two-stage R1	Process 58 Two-stage R2	Process 58 Two-stage R3.4	Process 65 Catalytic Steam Methanation
Coal (4.7% Moisture)	ton/day	12,860	13,410	12,280	16,280	13,330	13,600	15,340	15,600	12,170	11,900	12,240	11,700	11,420	10,460
Oxygen	ton/day	5,700	8,050	5,870	10,000	10,700	9,880	15,100	11,680	4,440	4,650	7,200	5,910	5,020	2,080
Cooling Water Circulated	gpm	263,000	356,000	241,000	411,000	386,000	362,000	501,000	407,000	225,000	208,000	301,000	263,000	235,000	143,000
Cooling Water Makeup	gpm	13,000	17,500	12,000	21,000	19,000	18,000	25,000	20,000	11,000	10,000	15,000	13,000	12,000	7,100
Boiler Feed Water Makeup	gpm	5,300	3,900	3,400	4,300	3,900	3,900	4,600	3,200	4,700	2,900	3,000	2,600	2,500	2,500
Fired Boiler Steam Production	M lb/hr	1,470	693	493	1,650	364	1,102	327	---	1,290	837	504	570	620	1,270
Total Steam Production	M lb/hr	4,765	3,103	3,740	4,340	4,910	4,890	6,110	4,575	4,265	3,375	3,850	3,610	3,300	2,400
Dry Gas to CO Shift	MM scfd	310	785	606	1,030	1,080	748	1,060	728	322	495	859	748	669	318
Dry Raw Synthesis Gas	MM scfd	883	802	807	1,030	1,080	748	1,060	1,039	804	829	859	748	669	578
Methane by Synthesis	MM scfd	133	166	128	205	225	139.5	225	218	119	94	146	123.5	109	37
CO_2 Removal	ton/day	16,050	18,500	15,600	19,100	21,750	20,400	26,100	22,450	14,550	14,450	17,600	16,050	15,100	11,350
Sulfur	ton/yr	49,000	59,000	89,000	59,000	105,000	99,000	122,000	115,000	46,000	89,000	91,000	86,000	84,000	69,000
Ammonium Sulfate	ton/yr	187,500	187,500	---	---	---	---	---	---	175,000	---	---	---	---	59,000
Raw Phenols	ton/yr	20,500	15,200	---	---	---	---	---	---	19,000	---	---	---	---	---
Char	ton/day	---	---	---	1,850	---	---	---	---	---	---	---	---	---	---
Overall Efficiency Btu in Gas/Btu in Coal	Percent	66.9	64.1	70.1	57.7	64.4	63.2	56.0	55.1	70.7	72.3	69.6	73.1	75.0	82.1

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e. Gasification: The investment costs of the coal gasifiers are the most difficult to derive, and are perhaps subject to the greatest possibility for inaccuracy of any of the individual processing units.

For the Lurgi Dry-ash Gasifier, the total investment for the gasifiers and associated coal charging and ash handling equipment, controls, and gas quenching and cooling equipment is based on costs furnished by the Lurgi company for delivery of vessels and piping of German manufacture to the United States. An allowance has been made for the additional cost of engineering the German equipment to American ASME and ASA standards. Based on experience of American companies for this type of unit, further allowances for freight and duty and erection costs have been added to the quotations from Lurgi.

The estimate for the Lurgi Slagging Gasifier provides for approximately half the number of gasifiers used in the Lurgi dry-ash estimate. On the other hand, a substantial part of the equipment such as the gas-quench towers, coal bunkers, and ash handling equipment will not be reduced in size or cost for the Lurgi Slagging Gasifier. The Lurgi slagging gasification investment is thus estimated in proportion to the investment for the Lurgi dry-ash process. This is considered to be the best approximation possible at this time.

The cost of the Hydrocarbon Research gasification system includes the gasifiers, their preheaters, and the associated waste heat boilers as well as coal charging and ash removal equipment and controls. For the total plant 15 gasifiers are provided, each about 40 ft high and 11 ft 4 in. ID and suitable for operation at 450 psig.

The unit cost for the gasifiers, including coal and ash handling and preheating equipment and controls, is assumed to be about $3 million per gasifier; this is about 40 percent higher per gasifier than for the Lurgi dry-ash process. This increase in cost results from the greater length of each gasifier as compared with the Lurgi dry-ash units. This estimated cost is an approximate one, but for lack of specific information, it is the best cost available at the present time and is considered to be sufficiently accurate to lead to valid conclusions for process comparisons.

The Bamag-Winkler Atmospheric Fluid-bed Gasifier operates at a pressure of only 10 psig. The data supplied by Pintsch-Bamag indicate for the largest unit a diameter of 10 meters (33 ft). The capacity of this size of unit would be such that seven units, including one spare, are necessary to produce approximately one billion cubic feet per day of dry raw gas. Because of the low pressure and in spite of the large diameter, the gasifier is of simpler design with respect to coal feeding and ash discharging operations; therefore, a cost approximately 85 percent of that used for the Hydrocarbon Research Gasifier has been used for the Bamag-Winkler Atmospheric Pressure Gasifiers. This leads to a cost of about $5.5 million per 33 ft ID unit, including accessories, as compared to $3 million per 13 ft ID pressurized Hydrocarbon Research Gasifier, including accessories.

Data for the Rummel single-shaft processes are based on investment cost data supplied by Dr. Otto & Comp. of Bochum, Germany. The cost data were based on a plant supplied with equipment manufactured in Germany for gas generation, coal charging, ash handling, and waste heat boilers. To the costs for the German equipment, there has been added the cost of bringing the equipment to the United States and erecting it under American conditions.

TABLE 6-5. LABOR SUMMARY FOR PROJECTED COMMERCIAL-SCALE PIPELINE GAS PLANTS

	Fixed-bed Processes				Fluidized-bed Processes				Entrained Processes							
	Process 11 Lurgi Dry-ash		Processes 18, 19, 20 Lurgi Slagging		Process 21 Hydrocarbon Research		Process 7 Koppers-Winkler Atmospheric		Process 61 Bummel Single-shaft Pressurized		Process 62 Bummel Modified Single-shaft Pressurized		Process 60 Koppers-Totzek Pressurized		Process 22 Texaco	
	Men/ Shift	Men/ Day	Men/ Shift	Men/ Day	Men/ Shift	Men/ Day	Men/ Shift	Men/ Day	Men/ Shift	Men/ Day	Men/ Shift	Men/ Day	Men/ Shift	Men/ Day	Men/ Shift	Men/ Day
Methane Synthesis	4	12	5	15	4	12	5	15	5	15	4	12	5	15	5	15
CO Shift and Waste Heat Boilers	1	3	2	6	2	6	2	6	2	6	2	6	2	6	2	6
Acid Gas Removal	5	15	6	18	5	15	7	21	7	21	7	21	8	24	7	21
Waste Heat Boiler and Dust Removal	2	6	1	3	1	3	1	3	1	3	1	3	1	3	1	3
Coal Charging	6	18	3	9	2	6	2	6	2	6	2	6	3	9	3	9
Ash Handling	4	12	2	6	2	6	4	12	2	6	2	6	3	9	2	6
Gasification Control Room	8	24	4	12	3	9	3	9	2	6	2	6	4	12	3	9
Raw Gas Quench	3	9	3	9	-	-	-	-	-	-	-	-		-	-	-
Gasification Auxiliaries	4	12	3	9	1	3	1	3	1	3	1	3	2	6	2	6
Oxygen Plant and Compression	8	24	10	30	8	24	12	36	13	39	12	36	17	51	14	42
Raw Gas Compression	-	-	-	-	-	-	1	3	-	-	-	-	-	-	-	-
Coal Storage	1	3	1	3	1	3	1	3	1	3	1	3	1	3	1	3
Coal Crushing and Screening	2	6	2	6	-	-	-	-	-	-	-	-	-	-	-	-
Coal Pulverizing and Drying	-	-	-	-	6	18	9	27	6	18	6	18	9	27	9	27
Dephenolization	4	12	4	12	-	-	-	-	-	-	-	-	-	-	-	-
Ammonium Sulfate	5	15	5	15	-	-	-	-	-	-	-	-	-	-	-	-
Sulfuric Acid	2	6	2	6	-	-	-	-	-	-	-	-	-	-	-	-
Power Plant and Boiler Feed Water Preparation	7	21	7	21	7	21	7	21	7	21	7	21	7	21	4	12
Electric Distribution	1	3	1	3	1	3	1	3	1	3	1	3	1	3	1	3
Cooling Water System	2	6	2	6	2	6	2	6	2	6	2	6	2	6	2	6
Labs, Gates, Janitors, Etc.	10	26	10	26	10	26	10	26	10	26	10	26	10	26	10	26
Others	8	24	8	24	8	24	8	24	8	24	8	24	8	24	8	24
Foremen	7	21	7	21	7	21	7	21	7	21	7	21	7	21	7	21
TOTAL	94	278	88	260	70	206	83	245	77	227	75	221	90	266	81	239

