RESEARCH AND DEVELOPMENT, SYNTHESIS-GAS LABORATORIES AND PILOT PLANTS, MORGANTOWN, W. VA., AND GORGAS, ALA.

Experimental Work on Synthesis-Gas Production

Advance estimates of the costs of producing synthetic liquid fuels from coal emphasize the importance of further experimental development and improvement of the technology of coal gasification. Experimentation in coal gasification has continued on pilot-plant scale at Morgantown, W. Va., and the field-scale experimentation in underground gasification has continued at Gorgas, Ala. During the past year, Congress authorized the construction at Morgantown of improved facilities for process development work in coal gasification.

The work at Gorgas is done in cooperation with the Alabama Power Co.; that at Morgantown in cooperation with West Virginia University. The Southern Natural Gas Co. maintains a resident engineer at Morgantown who does cooperative work. The work on pressure gasification is being done in cooperation with the Babcock & Wilcox Co.

Pulverized Coal Gasification Pilot Plant, Morgantown, W. Va.

During the past year experimental work has continued on the pilot plant described in the Annual Report of the Secretary of the Interior for 1949[1] and more completely in Report of Investigations 4733.[2]

Early in the year, construction of a new dust removal system was completed. Test runs have demonstrated that the gas can be made clean enough for synthesis use at reasonable costs and with standard equipment. The maximum throughput or capacity of the generator was found to be larger than that of available materials handling equipment, so the interior volume of the generator was reduced to obtain better data. The effort for the past few months has been concentrated on developing a practical method for continuous tapping of the slag made. The work now is on determining capacity relationships and data on reactant ratios using steam at 300° F. and 1,200° F.

A Report of Investigations covering the superheating of steam to high temperatures in the Royster pebble heaters is in process of preparation.

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[1] Bureau of Mines, Synthetic Liquid Fuels, Annual Report of the Secretary of the Interior for 1949, Part 1 - Oil from Coal: Rept. of Investigations 4651, 1950, p. 49.

[2] Strimbeck, G. R., Holden, J. H., Rockenbach, L. P., Cordiner, J. B., Jr., and Schmidt, L. D., Pilot Plant Gasification of Pulverized Coal with Oxygen and Highly Superheated Steam: Bureau of Mines Report of Investigations 4733, November 1950, 41 pp.

Tables 16 to 18 present the essential data obtained for the first 31 runs. The very considerable reduction in oxygen requirements obtained by the use of the highly superheated steam is clearly demonstrated. In figure 47 the calculated effect of using steam at 240° F. on the oxygen requirement is shown by the solid line. The experimental data from runs 15 A to 25 B are shown by the circles. Circles above the line indicate gasification percentages less than 90 percent. Figure 48 shows the effect of the steam:coal ratio on the hydrogen:carbon monoxide ratios in the synthesis gas. Figure 49 shows, for the various steam temperatures and ratios of steam to coal, the effect on the carbon dioxide content of the gas. Taken together, they illustrate the flexibility of the process in controlling the hydrogen:carbon monoxide ratios. This flexibility, taken with the fact that the process can use widely different coals, indicates that it is applicable to a wide variety of synthesis. Work to be reported later has shown that by substituting air heated to about 2,500° F. for oxygen, and with superheated steam, a gas suitable for ammonia synthesis can be made.

Figure 50 shows the increase in the steam decomposition per pound of coal with an increase in the steam:coal ratio. It may be seen that the steam decomposition varies linearly, and that the rate of increase corresponds to a steam decomposition of 25 percent for all three ranges of steam temperature. It is probable that the same relationship would not be found if steam temperature and oxygen:coal ratio were kept constant. The value for steam decomposition found agrees with that obtained by a heat balance. Thus, in the high-temperature steam runs, 3 pounds of steam is used to supply heat for the reaction per pound of the steam taking part in the reaction.

