APPENDIX 3.1

LITERATURE SEARCH - ANNOTATED BIBLIOGRAPHY

See Volume II

APPENDIX 3.2

LITERATURE SEARCH - ADDITIONAL REFERENCES

See Volume II

APPENDIX 3.3

ORGANIZATIONS CONTACTED ON SURVEY OF DEVELOPMENTS IN COAL GASIFICATION

UNITED STATES (by Field Interview)

Babcock & Wilcox Company	Research Center	Alliance, Ohio
Battelle Memorial Institute		Columbus, Ohio
Chemcoke, Inc.	Pilot Plant	Columbia, Tenn.
Con-Gas Service Corporation		Cleveland, Ohio
Consolidation Coal Company	Research & Development Lab.	Library, Pa.
Hydrocarbon Research, Inc.	Headquarters Research Laboratory	New York, N. Y. Trenton, N. J.
Illinois Institute of Gas Technology		Chicago, Ill.
McDowell-Wellman Engineering Company	Headquarters and Research Laboratory	Cleveland, Ohio
M. W. Kellogg Company	Headquarters	New York, N. Y.
Riley Stoker Corporation		Worcester, Mass.
Salem-Brosius, Inc.		Carnegie, Pa.
Texaco Research and Development Corp.	Headquarters	New York, N. Y.
U.S. Bureau of Mines, Headquarters Morgantown Coal Research Center		Washington, D. C. Morgantown, W. Va.

ENGLAND (By Field Interview)

British Coal Utilisation Research Association		Leatherhead, Surrey
Central Electricity Generating Board, Marchwood Engineering Laboratories		Southampton
Constructors John Brown Limited		London
Humphreys & Glasgow Limited		London
International Furnace and Equipment Co. Ltd.		Brierley Hill
Ministry of Power, Chief Scientist's Division		London
National Coal Board,	Headquarters	London
	Coal Products Division; Research and Development; Coal Research Establishment	Stoke Orchard
	Midlands Regional Headquarters	Dudley
	Coleshill Gas Works	Coleshill
	Coal Briquetting Pilot Plant	Birch Coppice
The Gas Council,	London Research Station	London
	Midlands Research Station	Solihull
Woodall-Duckham Construction Co. Ltd.		London

ENGLAND (By Mail)

Imperial Chemical Industries Limited	London
The Power-Gas Corporation Ltd.	Stockton-on-Tees

GERMANY (By Field Interview)

Badische Anilin-& Soda-Fabrik A.G.	Ludwigshafen
Bergbau-Forschung GmbH	Essen-Kray
Demag, A.G.	Duisburg
Dr. C. Otto and Company GmbH	Bochum
Heinrich Koppers GmbH	Essen
Lurgi Gesellschaft fur Warmetechnik mbH	Frankfurt
Pintsch Bamag A.G.	Butzbach
Ruhrgas, A.G.	Essen
Steinkohlengas A.G. Dorsten Gas Works	Dorsten
Union Rheinische Braunkohlen Kraftstoff, A.G.	Wesseling

FRANCE (By Mail)

Centre d' Etudes et Recherches des Charbonnages de France (CERCHAR)	Paris
Gaz a l'eau et Gaz Industriel (GEGI)	Montrouge (Seine)

NETHERLANDS (By Mail)

Staatsmijnen in Limburg	Geleen

AUSTRIA (By Mail)

Vergasungs Industrie A.G. (VIAG)	Vienna

APPENDIX 3.4

SUMMARY OF EUROPEAN SURVEY TRIP

The highlights of the European Survey Trip are summarized in the form of double page tables. The dates, places, organizations, and people interviewed are listed on the left-hand page, and remarks are listed on the right-hand.

SUMMARY OF EUROPEAN SURVEY TRIP

Date	Place	Organization	People
July 20 and 21	Leatherhead	BCURA Basic Research	Allan, Badzioch, Bond, Brown, Davidson, Dryden, Hicks, Hoy, Jackson, Sparham, Wilkins
July 22	London	The Gas Council London Research Station	Hopton, MacCormac, Wrobel
July 23	London	N. C. B. Hobart House	Brown, Greenwood, Skinner, Wynn
July 23	London	Woodall-Duckham	Randall, Richards, Taylor
July 24	London	Constructors John Brown	Guter, Masterman (Chief Scientist Div., M.o.P.) Stanislas
July 24	London	Humphreys & Glasgow	Balfour, Phillips
July 27	Coleshill	N. C. B. Gas Plant	Harvey
July 27	Solihull	Midlands Research Station The Gas Council	Hebden, Moignard, Moseley
July 28	Brierley Hill	International Furnace Equipment Co., Ltd.	Gee, Keane, Temple

SUMMARY OF EUROPEAN SURVEY TRIP
REMARKS

Operation of slagging gasifier for Ministry of Power terminated and work summarized in technical papers and in series of seven reports which have been distributed to USBM, OCR, BCR, and others. Results are a major contribution to our understanding slag behavior in gasifiers. Work on slagging units transferred to Midlands Research Station of The Gas Council. Work continues on pressurized devolatilization of coal in cooperation with Central Electricity Generating Board. Work expected to begin on air preheater for MHD with support by Ministry of Power. Considerable amount of basic research in progress.

Work on the Rummel double-shaft slagging gasifier terminated. Only 60 percent efficiency expected under most optimistic conditions. Heat losses too great--results to be given in paper before Institution of Gas Engineers this fall. No further work planned.

Plans for additional Lurgi plants shelved in face of increasing amounts of imported natural gas, and of cheap petroleum feedstocks for reforming into town gas. A major effort is being exerted to manufacture and merchandise smokeless briquets made from coal for open home fires. Coal production apparently is to be maintained at 200 MM tons per year as a matter of national policy.

Woodall-Duckham built the Coleshill Lurgi plant; one of three to prepare bids for proposed Eustead Lurgi plant of NCB. Has conducted no research nor development. W-D considers a super pressure Lurgi worthy of consideration.

Constructors John Brown built and operated slagging gasifier at Leatherhead for Ministry of Power. Has conducted limited experimental program of its own. CJB prepared one of the three bids for proposed Eustead Lurgi plant of NCB. No further work planned.

Humphreys & Glasgow has lost its mechanical gas producer business and with it its interest in coal gasification. H & G is a licensee for processes developed by Midlands Research Station of The Gas Council. H & G built the Westfield gasification plant, and was one of three who submitted bids on proposed Eustead plant.

Coleshill Lurgi gasification plant now operating reasonably satisfactorily after considerable start up trouble. There is no net production of tar. Gas production can be sustained at 110 percent of design, being limited by shift conversion plant. Costs for next plant could be cut considerably.

Program on pressurized slagging gasifier to be curtailed end of this year. A unit three feet six inches ID has been operated at 8 MM scf/day. Work on naphtha reforming is being given top priority. Fluidized bed hydrogenation studies will be continued after the urgency of the naphtha reforming program has passed. Some work is being done at super pressures and temperatures--500 atm and 1000 C.

Several automated two-stage gas producers are in operation at Sheffield. Costs are estimated to be lower than single stage units and gasification rate to be higher. Cost estimate of a large unit will be supplied.

July 28	Birch Coppice	N. C. B. Briqueting Pilot Plant	Jenkins, Nicole, Urquhart
July 29	Stoke Orchard	N. C. B. C. R. E.	Dainton, Gordon, Gregory, Kingsmill, Rhys-Jones, Standing, Troughton, Watson
July 31	Marchwood	CEGB Marchwood Engineering Laboratories	Elliott, Hughes, Johnson, Myerscough
August 3	Dorsten	Steinkohlengas A.G.	Bieger, Just, Schorfl, Weittenhiller
August 4	Essen	Ruhrgas A.G.	Brecht, Doering, Just, Peters, Sommers
August 5	Essen-Kray	Bergbau-Forschung	Beck, Koelling, Peters, Reerink

Pilot plant now operating at 350 tons briquets per week. Commercial plant is under construction at Coventry, and plant in Yorkshire is to be doubled. Roll presses are eventually to replace extrusion presses. Briqueting represents major costs of process.

The NCB Coal Research Establishment employs some 400 people, including 70 professional personnel, and operates on a yearly budget of 600,000 pounds sterling or $1,700,000. Facilities are available for preparing, devolatilizing and briqueting coals, as well as for evaluating resultant briquets in actual use in commercial home appliances. Several pilot units ranging in size from 8 inches to 48 inches in diameter are used. A major program of assaying available coals is in progress. Pilot unit using hot char recycle for heat is being operated to study optimum production of gas and char for briqueting. Bench-scale studies on kinetics of devolatilization are in progress. CRE has developed a fluidized unit for feeding coal under pressure.

Fluidized carbonization of coal at 5 atm in 8-inch ID unit is being studied to supply feed for gas turbine and char for steam boilers. Plans underway for 30-inch ID unit. An increase from 38 to 43 percent in overall efficiency of power station is expected. Bench-scale studies on non-slagging combustion of coal in fluidized bed indicate very high heat release rates are possible; sulfur retained in refuse.

Lurgi coal gasifiers being converted to naphtha-reforming units. Gas production is controlled by availability of gas from oven coke plant. Maximum capacity of plant never used. One unit is partially automated. Quality of coal used is quite variable. Usual problems connected with tar are encountered. Reinjection of tar and liquor are not practiced.

Ruhrgas stopped all pilot work on coal gasification about two years ago. Pilot units in stand by at Herten include the first and second vortex gasifier units, and the second unit for the LR Process. A larger LR Process unit was operated at the power plant of the Dorsten colliery. Cost estimates on LR Process were made available.

Bergbau-Forschung employs 600 people, including 110 professionals, with support primarily from coal producers; contract research is accepted with most of it in the Mining Department and coming from private suppliers and hydraulic equipment manufacturers. Discussions and tour of facilities were limited to those items of direct interest to the gasification program. Effect of catalysts on the reactivity of cokes, catalytic reforming of tar in raw gases from gasification processes and factors affecting spontaneous combustion of coal in underground mines and in storage piles, all were topics of investigation and discussion. B-F works closely with a subsidiary in the commercial scale development of processes originating at B-F. Pilot plant currently being erected for production of briquets by blending hot char with coking coals; project financed by the government. Pilot plant for studying briqueting problems of private industry is widely used. Pilot plant is also being used to study manufacture of active carbon.

August 6	Essen	Heinrich Koppers G.m.b.H.	Blank, Boenneman, Daniels, Fitz, Totzek
August 7	Bochum	Dr. C. Otto & Comp. G.m.b.H.	Domann, Marwig
August 10	Duisburg	Demag A.G.	von Neudeck
August 11	Wesseling	Union Reinische Braunkohlen Kaftstoff A.G.	Huettner, Penning, Wissel
August 12	Ludwigshafen	Badische Anilin-& Soda-Fabric A.G.	Jackh, Knobloch, Schreck
August 13	Frankfurt	Lurgi Gesellschaft fur Warmetechnik m.b.H	Hager, Kapp, Rudolph, Schmalfeld
August 14	Butzback	Pintsch Bamag A.G.	Doering, Riedel, Schilling, Senner

Heinrich Koppers currently designing and building a slagging Koppers-Totzek gasifier for 30 atm operation in mid-1965. H. K. has developed extrusion feeder for introducing coal under pressure. Slagging Koppers-Totzek now operating in Spain, Greece, and Japan. No data available for pressure operation.

Otto has license for Rummel gasifier from Union Kartstoff. Rummel single shaft gasifier 1.9 m ID was operated on oxygen sufficiently to eliminate operational problems. Costs are reported to be below Lurgi or Koppers-Totzek. Will supply cost estimate to BCR.

Demag has developed gas producer for caking bituminous coal; three adjustable compactors are mounted on top of producer for depressing the surface of the fuel bed. In 1955 a cost study on BASF Flesch-Demag, Ruhrgas Vortex, and Demag P-H units was made; results were made available.

U-K is now operating 12 Demag P-H units on Braunkohle and two Winkler dry bottom units on Braunkohle fines. These will be shut down in near future. Coal is being abandoned in favor of petroleum naphtha.

BASF is now operating four slagging gasifiers on coke supplemented by naphtha, but expect to replace them in two years by Texaco high pressure oil gasifiers. BASF has no further interest in coal at present.

Lurgi has conducted considerable research on continuous devolatilization of coal with recycled coke as heat carrier. The process, known as the LR Process, could be practical for supplying low yields of 600-800 Btu/cu ft gas together with high yields of tar. No pilot plant research in progress at present. Will supply cost estimate for Lurgi plant to BCR.

Pintsch Bamag has stopped coal gasification research. P. B. has built and operated several Winkler plants on bituminous coal and believes pressure operation possible. Will supply cost estimate for large Winkler units to BCR.

APPENDIX 3.5

SURVEY DATA AND INFORMATION FILE ON GASIFICATION PROCESSES

See Volume II

APPENDIX 4.1

ESTIMATED COST OF FEEDING COAL INTO PRESSURE GASIFIERS

The principal means for feeding coal into pressure gasifiers is by a lock hopper system or by pumping coal as a slurry. Costs for these two systems have been estimated in the present study.

A. Cost of Feeding Coal by Lock Hopper System

The main cost items in a lock hopper system are cost of lock hopper gas, cost of compressors, and cost of power.

1. Cost of Lock Hopper Gas: For Lurgi plants operating at 25 atm pressure, the Lurgi company stated that 2.5 percent of the total gas is being used for pressurizing the lock hoppers. This gas is then available at atmospheric pressure for use as fuel gas or for recompression and reuse as lock hopper gas.

Using 119 lb coal per MM Btu, and 300 Btu/scf in the raw gas, one obtains the following for the Lurgi process: one ton coal gives 56,000 scf raw gas and 2.5 percent of 56,000 equals 1400 scf lock hopper gas. At 25 atm, this quantity of lock hopper gas then amounts to 56 actual cubic feet of gas per ton of coal.

A coal density in the lock hopper is assumed as 75 percent of the actual coal bulk density; this corresponds to 36 lb/scf (75 percent of 48 lb/cu ft). With a coal specific gravity of 1.33, or 83 lb/cu ft, the lock hopper content is 43 volume percent coal and 57 volume percent gas. The volume of gas in the lock hopper is proportional to the operating pressure.

Thus, the data in the first two columns of Table 4.1-1 on the quantity of needed lock hopper gas are obtained.

The cost of the discharged lock hopper gas in column 3, Table 4.1-1, is charged at 25 cents per MM Btu, the difference between an assumed raw gas cost of 40 cents per MM Btu and coal cost of 15 cents per MM Btu ($4.00 per ton).

An alternative to using the lock hopper gas as fuel is to recompress it for recycle. Column 4, Table 4.1-1, shows first, the amount of power required for compression from 1 atm to the operating pressure, and then second, the cost of that power charged at 0.8 cent per kwh.

At the low value used for the raw gas (40 cents per MM Btu), the cost of power for recompression of the spent lock hopper gas, or compression of another gas, is comparatively high. Use of an inert gas (N_2, CO_2, etc.) especially if already available at elevated pressure, could be advantageous, particularly in cases where a comparatively low Btu gas is not produced.

The relatively high cost of gas for the pressurizing of lock hoppers suggests the use of steam for this purpose. In Table 4.1-2, the cost elements of such a system--steam cost, condensing equipment cost, and cooling water--are tabulated. The cost, using steam at 35 cents per M lb and cooling water at 2 cents per M gal are, except for operations at 10 atm, somewhat lower than

TABLE 4.1-1 SUMMARY OF LOCK HOPPER GAS COSTS

Operating Pressure		Volume of Lock Hopper Gas	Cost of Lock Hopper Gas		Compression Power Cost Data		
Atm	Psi	Scf/Ton Coal	Cents/Ton Coal	Cents/MM Btu in Gas	Kwh/Ton Coal	Cents/Ton Coal	Cents/MM Btu in Gas
10	147	560	4.2	0.25	1.7	1.4	0.08
25	367	1,400	10.5	0.63	6.0	4.8	0.29
50	735	2,800	21.0	1.25	15.0	12.0	0.72
75	1,100	4,200	31.5	1.87	25.2	20.2	1.21
100	1,470	5,600	42.0	2.50	36.8	29.5	1.75

TABLE 4.1-2 COSTS FOR USE OF STEAM IN LOCK HOPPERS
FOR 15,000 TON PER DAY GASIFICATION PLANT

1.	Operating Pressure,					
	atm	10	25	50	75	100
	psi	147	367	735	1,100	1,470
2.	Saturated Steam, F	364	437	508	556	594
3.	Steam Density, cu ft/lb	2.83	1.24	0.62	0.40	0.28
4.	Steam Required per ton coal, lb	20	45	90	140	200
5.	Steam Costs at 35 Cents/M lb					
	cents/ton coal	0.7	1.6	3.2	4.9	7.0
	cents/MM Btu in gas	0.04	0.08	0.16	0.25	0.35
6.	Steam Flow, M lb/hr	12.5	27	56	87	125
7.	Condenser Investment, M dollars*	25	37	70	95	135
8.	Condenser Operation Cost at 15% of Capital, dollars/hr	0.50	0.69	1.31	1.78	2.53
9.	Condenser Operating Cost, cents/ton coal	0.08	0.11	0.21	0.28	0.41
10.	Cooling Water** cents/ton coal	1.20	2.73	5.40	8.40	12.00
11.	Total (Items 5, 9, 10) Excluding Coal Preheating and Labor,					
	cents/ton coal	1.98	4.44	8.81	13.58	19.41
	cents/MM Btu in gas	0.10	0.22	0.44	0.68	0.97

* For vented steam, see Appendix 5.1, Figure 5.1-10.

** Use 30 F rise = 30 gal/lb steam at 2 cents/M gal.

compression cost for gas. The costs do not include preheating of the coal. The saving in cost by the use of steam does not seem to warrant its consideration as a general replacement for lock hopper gas, considering the complications introduced by the handling of preheated coal, especially in case of interruptions to normal operation. However, in special cases, e.g., if fixed-bed, countercurrent processes should prove advantageous, use of steam without coal preheat should be studied in more detail.

2. Capital Cost of Lock Hoppers: The economic analysis of the CO_2 acceptor and the hydrogasification processes by the Bureau of Mines (1) contains data from which costs have been derived for the feeder and lock hopper system. (See Table 4.1-3.)

3. Cost of Compression: The Bureau of Mines estimate (1) also includes compressors for the pressurizing of the feed and lock hoppers. The data are given in Table 4.1-4, together with calculated operating costs.

4. Overall Costs: The cost of feed and lock hoppers, Bailey feeders, gas compression at 15 percent of capital cost, power requirements at 0.8 cent per kwh corrected to a coal bulk density of 36 lb per cu ft, but without labor cost, are summarized in Table 4.1-5 as derived from Tables 4.1-2 and 4.1-3.

The coal feeding costs, as they depend on operating pressure, are shown in Figure 4.1-1. Straight line interpolation appears reasonably accurate, comparing the data with the trend of power consumption from Table 4.1-2.

B. Cost of Pumping Coal as Slurry

The volume of a 50 weight percent coal slurry containing 1 ton of coal is calculated as follows:

$$1 \text{ ton coal} = \frac{2{,}000 \text{ lb}}{83 \text{ lb/cu ft}} = \underline{24.1} \text{ cu ft}$$

$$1 \text{ ton } H_2O = \frac{2{,}000 \text{ lb}}{62.4 \text{ lb/cu ft}} = \underline{32.1} \text{ cu ft}$$

$$\text{Slurry} \qquad 56.2 \text{ cu ft}$$

Pumping costs of a 50 percent slurry using 75 percent pump efficiency are listed in Table 4.1-6.

To the pumping cost, a cost of the special preheater vaporizer for the coal slurry must be added. If it is assumed that units of 100 tons per hour of coal throughput are used, then 75 tons per hour of water must be vaporized; and, using $1,200,000 for the cost of such a vaporizer (2) and 15 percent of this as

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(1) Anon., "The utilization of lignite as a source of pipeline gas--An economic analysis," U.S. Bur. Mines, Process Evaluation, Morgantown Research Center, January 1964; Table 12, p 25A; Table 5, p 15B.

(2) The auxiliary plant cost tables in Appendix 5.1 indicate this as the cost of a 150,000 lb per hour steam boiler for 600 psi, 700 F steam. For this preliminary order-of-magnitude estimate, this figure will be used.

TABLE 4.1-3 LOCK HOPPER INVESTMENT COSTS

	Process	
	CO_2-Acceptor	Hydrogasification
Pressure, atm (psi)	20 (285)	70 (1,000)
Dried Lignite, ton/hr	213.3	253.7
Number of Lock Hoppers	10	6
Number of Feed Hoppers	5	12
Total Volume, cu ft	2,850 + 4,080	2,600 + 3,500
Feed and Lock Hopper Materials Cost, dollars	47,000 + 16,500	145,800 + 216,000
Factor: Complete Plant/Materials	3.7	3.6
Total Feed and Lock Hopper Cost, dollars	232,000	1,305,000
Total Feed and Lock Hopper Cost, dollars/hourly ton lignite	1,090	5,130
Total Feed and Lock Hopper Cost, cents/annual ton lignite	13.6	64
Capital Connected Cost at 15 Percent, cents/annual ton lignite	2.0	9.7
Cost of Bailey Feeders for Continuous Feed; Materials	---	74,400
Total Cost	---	270,000
Total Cost: dollars/hourly ton lignite	---	1,270
Total Cost: cents/annual ton lignite	---	13.3
Capital Connected Cost, 15 Percent of Total Cost: cents/annual ton lignite	---	2.0

TABLE 4.1-4 LOCK HOPPER COMPRESSOR COSTS

	Process	
	CO_2-Acceptor	Hydrogasification
Feeder Compressors, number of	5	6
Materials Cost, dollars	12,000	432,000
Total Cost, dollars	45,000	1,560,000
Cost, dollars/hourly ton lignite	211	6,130
Cost, cents/annual ton lignite	2.6	77
Capital Connected Cost at 15 Percent, cents/annual ton lignite	0.4	11.6
Compressor Size, hp	30	900
Power Requirements for all Compressors, kw	112	4,030
Unit Power Consumption, assumed kwh/M scf	52.0	77.6
Compressor Capacity, scfh	28,000	695,000
Actual Capacity at Pressure, cu ft/hr	1,400	9,900
Actual Capacity, cu ft/ton lignite	6.6	39
Bulk Density of Lock Hopper Material, lb/cu ft	303	51
Compression Power, kwh/ton coal	0.53*	16
Compression Power, cents/ton coal (Power 0.8 cent/kwh)	---	12.8
Total Compressed Gas Cost, cents/ton coal	---	24.4
Total Compressed Gas Cost, cents/M cu ft gas	---	9.0

* This figure seems to be based on other compressed gas supply being available

TABLE 4.1-5 SUMMARY OF COAL FEEDING COSTS

Item	Process Pressure, atm (psi): CO_2-Acceptor 20 (285)		Hydrogasification 70 (1,000)	
	Cents/ton Lignite	Cents/MM Btu in Gas**	Cents/ton Lignite	Cents/MM Btu in Gas**
Feed and Lock Hopper	2.0	0.10	9.7	0.48
Bailey Feeder	-	-	2.0	0.10
Compressors	3.4*	0.17	16.5*	0.83
Compressor Power	3.5*	0.18	18.2*	0.91
Total	8.9	0.45	46.4	2.32
Power Cost, Percent of Total	39		39	