	Super-pressure Processes											
	Process 56 Fixed-bed		Process 57 Fluidized-bed		Process 58 Two-stage R1		Process 58 Two-stage R2		Process 58 Two-stage R3.4		Process 65 Catalytic Steam Methanation	
	Men/ Shift	Men/ Day	Men/ Shift	Men/ Day	Men/ Shift	Men/ Day	Men/ Shift	Men/ Day	Men/ Shift	Men/ Day	Men/ Shift	Men/ Day
Methane Synthesis	4	12	3	9	4	12	4	12	3	9	2	6
CO Shift and Waste Heat Boilers	1	3	1	3	2	6	2	6	2	6	1	3
Acid Gas Removal	5	15	5	15	6	18	5	15	5	15	5	15
Waste Heat Boiler and Dust Removal	2	6	1	3	1	3	1	3	1	3	1	3
Coal Charging	4	12	2	6	2	6	2	6	2	6	2	6
Ash Handling	3	9	2	6	2	6	2	6	2	6	2	6
Gasification Control Room	5	15	3	9	2	6	2	6	2	6	2	6
Raw Gas Quench	3	9	-	-	-	-	-	-	-	-	-	-
Gasification Auxiliaries	4	12	1	3	1	3	1	3	1	3	1	3
Oxygen Plant and Compression	7	21	7	21	9	27	8	24	7	21	5	15
Raw Gas Compression	-	-	-	-	-	-	-	-	-	-	-	-
Coal Storage	1	3	1	3	1	3	1	3	1	3	1	3
Coal Crushing and Screening	2	6	-	-	-	-	-	-	-	-	-	-
Coal Pulverizing and Drying	-	-	6	18	6	18	6	18	6	18	6	18
Dephenolization	4	12	-	-	-	-	-	-	-	-	-	-
Ammonium Sulfate	5	15	-	-	-	-	-	-	-	-	2	6
Sulfuric Acid	2	6	-	-	-	-	-	-	-	-	2	6
Power Plant and Boiler Feed Water Preparation	7	21	7	21	7	21	7	21	7	21	7	21
Electric Distribution	1	3	1	3	1	3	1	3	1	3	1	3
Cooling Water System	2	6	2	6	2	6	2	6	2	6	2	6
Labs, Gates, Janitors, Etc.	10	26	10	26	10	26	10	26	10	26	10	26
Others	8	24	8	24	8	24	8	24	8	24	8	24
Foremen	7	21	7	21	7	21	7	21	7	21	7	21
TOTAL	87	257	61	197	71	209	69	203	67	197	67	197

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For the Rummel modified single-shaft process, the cost for six gasifiers, including one spare, was estimated to be $18 million. Because the Rummel Single-shaft Gasifier in unmodified form produces very little methane, the raw gas capacity required in this system is substantially greater than that required for the Rummel modified single-shaft process. By proportioning the number of gasifiers on the basis of raw gas volume, cost for the unmodified Rummel single-shaft gasification system is estimated as $26 million.

The Koppers-Totzek Gasifiers were estimated on the basis of capacities and costs obtained from Heinrich Koppers GmbH in Essen. The capacities and costs obtained for atmospheric pressure units were prorated to show increased capacity and increased costs for higher pressure operation. A capacity increase of approximately 17 times was assumed for the 450 psig operation, in comparison to that for the near atmospheric pressure operation. Thus, only three operating units plus one spare are required. Four atmospheric gasifiers with all accessories cost DM 14 million; four units with 17 times the capacity are estimated to cost $27 million, erected in the USA.

It should be emphasized that for the purposes of this report the investment costs for gasification processes, such as Winkler atmospheric, Koppers-Totzek, Rummel single-shaft, and Texaco, do not need to be derived with great accuracy, because the operating cost without gasification investment for these systems is higher than the total cost of gas production by the processes using fluidized or fixed coal beds. The main reasons for the higher cost of these processes are the higher oxygen consumption and the larger CO_2 removal systems.

For the Texaco Gasifier, an approximate estimate of gas generator costs, based on the entirely different coal charging system, gives a gasification investment of $19 million for nine gasifiers, plus approximately $4 million for coal water slurry preheaters, for a total investment of $23 million.

The super-pressure processes have been estimated without the benefit of quotations for 1050 psig gasification systems.

Based on 16 gasifiers for the fixed-bed super-pressure at 1050 psig including 2 spares, an estimated cost of $43 million was calculated. This includes coal charging, ash discharging, gas quenching, and waste heat boiler costs based on the estimates previously made for the Lurgi dry-ash and Lurgi slagging gasification systems. It is expected that this figure is high, but the uncertainties of the expense of fabricating the gasifiers and their lock hoppers for 1050 psig operation were deemed reasons for a conservative cost figure.

For the fluidized-bed super-pressure process a substantial reduction in the number of units was assumed as discussed under "Basis for Evaluations." The estimated cost of a system of fluidized-bed gasifiers at 1050 psig is $36 million. This figure was obtained by ratioing investment costs, based on the knowledge of the equipment involved. As stated for the fixed-bed super-pressure system, this figure is probably high.

The gasification investment for the two-stage super-pressure R1 system was estimated from cost data supplied by the Dr. Otto & Comp. for the Rummel Modified Gasifiers. Based on the conservative proportioning of number of gasifiers to the square root of the absolute pressure, as stated under "Basis for Evaluations," the number of gasifiers required for this super-pressure system is four in

TABLE 6-6. INVESTMENT SUMMARY FOR PROJECTED COMMERCIAL-SCALE PIPELINE GAS PLANTS (THOUSANDS OF DOLLARS)

Item	Fixed-bed Processes		Fluidized-bed Processes		Entrained Processes				Super-pressure Processes					
	Process 11 Lurgi Dry-ash	Processes 18,19,20 Lurgi Slagging	Process 21 Hydrocarbon Research	Process 7 Bamag-Winkler Atmospheric	Process 61 Rummel Single-shaft Pressurized	Process 62 Rummel Modified Single-shaft Pressurized	Process 60 Koppers-Totzek Pressurized	Process 22 Texaco	Process 56 Fixed-bed	Process 57 Fluidized-bed	Process 58 Two-stage R1	Process 58 Two-stage R2	Process 58 Two-stage R3.4	Process 65 Catalytic Steam Methanation
Methanation	4,000	5,000	3,900	6,200	6,800	4,200	6,800	6,600	5,000	4,000	6,100	5,200	4,600	2,800
Pipeline Gas and CO_2 Compression	2,500	2,500	2,500	2,500	2,500	2,500	2,500	2,500	2,000	2,000	2,000	2,000	2,000	2,000
Shift Conversion	4,000	4,600	4,400	5,000	5,000	4,600	5,000	4,600	4,000	4,200	4,800	4,600	4,600	1,800
Acid Gas Removal	35,000	40,000	34,000	38,000	46,000	43,000	55,000	48,000	32,000	32,000	38,000	35,000	33,000	26,000
Gasification*	54,500	(35,500)	(45,000)	(38,000)	(26,000)	18,000	(27,000)	(23,000)	(43,000)	(36,000)	(18,000)	(17,700)	(17,400)	(27,000)
Oxygen Plant	34,100	43,200	34,100	53,000	58,000	53,000	77,000	62,000	29,400	29,400	38,600	34,100	29,400	12,200
Oxygen Compression	4,200	5,800	4,300	1,000	7,500	7,100	10,900	8,500	4,300	4,500	7,000	5,800	4,800	2,100
Raw Gas Compression	---	---	---	31,000	---	---	---	---	---	---	---	---	---	---
Coal Preparation and Storage	3,500	3,500	6,800	9,700	6,800	6,800	10,900	9,700	3,500	6,800	6,800	6,800	6,800	6,800
Dephenolization	4,500	3,500	---	---	---	---	---	---	4,100	---	---	---	---	---
Sulfuric Acid Plant	1,300	1,300	---	---	---	---	---	---	1,200	---	---	---	---	600
Ammonium Sulfate Plant	3,600	3,600	---	---	---	---	---	---	3,400	---	---	---	---	1,600
Boiler Plant and Electric System	9,400	6,400	5,400	10,000	4,600	8,300	4,300	2,800	9,600	7,800	5,600	6,000	6,300	9,600
Boiler Feed Water Preparation	1,400	1,100	1,000	1,300	1,300	1,200	1,600	1,100	1,200	900	1,000	900	800	500
Cooling Water System	7,900	10,700	7,200	12,300	11,600	10,800	15,000	12,200	6,700	6,200	9,000	7,900	7,000	3,200
Off Site Facilities	25,000	24,000	21,000	30,000	26,000	23,000	31,000	26,000	22,000	19,000	20,000	18,000	17,000	13,800
NET FIXED INVESTMENT	194,900	190,700	169,600	238,000	202,100	182,500	247,000	207,000	171,400	152,800	156,900	144,000	133,700	110,000
Interest on Capital During Construction	9,700	9,500	8,500	11,900	10,100	9,100	12,300	10,400	8,600	7,600	7,800	7,200	6,700	5,500
TOTAL FIXED INVESTMENT	204,600	200,200	178,100	249,900	212,200	191,600	259,300	217,400	180,000	160,400	164,700	151,200	140,400	115,500
WORKING CAPITAL	5,400	5,300	5,000	6,900	5,800	5,400	6,900	6,200	4,800	4,700	4,800	4,500	4,400	4,000
TOTAL CAPITAL	210,000	205,500	183,100	256,800	218,000	197,000	266,200	223,600	184,800	165,100	169,500	155,700	144,800	119,500

*Gasification system costs in parentheses have been estimated without quotations from gasifier suppliers.

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operation, plus one spare. This is one less gasifier than used in the Rummel modified system. It has been assumed that a system utilizing five gasifiers at 1050 psig costs approximately the same as a system utilizing six gasifiers at 450 psig.