If it is assumed that only three reactions take place in the generator, namely, the reaction of carbon with oxygen to form carbon monoxide, that of carbon with steam to form carbon monoxide and hydrogen, and the water gas shift reaction, then it is possible to estimate the amount of steam actually reacting with carbon - that is, separately from that taking part in the shift. Thus, the mols of steam decomposed by the shift are equal to the mols of carbon dioxide formed. With those assumptions, the mols of hydrogen in the gas minus the mols of carbon dioxide are equal to the mols of steam reacting with carbon of the coal to produce carbon monoxide and hydrogen. Thus, it is possible, by this rather over-simplified method of analysis, to estimate the fraction of the steam decomposed that actually produces carbon monoxide and hydrogen. This fraction of steam decomposed, averaged for the high-temperature steam runs, is approximately 0.4. On the above basis, for each pound of coal gasified the pounds of steam reacting with carbon to produce synthesis gas are about 0.3 pound greater for the high-temperature steam runs than for the low-temperature steam runs.

Operation of the Royster stoves in the early runs had shown that it would be advisable to move the pebble beds periodically. Consequently, the stoves were rebuilt as shown in figure 51. The bed could be dropped a few inches by using the pebble outlet unit, and new pebbles could be added through the side opening. While these changes were being made in the stoves, a temporary steam superheater, shown in figures 52 and 53, was used to prepare steam at about 2,000° F. for runs 26A through 31B.

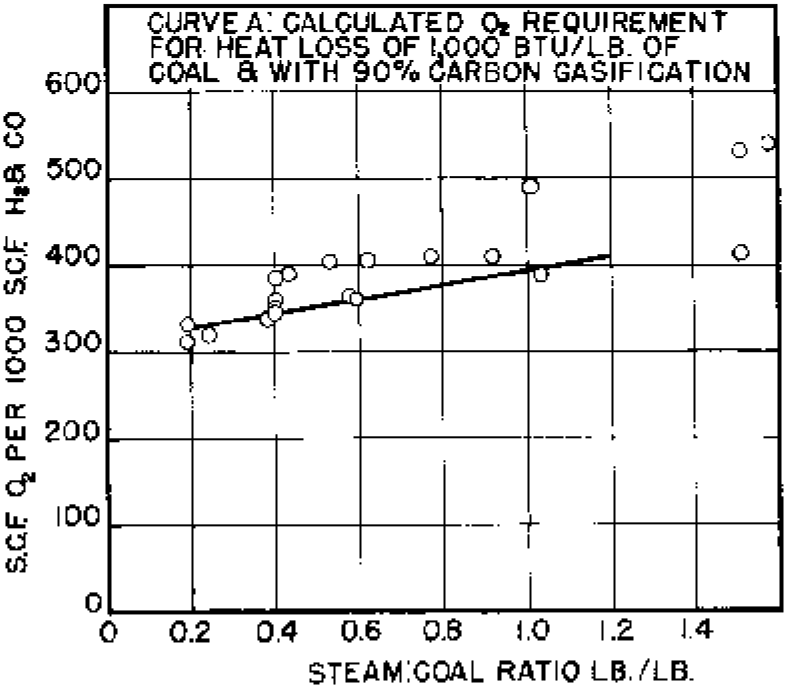

Figure 47. - Effect of steam:coal ratio on oxygen requirement for low-temperature steam runs.