* Data from Table 4.1-3 increased to attain 36 lb/cu ft bulk density

** Assume 12,500 Btu/lb coal and 80 percent cold gas efficiency

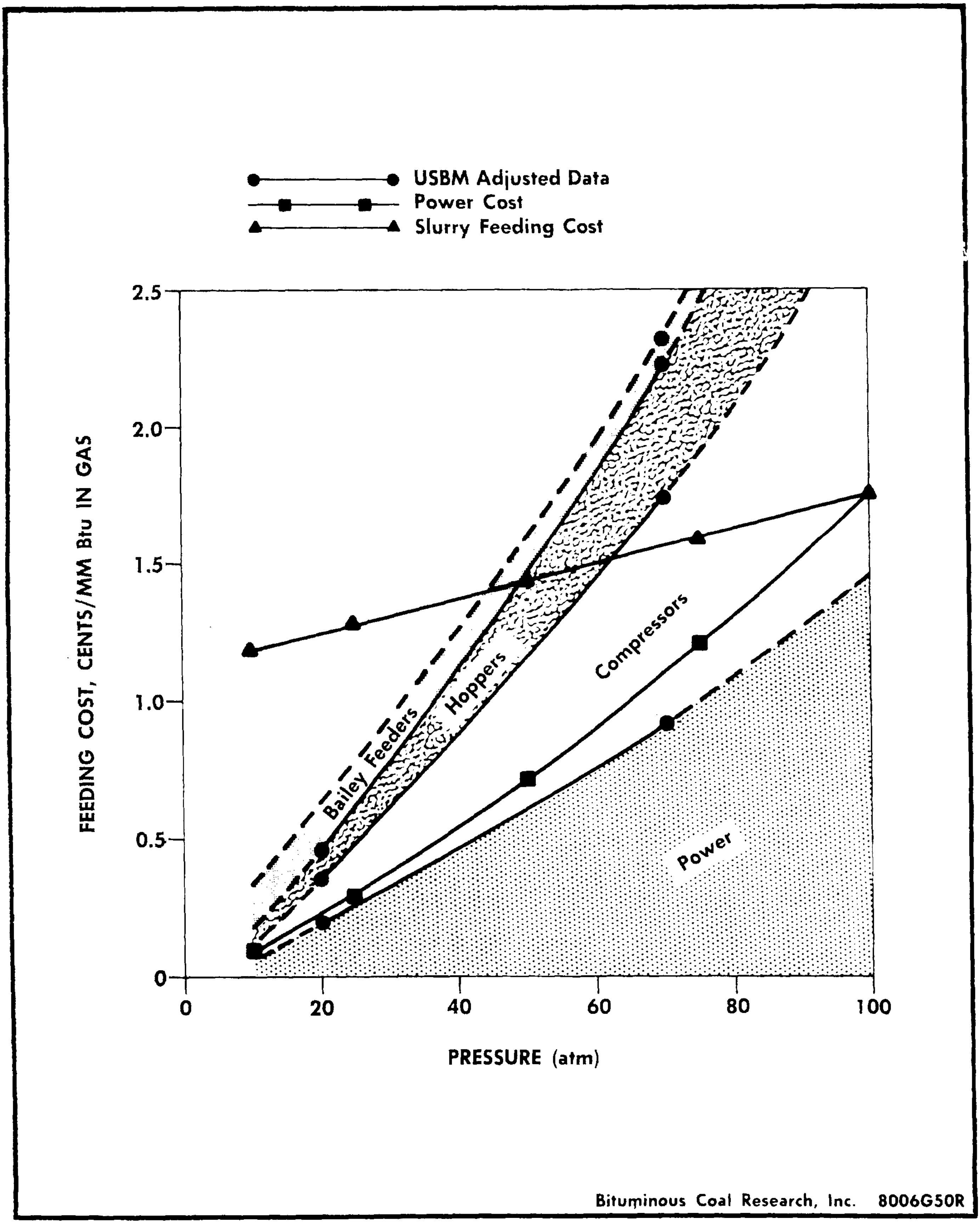

Figure 4.1-1 Effect of Pressure on Operating Costs of Coal Feeding

TABLE 4.1-6 COST OF COAL SLURRY PUMPING

Operating Pressure		Power			Pumping Cost* Cents/MM Btu in Gas	Vaporizer Operating Cost Cents/MM Btu in Gas	Total Coal Feeding Cost Cents/MM Btu in Gas
Atm	Psi	Kwh per Ton Coal	Cents/Ton Coal	Cents/MM Btu in Gas			
10	147	0.6	0.5	0.02	0.06	1.13	1.19
25	367	1.5	1.2	0.06	0.16	1.13	1.29
50	735	3.0	2.4	0.12	0.31	1.13	1.44
75	1,000	4.5	3.6	0.18	0.46	1.13	1.59
100	1,470	6.0	4.8	0.24	0.62	1.13	1.75

* Calculated on basis that power costs represent 39 percent of total pumping costs

operating cost, and disregarding fuel requirements, one then obtains

$\frac{1,200,000 \times 0.15}{100 \times 800}$ = \$0.225 per ton of coal, or 1.13 cents per MM Btu in the gas as a rough approximation. By adding this to the pumping cost in Table 4.1-5 the total feeding costs in column 7 are obtained. Comparison of these data with those in Figure 4.1-1 shows that lock hopper feeding is cheaper than slurry feeding at pressures below 50 atm (750 psi), while at higher pressures, slurry feeding has lower cost.

The problems connected with the vaporization of the water in a coal slurry are considerable. Flash vaporization by injecting the slurry into the hot gas from the coal gasifier has been proposed as a way to avoid this problem.(3)

The coal is recovered from the gas stream and reinjected without depressuring into the oxygen blown gasification zone. It is easy to show that at least the heat of vaporization of the water in the coal slurry is lost to high pressure steam generation in a waste heat boiler. For the slurry concentration used in this present study, this amounts to 0.75 tons or 1500 lb steam per ton of coal. Using a price of 35 cents per 1000 lb of steam, the cost is 52.5 cents per ton of coal or 2.63 cents per MM Btu in the gas. Thus, this method of coal feeding is uneconomical.

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(3) Steever, A. B. (to The Babcock & Wilcox Co.), "Comminuted solid fuel introduction into high pressure reaction zone," U.S. 2,961,310 (Nov. 22, 1960).

APPENDIX 4.2

DESIGN AND ESTIMATED OPERATING COST OF A NEW CONCEPTUAL COAL FEEDING DEVICE

A schematic drawing of a new conceptual coal feeding device is shown in Figure 4.2-1; a plunger type element is used for purposes of illustration.

However, before describing the operation of this device, the basic objection to piston feeders for solids will be discussed; namely, the possibility of excessive and unsafe pressures being developed in case deposits of solids occur. This can be overcome completely by using either a direct steam drive without crankshaft or fly-wheel, or by using, as in some diaphragm compressors, an intermediate hydraulic drive fluid, together with pressure safety valves.

The operating cycle of the pump in Figure 4.2-1 is as follows:

> Starting with the piston in the down position, Vessel B is filled with coal and then with gas to 3 atm. Then as the piston moves up, coal and gas at 3 atm pressure is moved from Vessel B into the pump cylinder (Vessel A), with additional gas supplied through Valve 5. As Vessel B becomes empty, Valve 2 is flushed by a continued flow of 3 atm gas and then closed. Vessel A is then pressurized to 75 atm by raw coal product gas through Valve 6, and feeding of coal into the gasifier begins as the piston moves down and Valve 1 opens. As the piston approaches its down position, a flow of 75 atm gas from Valve 6 is used to flush Valve 1 before it closes. The gas remaining in Vessel A at 75 atm is released via Valve 7 into the 3 atm line, and the cycle is repeated. During the down-stroke of the piston, Vessel B is depressured through Valve 4, and coal is fed into it from the bin via Valve 3.

<u>Gas Costs</u>: For calculation of the gas and power requirements for the feeding of coal at 75 atm pressure, a bulk density in Vessel B of 36 cu ft and in Vessel A of 32.4 lb per cu ft is assumed. The volume of Vessel B is 90 percent of the displacement per stroke of the piston in Vessel A. Vessel A contains, for each cubic foot of its volume, 32.4 lb of coal equal to

$\frac{32.4 \text{ lb}}{83 \text{ lb/cu ft}} = 0.39$ cu ft. It is further assumed that the piston, during each

stroke, displaces 95 percent of the total volume of Vessel A. From these assumptions then, the gas requirements, gas losses, and piston displacement per cubic foot of Vessel A are calculated as shown in Table 4.2-1, together with the same data related to 1 ton of coal and to 1 MM Btu in the gas.

The data in column 6 of Table 4.2-1 were used to obtain cost data shown in Table 4.2-2. For the raw gas, a heating value of 300 Btu per scf is assumed, together with a cost of 40 cents per MM Btu. For the 3 atm gas, a heating value credit of 15 cents per MM Btu (equals $4 per ton of coal) is used, making the

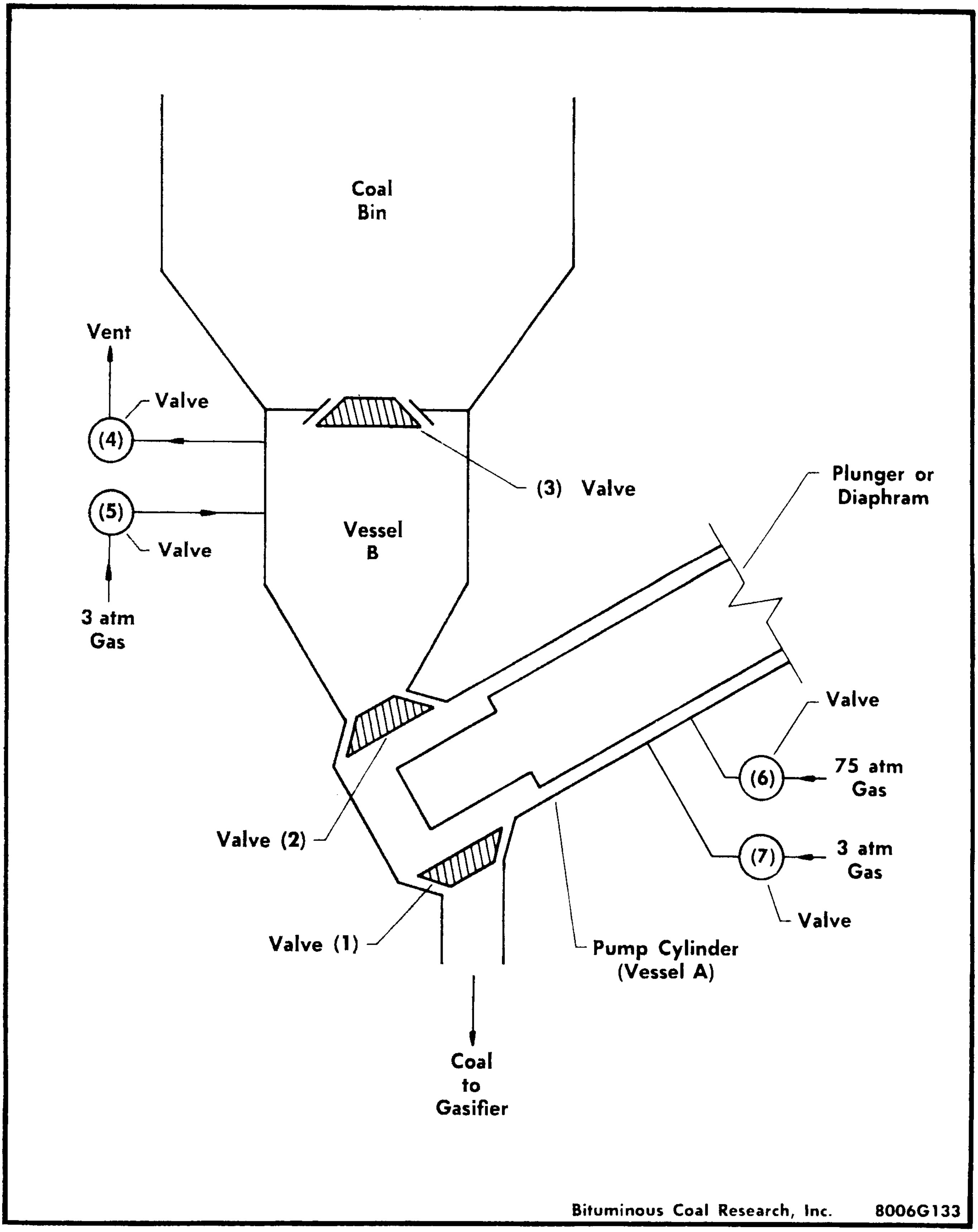

Figure 4.2-1 Diagram of New Conceptual Coal Feeding Device

TABLE 4.2-1 GAS REQUIREMENTS FOR FEEDING OF COAL AT 75 ATMOSPHERES

	Scf Gas/Cu Ft of Vessel*
Requirement for 3 atm gas	
Gas for 32.4 lb coal in Vessel B, including 5 percent flushing gas (1.0 x 3)	3.0
Gas available from Vessel A at end of stroke (0.05 x 75 - 3)	3.6
Excess 3 atm gas	0.6
Requirement for 75 atm gas (feeding gas)	
Gas volume Vessel A (1 - 0.39 = 0.61 cu ft)	45.8
Flushing gas, 10 percent	4.6
Gas present at end of stroke at 3 atm before admitting feeding gas at 75 atm (0.61 x 3)	-1.8
Total feeding gas	48.6
Gas loss	
Gas displaced from Vessel B to coal bin	0.4

Summarized Data	
For 1 ton coal:	
Gas loss at 3 atm	37 scf
Gas loss at 1 atm	25 scf
Feeding gas (recycled)	3,000 scf
Piston displacement	58.7 cu ft
For 1 MM Btu in gas:	
Gas loss at 3 atm	1.9 scf
Gas loss at 1 atm	1.2 scf
Feeding gas (recycled)	150 scf
Piston displacement	2.94 cu ft

* Calculated with piston displacement of 0.95 cu ft

TABLE 4.2-2 COST OF FEEDING COAL AT 75 ATMOSPHERES

Item	Cents/MM Btu in Gas
1.2 scf of 360 Btu gas lost at 40 cents/MM	0.02
1.9 scf of 570 Btu in 3 atm gas at 25 cents/MM Btu	0.02
150 scf feeding gas recycled (70 to 77 atm at 0.08 kwh per M scf and 0.8 cents/kwh)	0.1
8.4 lb steam for power* $(2.94 \times \frac{75}{41} \times \frac{1}{0.7 \times 0.92})$	0.29
Low pressure steam credit 80 percent	-0.23
Total Gas and Power Cost	0.20

* Steam drive at 70 percent volumetric efficiency, using 600 psi (41 atm), 600 F steam of 0.92 cu ft/lb and priced at 35 cents per M lb, together with a piston displacement of 2.94 cu ft.

cost of this gas loss 25 cents per MM Btu. On this basis, a total gas and power cost of the piston feeder of 0.2 cents per MM Btu in the gas is obtained. This is much lower than the cost of gas used in lock hopper feeding and is comparable to the power cost of slurry feeding.

Equipment Cost: For an approximation of the cost of such a piston feeder, the weight of such a feeding device for a 250 MM scf per day pipeline gas plant has been estimated. Such a plant will require 12,000 tons per day of coal, or 500 tons per hour. Assuming that this amount of coal will be processed in 5 gasifiers using 20 percent spare capacity, each gasifier will have a maximum coal throughput of 625 tons per hour of coal. This rate has been used as the throughput of 1 coal pump.

A pump with a cylinder of 8 in. ID and 48 in. stroke has a volume of 1.4 cu ft corresponding to $32.4 \times 1.4 = 45$ lb coal per double stroke; at 4 seconds per double stroke, 40,500 lb per hour or 20.2 tons per hour will be pumped.

This permits an estimate of the valve size required to permit passage of this amount of coal. Forty-five pounds of coal must flow in 2 seconds through the valves; this corresponds to 22.5 lb per second. Suspended coal at a density of 21 lb per cu ft can be fed at velocities from 10 to 20 ft per second.(4) Thus, for 22.5 lb of coal per second, a circular opening of 0.054 to 0.108 sq ft, or 7.8 to 15.4 sq in. cross section is required. This cross section can be accommodated by a valve of 5 in. in diameter lifted 0.5 to 1.0 in.

Thus, 6 cylinders will be needed for each pump. The cross section of a 600 psi (41 atm) steam drive cylinder at 70 percent efficiency is

$\frac{75}{41 \times 0.7} = 2.61$ times that of the driven piston, or 13 in. ID. The indicated

outside diameter (5) of the coal cylinder at 100 F and 2160 psi is 11.75 in., and the steam cylinder at 600 F and 1110 psi is 16.75 in. This gives the following weights:

Coal cylinder	810 lb
Steam cylinder	1,210 lb
Total	2,020 lb

Assuming, roughly, the same figure for weight of pistons and closures, and in addition, the same for supports, the weight of the total machine with 6 cylinders would be $2,020 \times 3 \times 6 = 36,400$ lb. Using, for estimating purposes, a

(4) Huff, W. R., et al, "A pilot-scale fluidized coal feeder utilizing zone fluidization," U.S. Bur. Mines, Rept. Invest. 6488 (1964).

(5) "Chemical Engineers' Handbook," Perry, J. H., ed., New York: McGraw-Hill, 1963. pp 24-36.

Ladish Catalog, p 180 ff.

price for such a machine of $6 per lb, the cost of one 625 ton per hour feeder would be 36,400 x 6 = $218,000. Arbitrarily, doubling this cost to include installation, and using 15 percent annual charges, one obtains 218,000 x 2 x 0.15 = $66,000 per gasifier per year, or $190 per gasifier per day; this corresponds to

$\frac{19,000}{50,000} = 0.38$ cent per MM Btu in the gas. Adding this to the gas and power cost of Table 4.2-1, a coal feeding cost of 0.68 cent per MM Btu in the gas is obtained. This is much lower than the cost of lock hopper or slurry feeding of 2.3 and 1.7 cents, respectively.

APPENDIX 5.1

COST DATA FOR AUXILIARY EQUIPMENT AND SERVICES

A series of graphs was derived by Blaw-Knox for use in determining the cost of oxygen, steam, and oxygen compression directly. (See Figures 5.1-1 through 5.1-14.) They also were used to arrive at part of the plant steam balance, using curves for methanation steam production, shift steam requirements, and turbine steam rates for mechanical drive turbines with various steam inlet conditions and back pressure conditions.

Brief statements concerning the nature of these graphs follow:

1. Atmospheric Oxygen Cost

Investment costs were obtained from Figure 5.1-4 - Oxygen Plant Installed Cost. Utility requirements were from the Clark Brothers Co., Division of Dresser Industries, Inc., for the compressors; General Electric Co., for the drive turbines; and Chemical Plants Division of Blaw-Knox Co., for the miscellaneous requirements for lighting, controls, expansion turbine, etc. Labor requirements were based on two operators per shift, plus 15 percent for supervision for plants up to 1,000 tons per day. For larger plants the labor requirements increased in proportion to the number of 1,000 tons per day plants required.

2. Steam Cost

Investment costs were obtained from Figure 5.1-6 - Steam Generator Installed Cost. Fuel costs were based on 15.4¢/MM Btu coal costs ($4/ton coal at 26 MM Btu/ton), and representative boiler efficiencies from the Riley Stoker Corporation and the Foster Wheeler Corporation. Miscellaneous utility costs for boiler feed water pumps, feed water treating, lighting, and controls were estimated from the experience of Chemical Plants Division of Blaw-Knox Co. Labor requirements are based on the requirements of the Foster Wheeler Co., for direct labor, plus 15 percent for supervision.

3. Oxygen Compression Cost

Investment costs were obtained from Figure 5.1-5 - Oxygen Compression System Installed Cost. Utility requirements were based on horsepower and cooling water requirements from the Clark Brothers Co., Division of Dresser Industries, Inc., plus steam requirements from Figure 5.1-12 - Turbine Steam Rates. The oxygen compressors are assumed to be located in the oxygen production plant area, so that no additional labor is required for oxygen compression beyond that labor already in use in the oxygen plant.

4. Oxygen Plant Installed Cost

Prices were obtained from Figure 5.1-7 - "Cold Box" Equipment Prices; Figure 5.1-8 - Centrifugal Compressor Prices; Figure 5.1-9 - Turbine and Motor Prices; and Figure 5.1-10 - Steam Turbine Auxiliaries Prices. These major equipment prices were totaled and factored up by adding miscellaneous materials and equipment costs, construction costs, engineering and start-up costs, general administrative expense, and fee. The factors used were developed from the experience of Chemical Plants Division of Blaw-Knox Co.

5. Oxygen Compression System Installed Cost

Prices were obtained from Figure 5.1-8 - Centrifugal Compressor Prices; Figure 5.1-9 - Turbine and Motor Prices; and Figure 5.1-10 - Steam Turbine Auxiliaries Prices. These prices were factored by the same method outlined for Figure 5.1-4 above.

6. Steam Generator Installed Cost

Budget prices for complete installed outdoor steam generating units from Foster Wheeler Corporation and Riley Stoker Corporation were used to update cost curves originally made in July of 1958 by Chemical Plants Division of Blaw-Knox Co. for a Lurgi gasification plant study.

7. "Cold Box" Equipment Prices

Budget prices were obtained on complete equipment packages, including cold box steel, insulation, expansion turbine, aftercooler/scrubber, etc., from American Air Liquids.

8. Centrifugal Compressor Prices

Budget prices were obtained on turbo compressors for air and oxygen from the Clark Brothers Co., Division of Dresser Industries, Inc.

9. Turbine and Motor Prices

General Electric Co. provided prices at Blaw-Knox Co. resale discount for turbines, motors, and exciters.

10. Steam Turbine Auxiliaries Prices

Prices for surface condensers, condensate pumps, and vacuum equipment were updated from a curve, originally made in July 1958, by Chemical Plants Division of Blaw-Knox Co. for a Lurgi gasification plant study, by using the Marshall and Stevens cost index.

11. Gear Prices

Prices for gears were obtained from the General Electric Co. and the Western Gear Corporation. Since gears represent such a minor portion of the total cost of a compressor/gear/motor combination (less than 10 percent), no attempt at an accurate correlation of gear cost vs. speed, ratio, and horsepower was made. The cost vs. horsepower curve shown is accurate to ± 20 percent and is adequate for the purpose of the present study.

12. Turbine Steam Rates

General Electric Co. quoted the actual steam rates shown for mechanical drive turbines to be direct coupled to large turbo compressors.

13. Waste Heat Steam from Methanation

The steam production shown is based on the use of a methanation process using a once-through tube wall catalytic reactor as described by Field, et al, in the paper "Development of Catalysts and Reactor Systems for Methanation," presented at the ACS Symposium on Gas Generation, April 5-10, 1964.

14. Shift Requirements to Produce $3H_2/1CO$

Approximate values for water vapor in the feed gas and quench water between catalyst layers are shown for the water gas shift using a catalyst such as "Catalysts and Chemicals Inc. #C12."

Equilibrium constants and heats of reaction were taken from Table III of the 1954 National Cylinder Gas Co. Girdler Catalyst Book on "Physical and Thermodynamic Properties of Various Elements and Compounds and Other Useful Information."