The gasification investments for the two-stage super-pressure R2 and R3.4 systems were estimated by allowing for a slight reduction in gasifier size, while using the same number of gasifiers as for the R2 system.

The investment for the Catalytic Steam Methanation gasification system, utilizing six gasifiers, was estimated to be $27 million, by ratioing from costs of the fluidized-bed super-pressure system.

f. Oxygen Plant: The total investment for the oxygen plant has been estimated from quotations obtained from American Air Liquide, Lotepro Corporation, Clark Compressor, Inc., Division of Dresser Industries, Allis Chalmers, General Electric, and Western Gear.

The prices for the low temperature separation plant, the associated defrosting equipment, the expansion turbines and generators, the automatic controls, the direct contact air cooler, the centrifugal and axial air compressors, the gears, the condensing turbine drives for the compressors, the surface condensers, and the building and foundations for the equipment have all been consolidated into the total oxygen plant cost. It must be realized that the size of the contemplated oxygen production units is several times that of the largest oxygen unit in existence at the present time. It is expected that detailed studies of the economics for an oxygen production facility of this size would result in costs lower than those used in this present study.

g. Oxygen Compression: The costs of oxygen compression are shown for the oxygen compressor, a separate drive turbine for the compressor, and a condenser for the turbine. It is quite possible that combinations of oxygen compressors and air compressors with drive-through shafts and gears could be arranged to give an entirely integrated oxygen production and compression facility with oxygen compression costs lower than those used in the present estimates. All mechanical equipment has been estimated using quotes by domestic suppliers. Quotations received from a Swiss firm, Escher-Wyss, indicate that up to 50 percent of the delivered cost of centrifugal and axial flow compressors could be saved if foreign made units could be purchased.

h. Raw Gas Compression: Only the Winkler atmospheric system requires compression of the raw gas produced by gasification before it is further processed. The cost of this raw gas compression to 450 psig is shown. This investment is based on the cost for the oxygen compression plant, with an allowance made for the fact that the less critical material of construction for the raw gas compressors together with their larger size would allow for overall raw gas compression costs somewhat lower than those for oxygen compression.

i. Coal Preparation and Storage: The costs for coal preparation and storage have been estimated using $1 million for the cost of equipment necessary to convey the coal to and from the storage area, to store it, and to take it out of storage, plus the costs for pulverization and drying. The latter have been based on information reported by Katell (55) for plants of approximately 250 tons per hour or 6000 tons per day coal capacity.

j. Dephenolization: The costs for plants to remove phenols from the effluent streams for the fixed-bed gasifiers are based upon the Phenosolvan process. These costs were derived from quotations on similar processes that the Blaw-Knox Chemical Plants Division has made in the past. The cost of a biological oxidation unit for removing the few ppm of residual phenol from the Phenosolvan plant effluent streams is also included in the dephenolization costs. A careful study of the phenol recovery and removal system might show that the biological oxidation of the total phenols would be the process to use for this size plant. However, the present phenol system gives a conservative investment cost.

k. Sulfuric Acid Plant: Investment for a sulfuric acid plant is required for processes having aqueous ammonia as a byproduct, in order to provide sulfuric acid for an ammonium sulfate facility. The sulfuric acid would be produced in a plant burning elemental sulfur obtained from the Vetrocoke hydrogen sulfide removal unit. The investment was obtained from published costs for sulfuric acid "package" plants.

l. Ammonium Sulfate Plant: The ammonium sulfate plant is designed to utilize the ammonia recovered from the raw gas condensate, and to form ammonium sulfate by combining it with sulfuric acid. The costs for the ammonium sulfate plant are based on quotations previously given to Blaw-Knox by suppliers of these plants.

m. Boiler Plant and Electric System: The costs for the fired boiler plant and the electric generating and distribution system were derived by assuming first, based on approximate calculations, that 10,000 kilowatts of electric power would be required for in-plant generation, and that the cost of generating and distributing this power throughout the plant would be $2 million, and second, that each of the gasification systems except Texaco required a pair of fired boilers for steam production at either 600 psig or 1100 psig and 750 F. Costs for these boilers were obtained from investments by Foster Wheeler and Riley Stoker previously furnished to Blaw-Knox, and checked against data recently published on fired boiler costs.

An exception to the stated costs for boilers is the Texaco gasification system which does not normally require a fired boiler and, therefore, has only a single 100,000 lb/hr boiler for start-up.

n. Boiler Feed Water Preparation: The cost of boiler feed water preparation has been estimated using the hot lime process to soften the available raw water for boiler feed water use, and adding the cost of boiler feed water pumps and drive equipment for pumping the feed water to the required pressure. These data are commonly available from standard cost estimating guides and no quotations were used for estimating the investment for these items of equipment. For super-pressure boilers the cost for demineralized water was used.

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(55) Katell, S. and Joyce, T. J., "What pulverizing costs," Coal Age 66 (6), 92-3 (1961).

o. Cooling Water System: The cooling water system investment costs have been estimated based on systems generally substantially smaller than the one required here; again it is expected that a detailed study of such systems would result in somewhat lower investment costs. The cooling water system costs include the cost for the supply and return piping for circulating water within the plant site.

p. Off Site Facilities: An estimate of the costs for auxiliaries such as administration buildings, shops, laboratories, steam and fresh water distribution, yard and road lighting, fire prevention equipment, sanitary facilities, railroads, roads, fences, communications systems, etc., was made by taking 14.5 percent of the total of all other investment costs.

q. Fixed Investment: A total of the above costs gives a net fixed investment to which interest on capital during construction is added. For this interest, 5 percent of the net fixed investment is obtained based on a 6 percent annual interest rate and disbursement of construction cost at the rate assumed by the Atomic Energy Commission in their report for gasification using nuclear heat.(56) Net fixed investment and interest during construction combine to give the total fixed investment, which is used as the basis for calculating maintenance costs and fixed charges.

r. Capital: Working capital was calculated on the basis of a 5-day coal supply, and 30 days sales for the gas produced, plus the cost of a 30-day supply of catalyst, and chemicals. The working capital added to the total fixed investment gives the total capital requirements for the plant.

5. Operating Cost Summary: The operating cost summary in Table 6-7 utilizes all of the economically significant statistics previously tabulated on the other summary sheets and presents total pipeline gas cost in thousands of dollars per year and in cents per M scf.

The significant operating cost items are: (a) coal at $4 per ton at the plant; (b) make-up water at 10 cents per M gal for pumped and strained river water; (c) catalyst and chemical costs, including the cost of Raney nickel methanation catalyst, gasification catalyst for the Catalytic Steam Methanation, and miscellaneous chemicals used in by-product plants and waste treatment plants; (d) limestone for slag fluxing at $2.50 per ton; and (e) operating labor at an average rate of $2.75 per hour, plus 10 percent for supervision other than shift foremen, plus 60 percent of labor and supervision for payroll overload.

The total Raney nickel catalyst replacement cost is 1 cent per M scf of methane formed in methanation. In addition, all of the processes have been charged with the same miscellaneous chemical cost of 1.36 cents per M scf of pipeline gas for chemicals for acid-gas removal, for shift catalyst, for laboratory reagents and chemicals, for make-up carbon for the activated carbon beds, for desiccant for the final gas drying, and for other unspecified chemical expense. The gasification catalyst mixture assumed for the Catalytic Steam Methanation consists of soda ash, limestone, and iron ore at an average of $7.35 per ton.

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(56) Pieroni, L. J., et al, "A technical and economic evaluation of solid fuel gasification using nuclear heat," U.S. Atomic Energy Commission, Rept. NYO-10301, prepared by The M. W. Kellogg Co., November 30, 1962.

TABLE 6-7. OPERATING COST SUMMARY FOR PROJECTED COMMERCIAL-SCALE PIPELINE GAS PLANTS (THOUSANDS OF DOLLARS PER YEAR)