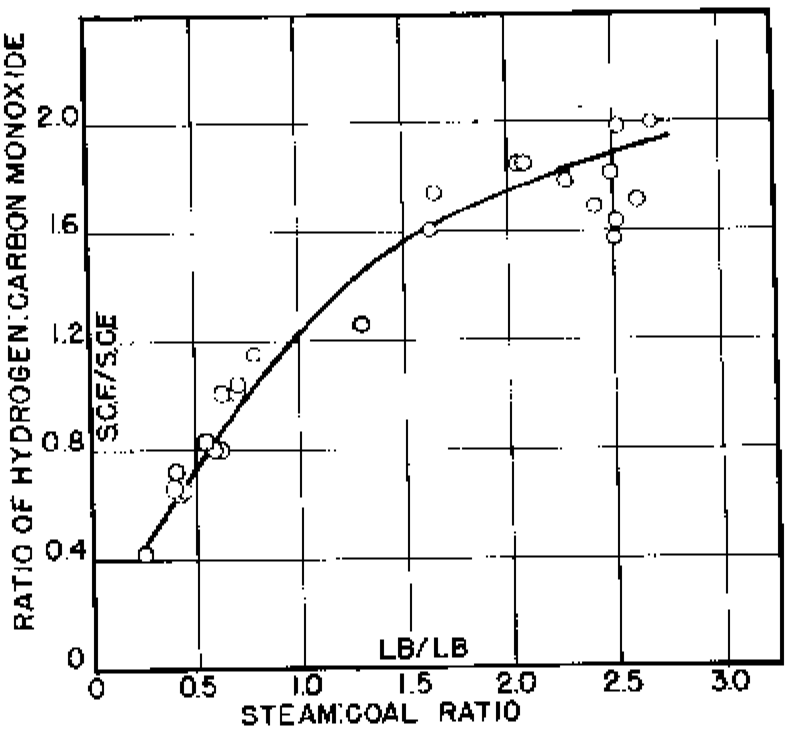

Figure 48. - Effect of steam:coal ratio on ratio of hydrogen to carbon monoxide.

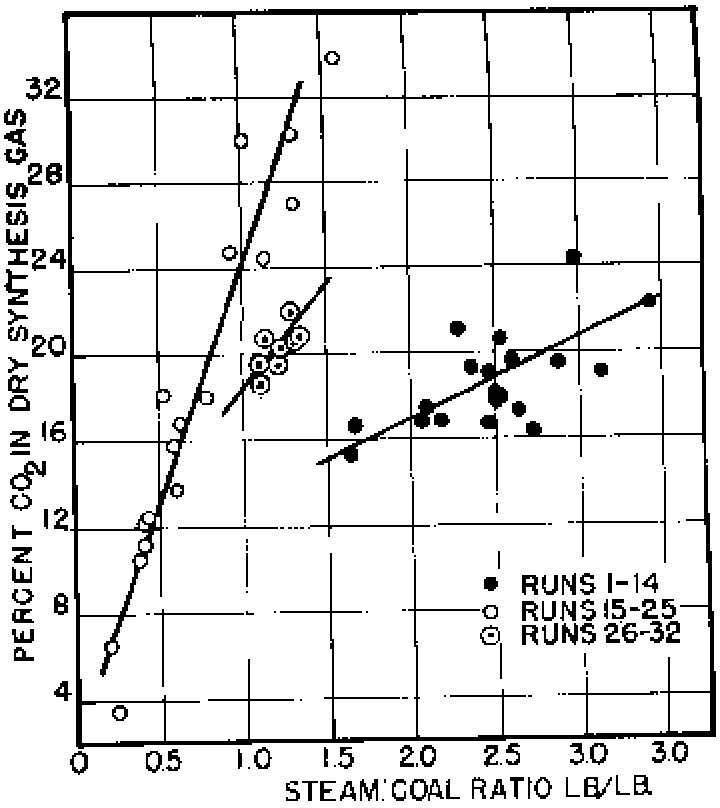

Figure 49. - Effect of steam:coal ratio on percent carbon dioxide in dry synthesis gas.