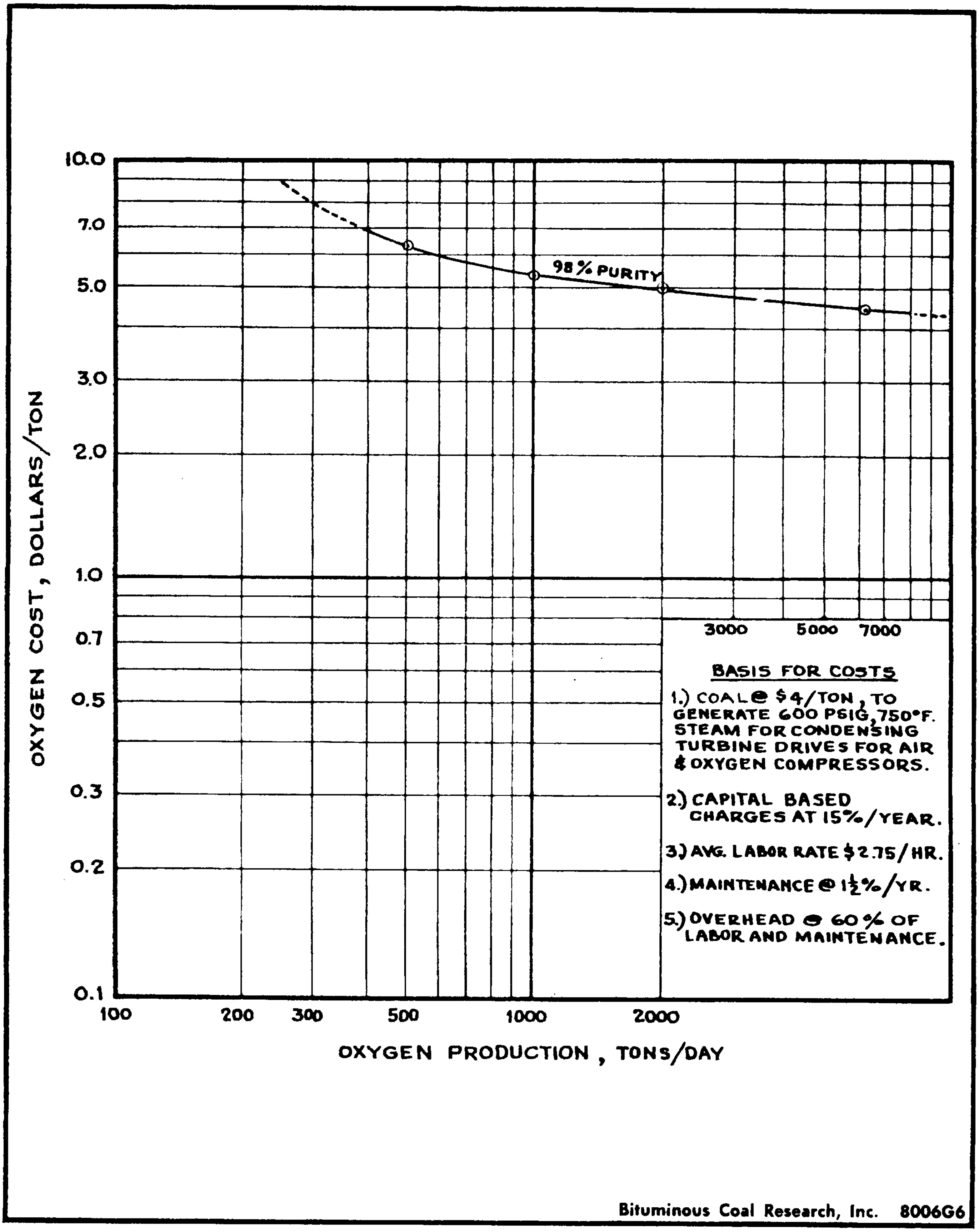

Figure 5.1-1 Auxiliary Equipment and Service Cost Data: Atmospheric Oxygen

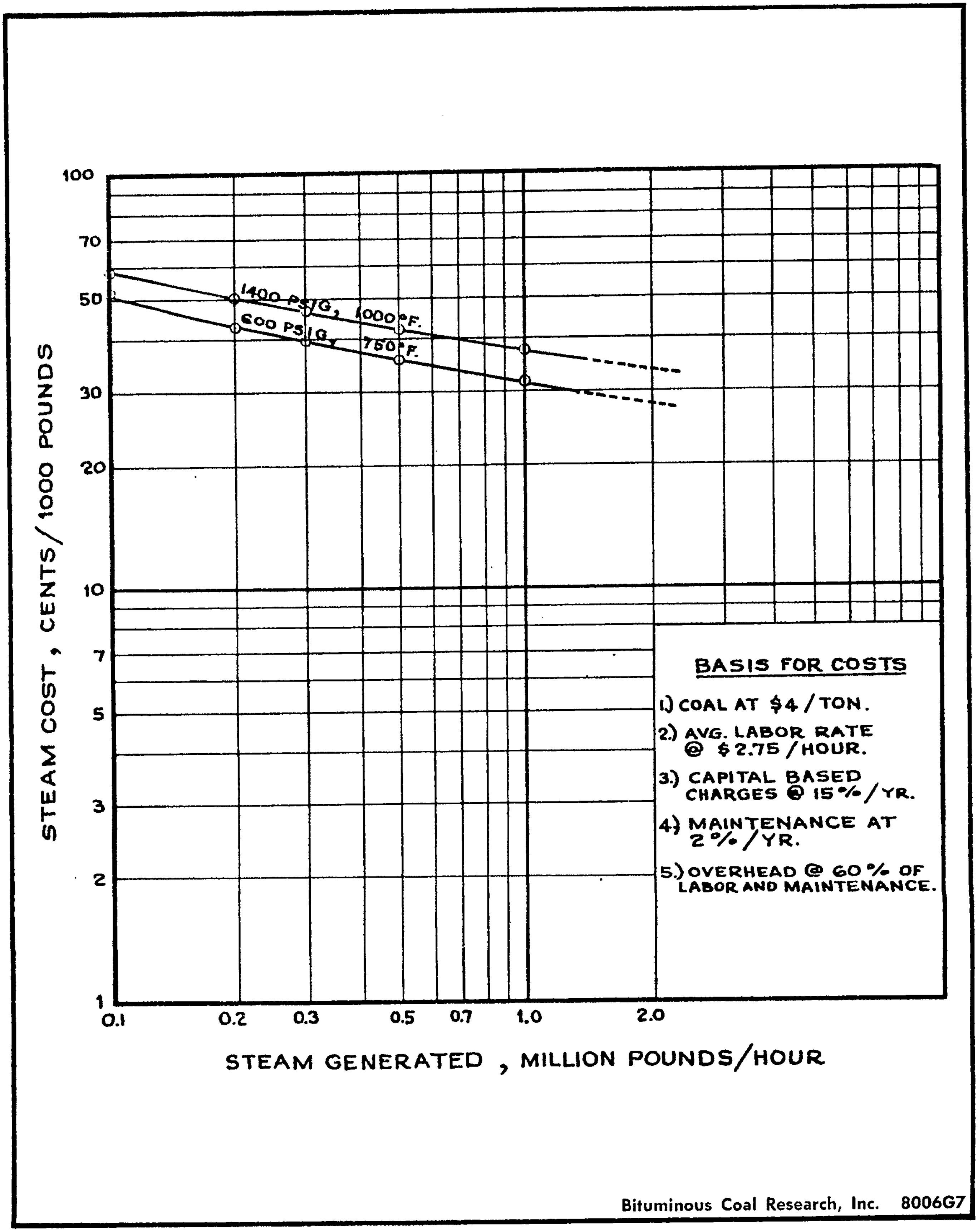

Figure 5.1-2 Auxiliary Equipment and Service Cost Data: Steam

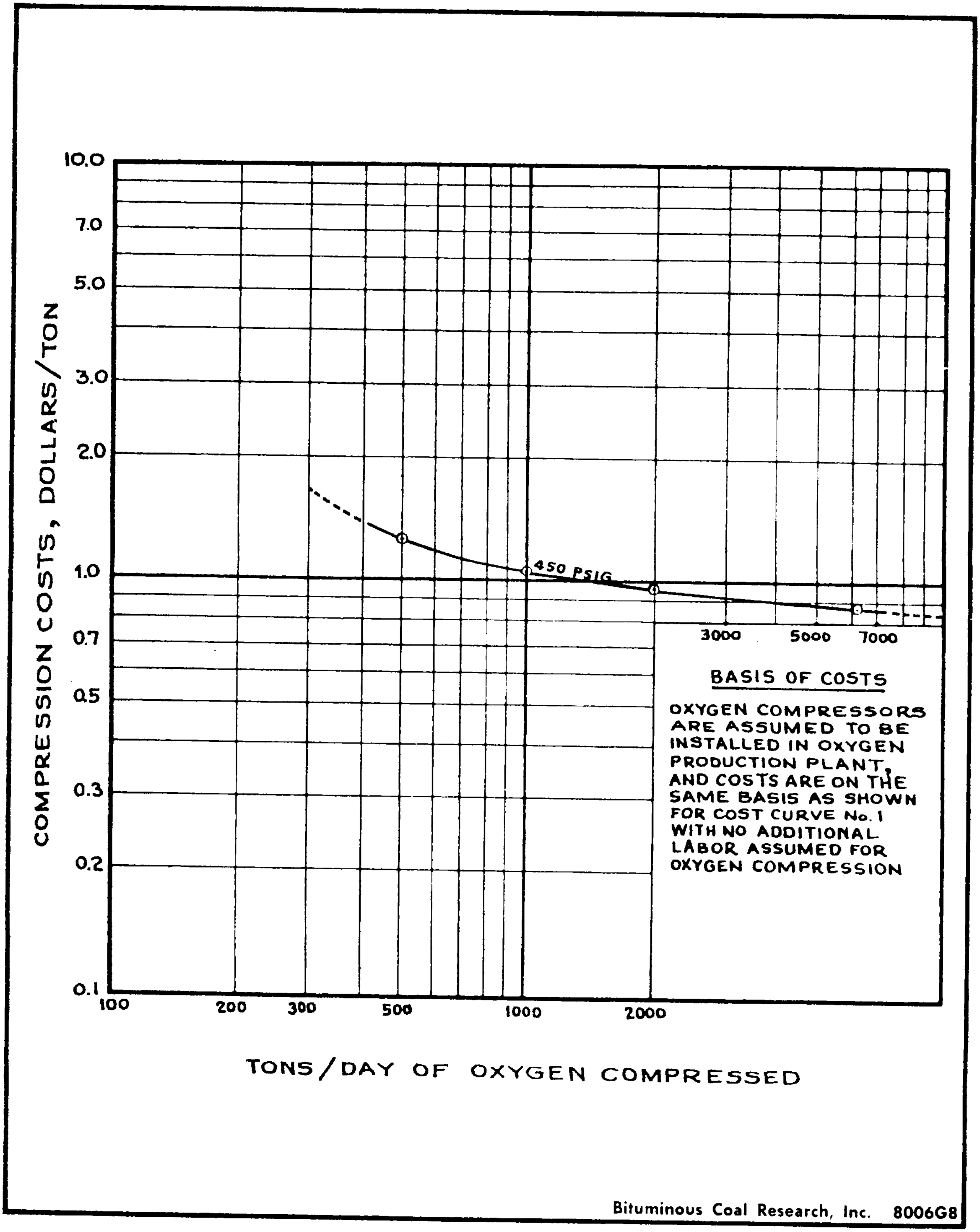

Figure 5.1-3 Auxiliary Equipment and Service Cost Data: Oxygen Compression

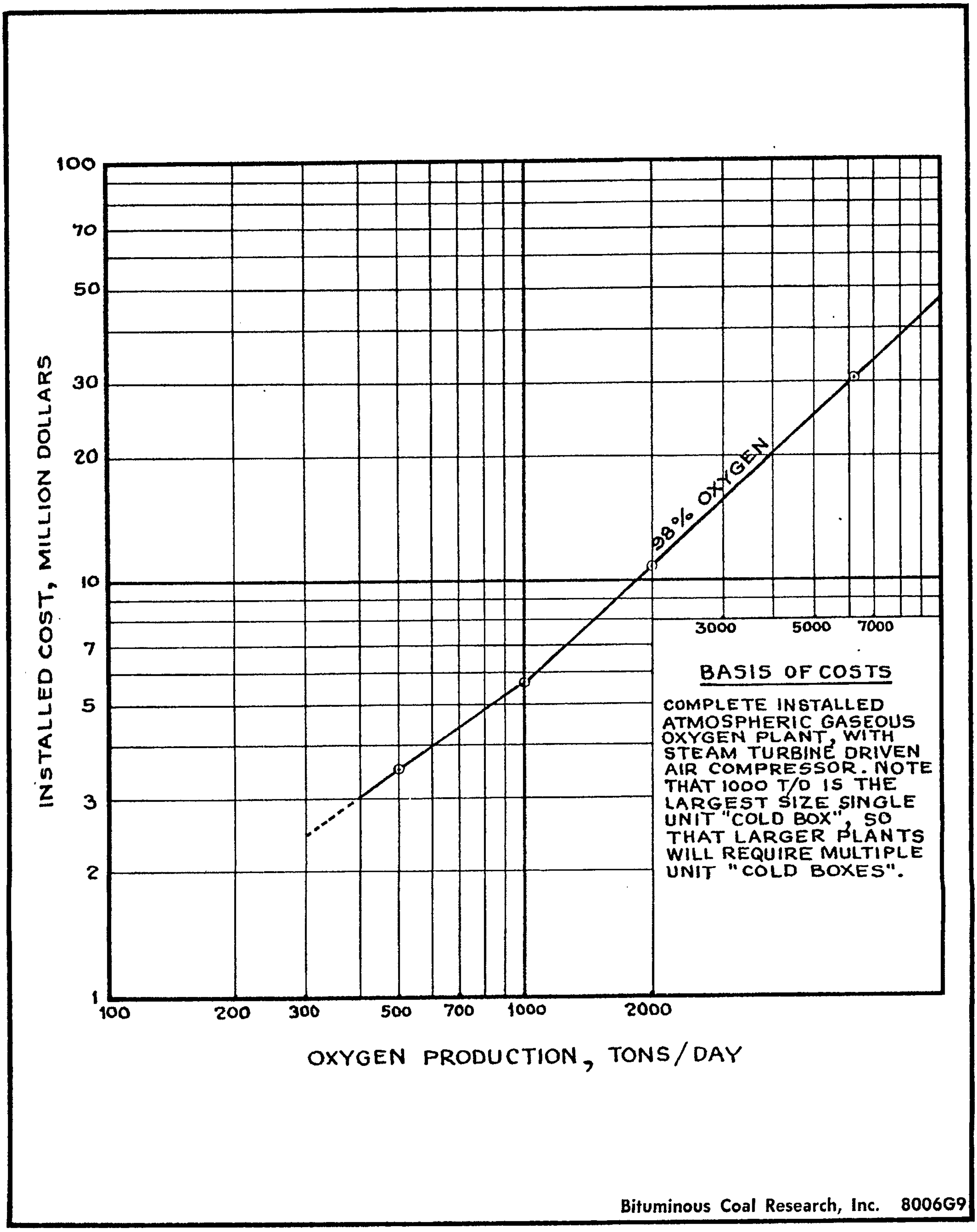

Figure 5.1-4 Auxiliary Equipment and Service Cost Data: Oxygen Plant

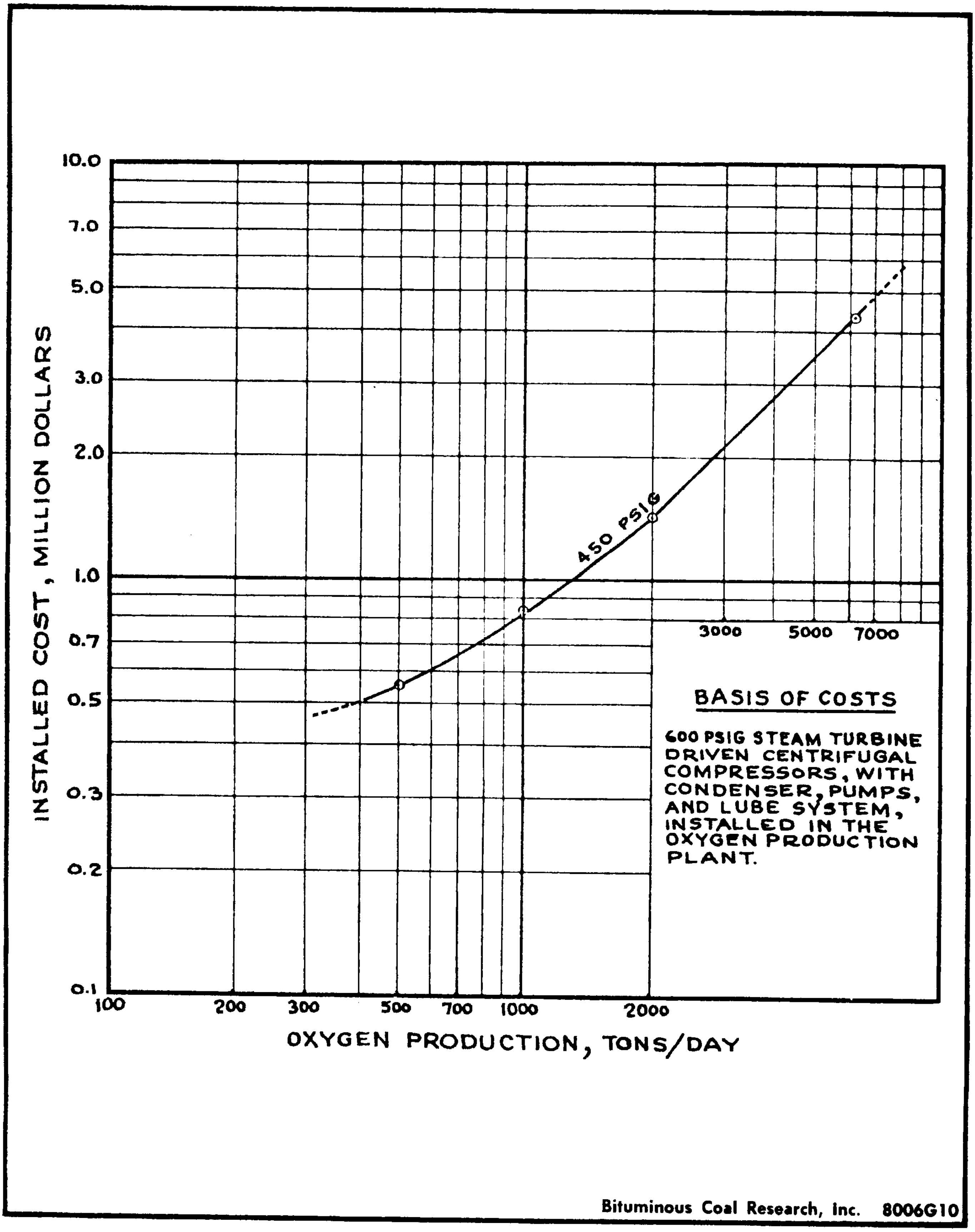

Figure 5.1-5 Auxiliary Equipment and Service Cost Data: Oxygen Compression System