		Fixed-bed Processes		Fluidized-bed Processes		Entrained Processes				Super-pressure Processes					
		Process 11	Processes 18,19,20	Process 21	Process 7	Process 61	Process 62	Process 60	Process 22	Process 56	Process 57	Process 58	Process 58	Process 58	Process 65
	Unit Price	Lurgi Dry-ash	Lurgi Slagging	Hydrocarbon Research	Bamag-Winkler Atmospheric	Rummel Single-shaft Pressurized	Rummel Modified Single-shaft Pressurized	Koppers-Totzek Pressurized	Texaco	Fixed-bed	Fluidized-bed	Two-stage R1	Two-stage R2	Two-stage R3.4	Catalytic Steam Methanation
Charges:															
Coal	$4/ton	17,850	18,600	17,050	22,600	18,500	18,850	21,300	21,650	16,900	16,500	17,000	16,250	15,850	14,500
Water Makeup	10¢/M gal	910	1,070	770	1,260	1,140	1,090	1,480	1,160	780	640	900	780	720	480
Catalysts, Chemicals	---	1,900	2,030	1,680	1,980	2,060	1,730	2,060	2,030	1,850	1,550	1,750	1,660	1,610	2,610
Limestone	$2.50/ton	---	830	---	---	---	---	---	---	---	---	---	---	---	---
Labor, Plus 10% Supervision and 60% Overhead	$2.75/hr	3,930	3,670	2,910	3,460	3,210	3,120	3,750	3,380	3,630	2,780	2,950	2,870	2,780	2,780
Operating Cost Sub-Total	$1,000/yr	24,590	26,200	22,410	29,300	24,910	24,790	28,590	28,220	23,160	21,470	22,600	21,560	20,960	20,370
Credits:															
Char	$4/27 MM Btu	---	---	---	1,900	---	---	---	---	---	---	---	---	---	---
Sulfur	$20/ton	980	1,180	1,780	1,180	2,100	1,980	2,440	2,300	920	1,780	1,820	1,720	1,680	1,380
Raw Phenols	4¢/lb	1,640	1,220	---	---	---	---	---	---	1,520	---	---	---	---	---
Sulfate	$24/ton	4,500	4,500	---	---	---	---	---	---	4,200	---	---	---	---	1,420
Credit Sub-Total	$1,000/yr	7,120	6,900	1,780	3,080	2,100	1,980	2,440	2,300	6,640	1,780	1,820	1,720	1,680	2,800
Net Operating Costs	$1,000/yr	17,470	19,300	20,630	26,220	22,810	22,810	26,150	25,920	16,520	19,690	20,780	19,840	19,280	17,570
Net Operating Costs	¢/M scf of Pipeline Gas	20.2	22.3	23.8	30.2	26.3	26.3	30.2	29.9	19.1	22.7	23.9	22.9	22.2	20.2
Maintenance (Includes Materials and Labor Plus Supervision and Overhead for Labor)	4.18%/yr of Fixed Investment	8,550	8,360	7,440	10,430	8,870	8,000	10,830	9,070	7,520	6,700	6,870	6,320	5,860	4,800
Fixed Charges (Depreciation, R. E. Taxes and Insurance, and Return on Investment)	15%/yr of Fixed Investment	30,690	30,060	26,710	37,500	31,840	28,760	38,880	32,580	27,000	24,060	24,700	22,700	21,070	17,250
Interest on Working Capital	6%/yr of Working Capital	320	320	300	410	350	320	410	370	290	280	290	270	260	240
Maintenance and Fixed Charges, Total	$1,000/yr	39,560	38,710	34,450	48,340	41,060	37,080	50,120	42,020	34,810	31,040	31,860	29,290	27,190	22,290
GRAND TOTAL	$1,000/yr	57,030	58,010	55,080	74,560	63,870	59,890	76,270	67,940	51,330	50,730	52,640	49,130	46,470	39,860
Total Pipeline Gas Cost	¢/M scf	65.7	66.9	63.5	85.9	73.6	69.0	88.0	78.3	59.1	58.5	60.7	56.7	53.6	45.9

A

B

Each of the processes makes byproducts which are sold for credit: (a) the saleable char produced in the Bamag-Winkler Atmospheric plant is credited at \$4 per 27 million Btu which equals \$2.96 per ton of char; (b) each plant produces elemental sulfur in the Vetrocoke hydrogen sulfide removal system; this is credited at \$20 per short ton; (c) the raw phenols from the Phenosolvan recovery plant are assumed to be saleable at 4 cents per pound. This credit is not based on an evaluation of the market, but was chosen to be substantially below the 10 to 11 cents per pound sale price for pure phenol; (d) ammonium sulfate has been credited at \$24 per short ton; this is also substantially below the present market price of approximately \$35 per short ton.

These credits were sub-totaled for each of the processes and subtracted from the operating cost sub-total to yield a net operating cost per year. Then these net operating costs were converted to the basis of cents per M scf of pipeline gas; this makes readily apparent that portion of the total pipeline gas cost represented by the cost of raw materials, chemicals, and labor.

For the commercial Lurgi gasification system, systems maintenance costs, consisting of direct maintenance labor and materials, were known as a percentage of the investment cost of the various plants and gas processing operations. These percentages were used in the present estimate of the Lurgi dry-ash gasification plant. Then, the labor portion of this cost was charged with 10 percent supervision and 60 percent payroll overhead. This gives an average for the Lurgi dry-ash plant of 4.18 percent annual maintenance cost based on the total fixed plant investment. This same 4.18 percent figure was used in the estimates of plants based on all the other processes.

The annual fixed charges for depreciation, real estate taxes, insurance, and return on investment, were calculated at 15 percent of the total fixed investment. A possible breakdown of these 15 percent annual fixed charges can be assumed as follows:

5 Percent	- Depreciation (20 Years)
2 Percent	- Real Estate and Personal Property Taxes and Insurance
8 Percent	- Return on Investment and Interest on Debt
15 Percent	- Total Annual Fixed Charges

If a capital structure comprised of 65 percent borrowed capital and 35 percent equity capital is assumed, with capital available at 5 percent, then 5 percent of the 65 percent equals 3.25 percent interest on borrowed capital. Then, 8 percent minus 3.25 percent leaves 4.75 percent of total capital remaining for return on investment. This 4.75 percent corresponds to 13.6 percent gross annual return on equity before taxes.

Deduction of federal income tax of 48 percent from this return on equity leaves 0.52 x 4.75 or 2.47 percent of total capital as net annual profit after taxes. This 2.47 percent is equivalent to $\frac{2.47}{0.35}$ or 7.07 percent net profit on equity.

6. <u>Pipeline Gas Cost by Lurgi Dry-ash Gasification</u>: The cost of producing pipeline gas by the Lurgi process has been previously estimated by others. Comparison of previously obtained cost data with those of this report will be pertinent. Investment costs from a recent cost estimate for the U.S. Bureau of Mines by The M. W. Kellogg Co. (57) for the gasification of anthracite are given in Table 6-8, together with similar data from the present study. The present plant having a capacity of 250 MM scfd is 2.78 times that of the Kellogg plant. This increase in capacity is accompanied by a 220 percent increase in investment cost, corresponding to a cost increase in proportion to the 0.77 power of the plant size. Not taken into consideration in this comparison of investment cost is the fact that the gasification of anthracite requires 13 percent more oxygen and 60 percent more coal per M scf of gas. The higher oxygen requirement is due to a lower methane content, 6.7 percent versus 10 percent, in the raw Lurgi gas made from process anthracite. The higher coal consumption results from both a lower methane content of the gas and higher ash content in the anthracite, that is, 25 percent versus 7.1 percent in the bituminous coal.

Similarly, the operating cost from the two estimates is shown in Table 6-9. In addition, for comparison, the Kellogg data for anthracite gasification have been adjusted to the same coal cost on a Btu basis and to the method of capital cost calculation used in this report. Even after this adjustment, the coal cost for anthracite gasification is still about 5 cents per M scf higher than that for the bituminous coal gasification. This again is a reflection of the lower methane content of the gas and the higher carbon loss in a larger quantity of ash. It may be repeated here that the coal consumption in this report is based on data from the Lurgi company and is in agreement with operating results in commercial plants.

The difference of 5.6 cents per M scf in labor cost is due to the larger plant size used in the present study. Specifically, the larger plant does not require an increase in labor for many operations; it utilizes the increased gasifier capacity recently indicated by Lurgi and a higher degree of automation.

The absence of by-product credits for the anthracite gasification is the direct result of the differences in the raw material. For the anthracite gasification, no sulfur recovery is provided, which in turn makes ammonium sulfate production uneconomical. Anthracite produces very little tar and phenols; therefore, no recovery of these is provided.

The differences in maintenance costs and fixed charges of 9.2 cents per M scf are directly attributable to the differences in capital investment costs.

In summary, the lower costs indicated in this study for the Lurgi dry-ash process in comparison to that previously given by Kellogg are attributed to:

(a) large plant size, 250 MM versus 90 MM scfd,

(b) technological progress in equipment, lower cost acid gas removal and methanation plants, higher gasifier throughput, greater automation,

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(57) "Pipeline gas and hydrogen from anthracite coal," The M. W. Kellogg Co., Rept. CE-58-189, September 19, 1958.

TABLE 6-8 INVESTMENT COST COMPARISON: PIPELINE GAS PRODUCTION BY LURGI DRY-ASH PROCESS

Source of Data	Millions of Dollars This Report	Kellogg Report	
Raw Material	Bituminous Coal	Anthracite Coal	Ratio
Plant Capacity, tons/day coal	12,860	7,400	1.74
MM scf/day gas	250	90	2.78
Coal Preparation and Storage	3.5	3.01	1.16
Gasification	54.5	25.72	2.12
Oxygen Plant	38.3	16.05	2.39
Shift Conversion	4.0	1.70	2.35
Acid Gas Removal	35.0	19.05	1.84
Methanation	4.0	1.99	2.01
Compression	2.5	1.61	1.55
Off Site Facilities and Auxiliary Plants	53.1	12.69	4.2
Net Fixed Investment	194.9	81.82	2.38
Contractors Fee	--	4.5	--
Interest During Construction	9.7	4.75	2.04
Working Capital	5.4	4.08	1.32
Total Capital	210.0	95.15	2.20

TABLE 6-9 OPERATING COST COMPARISON: PRODUCTION OF PIPELINE GAS BY LURGI DRY-ASH PROCESS, CENTS/M SCF

Source of Data:	Kellogg Report Unadjusted	Kellogg Report Adjusted	This Study
Raw Material:	Anthracite Coal	Anthracite Coal	Bituminous Coal
Plant Capacity, MM scfd:	90	90	250
Coal	41.0	25.5	20.6
Labor	10.1	10.1	4.5
Catalysts, Chemicals	3.3	3.3	2.2
Water, Supplies	1.3	1.3	1.1
By-product Credits	--	--	(8.2)
Sub-total	55.7	40.2	20.2
Maintenance	13.3	11.9	9.8
Fixed Charges, Including Interest on Working Capital	48.1	44.6	35.7
Sub-total	61.4	56.5	45.5
Total Pipeline Gas Cost ¢/M scf	117.1	96.7	65.7

(c) better fuel, lower ash content, more methane in the primary gas, and

(d) by-product recovery.