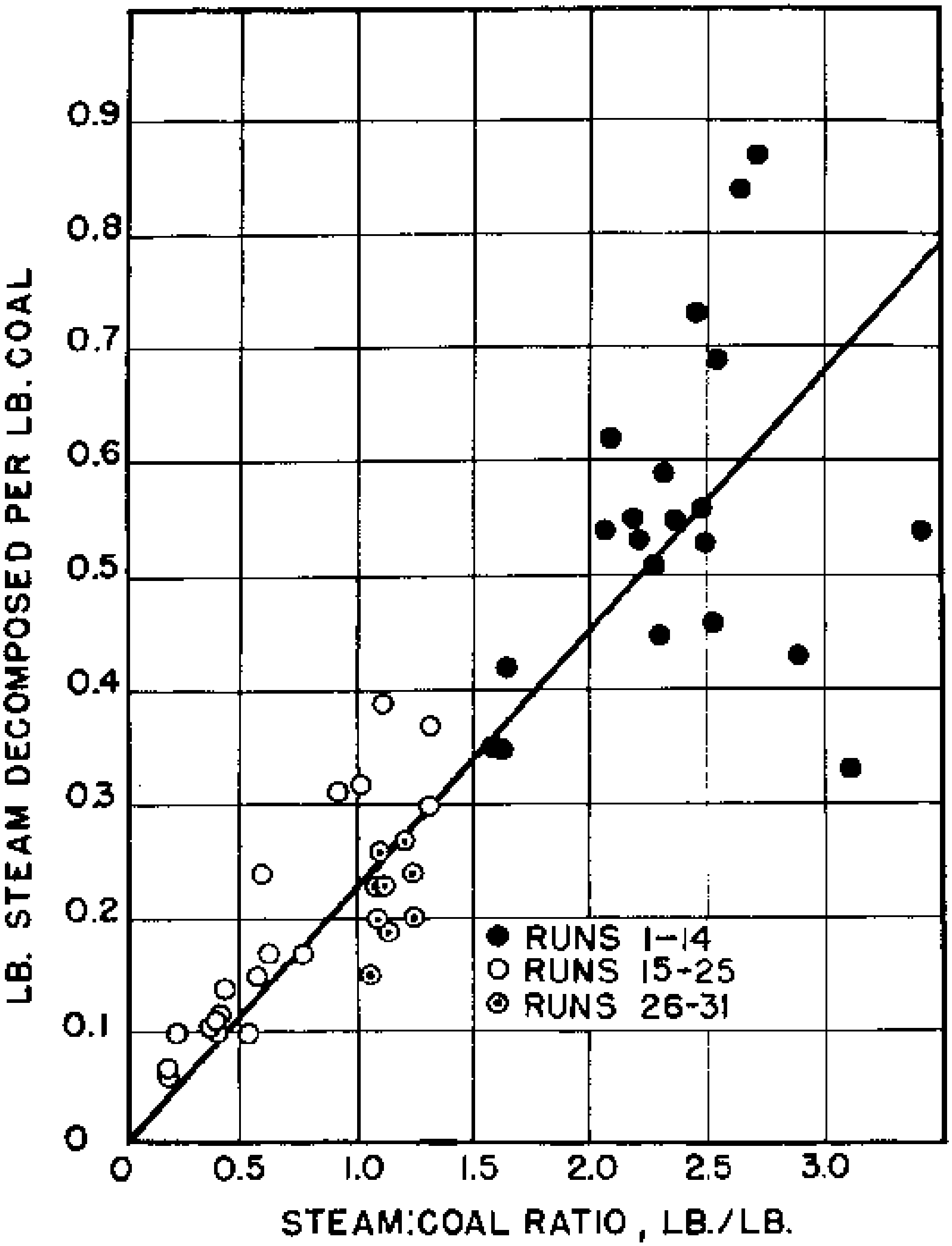

Figure 50. - Effect of steam:coal ratio on steam decomposition.