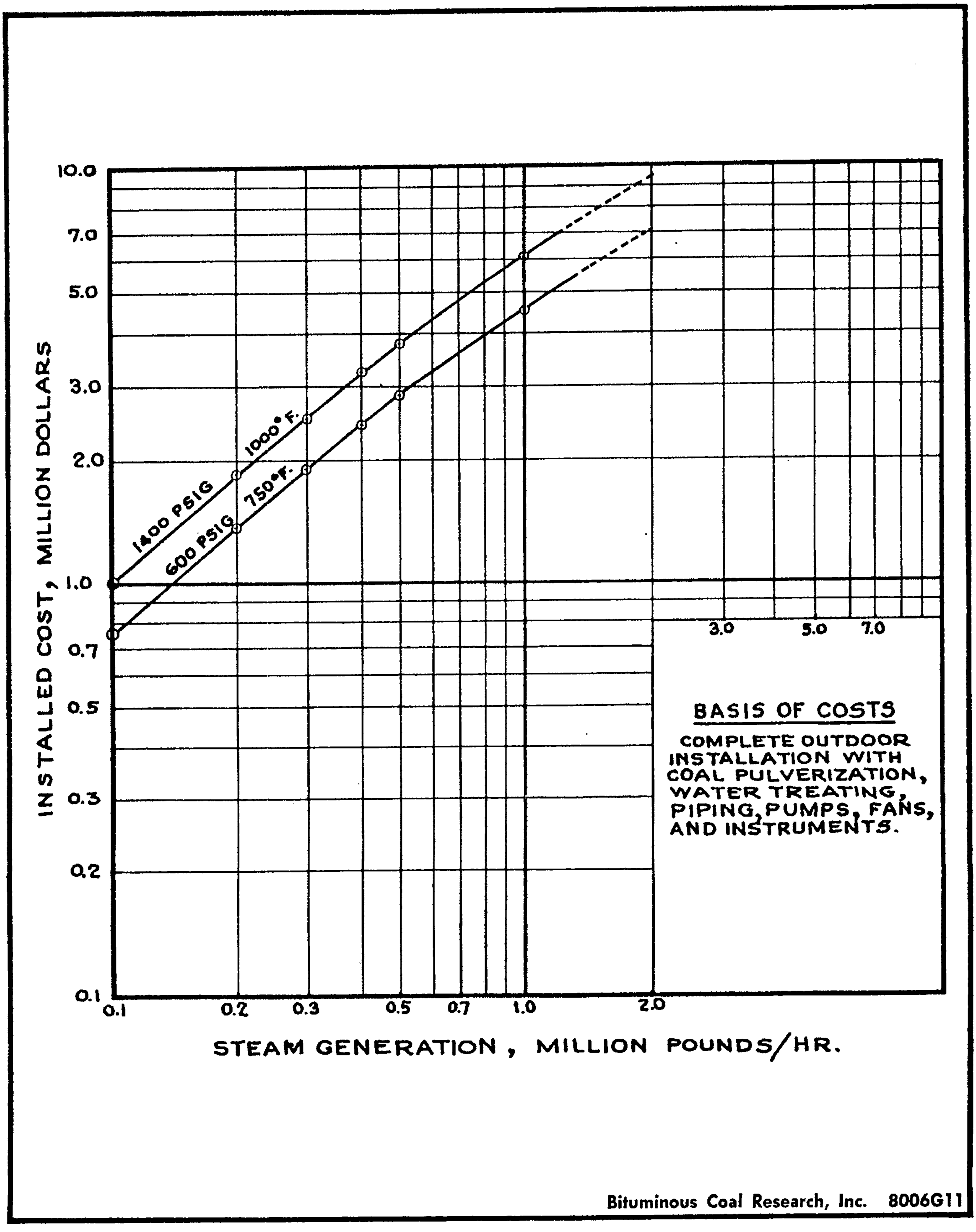

Figure 5.1-6 Auxiliary Equipment and Service Cost Data: Steam Generator

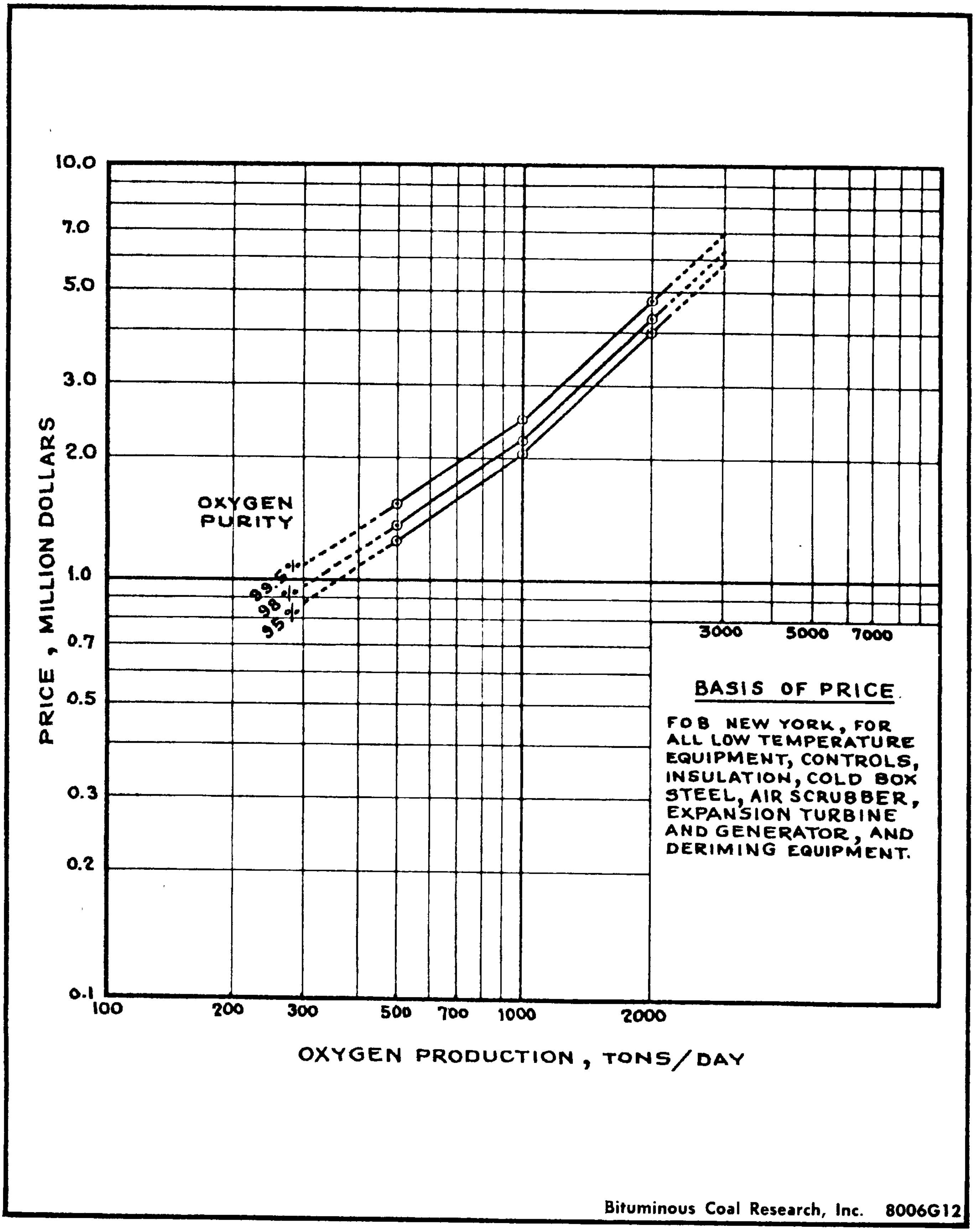

Figure 5.1-7 Auxiliary Equipment and Service Cost Data: "Cold Box" Equipment

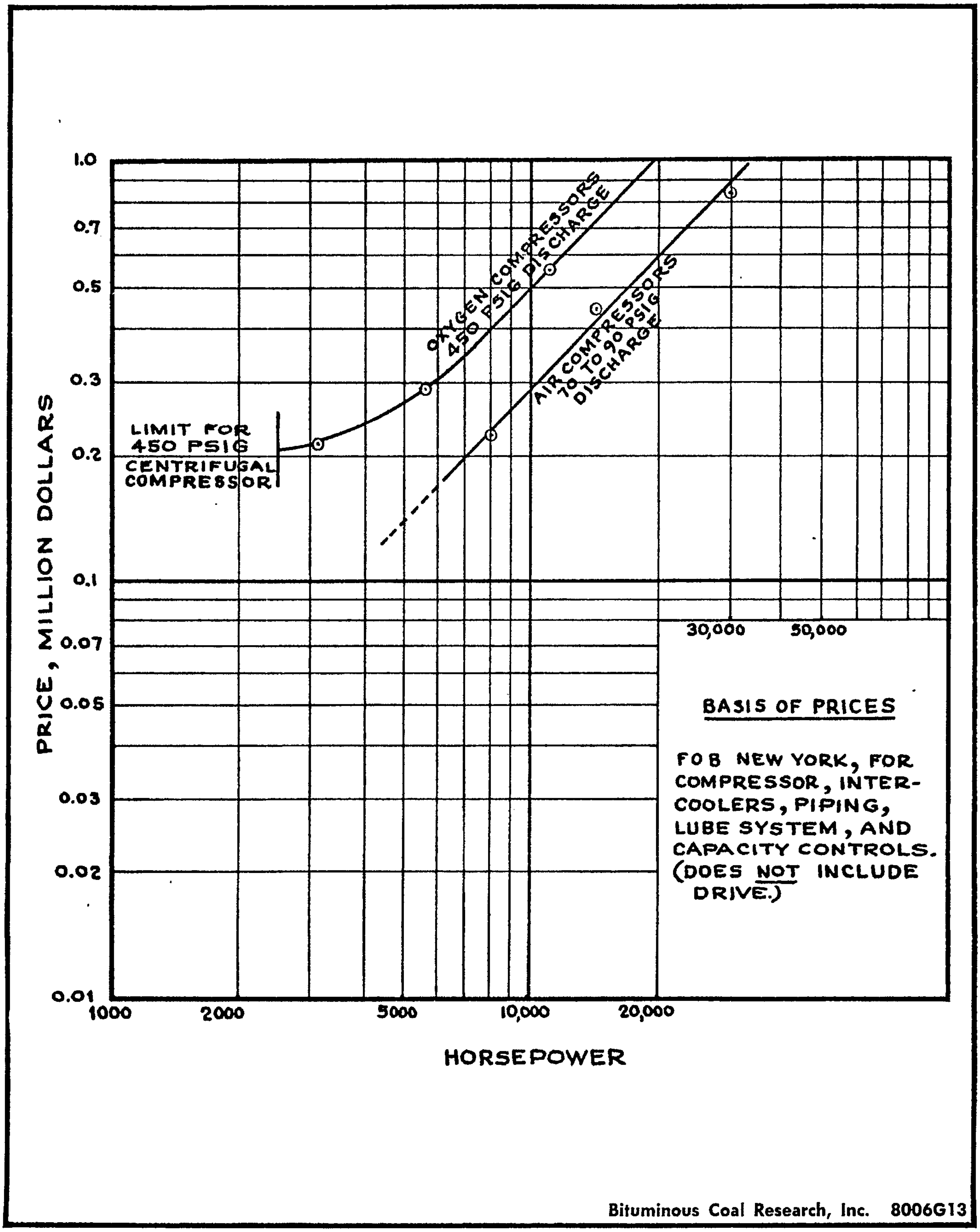

Figure 5.1-8 Auxiliary Equipment and Service Cost Data: Centrifugal Compressors

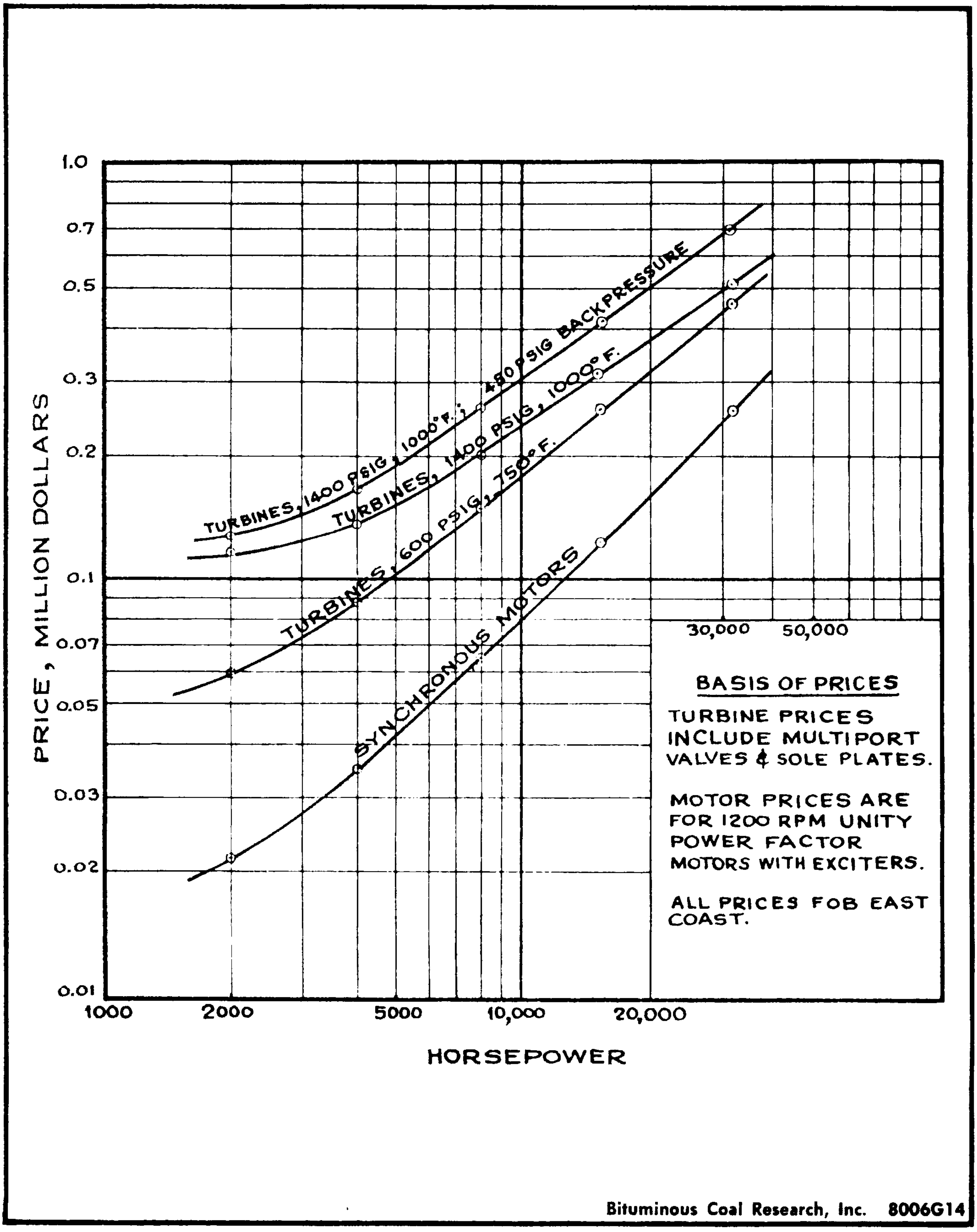

Figure 5.1-9 Auxiliary Equipment and Service Cost Data: Turbines and Motors

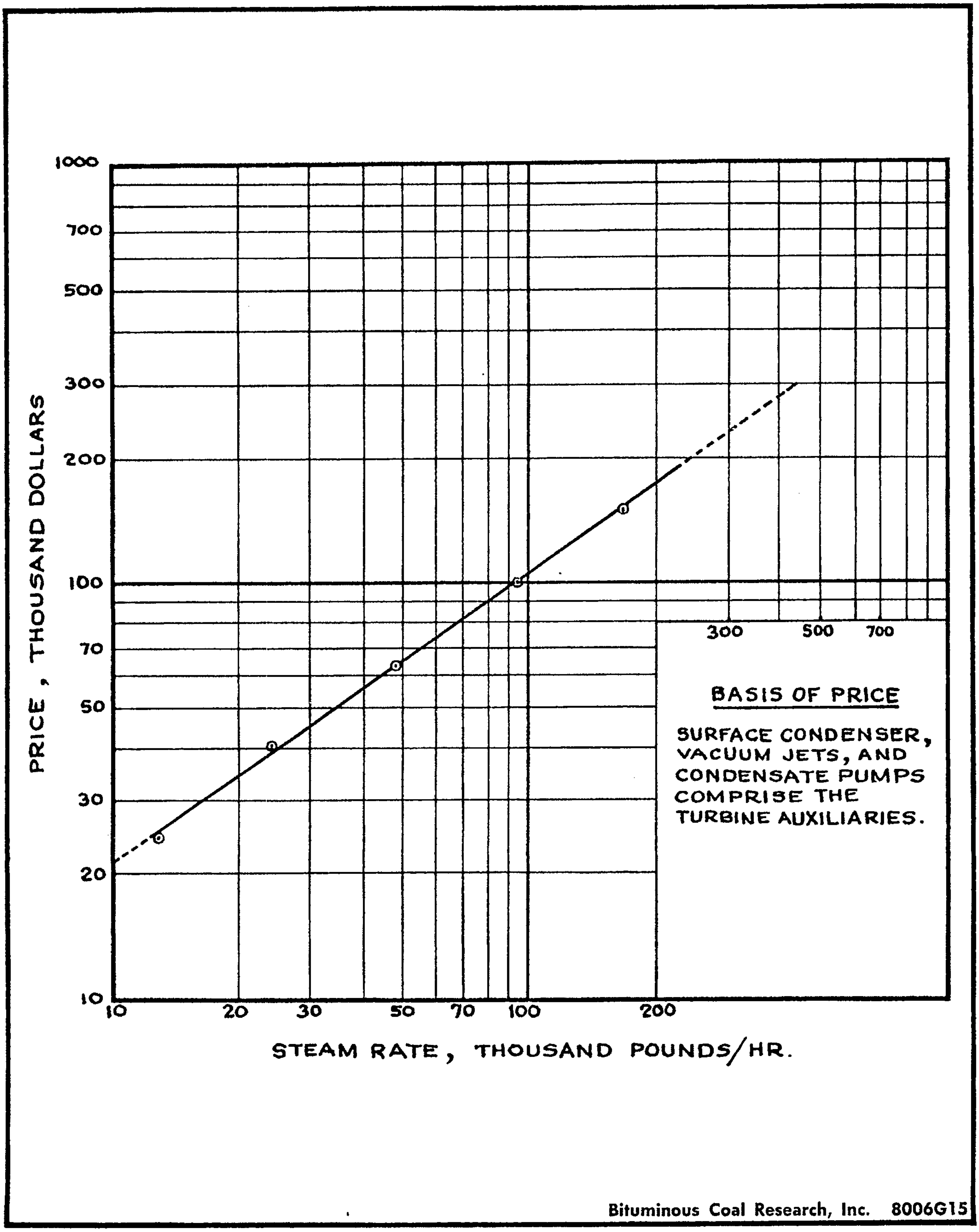

Figure 5.1-10 Auxiliary Equipment and Service Cost Data: Steam Turbine Auxiliaries

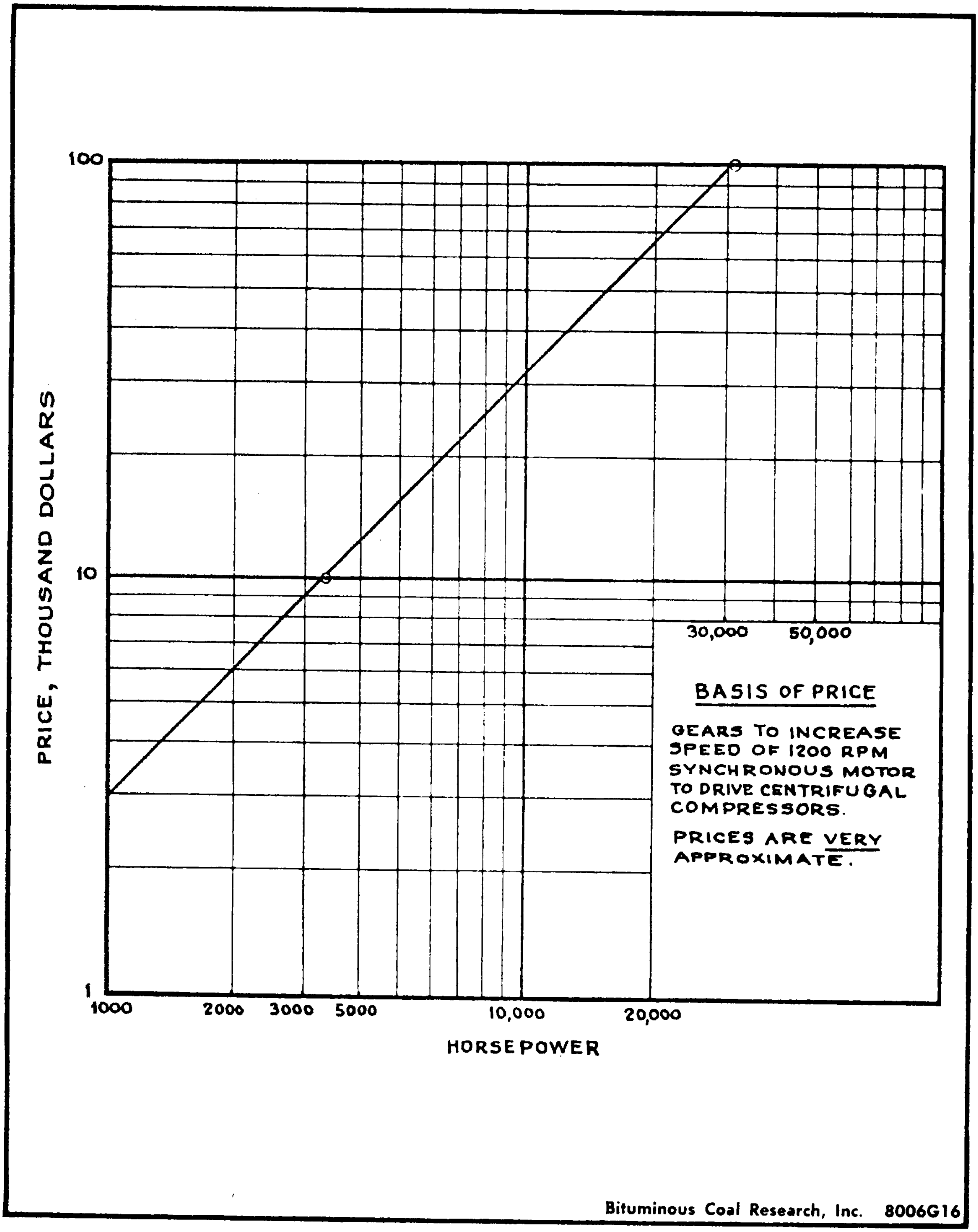

Figure 5.1-11 Auxiliary Equipment and Service Cost Data: Gears

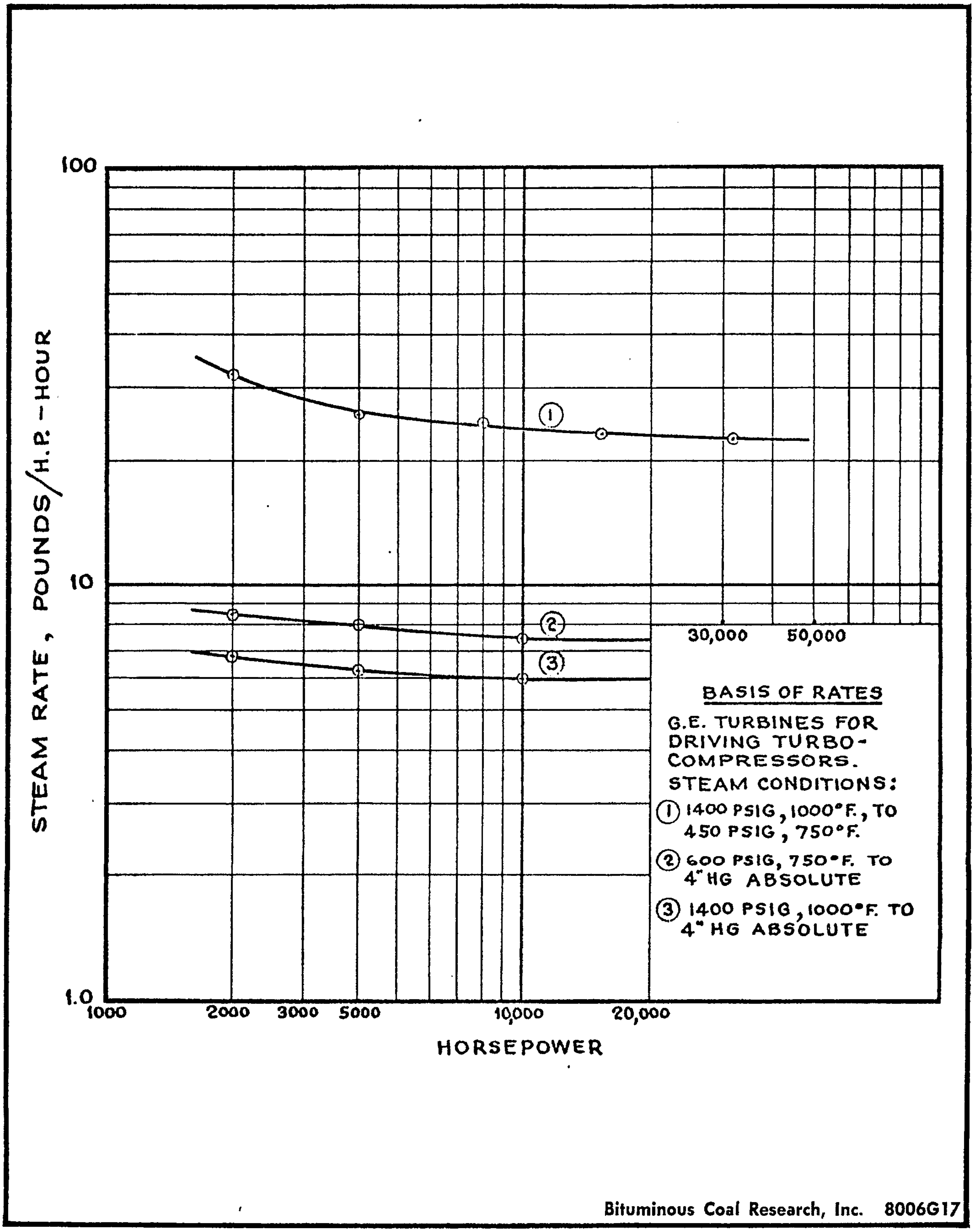

Figure 5.1-12 Auxiliary Equipment and Service Cost Data: Turbine Steam

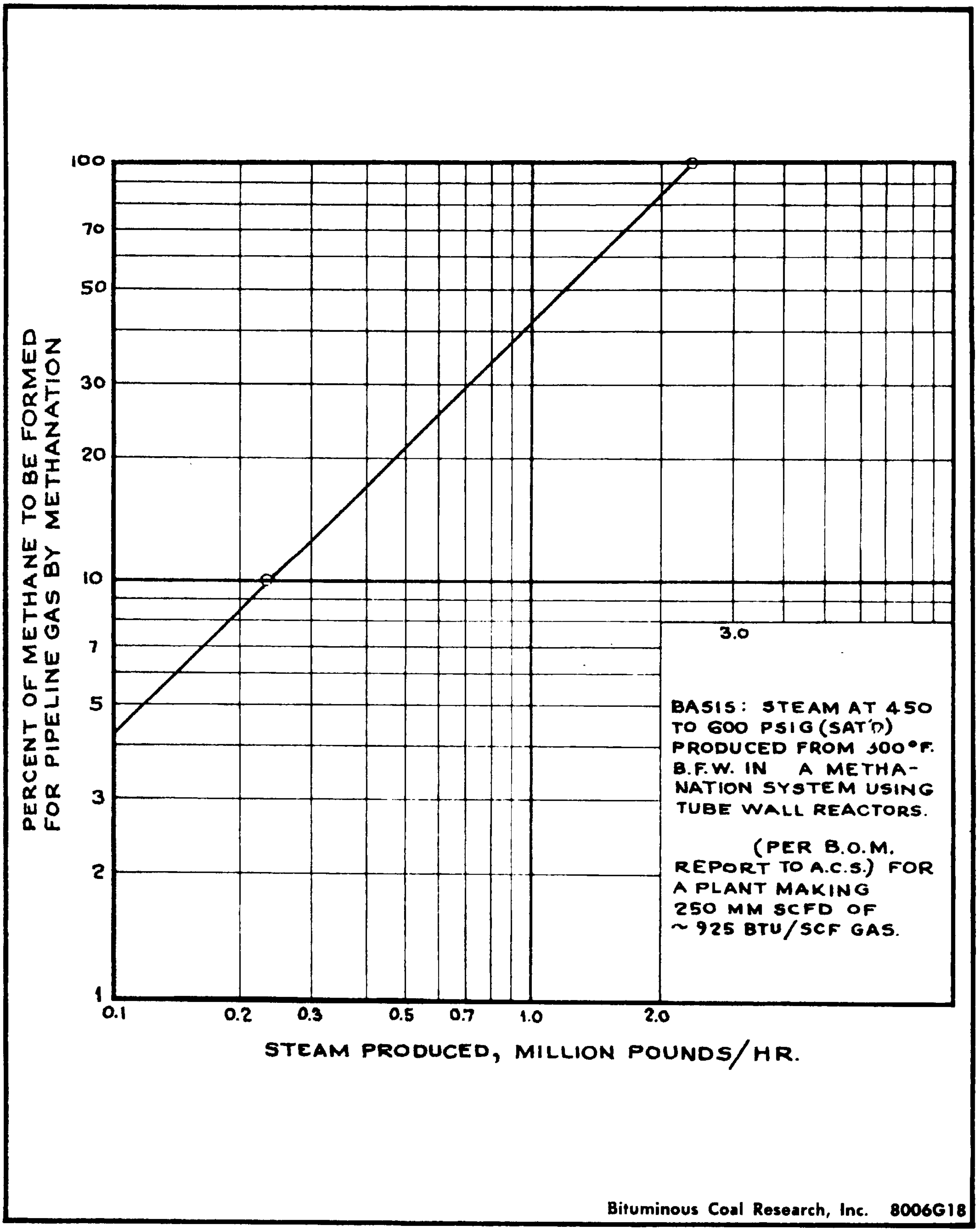

Figure 5.1-13 Auxiliary Equipment and Service Cost Data: Waste Heat Steam from Methanation