7. Relative Costs of Pipeline Gas by Various Processes:

a. Advantage of Pressure Operation: Evaluation of available commercial coal gasification processes shows that operation at atmospheric pressure and compression of the raw gas to pipeline pressure is more costly than gasification at elevated pressure. This is illustrated in detail in the data for the two fluidized-bed processes--the Bamag-Winkler process operating at atmospheric pressure, and the Hydrocarbon Research process operating at 450 psi. (See Table 6-7.) For the former process, data from many commercial plants are available and extrapolated costs of units larger than used heretofore have been obtained from the Bamag company. For the Hydrocarbon Research process, data from a large pilot plant producing 650 M scfd of gas at pressures as high as 245 psi were used as a basis.

The main reasons for the greater economy of elevated pressure operation are that (a) the costly raw gas compression is avoided, and (b) the direct exothermic formation of methane at elevated pressure in the fluidized bed leads to a drastically decreased oxygen consumption.

The savings that are due to these factors lead to greatly decreased capital investment and operating cost for gasification at 450 psi pressure as shown graphically in Figures 6-3, 6-4, and 6-5, showing investment costs of $183 and $257 million, and operating costs of 63.5 and 85.9 cents per M scf of 928 Btu per scf pipeline gas, respectively.(58)

b. Selection of Lurgi as Bench Mark: The atmospheric pressure Winkler gasifier uses a fluidized fuel bed and was the first process used on a large scale for the gasification of fine coal with oxygen. The next process used commercially was the Lurgi pressure gasifier using coal in a fixed bed. This is the only coal gasification process which has been and is being used commercially at pressures up to 450 psi. For this process, data based on cost of actually built gasifiers, from Lurgi, Frankfurt, were obtained. For this reason, the costs of commercial Lurgi dry-ash process were studied in considerable detail and used as a bench mark, and to some extent as a basis for the cost of the other processes. The data from the present evaluation show that with this commercial process, pipeline gas could be produced at a cost of 65.7 cents per M scf in a plant costing $210 million.

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(58) Basis of cost evaluation, see Appendix 5.2; it is briefly:

Plant Size: 250 MM scf gas per day of 928 Btu per scf
Coal Cost: $4 per ton (15 cents per MM Btu)
Fixed Charges: 15 percent of total fixed investment
Labor Cost: $2.75 per hour plus 70 percent supervision and payroll overhead
Make-up Water: 10 cents per M gallons
Load Factor: 95 percent (347 operating days per year)

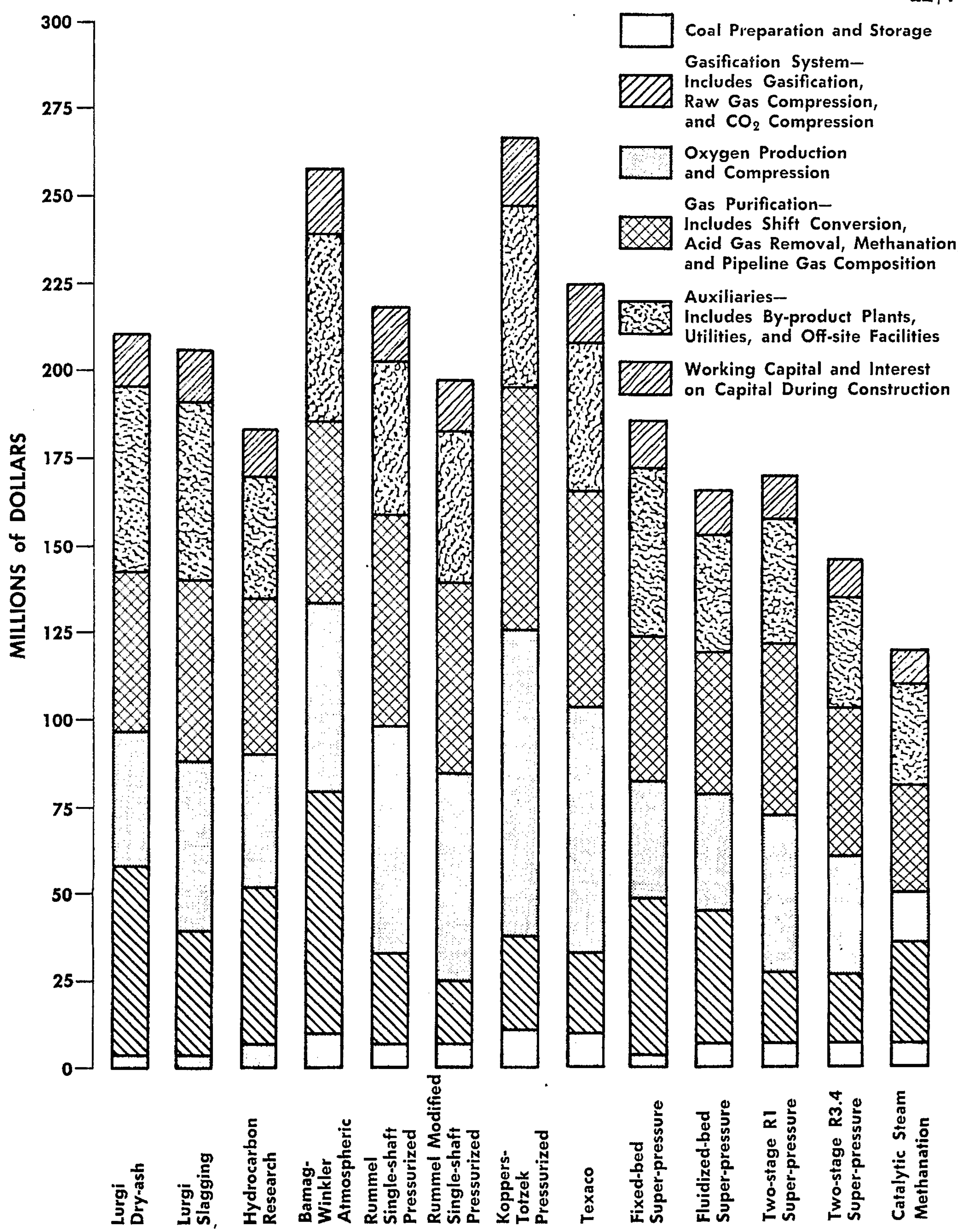

Figure 6-3 Comparative Investment Costs for 250 MM scfd Pipeline Gas Plants Based on Coal (BCR 8006G167)

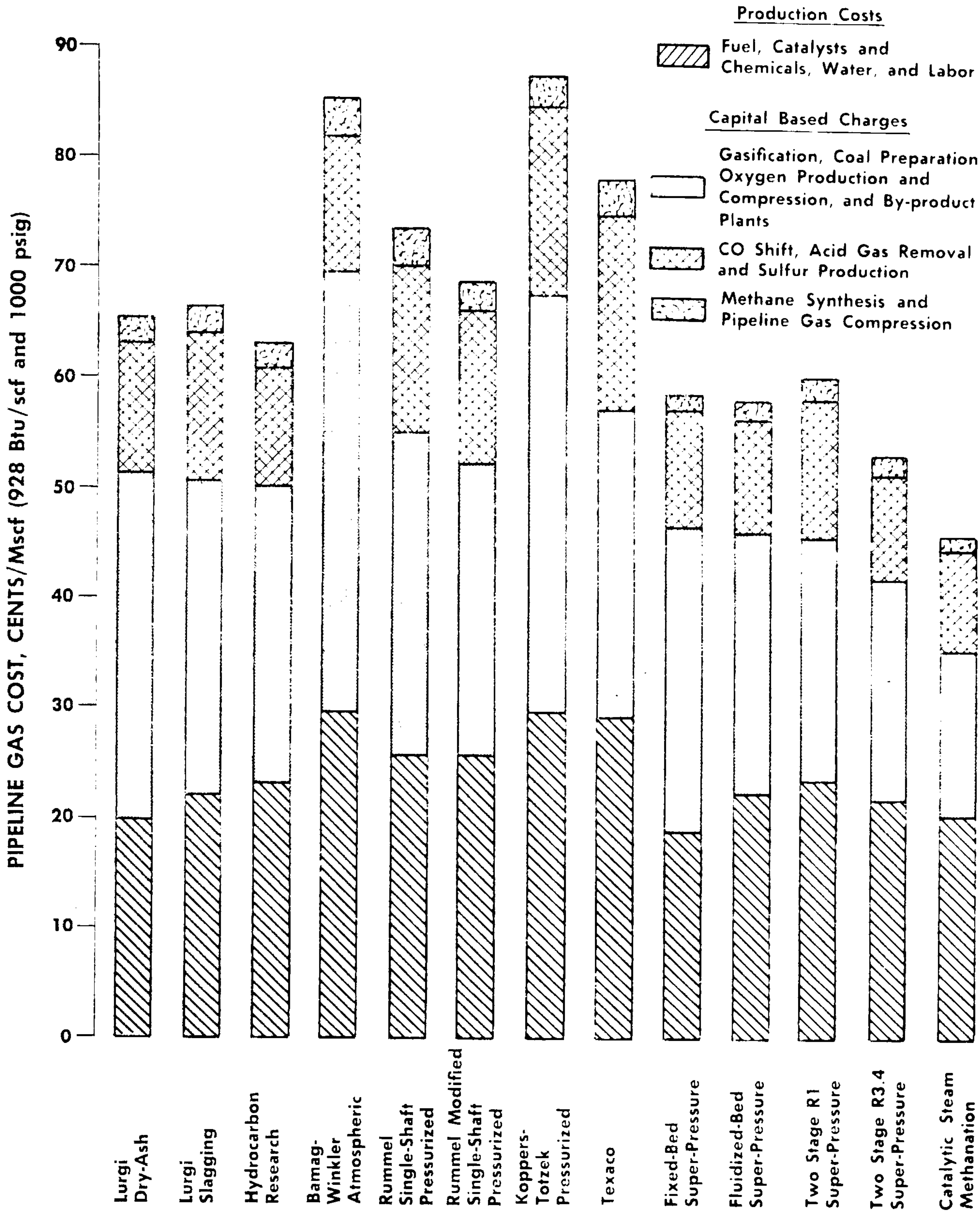

Figure 6-4 Comparative Costs for 250 MM scfd Pipeline Gas from Coal by Selected Processes Showing Purification and Methanation Components (BCR 8006G127)