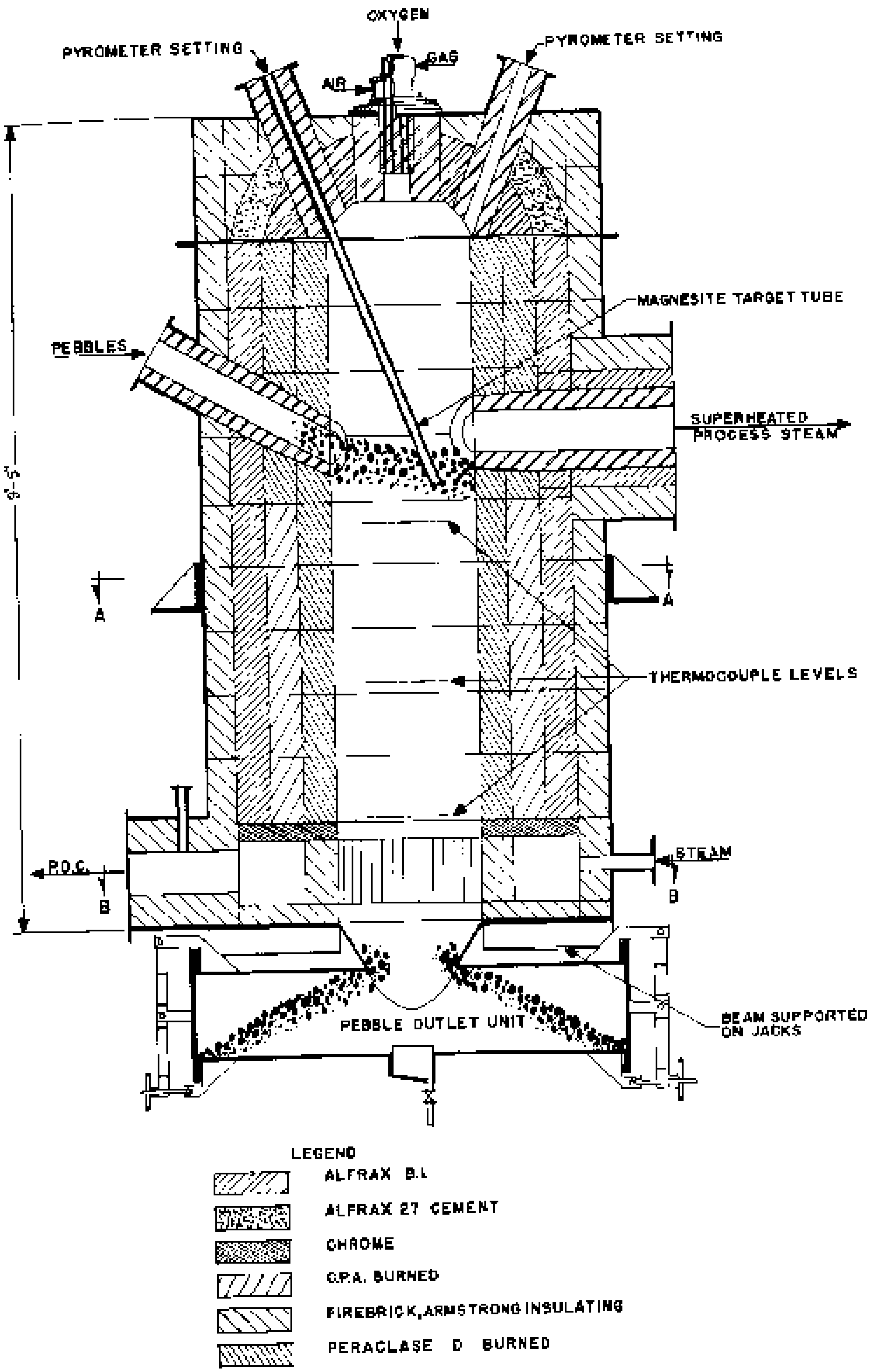

Figure 51. - Royster stove, design 2.