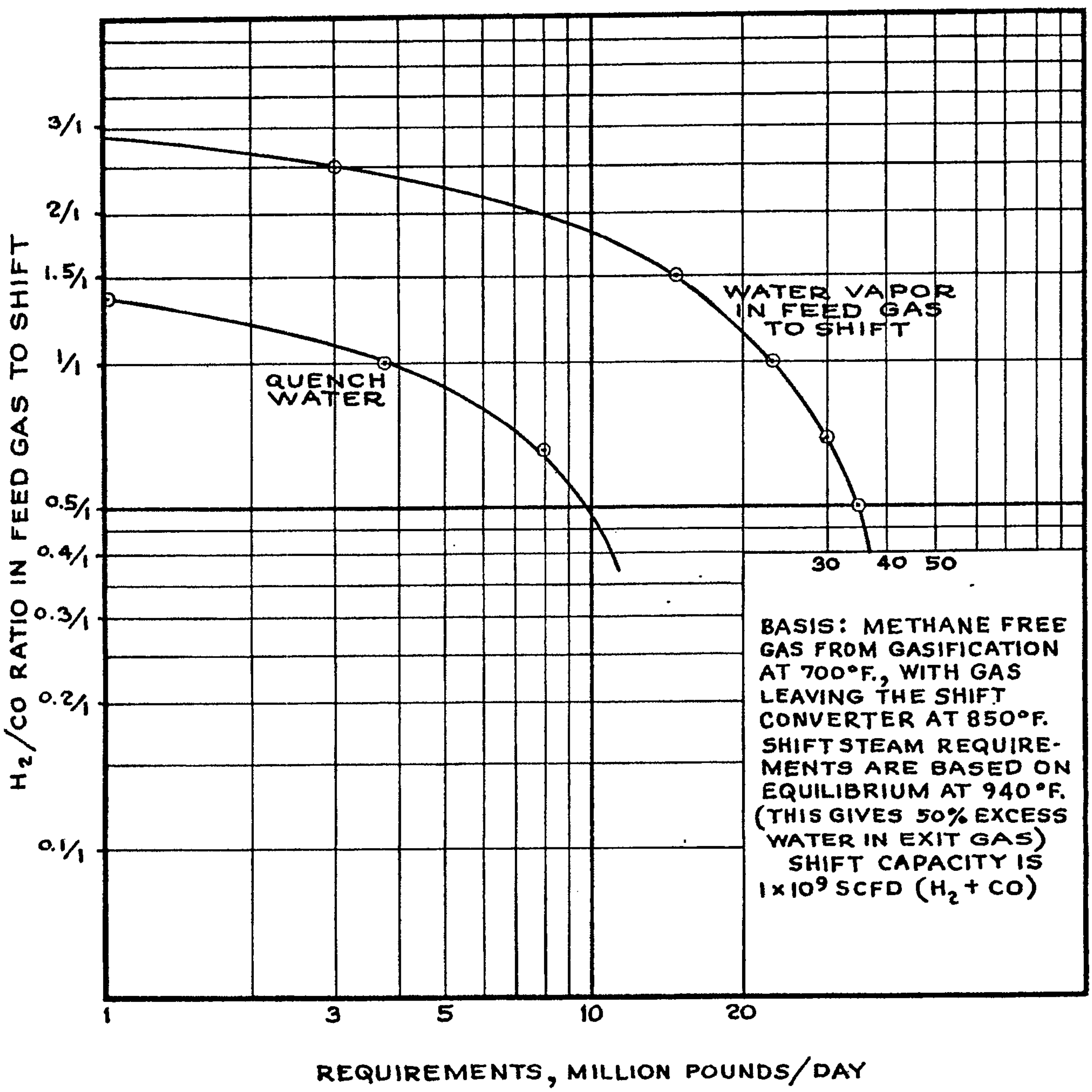

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Figure 5.1-14 Auxiliary Equipment and Service Cost Data: Shift Requirements

APPENDIX 5.2

COST ESTIMATES FOR ELECTRIC GASIFICATION OF COAL

Internal electric heating, according to a recent proposal (6) would be suitable for the gasification of coal. The data in Table 5.2-1 have been calculated, assuming complete reaction of highly preheated steam with coal, with complete carbon utilization. The cost of generating power in a base load plant has been calculated as 0.441 cents per kwh. (See Table 5.2-2.) From the data in these two tables, raw material and power cost have been calculated and are shown in Table 5.2-3. It can be assumed that Case A (0 percent CH_4) would correspond to atmospheric pressure gasification, while Case B (8 percent CH_4) would correspond to pressure gasification, if such should be feasible. For comparative purposes, similar data for Lurgi gasification are also included in Table 5.2-3.

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

TABLE 5.2-1 BASIC PROCESS DATA FOR ELECTRIC GASIFICATION OF COAL

Coal Analysis, Percent Dry Basis:

Carbon	77.3
Hydrogen	5.4
Nitrogen	1.4
Sulfur	2.5
Oxygen (by diff.)	6.3
Ash	7.1

Coal Heating Value: 13,990 Btu/lb

Steam for Gasification:

1.02 lb/lb coal, Preheated by exchange to 1380 F

Gasification Reaction:

100 Percent C gasification,
500 Btu heat loss/lb coal

1650 F gas exit temperature

Carbon Distribution:	A	B
C in Coal to CO, Percent	100	84.5
C in Coal to CH_4, Percent	0	15.5
Gas Analysis, Percent:		
Carbon Dioxide	-	2.9
Carbon Monoxide	42.3	40.3
Hydrogen	56.9	47.8
Methane	-	8.0
Nitrogen	0.3	0.4
Hydrogen Sulfide	0.5	0.6
Gas Heating Value, Btu/scf:	321	363
Process Requirements per MM Btu in gas:		
Coal, MM Btu	0.82	0.87
Steam, lb	59	62
Power, kwh	61.2	47.2

TABLE 5.2-2 POWER COST FOR ELECTRIC GASIFICATION OF COAL

	¢/kwh
Fuel (a)	0.133
Operating Cost (b)	0.047
Investment Connected (c) Cost at 15 Percent	0.261
TOTAL	0.441

(a) Based on coal at 15.4¢/MM Btu and a plant thermal efficiency of 40 percent, or 8,600 Btu/kwh

(b) Based on an annual operating cost of $3.50/kw and an operating factor of 85 percent equivalent to 7,440 hr/yr

(c) Based on an investment cost of $130/kw

TABLE 5.2-3 RAW MATERIAL AND OPERATING COST FOR ELECTRIC GASIFICATION OF COAL

	Electric Gasification				Lurgi	
	Case A 0% CH_4 in Gas		Case B 8% CH_4 in Gas			
	Amount	¢/MM Btu in Gas	Amount	¢/MM Btu in Gas	Amount	¢/MM Btu in Gas
Coal (15.4¢/MM Btu	0.82 MM Btu	12.6	0.87 MM Btu	13.4	1.15 MM Btu	17.8
Steam (35¢/1,000 lb)	59 lb	2.1	62 lb	2.2	149 lb	5.2
Oxygen ($5/ton)	--	--	--	--	39.5 lb	9.8
Power (0.441¢/kwh)	61.2 kwh	27.0	47.2 kwh	20.8	--	--
TOTAL COST		41.7		36.4		32.8

APPENDIX 5.3

PROCESS DATA FOR PROPOSED TWO-STAGE SUPER-PRESSURE ENTRAINED GASIFIER (7)

Heat and material balances needed for an evaluation of the two-stage entrained slagging gasifier are presented and discussed below.

As a basis for the calculations, 100 lb of moisture and ash-free (maf) coal was selected. Net heating values were used throughout, and the various assumptions made are as follows:

1. The feed coal is Pittsburgh seam coal, a high volatile A bituminous coal, with proximate and ultimate analysis as given in Table 6-1.

2. The volatile matter (VM) from the coal reacts in Stage II and contains all of the N, H, O, and S, together with sufficient carbon (27.3 lb) to make a total weight of 42.9 lb.

3. The char from the char separator is fed to Stage I -- the slagging section of the gasifier -- with the addition of 0.5 lb steam per pound of char gasified, i.e., a total of 32.4 lb steam per pound char.

4. In Stage I, there is 61 percent char utilization (8) on a once-through basis with the remainder of the char being recycled to obtain complete carbon gasification.

5. Oxygen added with the char is equivalent to that used for gasification of anthracite (8) and converts part of the carbon to carbon monoxide. The rest of the carbon reacts with steam. This assumption is conservative since the pilot plant in which these results were obtained had a heat loss of 925 Btu/lb coal.

6. The ash is slagged by being heated to about 2700 F.

7. In the heat balance for Stage I, an allowance is made for a heat loss of 250 Btu/lb coal.

8. The heat content of the product gas above 1700 F from Stage I is used in Stage II to gasify the volatile matter of the preheated coal with: (a) 75 percent of the hydrogen in the volatile matter forming methane (CH_4); and, (b) the rest of the carbon in the volatile matter, which is not used to form methane, forming carbon monoxide (CO).

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(7) See Process No. 58, Table 3-2 and Appendix 3.5

(8) Strimbeck, G. R. et al., "Gasification of pulverized coal at atmospheric pressure," U.S. Bur. Mines, Rept. Invest. 5559 (1960). See Table 1 on p 3: Anthracite 61 percent carbon utilization; 465 scf oxygen per Mcf of CO + H_2 or 122/lb oxygen per MM Btu in gas.

9. In Stage II, one pound of steam per pound of coal is used; the steam is preheated to 932 F and the coal is preheated to 212 F.

10. The homogeneous water-gas shift reaction comes to equilibrium at 1700 F for the individual reaction products in Stages I and II and also for the combined final products of both Stages I and II.

11. The equilibrium partial pressures of the final gaseous products is calculated for the reaction (9)

$$CH_4 + CO_2 \rightleftharpoons 2CO + 2H_2 - 44{,}820 \text{ Btu}$$

Stage I

Figure 5.3-1 shows the composition and temperature of the various streams -- gas, solids, and heat loss -- entering and leaving Stage I of the gasifier. The gas composition is given on (a) a molal basis before and after the shift reaction, and (b) a molal or volume percentage basis at the exit section (Point A) of Stage I after the shift reaction based on a 1700 F equilibrium temperature.

Table 5.3-1 presents the heat balance for Stage I. The difference between heat input and output of 147,600 Btu gives the excess heat available above 1700 F for use in Stage II.

The maximum temperature of the product gases and char in Stage I, assuming the slag leaves at 2700 F, is calculated to be approximately 3500 F. Thus, the temperature level is sufficiently high to insure slagging of the ash.

Stage II

Figure 5.3-2 shows the composition and temperature of the various streams entering and leaving Stage II. The composition of the gases at the exit section of Stage II (Point B) is shown in the same manner as that for Stage I.

Table 5.3-2 presents the heat balance for Stage II. The total heat input is only 8,360 Btu greater than the output and this amount is available to provide for an additional heat loss. Experimentation is proposed to determine the gasification reactions of the volatile matter of coal under the related conditions.

Figure 5.3-3 presents the overall system flow sheet, and Table 5.3-3 lists gas compositions for Stages I and II after shift, and for the final gas before and after the shift reaction at the exit (Point C) of the second stage of the gasifier. The analysis of the final gas is also given in volume percent on both a dry and a dry, CO_2-free basis.

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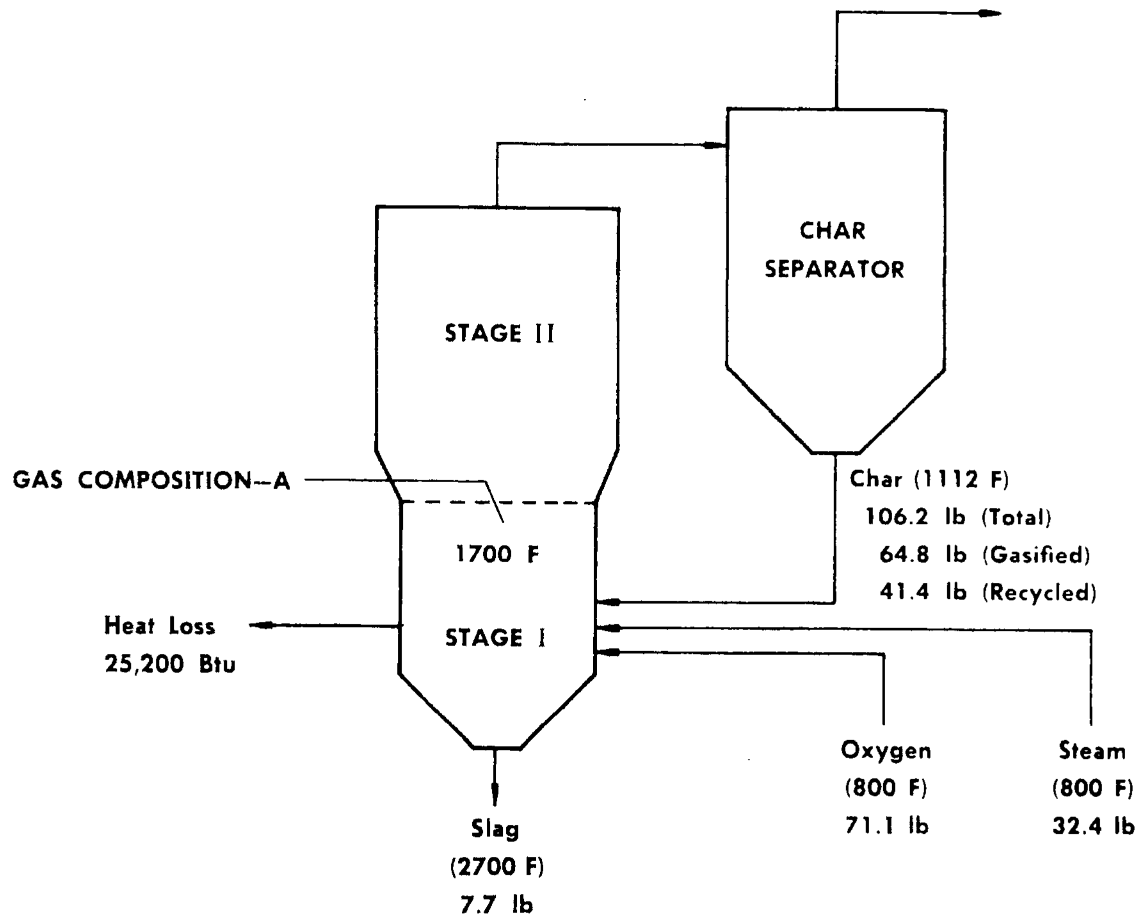
(9) Odell, W. W., "Gasification of solid fuels in Germany by the Lurgi, Winkler, and Leuna slagging-type gas-producer processes," U.S. Bur. Mines, Inform. Circ. 7415 (1947).
See also Wagman, D. D., et al, "Heats, free energies, and equilibrium constants of some reactions involving O_2, H_2, H_2O, C, CO, CO_2, and CH_4," Natl. Bur. Std. (U.S.), Res. Paper 1634, 143-61 (1945).



GAS COMPOSITION—A

	Mols*	Mols**	Mol %**
CO_2		1.0	15.4
H_2	0.3	1.3	20.0
CO	4.7	3.7	56.9
H_2O	1.5	0.5	7.7
Total	6.5	6.5	100.0

*Before Shift **After Shift

Basis: 100 lb daf Coal Input to Stage II

Heat Balance: Output Based on Stage I Product of 1700 F Exit Temperature

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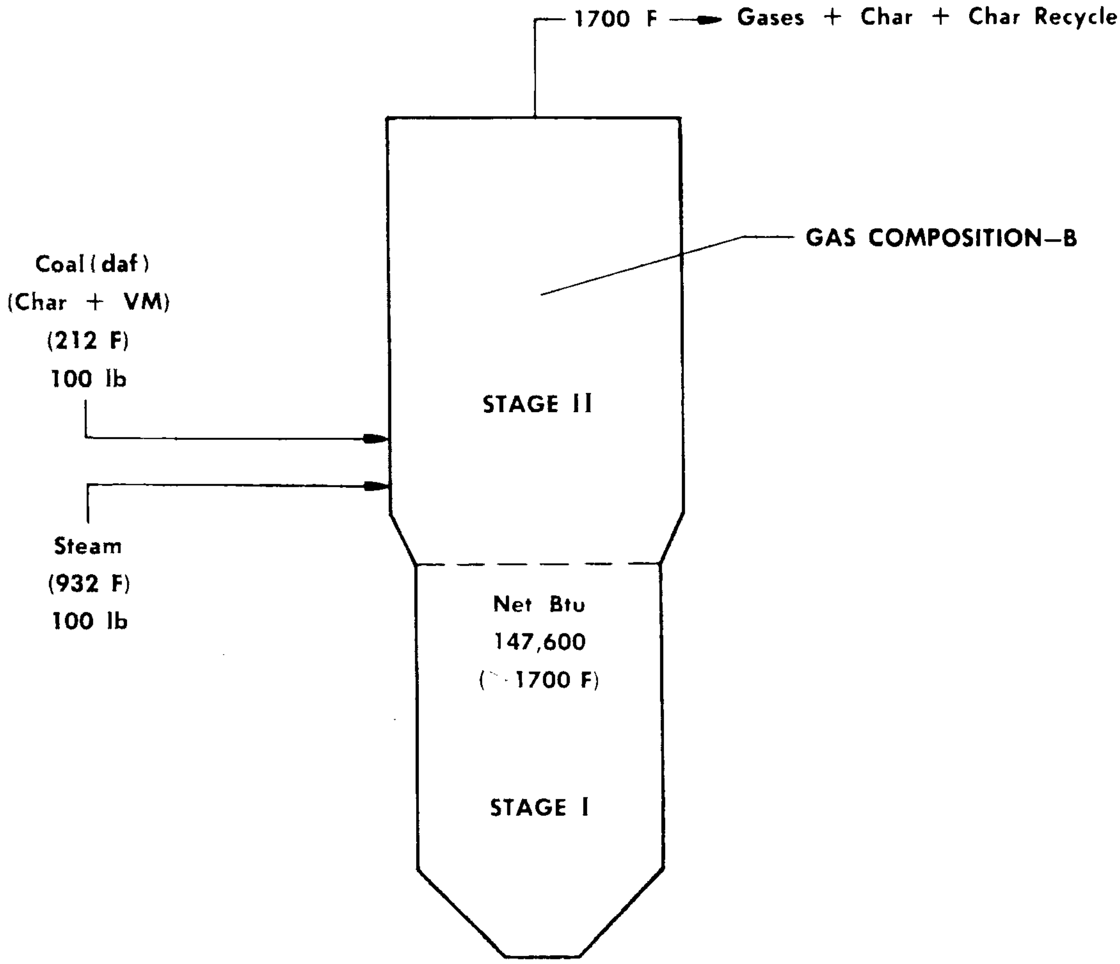
Figure 5.3-1 Steam Temperatures and Gas Composition for Stage I — Gasification of Char in Two-stage Super-pressure Entrained Gasifier

TABLE 5.3-1 HEAT BALANCE FOR TWO-STAGE SUPER-PRESSURE ENTRAINED GASIFIER: STAGE I

Input, Btu		Output Above 1700 F, Btu	
Char, Gasified, Heating Value	814,000	Char, Recycled, Heating Value	521,000
Char, Recycled, Heating Value	521,000	Char, Recycled, Sensible	24,300
Char, Gasified, Sensible	27,000	Slag, Sensible	5,580
Char, Recycled, Sensible	14,050	Gas, Heating Value	587,200
Oxygen, Sensible	11,890	Gas, Sensible	87,900
Steam, Sensible	10,900	Heat Loss	25,200
		Excess Heat Above 1700 F for Stage II	147,660
	1,398,840		1,398,840



GAS COMPOSITION–B

(Stage II Products of VM only)

	Mols*	Mols**	Mol %**
CO_2		0.70	8.0
H_2	1.55	2.25	25.7
CO	1.25	0.55	6.3
CH_4	1.05	1.05	12.0
N_2	0.05	0.05	0.6
H_2S	0.10	0.10	1.1
H_2O	4.75	4.05	46.3
Total	8.75	8.75	100.0

*Before Shift **After Shift

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Figure 5.3-2 Steam Temperatures and Gas Composition for Stage II — Gasification of Volatile Matter in Two-stage Super-pressure Entrained Gasifier

TABLE 5.3-2 HEAT BALANCE FOR TWO-STAGE SUPER-PRESSURE ENTRAINED GASIFIER: STAGE II

Input, Btu		Output, Btu*	
Char, Heating Value	814,000	Char, Heating Value	814,000
Volatile Matter, Net Heating Value	658,800	Char, Sensible	36,200
Coal, Sensible	5,760	Gas, Net Heating Value	675,000
Steam, Sensible	40,800	Gas, Sensible	133,400
Net Heat from Stage I (above 1700 F)	147,600	Unaccounted for	8,360
	1,666,960		1,666,960

*Sensible heat calculated at 1700 F

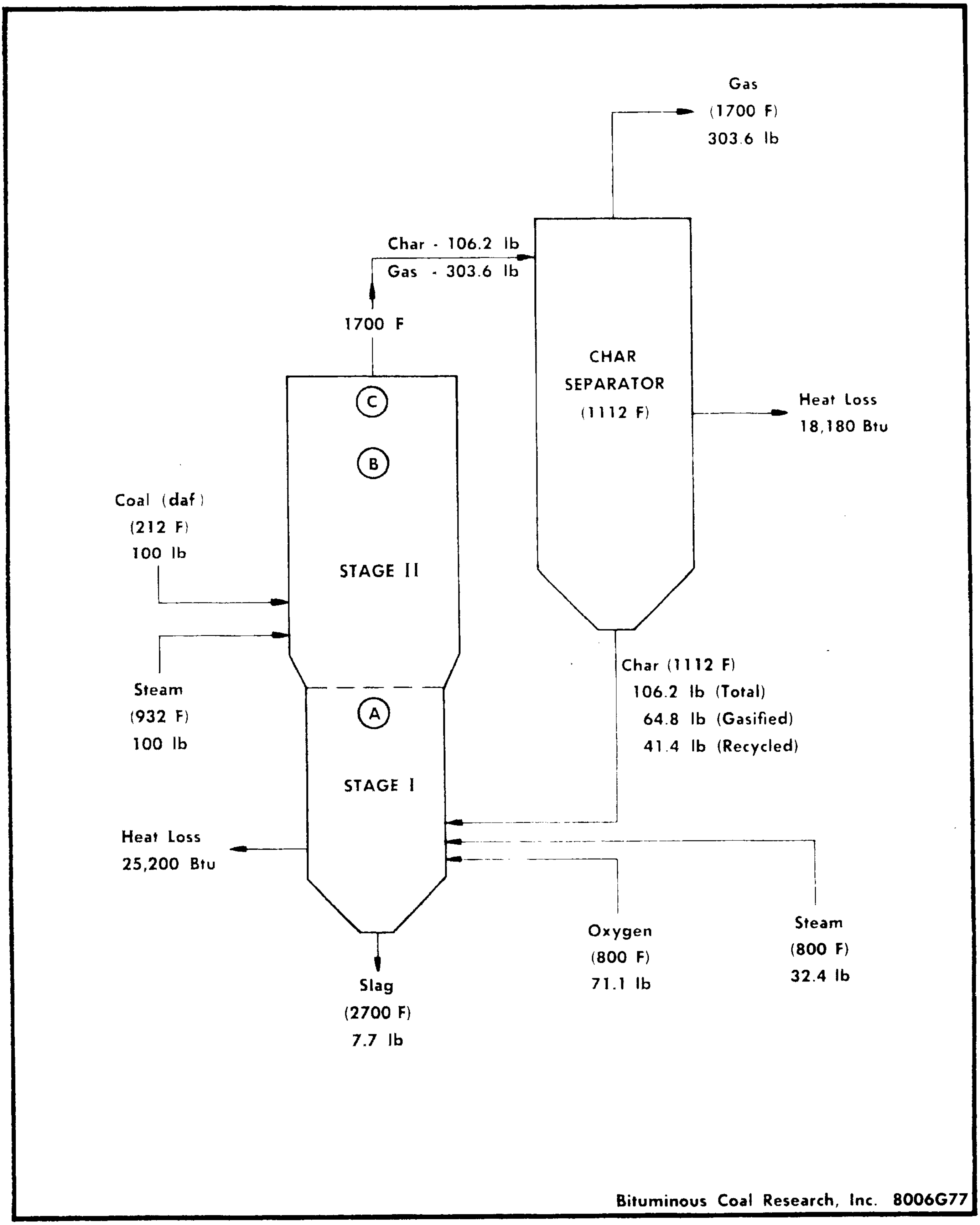

Figure 5.3-3 Steam Temperatures and Composition for All Stages in Two-stage Super-pressure Entrained Gasifier

TABLE 5.3-3 GAS COMPOSITION--STAGE I, STAGE II, AND FINAL GAS FROM TWO-STAGE SUPER-PRESSURE ENTRAINED GASIFIER

Stage	I	II	Final Gas			
Location ***	A	B	C			
					Volume Percent	
	Mols **	Mols **	Mols *	Mols **	(Dry)	Dry, CO_2-Free
CO_2	1.0	0.7	1.70	2.40	21.1	-
H_2	1.3	2.25	3.55	4.25	37.2	47.2
CO	3.7	0.55	4.25	3.55	31.2	39.5
CH_4	-	1.05	1.05	1.05	9.2	11.7
N_2	-	0.05	0.05	0.05	0.4	0.5
H_2S	-	0.10	0.10	0.10	0.9	1.1
H_2O	0.5	4.05	4.55	3.85	-	-
	6.5	8.75	15.25	15.25	100.0	100.0

* Before Shift
** After Shift
*** See Figure 5.3-3

Table 5.3-4 presents an overall heat balance for the total system. The "unaccounted for" heat of about 0.4 percent is due to minor inaccuracies in the various components.