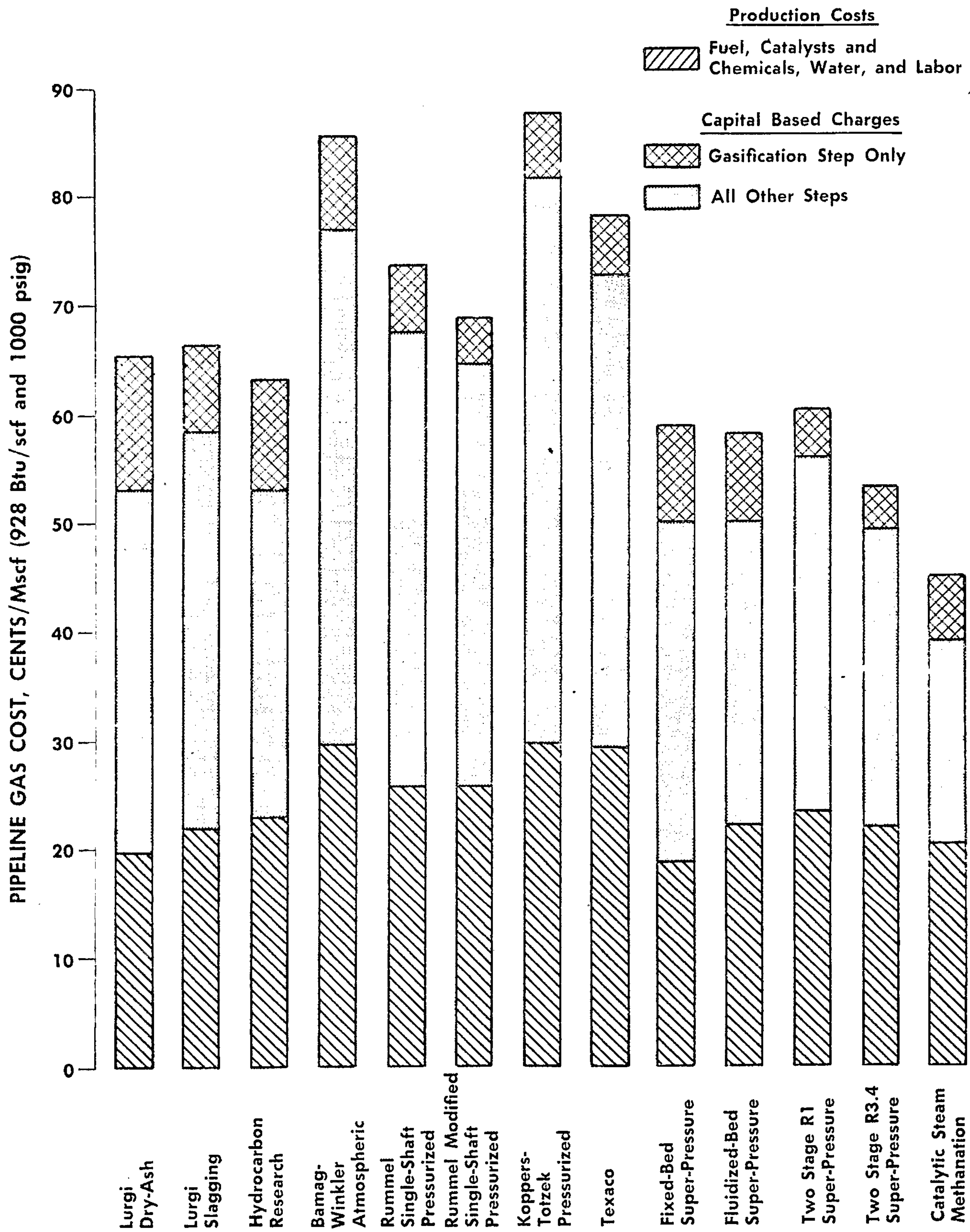

Figure 6-5 Comparative Costs for 250 MM scfd Pipeline Gas from Coal by Selected Processes Showing Gasification Component (BCR 8006G128)

c. Dry-ash Versus Slagging Operation: The Lurgi dry-ash process uses a large excess of steam over that needed for gasification to avoid melting of the coal ash. In slagging operation this excess steam is not needed and an increase in the gasifier capacity results. This leads to a lower investment cost of the gasifier. However, this saving is offset by higher oxygen consumption which leads to higher cost of the oxygen and acid gas removal plants. Thus, investment and operating costs of these two Lurgi process versions are close together. Selection of the individual process would depend upon properties of the coal ash. For coals with low ash melting point, the slagging process would be preferable.

d. Fixed-bed Operation with Caking Coal: The Lurgi process originally was developed for non-caking lump coals and was later adapted and found suitable for caking coals. With the highly swelling Pittsburgh seam coal, a short experiment was made in a commercial gasifier with slag added to the coal. In this short run, operation was satisfactory; however, performance data were not obtained. Thus, coal quality is a point that needs attention when the Lurgi process is contemplated for use.

e. Advantage of Entrained Gasification Processes: The gasification processes that use coal in suspension have the widest latitude as far as coal quality is concerned. Therefore, and because of the simple gasifier design that is suitable to the building of large units, entrained gasification systems have been investigated in greater detail. Commercial plants for atmospheric pressure operation have been built using the Koppers-Totzek process in several plants and the Babcock and Wilcox, duPont, and the Rummel processes in one plant each. Operation at elevated pressure has been demonstrated in pilot plants by the U.S. Bureau of Mines, the Institute of Gas Technology, and the Texaco Development Corporation. Common to all these processes is a short residence time of the coal in the gasifier. To attain the high reaction rate required, operation is at high temperature; this results in a low methane content in the gas and a high oxygen consumption. Thus, the gas from entrained processes is more expensive for the production of methane.

f. Operation at Super Pressure--1050 psi: In view of the greater economy of coal gasification at 450 versus 15 psig, the cost of operating at 1050 psi was investigated. This pressure was selected somewhat arbitrarily as suitable for direct delivery into a pipeline. Considerable reductions in investment and operating costs were obtained for fixed-bed and fluidized-bed operations. Both give a pipeline gas cost below 60 cents per M scf of 928 Btu per scf pipeline gas.

The costs of all entrained gasification processes as estimated for operation at 450 psi are higher than those obtained for the fixed-bed and fluidized-bed processes operated at the same pressure.

g. Two-stage Super-pressure Operation: The possibility of a decrease in oxygen consumption for entrained processes is indicated by two-stage operation in one gasifier unit. In the first stage, recycle char is gasified with oxygen. Into the hot gas stream coming from this stage, the fresh coal is injected and thus, the volatile matter is gasified rapidly. Since the fresh coal does not pass through the high temperature zone in the presence of oxygen, a gas containing methane is obtained in the second stage. This leads to smaller oxygen and acid gas removal plants and to a cost reduction.

The Otto company supplied a cost estimate for operation of such an entrained gasifier. Those data are the basis for the cost given for the Rummel modified single-shaft pressurized process.

h. Methane Formation in Primary Gas: The concentration of methane in the gas from the primary gasification step increases not only with increasing pressure and decreasing temperature, but also with increasing activity of the carbon being gasified. Squires has correlated the data obtained by various investigators (59) and his "average" curve is shown in Figure 6-6. The carbon activity of anthracite in fluidized-bed gasification was found by Squires to be about 3.4 times that for beta graphite.