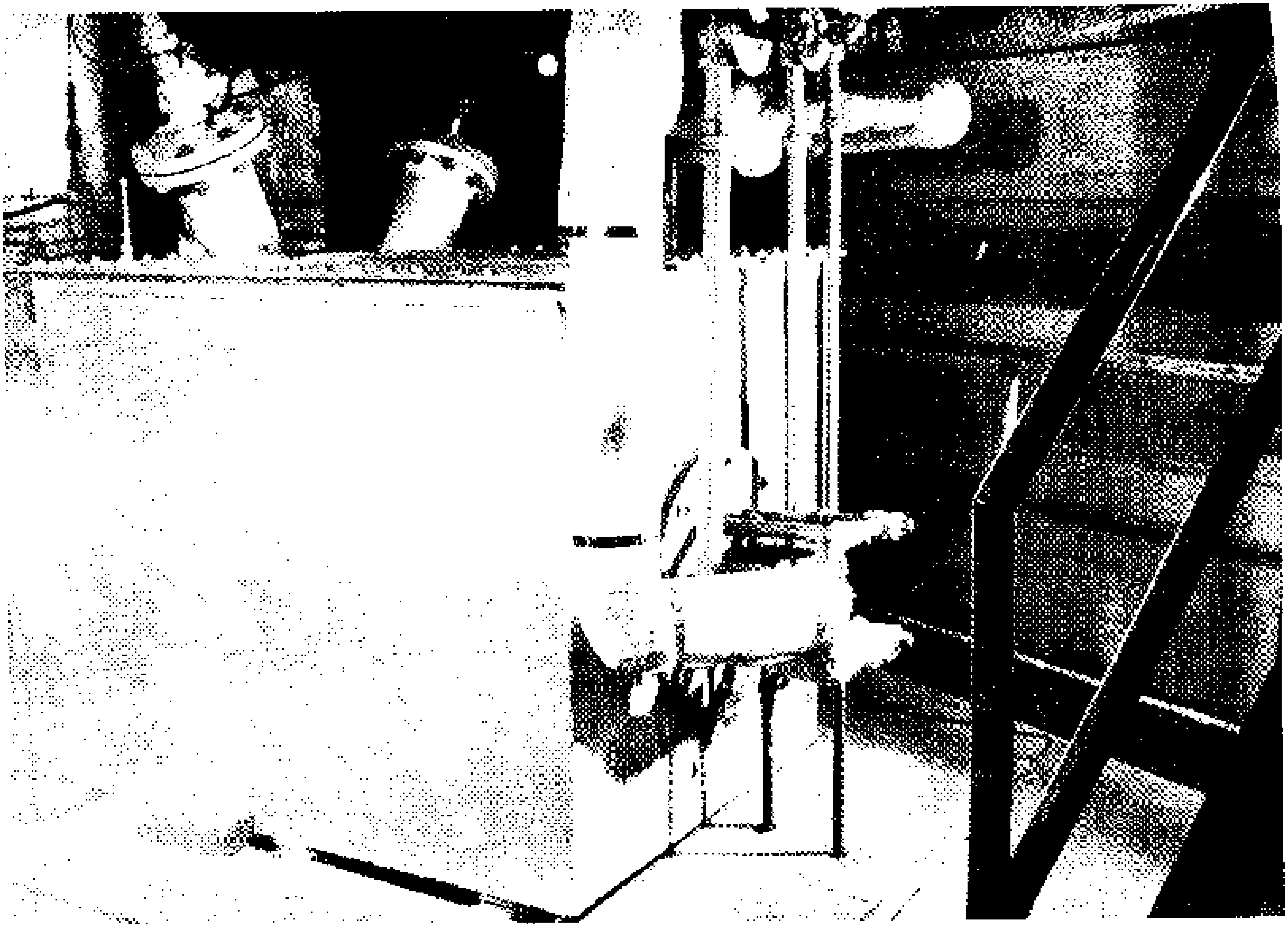

Figure 52. - View of steam superheater used in runs 26-31.