Calculated values for the heat content of the final gas, the thermal efficiency of the overall process, and the other process requirements, are given in Table 5.3-5.

Table 5.3-6 presents an overall material balance for the total system.

The calculated equilibrium pressure required to maintain a methane content of 11.7 percent in the final dry, CO_2-free gas at 1700 F is about 70 atm or 1050 psi using graphite as the solid phase entering the reaction.

$$C + H_2 = CH_4$$

The CH_4 content of gases obtained in fluidized gasification at elevated pressure is consistent with a carbon activity higher than one for the carbon entering this reaction.(10) This high "activity" may be a true thermodynamic property of the carbon entering the reaction or it may be a function of the volatile matter of coal being connected directly into methane. In the two-stage gasifier, this methane is removed from the reactor before it has time to decompose. It is believed that similar conditions of carbon "activity" and/or volatile matter decomposition as found in the Hydrocarbon Research Gasifier will also prevail in the second stage of the Two-stage Super-pressure Gasifier. However, in an evaluation of the process as projected to full scale commercial operation, it would be well to prepare estimates for carbon activity the same as that for graphite and perhaps twice that of graphite. In practice it may be that activities will be observed that are even greater than the 3.4 value of Squires.

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

TABLE 5.3-4 OVERALL HEAT BALANCE FOR TOTAL TWO-STAGE SYSTEM

Input, Btu		Output, Btu	
Coal, Net Heating Value	1,472,800	Gas, Net Heating Value	1,256,000
Coal, Sensible	5,760	Gas, Sensible	228,700
Steam (Stage I)	10,900	Char, Sensible	41,100
Oxygen (Stage I)	11,890	Char (Recycle) Heating Value	521,000
Steam (Stage II)	40,800	Slag, Sensible	5,580
Char, Sensible	41,100	Heat Loss	25,200
Char (Recycle) Heating Value	521,000	Loss, Char Separator	18,180
		Unaccounted for	8,490
	2,104,250		2,104,250

TABLE 5.3-5 OVERALL PROCESS EVALUATION FOR TOTAL TWO-STAGE SYSTEM

1.	Gross heating value of dry, raw gas, Btu/scf	314
2.	Gross heating value of dry, CO_2-free gas, Btu/scf	399
3.	Volume of dry, raw gas per 100 lb maf coal, scf	4,300
4.	Gross heating value of dry, raw gas per 100 lb maf coal, Btu	1,350,000
5.	Thermal Efficiency	

$$\frac{\text{Gross output of gas}}{\text{Gross input of coal}} = \frac{1,350,000}{1,527,000} \times 100 = 88.4 \text{ percent}$$

6.	Oxygen (100 percent) required per MM Btu in gas, lb	52.6
7.	Steam required per MM Btu in gas, lb	98.2
8.	Steam decomposition, percent	59.7

TABLE 5.3-6 MATERIAL BALANCE FOR OVERALL TWO-STAGE SYSTEM

Input, lb		Output, lb	
Coal (maf)	100.0	Gas	303.6
Moisture	1.3	Slag	7.7
Ash	7.7		
Steam, Stage I	32.4		
Oxygen	71.1		
Steam, Stage II	100.0	Unaccounted for	1.2
	312.5		312.5

APPENDIX 5.4

CATALYTIC COAL GASIFICATION EQUILIBRIUM DATA

Gasification of coal with steam is an endothermic reaction. The supply of this heat to the reaction zone is a major cost factor. Means to avoid this endotherm should be investigated to determine whether they could lead to a simpler and lower cost coal gasification process.

The reaction shown in the equation

$$2C_s + 2H_2O_g = CH_{4g} + CO_{2g} \tag{1}$$

is only slightly endothermic. The endotherm amounts to 280 Btu/lb carbon (11) or only 6 percent of the 4470 Btu/lb carbon required for the normal carbon-steam gasification reaction

$$C + H_2O = CO + H_2 \tag{2}$$

The endotherm of reaction (1), slight as it is, decreases further as the reaction temperature increases and amounts to 204 Btu/lb carbon at 600 C (ca. 1100 F).

The advantages of a coal gasification process that is not endothermic obviously are considerable and, therefore, the calculation of the thermodynamic equilibrium of reaction (1) has been made to determine whether favorable conditions for this reaction exist in a pressure and temperature range in which reasonable reaction rates can be expected.

Literature data are shown in Table 5.4-1 for the equilibrium constants for the following equations:

$$C + 2H_2 = CH_4 \tag{3}$$

$$CO_2 + H_2 = CO + H_2O \tag{4}$$

$$C + CO_2 = 2CO \tag{5}$$

These data have been used to calculate the equilibrium constant for equation (1) at temperatures from 500 to 800 C.(12) Starting arbitrarily with water vapor partial pressures of 1, 10, 100, and 1000 atm, the equilibrium pressures of CH_4 and CO_2 have been calculated and are shown also in Table 5.4-1.

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(11) Wagman, D. D., et al, "Heats, free energies, and equilibrium constants of some reactions involving O_2, H_2, H_2O, C, CO, CO_2, and CH_4," Natl. Bur. Std., Res. Paper RP 1634 (1945).

(12) The equilibrium constants used are from Gumz, W., "Brennstoff and Feuerungstechnik," Berlin: Verlag Springer, 1942. p 269 ff.

TABLE 5.4-1 EQUILIBRIUM CONSTANT DATA FOR REACTION: $2C + 2H_2O = CH_4 + CO_2$

Equation	Equilibrium Formula	Temperature C 500	600	700	800
(3) $C + 2H_2 = CH_4$	$Kp = \frac{pCH_4}{p^2H_2}$	2.73	0.54	0.147	0.051
(4) $-2\left[CO_2 + H_2 = CO + H_2O\right]$	$Kp = \frac{pCO \cdot pH_2O}{pCO_2 \cdot pH_2}$	0.205	0.388	0.64	0.95
(5) $C + CO_2 = 2CO$	$Kp = \frac{p^2CO}{pCO_2}$	0.003	0.096	1.16	7.42
(1) $2C + 2H_2O = CH_4 + CO_2$	$Kp = \frac{pCH_4 \cdot pCO_2}{p^2H_2O}$	0.20[a]	0.35[b]	0.42[c]	0.42[d]

Assuming:

	500	600	700	800
$pH_2O =$ 1 atm; $pCH_4 = pCO_2$	0.45	0.6	0.65	0.65
$pH_2O =$ 10 atm; $pCH_4 = pCO_2$	4.5	6.0	6.5	6.5
$pH_2O =$ 100 atm; $pCH_4 = pCO_2$	45	60	65	65
$pH_2O =$ 1,000 atm; $pCH_4 = pCO_2$	450	600	650	650

(a) $\frac{2.73 \times 0.003}{0.205 \times 0.205}$ (b) $\frac{0.54 \times 0.096}{0.388 \times 0.388}$ (c) $\frac{0.147 \times 1.16}{0.64 \times 0.64}$ (d) $\frac{0.051 \times 7.42}{0.95 \times 0.95}$

The equilibrium of equation (1) is overlaid by the equilibria of the methane dissociation reaction

$$CH_{4g} = C_s + 2H_{2g} \tag{6}$$

and of the shift reaction

$$CO_g + H_2O_g = CO_{2g} + H_2H_g \tag{7}$$

The partial pressures from Table 5.4-1 have been used in Table 5.4-2 to calculate the equilibrium hydrogen pressures for reaction (6). The partial pressures for H_2O, CO_2, and H_2 from Tables 5.4-1 and 5.4-2, in turn, have been used in Table 5.4-3 to obtain the equilibrium pressures for CO in reaction (7).

It may be remarked that this straightforward procedure establishes an arbitrary H_2O/C ratio which probably is not the optimum ratio. Furthermore, the equations used are valid only for the ideal gas state. At the high pressures proposed for this process, deviations from the ideal gas laws must be expected. However, it may be assumed at this time that this factor and the use of coal instead of carbon will not lead to significant deviations.

Using the data from Tables 5.4-1 to 5.4-3, the total equilibrium pressures and gas composition in volume percent at the selected temperatures and steam partial pressures have been calculated and are shown in Table 5.4-4. From this, in turn, the dry gas compositions in Table 5.4-5 have been calculated.

The dry gas composition from Table 5.4-5 and the H_2O content in volume percent of the wet gas have been plotted against total pressure for 500, 600, 700, and 800 C in Figures 5.4-1, 5.4-2, 5.4-3, and 5.4-4, respectively. Suitable points from these figures have in turn been used to plot the composition of the raw dry gas in Figure 5.4-5. The equilibrium water vapor content of the product gas before water condensation is also shown in Figure 5.4-5 for temperatures from 500 to 800 C and for three pressures: 70, 200, and 500 atm.

The points from Figure 5.4-5 have been replotted in Figure 5.4-6 to give the gas composition after CO_2 removal.

The CH_4 content of the gas after CO_2 removal is also shown in Table 5.4-6. If equilibrium can be established at temperatures around 600 C and pressures in the 70-200 atm range, the bulk of the fuel gas will be CH_4. To achieve this equilibrium at an acceptable reaction rate, acceleration of the reaction by use of catalysts is required. At a temperature of 600 C, supply of the small amount of heat required should be possible through conventional heat transfer surfaces, and thus the use of oxygen could be eliminated. The temperature and pressure limit of this type of operation will be given by the properties, availability, and cost of high temperature steels.

Even at higher temperatures, e.g., 800 C and a pressure of 70 atm, a gas with 33 percent CH_4 is obtained. This methane content represents 60 percent of the Btu content of the gas and is much higher than that obtained in the Lurgi process. Gasification with oxygen at this temperature (800 C) and e.g., 70 atm pressure should be attractive because the oxygen consumption is lower and the Btu loss in the methanation reaction is reduced.

TABLE 5.4-2 EQUILIBRIUM CONSTANT DATA FOR REACTION: $C + 2H_2 = CH_4$

Temperature	500 C	600 C	700 C	800 C
$Kp = \frac{pCH_4}{p^2H_2} =$	2.73	0.54	0.147	0.051
A. For pCH_4 atm =	0.45	0.6	0.65	0.65
pH_2 atm =	0.41	1.05	2.1	3.6
B. For pCH_4 atm =	4.5	6.0	6.5	6.5
pH_2 atm =	1.3	3.3	6.6	11.3
C. For pCH_4 atm =	4.5	60	65	65
pH_2 atm =	4.1	10.5	21	36
D. For pCH_4 atm =	450	600	650	650
pH_2 atm =	13	33	66	113

TABLE 5.4-3 EQUILIBRIUM CONSTANT DATA FOR REACTION: $CO_2 + H_2 = CO + H_2O$

Temperature	500 C	600 C	700 C	800 C
$Kp = \frac{pCO \times pH_2O}{pCO_2 \times pH_2}$	0.205	0.388	0.64	0.95
For pH_2O = 1 atm; pCO =	0.40	0.24	0.87	2.2
For pH_2O = 10 atm; pCO =	0.12	0.77	2.8	7.0
For pH_2O = 100 atm; pCO =	0.38	2.4	8.7	22.0
For pH_2O = 1,000 atm; pCO =	1.2	7.7	28	70

TABLE 5.4-4 PARTIAL AND TOTAL EQUILIBRIUM PRESSURE AND GAS COMPOSITION DATA FOR THE REACTION:

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

		Temperature, C							
		500		600		700		800	
		atm	%	atm	%	atm	%	atm	%
A.	pH_2O	1	43	1	29	1	19	1	12
	pCO_2	0.45	19	0.6	17	0.65	12	0.65	8
	pH_2	0.41	17	1.05	30	2.1	40	3.6	45
	pCO	0.04	2	0.24	7	0.87	17	2.2	27
	pCH_4	0.45	19	0.6	17	0.65	12	0.65	8
	Total	2.35	100	3.49	100	5.27	100	8.1	100
B.	pH_2O	10	49	10	38	10	31	10	24
	pCO_2	4.5	22	6	23	6.5	20	6.5	16
	pH_2	1.3	6.5	3.3	13	6.6	20	11.3	27
	pCO	0.1	0.5	0.8	3	2.8	9	7.0	17
	pCH_4	4.5	22	6	23	6.5	20	6.5	16
	Total	20.4	100	26.1	100	32.4	100	41.3	100
C.	pH_2O	100	51.7	100	43	100	39	100	35
	pCO_2	45	23	60	26	65	25	65	23
	pH_2	4.1	2.1	10.5	4	21	8	36	12
	pCO	0.4	0.2	2.4	1	8.7	3	22	7
	pCH_4	45	23	60	26	65	25	65	23
	Total	194	100	233	100	260	100	288	100
D.	pH_2O	1,000	52	1,000	44	1,000	42	1,000	40
	pCO_2	450	23.6	600	27	650	27	650	26
	pH_2	13	0.7	33	1.6	66	3	113	5
	pCO	1	0.1	7.7	0.4	28	1	70	3
	pCH_4	450	23.6	600	27	650	27	650	26
	Total	1,914	100	2,241	100	2,394	100	2,483	100

TABLE 5.4-5 GAS COMPOSITION DATA FOR REACTION (1):
$2C + 2H_2O = CH_4 + CO_2$, PERCENT DRY BASIS

	Temperature, C			
	500	600	700	800
A. Total Pressure, atm	2.35	3.49	5.27	8.1
Carbon Dioxide	33	24	15	9
Hydrogen	31	42	50	51
Carbon Monoxide	3	10	20	31
Methane	33	24	15	9
B. Total Pressure, atm	20.4	26.1	32.4	41.3
Carbon Dioxide	43	37	29	21
Hydrogen	13	21	30	36
Carbon Monoxide	1	5	12	22
Methane	43	37	29	21
C. Total Pressure, atm	194	233	260	288
Carbon Dioxide	47	45	41	35
Hydrogen	4	8	13	19
Carbon Monoxide	1	2	5	11
Methane	48	45	41	35
D. Total Pressure, atm	1,914	2,241	2,394	2,483
Carbon Dioxide	49.3	48.3	46.7	44
Hydrogen	1.3	2.7	4.6	7
Carbon Monoxide	0.1	0.7	2.0	5
Methane	49.3	48.3	46.7	44