In the present studies, the carbon activity for high volatile bituminous coal has also been taken as equal to or greater than that for graphite, and cost data for two-stage gasification have been developed using activities of 1, 2, and 3.4 times that for beta graphite. Increasing the carbon activity from 1 to 3.4 decreases the cost of the final pipeline gas by about 7 cents per M scf as shown in Figure 6-7. An even greater carbon activity for high volatile bituminous coal may be observed experimentally; if so, then the pipeline gas cost will be decreased even further.

i. Cost of Coal: The effects of cost of coal on final cost of pipeline gas by two-stage super-pressure entrained gasification is shown in Figure 6-8. With coal at $3 per ton and a carbon activity of 3.4, pipeline gas would cost approximately 49 cents per M scf as compared to 53.5 cents for coal costing $4 per ton. Thus, a reduction in cost of coal of $1 per ton would reduce the cost of pipeline gas by 4.5 cents per M scf.

j. Fixed Charges: In the present studies, annual fixed charges are 15 percent of total fixed investment. Other rates for fixed charges have been used by others in estimating cost of pipeline gas from coal. The effect of different rates of computing fixed charges on the final cost of pipeline gas is shown in Figure 6-9; again as derived from the evaluation of two-stage super-pressure entrained gasification at two levels of carbon activity. A reduction in the annual fixed charges from 15 percent to 10 percent lowers the final cost of pipeline gas by about 9 cents per M scf.

k. Catalytic Gasification: An increase in the methane content of the primary coal gasification gas and thus, a further reduction in cost, is indicated for a process that would combine gasification temperatures lower than used heretofore with an increased rate of reaction between coal and steam to form methane directly. Assuming that a satisfactory catalyst can be found, a cost estimate has been prepared for the Catalytic Steam Methanation process, based on 1250 F reaction temperature and 1050 psi pressure.

The results indicate that the cost of pipeline gas by this process would be some 7 or 8 cents per M scf lower than by any of the other proposed processes; however, it must be stated that the validity of the estimate for the Catalytic Steam Methanation process depends entirely upon the successful development of a

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(59) Squires, A. M., "Steam-oxygen gasification of fine sizes of coal in a fluidised bed at elevated pressure," Trans. Inst. Chem. Engrs. 39, 3-27 (1961).

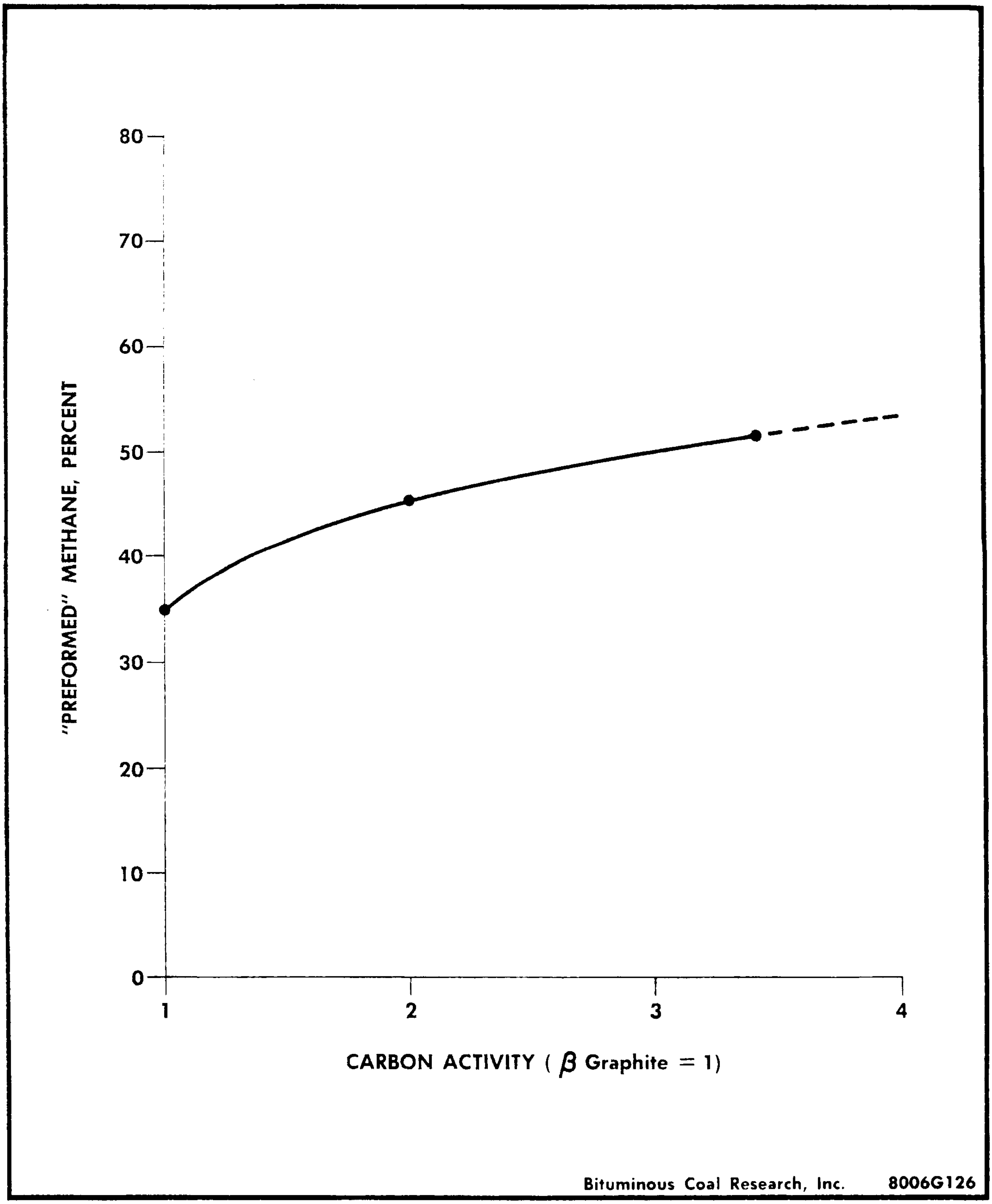

Figure 6-6 Effect of Carbon Activity on "Preformed" Methane Production in Two-stage Super-pressure Entrained Gasification

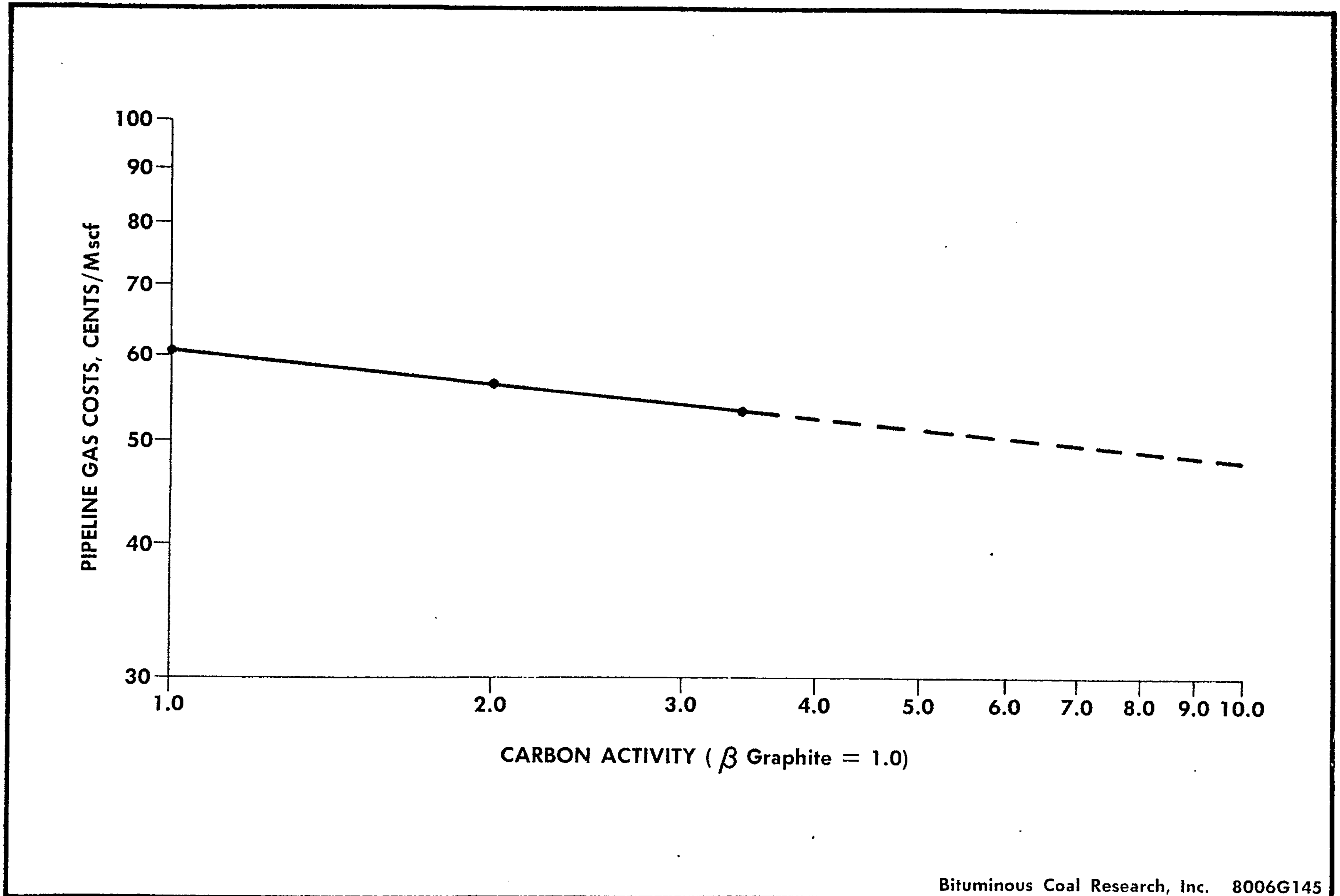

Figure 6-7 Effect of Carbon Activity on Cost of Pipeline Gas by Two-stage Super-pressure Entrained Gasification