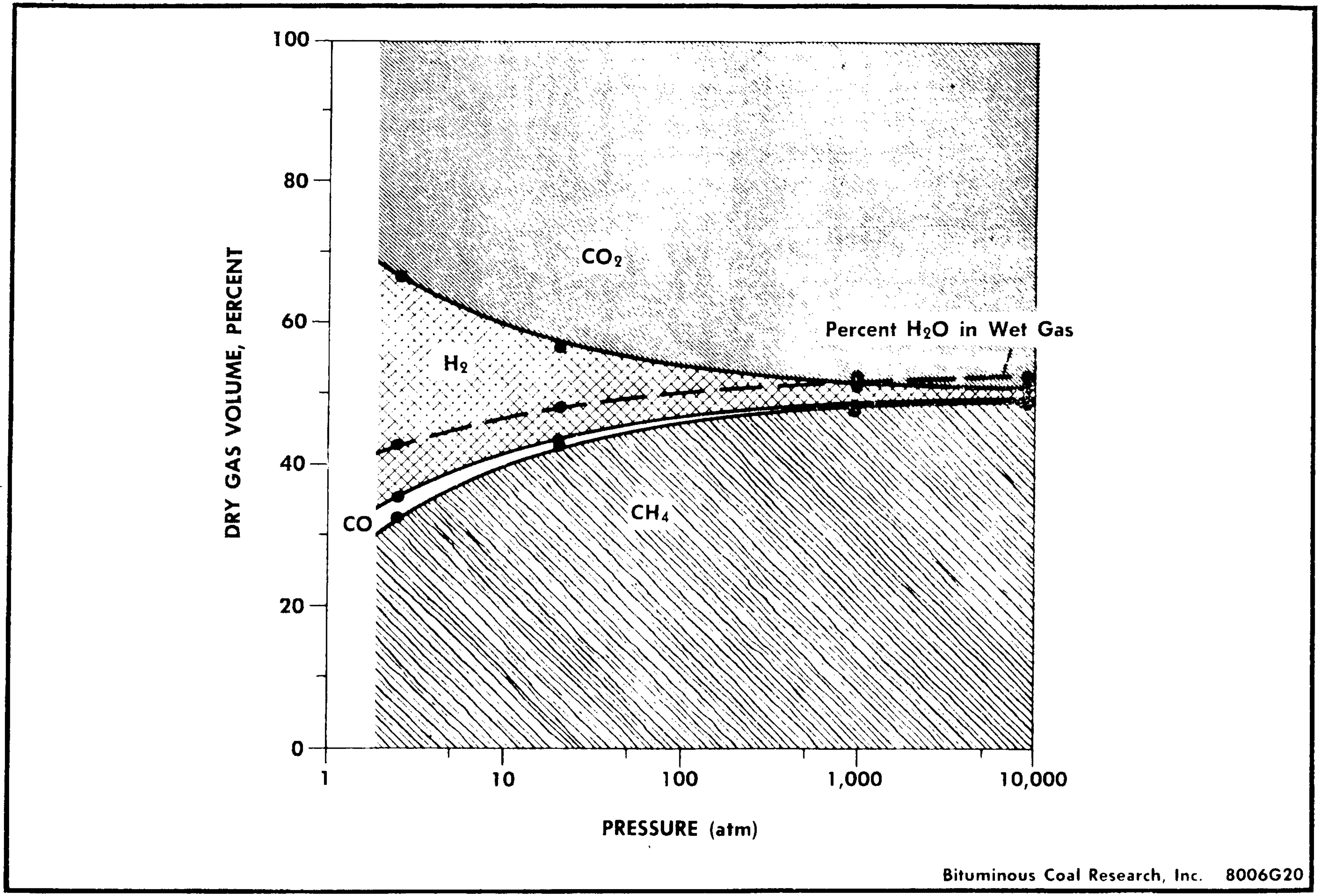

Figure 5.4-1 Catalytic Coal Gasification Data: Gas Composition at 500 C

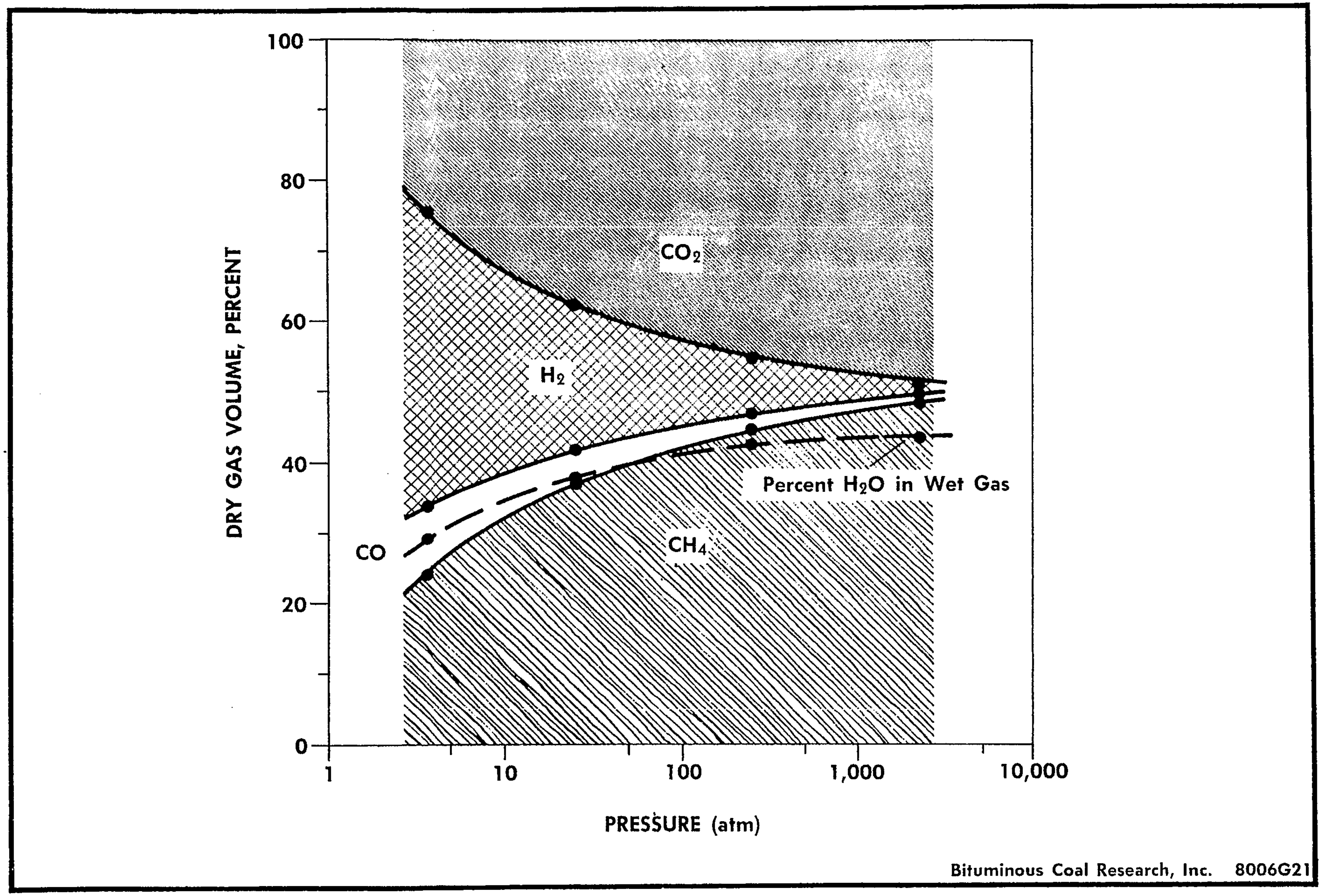

Figure 5.4-2 Catalytic Coal Gasification Data: Gas Composition at 600 C

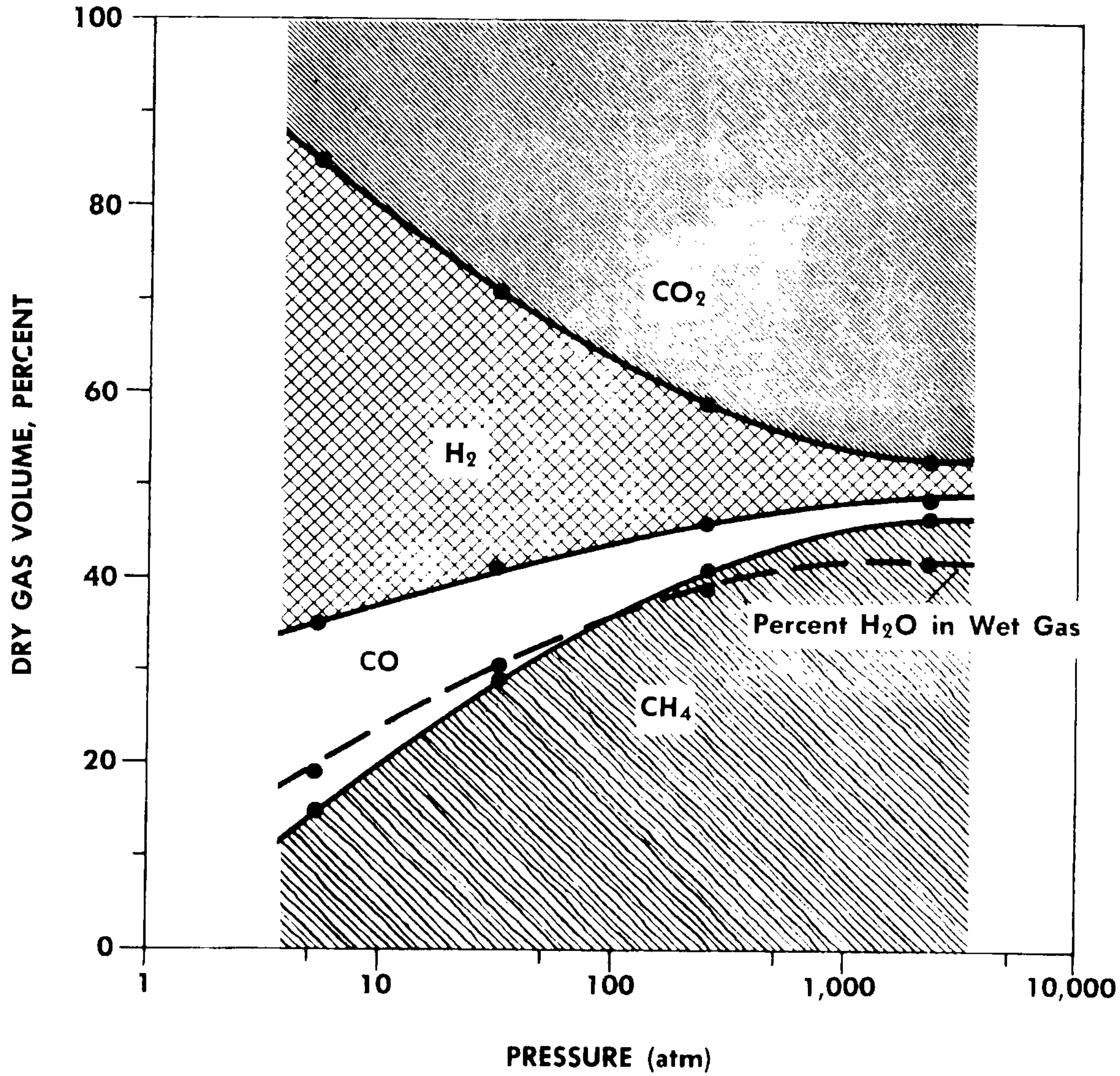

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Figure 5.4-3 Catalytic Coal Gasification Data: Gas Composition at 700 C

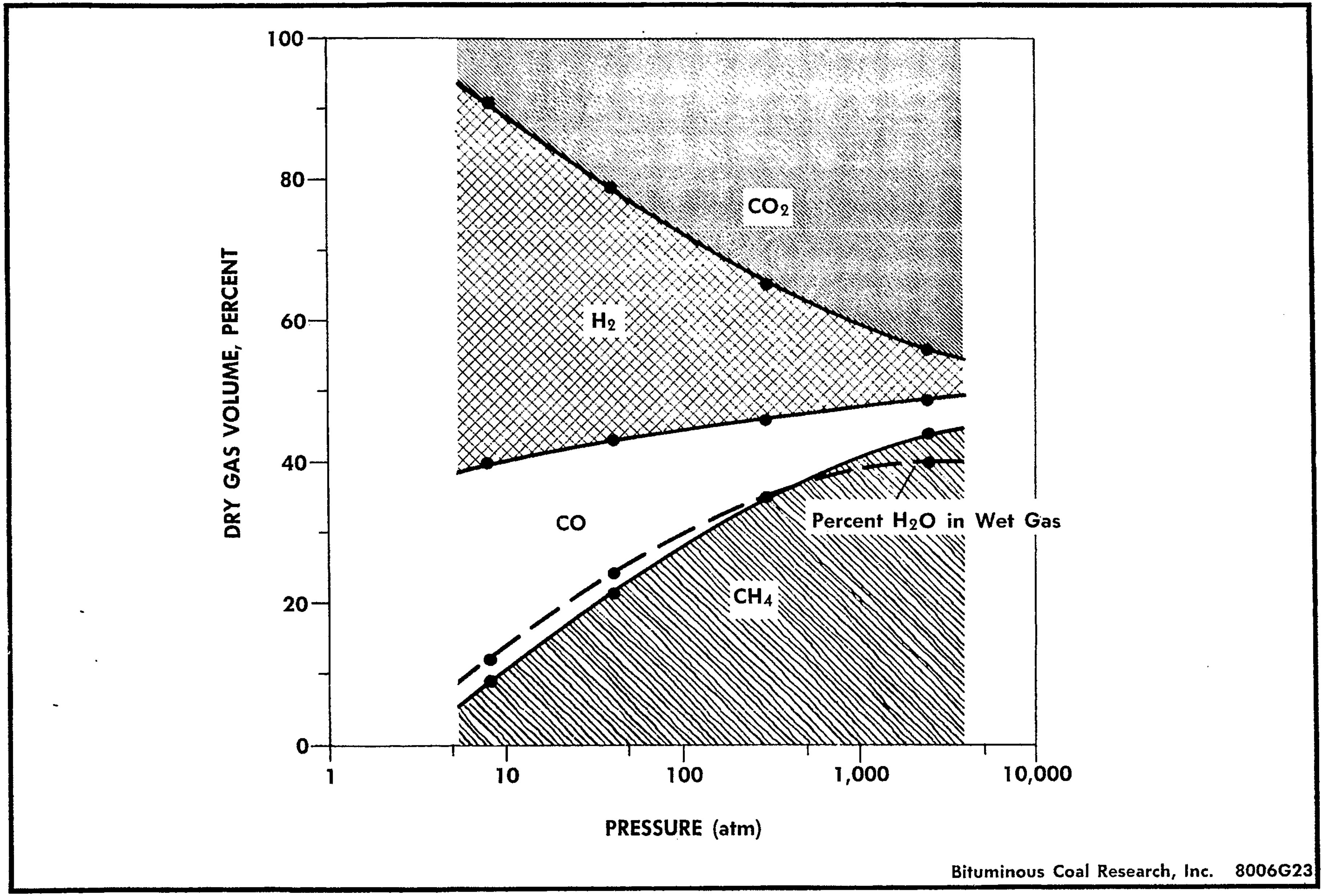

Figure 5.4-4 Catalytic Coal Gasification Data: Gas Composition at 800 C

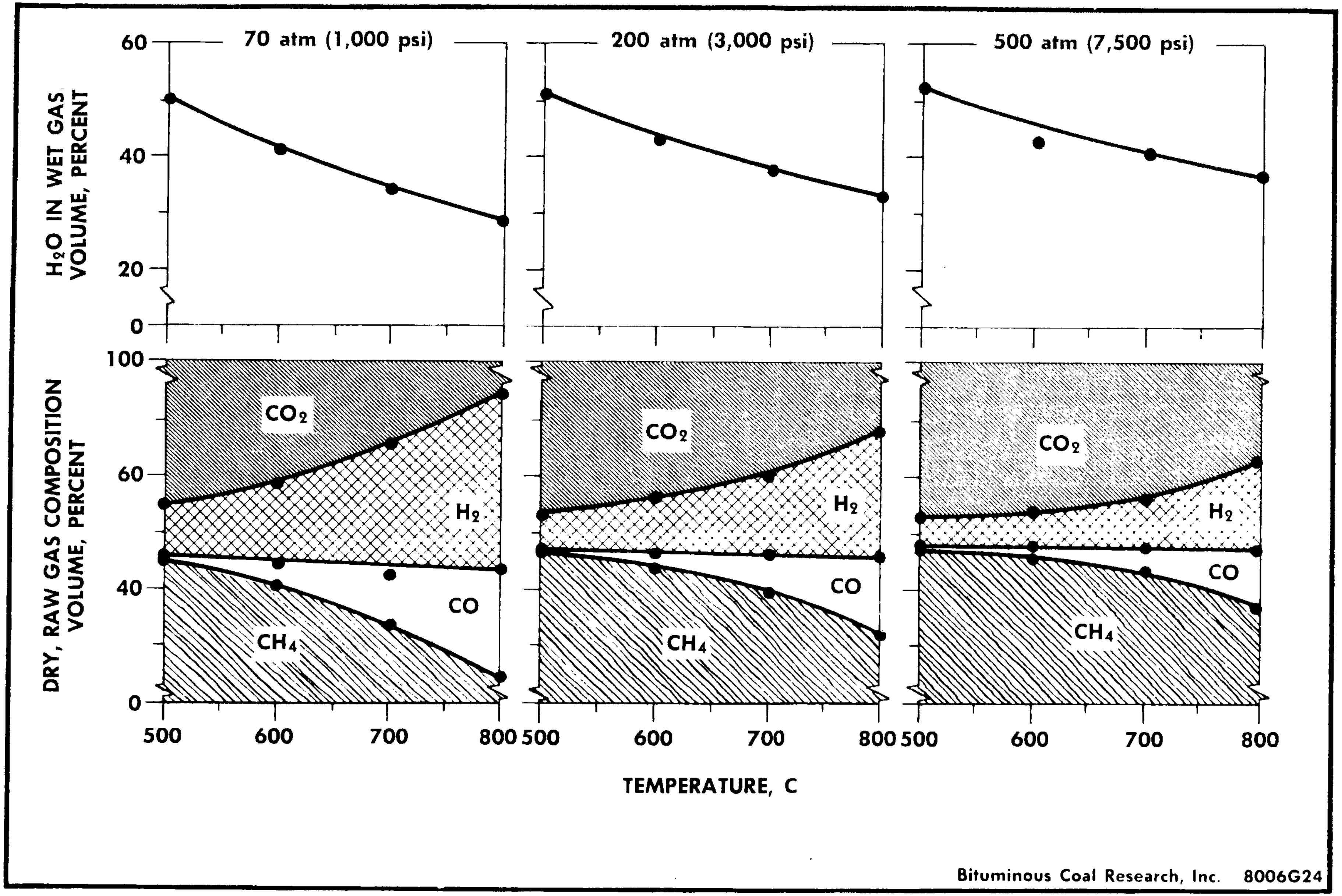

Figure 5.4-5 Catalytic Coal Gasification Data: Gas Composition; Total Pressure and Temperature

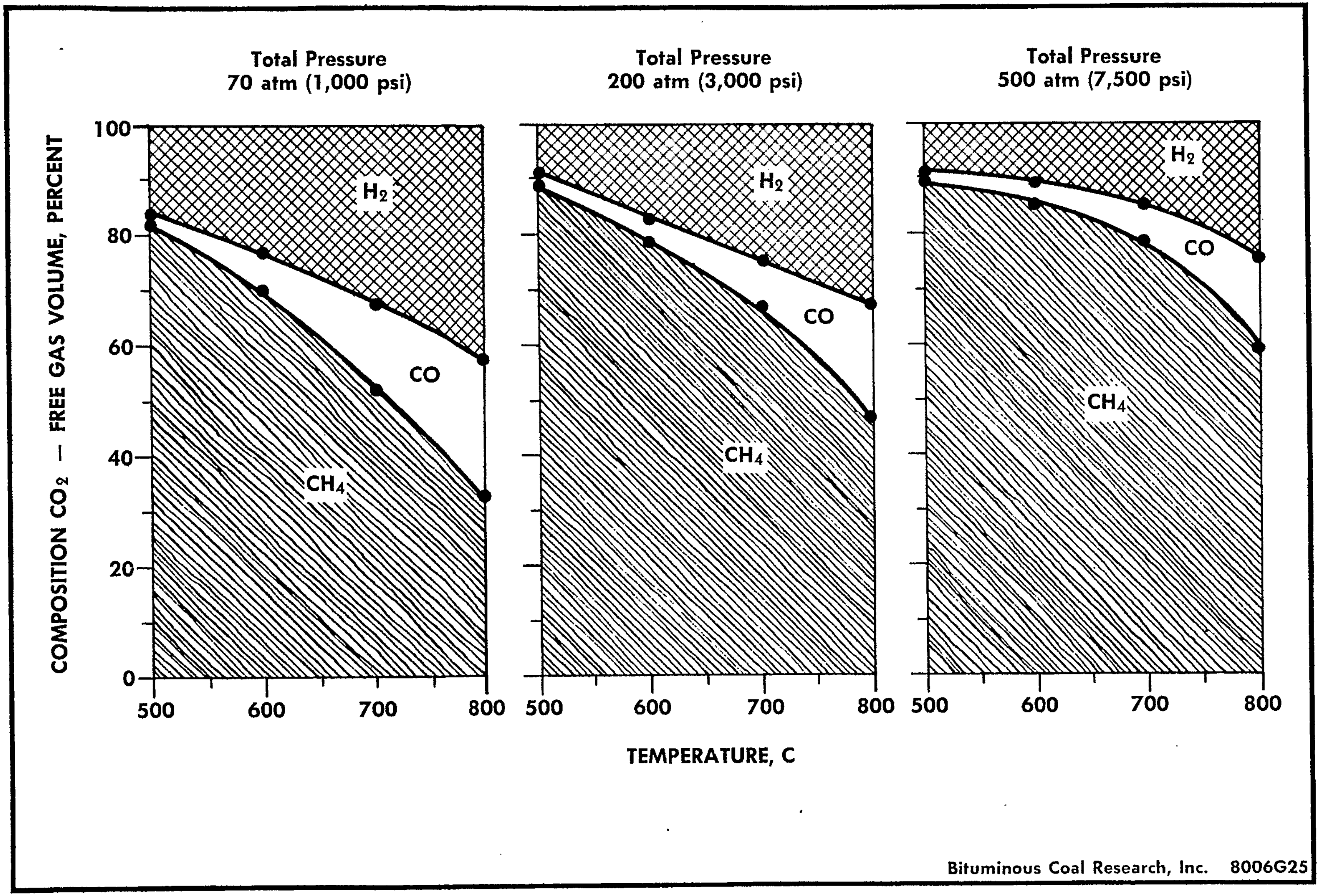

Figure 5.4-6 Catalytic Coal Gasification Data: Gas Composition, CO_2 — Free Gas

TABLE 5.4-6 CATALYTIC COAL GASIFICATION DATA: METHANE CONTENT OF GASES AFTER CARBON DIOXIDE REMOVAL

Temperature, C	Pressure, atm		
	70	200	500
500	82	89	90
600	70	79	85
700	47	66	79
800	33	47	60

APPENDIX 5.5

PROCESS DATA FOR TWO-STAGE FLUIDIZED SUPER-PRESSURE GAS PRODUCER (PROCESS 46)

Producer gas at high temperature and high pressure is required for a modified version of the steam-iron process for hydrogen manufacture. The hydrogen is intended for subsequent use in the hydrogasification of bituminous coal.(13) Hot char from the hydrogasification operation will be supplied to the gas producer which operates at 1500 psig and at temperatures of 2000 F and above.

Operation of the gas producer is based on two stages of gasification in the following manner:

1. The first stage of gasification is carried out according to the method used in the slagging Ruhrgas Vortex Gasifier.(14) The exit gases have a CO to CO_2 ratio of 22.8:5.1, and 30 percent of the vortex feed is recycled as dust containing 30 percent ash. Air is preheated to 1400 F, and all ash is removed in this stage as slag at 2700 F.

2. The second stage of gasification is carried out in a Hydrocarbon Research-type fluidized bed. Fresh hot char at 1200 F is injected with steam at 1400 F. Steam decomposition is 80 percent. About 50 percent carbon is maintained in the fuel bed with about 45 percent of the char being gasified.

3. Excess char from Stage II will be cycled as feed to Stage I.

4. A heat loss of 500 Btu per pound of fresh char fed to Stage I will be accounted for in Stage II.

5. The homogeneous water gas shift reaction ($CO + H_2O \rightleftharpoons CO_2 + H_2$) reaches equilibrium.

6. Methane is formed in accord with the reaction: ($CO + 3H_2 \rightarrow CH_4 + H_2O$).

7. Fresh char fed to Stage II has the composition as shown in Table 5.5-1, and a heating value as calculated by DuLong's formula.

8. Char requirements for the two-stage gas producer are based on the following:

   a. A hydrogasification plant producing 250 MM scfd of pipeline gas requires 250 MM scfd of usable ($CO + H_2$).

   b. The lower useful limit of ($CO + H_2$) in the combined ($CO + H_2 + CO_2 + H_2O$) in the producer gas

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(13) See Figure 8-3

(14) See Process No. 35, Table 3.1, and Appendix 3.5

TABLE 5.5-1 ANALYSES OF COAL (SIZE 14 MESH x 0) AND DERIVED CHAR AS USED IN TWO-STAGE FLUIDIZED SUPER-PRESSURE GAS PRODUCERS**

	Weight Percent, Dry Basis		
Component	Dry Coal	Dry Char	Daf Char
Carbon	71.8	72.0	89.0
Hydrogen	5.2	2.9	3.6
Oxygen	9.8	1.5	1.8
Nitrogen	2.1	2.1	2.6
Sulfur	2.0	2.4	3.0
Ash	9.1	19.1	---
Total	100.0	100.0	100.0

Gross Heating Value,* Btu/lb	15,140
Net Heating Value,* Btu/lb	14,800

* Calculated by use of DuLong's formula.

** Char corresponds to that to be expected from devolatilization of coal at about 1300 F.

is 29.5 percent by volume.(15)

c. The usable reduction gas ($CO + H_2$) in the producer gas is that amount in excess of the lower useful limit.

9. Material and heat balances (Tables 5.5-2 and 5.5-3) are based on:

a. 100 pound daf char,

b. net heating values, and

c. a base temperature of 32 F.

10. Calculations show that as the temperature of the fluidized bed in the second stage is increased in the range from 1700 to 2100 F, the overall gasification efficiency decreases slightly, while the yield of usable ($CO + H_2$) increases. Thus high gas exit temperatures are more favorable. This result is based on heat and material balance calculations and the assumption that the equilibria of the reactions

$$CO + H_2O = CO_2 + H_2$$
$$CO + 3H_2 = CH_4 + H_2O$$

are established at the exit temperature of Stage II. This assumption appears justified in view of the high operating temperature of 2000 F. Experimental verification of this and a determination of the carbon-steam reaction rate in Stage II will be necessary. Estimates of the carbon-steam reaction rate and the steam decomposition that can be expected at 2000 F, and the material and heat balance calculations lead to the following specific assumption for the cost estimate:

a. five gasifier units each with a capacity of 1,030 tons per day of daf char,

b. a first stage generator 6 ft 6 in. in diameter, and

c. a second stage generator 12 ft in diameter and 40 ft high with a fluidized bed 20 ft deep with about 10 lb per cu ft char.

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(15) Gasior, S. J., et al, "Production of synthesis gas and hydrogen by the steam-iron process: Pilot plant study of fluidized and free-falling beds," U.S. Bur. Mines, Rept. Invest. 5911 (1961).
Spent producer gas analysis, p 40.

H_2	7.6%
CO	8.1%
H_2O	17.9%
CO_2	19.6%
CH_4	0.3%
H_2S	0.2%
N_2	46.3%

$$\frac{H_2 + CO}{H_2 + CO + H_2O + CO_2} = 29.5\%$$

TABLE 5.5-2 MATERIAL BALANCE FOR TWO-STAGE FLUIDIZED SUPER-PRESSURE GAS PRODUCER

Operating Conditions: 2,000 F Outlet Gas Temperature
80 Percent Steam Decomposition

Basis: 100 lb daf char

	Input, lb		Output, lb
STAGE I			
Char		Gas	
Carbon	55.0	To Stage II	426.0
Ash	23.6		
Dust		Dust	
Carbon	23.5	Carbon	23.5
Ash	10.1	Ash	10.1
Air	371.0	Slag	23.6
TOTAL	483.2	TOTAL	483.2
STAGE II			
Char (100 lb)		Gas	534.7
Carbon	89.0		
H, O, N, S	11.0	Carbon	55.0
Ash	23.6		
		Ash	23.6
Steam	63.7		
Dust		Dust	
Carbon	23.5	Carbon	23.5
Ash	10.1	Ash	10.1
Gas			
From Stage I	426.0		
TOTAL	646.9	TOTAL	646.9

TABLE 5.5-3 HEAT BALANCE FOR TWO-STAGE FLUIDIZED SUPER-PRESSURE GAS PRODUCER

Operating Conditions: 2,000 F Outlet Gas
80 Percent Steam Decomposition

	Input, M Btu		Output, M Btu
STAGE I			
Char		Gas to Stage II	
Heating Value	797.00	Heating Value	459.00
Sensible Heat	47.70	Sensible Heat	226.65
Dust (1,900 F)		Dust (2,000 F)	
Heating Value	340.00	Heating Value	340.00
Sensible Heat	20.42	Sensible Heat	21.86
Air		Heat Loss	50.00
Sensible Heat	129.90	Slag Heat	18.05
		Exotherm	219.46*
TOTAL	1,335.02	TOTAL	1,335.02
STAGE II			
Fresh Char		Raw Gas	
Heating Value	1,480.00	Heating Value	1,258.60
Carbon, Sensible Heat	37.30	Sensible Heat	346.95
Ash, Sensible Heat	6.14		
Steam		Char to Stage I	
Sensible Heat	43.00	Heating Value	797.00
		Sensible Heat	51.50
Dust		Dust	
Heating Value	340.00	Heating Value	340.00
Sensible Heat	21.86	Sensible Heat	21.86
Gas from Stage I		Unaccounted	17.44
Heating Value	459.00		
Sensible Heat	226.65		
Exotherm			
From Stage I	219.40		
TOTAL	2,833.35	TOTAL	2,833.35

* This exotherm of 219.5 M Btu is available and sufficient to heat gaseous products to a temperature of about 3,500 F in the absence of steam.

Table 5.5-4 presents the composition of the final raw gas, and Table 5.5-5 shows the summary data.

Figure 5.5-1 shows the composition and temperature of the various streams--gas, solids, and heat loss--entering and leaving Stage I of the producer. The gas composition for the exit gases at point A is shown on both a mol and mol percent basis.

Figure 5.5-2 shows the composition and temperature of the various streams entering and leaving Stage II. The composition of the gases for point B is that only for the products resulting from reaction of the daf char (45 lb) with steam, and after methane formation and shift equilibrium have been established at 2000 F.

Figure 5.5-3 presents the overall system flow sheet.

TABLE 5.5-4 FINAL RAW GAS COMPOSITIONS (2000 F) FOR TWO-STAGE SUPER-PRESSURE GAS PRODUCER

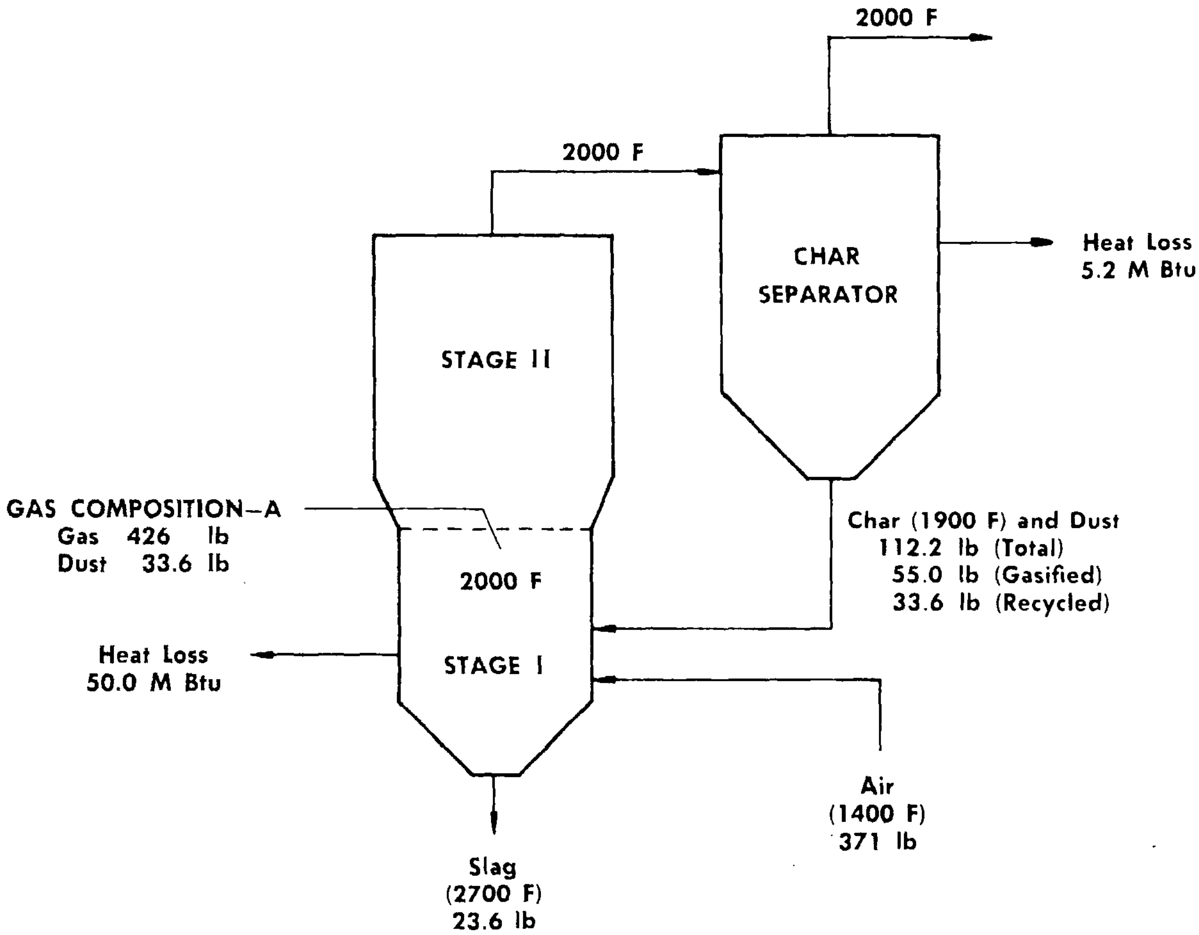
		Volume, Percent	
Component	Lb Mols*	Wet	Dry
Carbon Dioxide	0.87	3.9	4.1
Hydrogen	3.57	15.9	16.7
Carbon Monoxide	6.24	27.9	29.3
Methane	0.30	1.3	1.4
Nitrogen	10.25	45.8	48.1
Hydrogen Sulfide	0.10	0.4	0.4
Water	1.07	4.8	--
Total	22.40	100.0	100.0

*Based on 100 lb daf char

TABLE 5.5-5 SUMMARY DATA FOR TWO-STAGE FLUIDIZED SUPER-PRESSURE GAS PRODUCER

(Basis: 100 lb daf char and 2000 F)

1. Net Heating Value, Btu/scf	148
2. Thermal Efficiency, percent (Item 4 ÷ Item 5 x 100)	85.1
3. Volume of dry gas, scf	8,070
4. Net Heating Value of dry gas, M Btu	1,258.6
5. Net Heating Value of daf char, M Btu	1,480.0
6. Steam used, lb	63.7
7. ($CO + H_2$) produced, scf	3,720
8. ($CO + H_2$) usable, scf	2,410
9. Heat in spent gas, M Btu	524.6
10. Heat in spent gas, percent	41.7
11. Requirement for 250 MM scfd pipeline gas	
a. daf, char, tons/day	5,160
b. dry char, tons/day	6,330



GAS COMPOSITION–A

	Mols	Mol %
CO_2	0.83	5.6
CO	3.75	25.5
N_2	10.15	68.9
Total	14.73	100.0

Basis: 100 lb daf Char Input to Stage II

Heat Balance: Output Based on Stage I Product of 2000 F Exit Temperature

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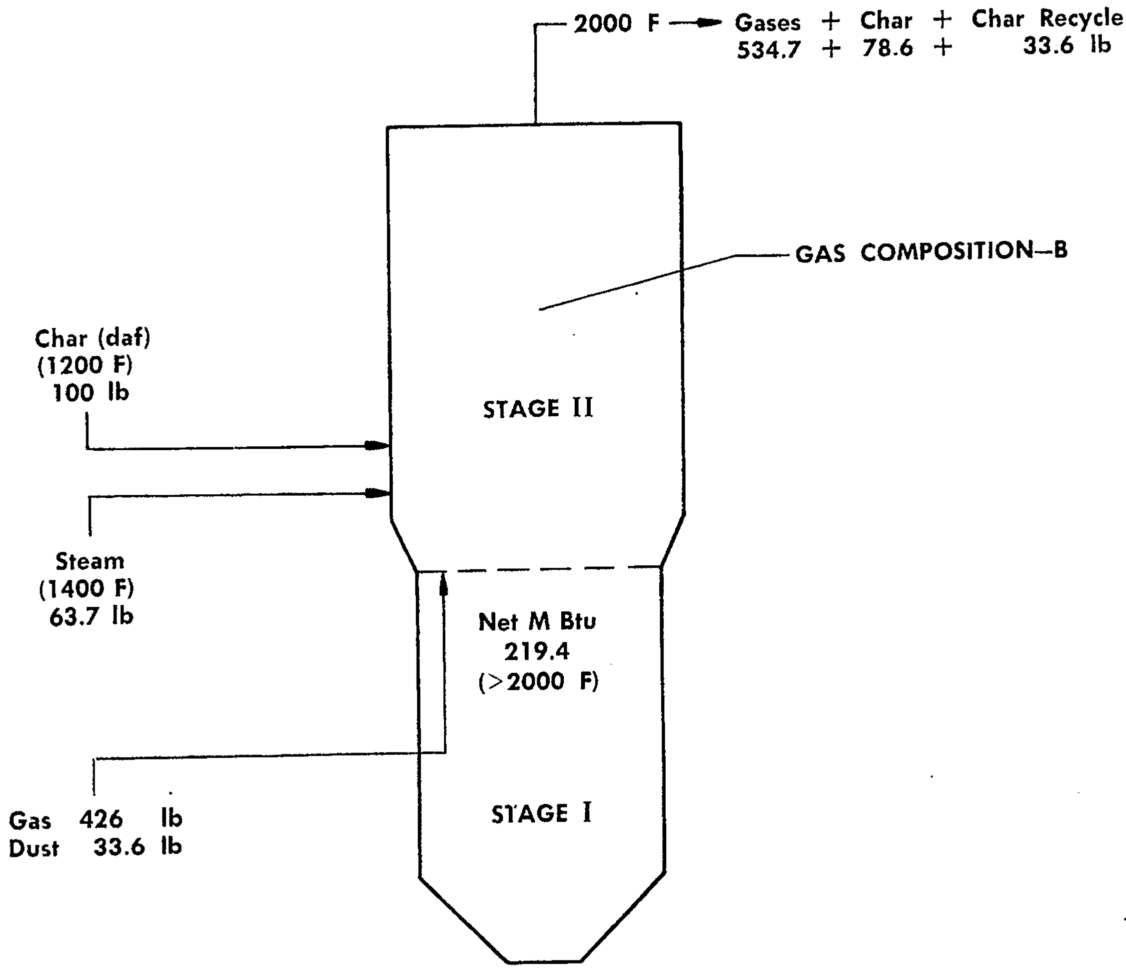
Figure 5.5-1 Steam Temperatures and Gas Composition for Stage I — Two-stage Fluidized Super-pressure Gas Producer



GAS COMPOSITION—B
(Stage II Products only)

	Mols	Mol %
CO_2	0.25	3.3
H_2	3.78	49.3
CO	2.28	29.7
CH_4	0.30	3.9
N_2	0.10	1.3
H_2S	0.10	1.3
H_2O	0.86	11.2
Total	7.67	100.0

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Figure 5.5-2 Steam Temperatures and Gas Composition for Stage II — Two-stage Fluidized Super-pressure Gas Producer

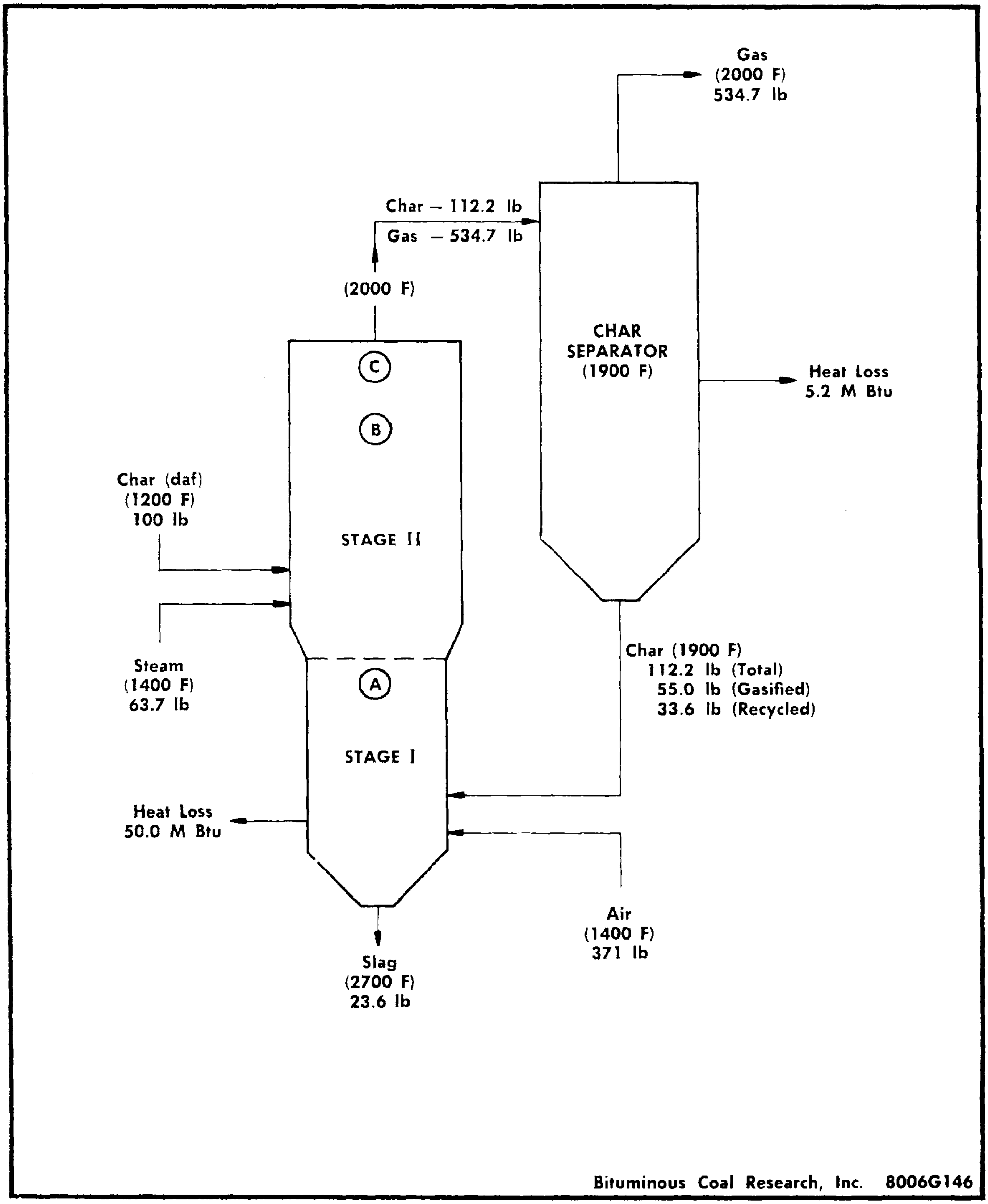

Figure 5.5-3 Overall Flow Sheet for Gasification of Char in Two-stge Fluidized Super-pressure Gas Producer

APPENDIX 5.6

PROCESS DATA FOR 150 PSIG AIR BLOWN LURGI GAS PRODUCER (PROCESS 37)

Heat and material balances and summary evaluations of the system requirements are presented for the production of gas from operation of an air blown Lurgi gasifier at 150 psig (10 atm) pressure. The product gas is intended for use as fuel for gas turbine operation. The basic data were obtained from information supplied by Lurgi.(16)

Briefly, the production of producer gas in an air blown Lurgi gasifier under pressure is based on data by Lurgi from a study conducted during start-up operations of 3.7 meter (12 ft 2 in.) ID units at the SASOL plant in South Africa. Data on the effects of pressure were obtained from operations at different pressures with air at low loads. Data were also obtained for air operation at 14 atmospheres.

A typical analysis for the dry, ammonia- and tar-free gas obtained during operation with air at 14 atm is given in Table 5.6-1. Tar and ammonia yields on a dry ash-free coal basis were 6.1 and 1.3 weight percent, respectively.

For the present calculations, the heating value of the tar was taken as 16,750 Btu/lb and the tar composition as 83 percent carbon, 9 percent hydrogen, and 8 percent oxygen.

As a basis for the present calculations, 100 lb of dry, ash-free coal was selected. Gross heating values were used throughout as well as a base temperature of 32 F. The feed coal was a high volatile A bituminous coal, namely, Pittsburgh seam coal, with the proximate and ultimate analysis as previously given in Table 6-1.

The various assumptions made and an outline of procedure follow:

1. Equivalent amounts of the individual elements required for tar, ammonia and hydrogen sulfide formation, based on the feed coal analyses, were subtracted from the weight of the feed coal (daf basis).

2. A portion of the net carbon of the coal was then reacted with oxygen of air to form carbon dioxide.

3. The air requirement was based on the ratio of N_2 to (CO_2 + CO + CH_4) as calculated from the analysis of the Lurgi producer gas.

4. The remaining carbon was then reacted with steam to form CO and H_2 in accord with the reaction:

$$C + H_2O \longrightarrow CO + H_2$$

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(16) See Process No. 37, Table 3-1, and Appendix 3-5

TABLE 5.6-1 COMPOSITION OF GAS FROM AIR BLOWN LURGI GAS PRODUCER OPERATIONS AT 14 ATM

(Dry, Ammonia- and Tar-free Basis)

Gas	Volume, Percent
Carbon Dioxide	16.0
Hydrogen	23.2
Carbon Monoxide	15.5
Methane	4.5
Nitrogen	40.8
Total	100.0

5. The analysis of the gas thus obtained was adjusted for methane formation in accord with:

$$CO + 3H_2 \rightleftharpoons CH_4 + H_2O$$

6. The resulting gas from the above operations was then adjusted using the shift reaction to give a CO_2 to CO ratio according to the Lurgi gas analysis, i.e., 16 to 15.5:

$$CO + H_2O \rightleftharpoons CO_2 + H_2$$

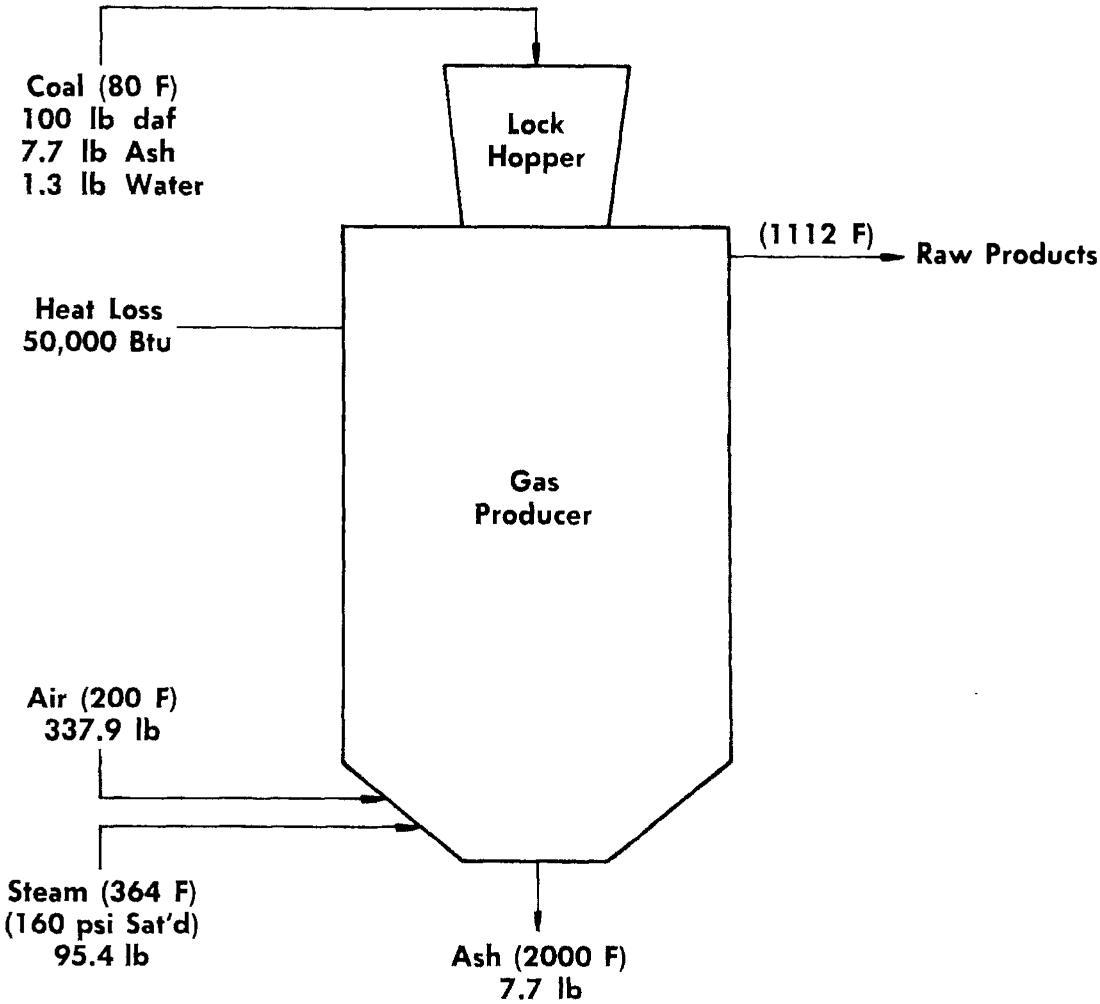
7. Enthalpies and heating values of the raw gas (or output products) were then calculated, slag and heat losses added, and then compared for balance with the enthalpy and heating values of the input material.

8. The raw gas capacity of the air blown 3.7 meter (12 ft 2 in.) ID Lurgi gasifier at 14 atm was 1.118 MM scfh. This capacity was increased by 32 percent to adjust for the capacity increase achieved with oxygen above the original design figure of Lurgi for full load conditions. Then, this capacity was further adjusted by a factor equivalent to the square root of the ratio of the operating pressures or 0.84 to compensate for reduced pressure operation, that is, from 14 atm to 10 atm. The calculated capacity at 10 atm was:

$$1.118 \text{ MM scfh} \times 1.32 \times 0.84 = 1.23 \text{ MM scfh}$$

Figure 5.6-1 shows the temperature, composition and weight of the various streams of materials entering and leaving the gasifier. Table 5.6-2 gives the material balance for the system and Table 5.6-3 gives the heat balance. The unaccounted for heat, being +61,200 Btu, or 3.7 percent of the total heat content of the system, is considered to be more than adequate to compensate for a possible carbon conversion lower than the assumed 100 percent. Table 5.6-4 shows an analysis of the dry, raw product gas and lists the heating value of the tar- and moisture-free gas as 158 Btu per scf. The heating value of the tar produced amounts to 6.7 percent of the heat content of the coal as fired. As shown in Table 5.6-5, the thermal efficiency of the cold gas is 80.5 percent; including tar, it is 87.2 percent. The hot gas efficiency is 91.0 percent.

Table 5.6-6 presents a summary evaluation for the air blown Lurgi gasifier. The capacity is 1.23 MM scfh, requiring a coal input of 8.58 tons per hour.



RAW GAS

	Mols %	Mols	Lb
CO_2	13.45	3.00	132.0
H_2	20.62	4.59	9.2
CO	12.92	2.88	80.6
CH_4	3.24	0.72	11.5
N_2	41.25	9.20	258.0
NH_3	0.45	0.10	1.7
H_2S	0.45	0.10	3.4
H_2O	7.62	1.72	31.0
Total	100.00	22.31	527.4
Tar			6.1
			533.5

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Figure 5.6-1 Steam Temperatures and Steam Compositions for Gas Turbine Fuel from Airblown Lurgi Gas Producer at 150 psi

TABLE 5.6-2 MATERIAL BALANCE FOR 150 PSIG AIR BLOWN LURGI GAS PRODUCER

(Basis 100 lb daf Coal)

Input	lb.	Output	lb.
Coal (daf)	100.0	Raw Gas	527.4
Water	1.3	Tar	6.1
Ash	7.7		
		Ash	7.7
Air	337.9		
Steam	95.4	Unaccounted for	1.1
Total	542.3	Total	542.3

TABLE 5.6-3 HEAT BALANCE FOR 150 PSIG AIR BLOWN 150 PSIG PRODUCER

(Basis 100 lb daf Coal)

Input, Btu		Output, Btu	
Coal, Gross Heating Value	1,527,000	Wet, Raw Gas, Gross Heating Value	1,230,000
Coal, Sensible	2,900	Tar, Gross Heating Value	102,500
Steam, Sensible and Latent	114,200	Wet, Raw Gas, Sensible	175,800
Air, Sensible	13.700	Wet, Raw Gas, Latent	34,700
		Ash	3,600
		Heat Loss	50,000
		Unaccounted for	61,200
Total	1,657,800	Total	1,657,800

TABLE 5.6-4 ANALYSIS AND HEATING VALUE OF DRY RAW GAS FROM 150 PSIG AIR BLOWN LURGI GAS PRODUCER

(Tar- and Moisture-Free)

Gas Composition	Volume, Percent
Carbon Dioxide	14.6
Hydrogen	22.3
Carbon Monoxide	14.0
Methane	3.5
Nitrogen	44.6
Ammonia	0.5
Hydrogen Sulfide	0.5
Total	100.0

Raw Gas Gross Heating Value,
Dry, Tar-free 158 Btu/scf

TABLE 5.6-5 EFFICIENCY VALUES FOR 150 PSIG AIR BLOWN LURGI GAS PRODUCER

1. Efficiency, Cold:

$$\frac{\text{Gross Heating Value Dry Raw Gas}}{\text{Gross Heating Value, Coal-fired}} = 80.5 \text{ percent}$$

2. Tar:

$$\frac{\text{Gross Heating Value Tar Produced}}{\text{Gross Heating Value, Coal-fired}} = 6.7 \text{ percent}$$

3. Efficiency, Cold Gas and Tar: = 87.2 percent

4. Efficiency, Hot:

$$\frac{\text{Gross Heating Value Wet Raw Gas + Sensible Heat + Heating Value Tar}}{\text{Total Heat Input to System}} = 91.0 \text{ percent}$$