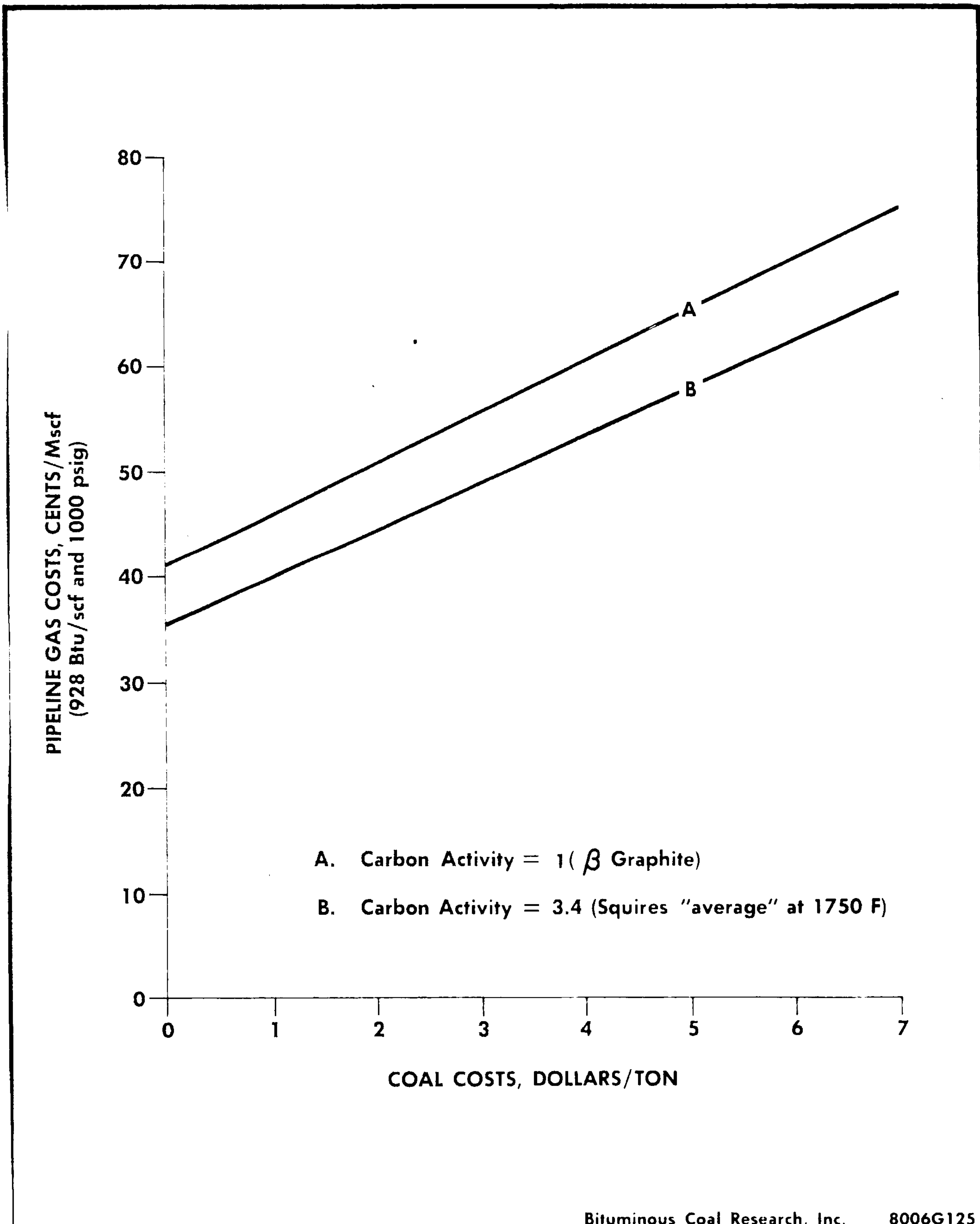

Figure 6-8 Effect of Coal Cost and Carbon Activity on Pipeline Gas Cost for Two-stage Super-pressure Entrained Gasification

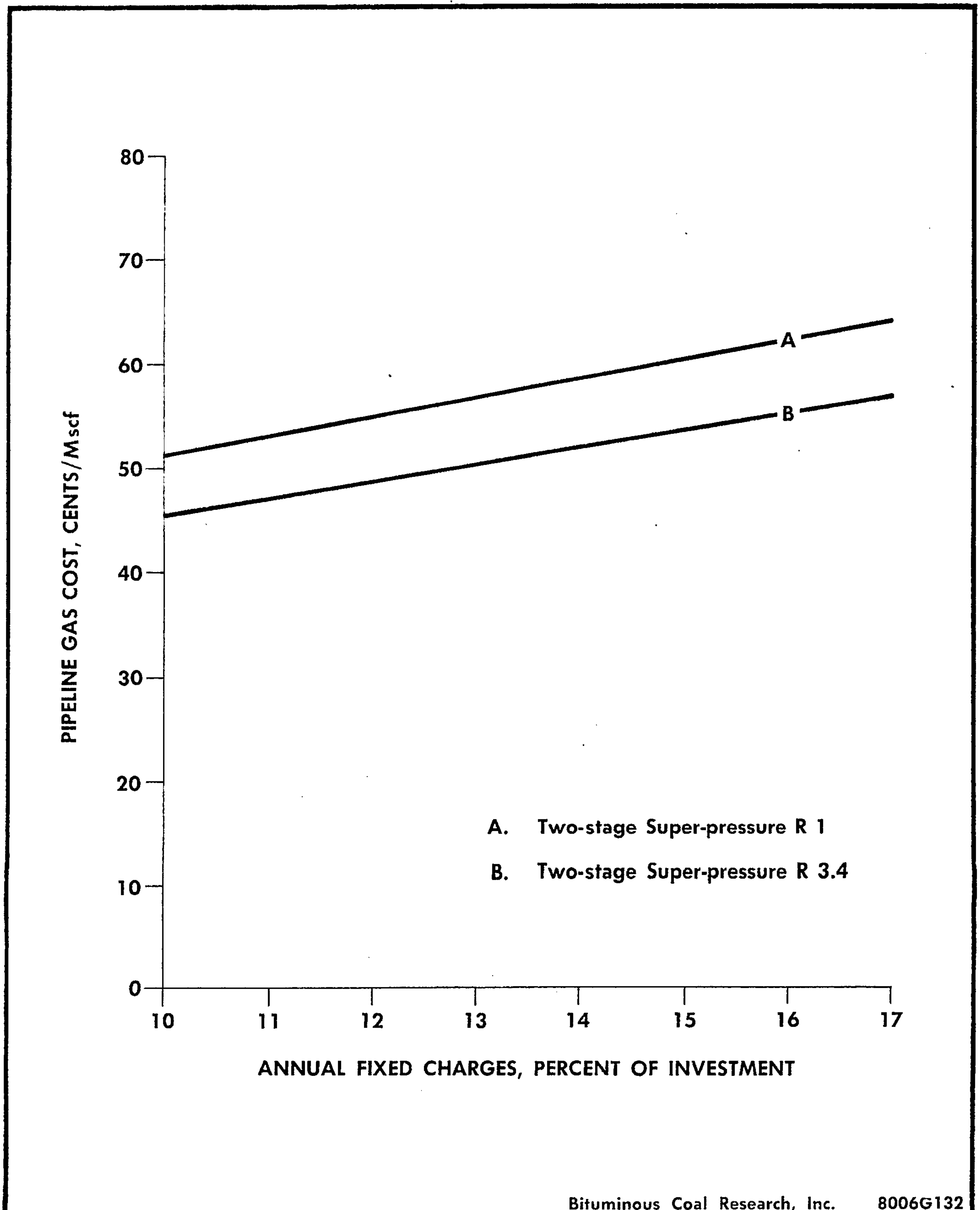

Figure 6-9 Effect of Annual Fixed Charges on Cost of Pipeline Gas from Coal Based on Two-stage Super-pressure Entrained Gasification

suitable catalyst or catalyst combination. It is thus a hypothetical case. By contrast, the two-stage super-pressure entrained gasification process is based upon the application and extrapolation of existing technology to higher pressure and larger units.

D. Conclusions

Certain conclusions are readily reached from a review of the economic data as assembled in the present studies on the various proposed schemes for large pipeline gas plants based on coal.

First, the processes which do not produce an appreciable amount of methane in the raw gas from gasification, such as the Bamag-Winkler Atmospheric, the Koppers-Totzek, the Rummel single-shaft, and the Texaco, cannot compete with those processes which do produce an appreciable amount of methane in the gasification step. The high cost of the non-methane producers is associated with high oxygen consumption for gasification, and is subsequently reflected in high acid-gas removal costs.

Second, it is cheaper to operate those gasification units which do produce appreciable methane at pipeline pressure rather than at some intermediate pressure such as 450 psig. This is due principally to the increased formation of methane at the higher pressure, and to the lower cost of the smaller number of units required to process the gas at the higher pressure.

Since economic pipeline production processes are those processes which operate at high pressure and which produce substantial amounts of methane in the raw gas from gasification, it follows that the greatest economic potential can be realized by a process that has high carbon conversion as well as high methane formation. Such a process is the two-stage process, which completely gasifies carbon in the lower stage under slagging conditions, and which forms methane by devolatilization of coal and by reaction between char and hydrogen in the upper stage.

It can also be concluded that if a catalyst is found with sufficiently low cost and sufficiently high activity to achieve the gasification parameters assumed for the Catalytic Steam Methanation Gasifier, then a potential exists for reducing the cost of pipeline gas below that by any of the other processes evaluated in this study.