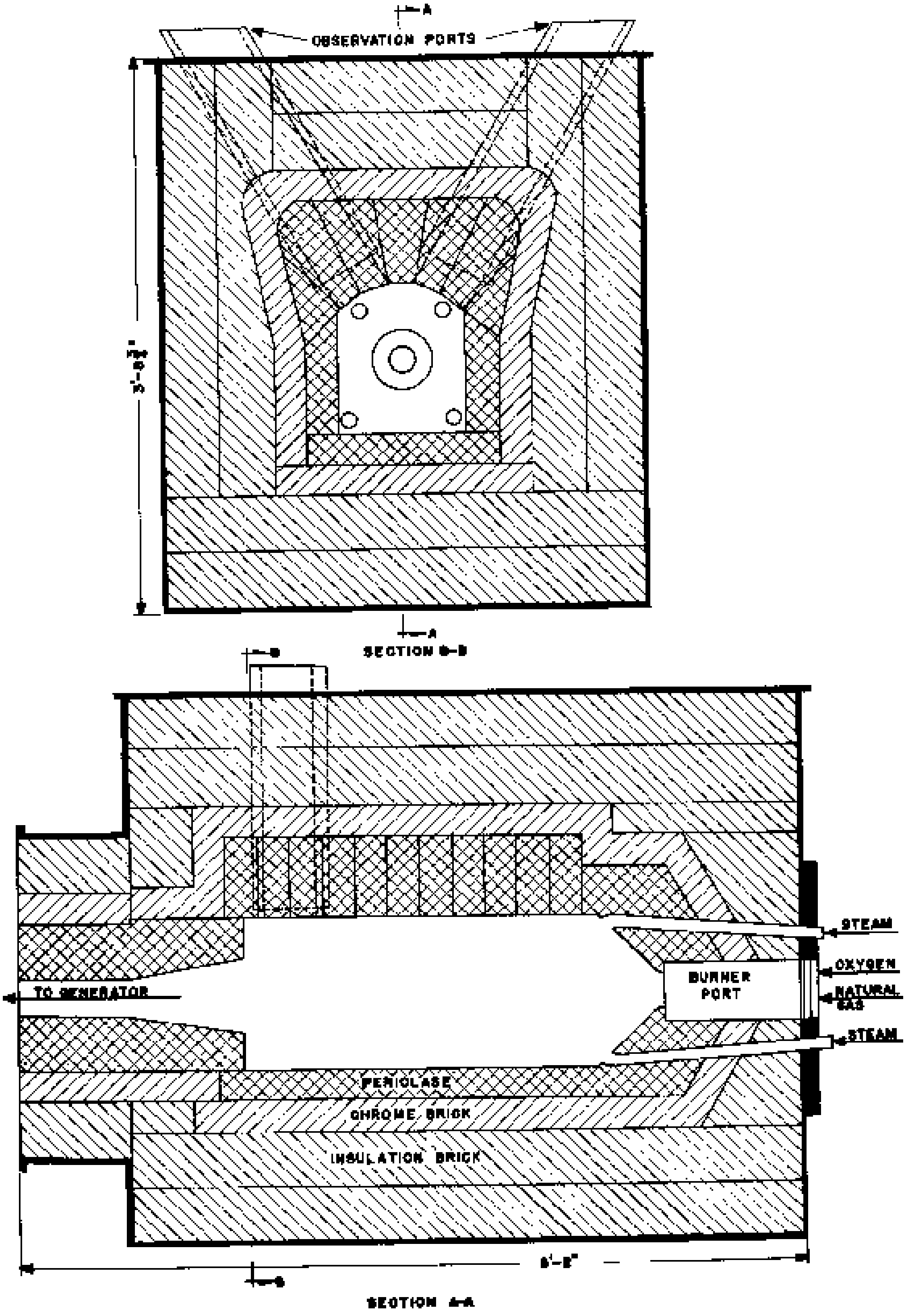

Figure 53. - Steam superheater in cross section.

TABLE 16. - **Principal results on pilot-plant runs, using high-temperature steam (runs 1-14)**

Run No.	Raw-coal rate, lb./hr.	Process-steam rate, lb./hr.	Process oxygen rate, S.C.F./hr.	Steam-inlet temp., °F.	Steam/lb. raw coal, lb.	Oxygen/lb. raw coal, S.C.F.	Synthesis-gas analysis, percent					H_2:CO ratio	Synthesis gas/lb. raw coal, S.C.F.	CO + H_2 made/lb. raw coal, S.C.F.	Total carbon gasified, percent[1]	Steam decomposed, percent
							CO_2	H_2	CO	CH_4 + Ill.	N_2					
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
1	211.5	483	1,180	2,960	2.26	5.58	21.1	51.4	28.8	2.6	-4.3	1.78	35.9	28.8	83.9	19.5
2A	200.0	593	1,116	3,043	2.97	5.58	24.4	46.2	21.1	2.7	4.3	2.19	41.7	28.1	89.5	18.7
2B	257.7	609	1,469	3,022	2.36	5.70	19.3	49.2	27.3	3.0	- .6	1.80	40.5	31.0	89.6	23.2
3	293.7	737	1,634	2,816	2.51	5.56	17.7	43.8	27.9	2.3	6.6	1.57	39.4	28.2	83.9	18.4
4A	309.0	674	1,664	2,772	2.18	5.39	16.8	44.6	26.9	2.0	7.5	1.66	42.0	30.1	85.6	25.4
4B	314.0	820	1,692	2,385	2.61	5.39	19.6	46.6	25.5	2.5	5.4	1.83	40.1	28.9	85.6	22.4
4C	354.0	866	1,887	2,850	2.45	5.33	16.7	43.5	27.4	2.1	7.7	1.59	41.6	29.5	85.6	21.7
5	299.0	752	1,785	2,889	2.52	5.97	20.6	46.3	28.3	1.6	1.6	1.64	39.6	29.5	88.0	20.2
6A	186.0	581	1,170	2,630	3.12	6.29	19.1	49.1	25.8	1.5	2.9	1.84	45.2	34.4	94.4	10.5
6B	232.0	667	1,460	2,638	2.88	6.29	19.5	47.6	27.5	1.1	3.8	1.74	43.5	32.8	91.4	14.9
7	235.0	815	1,740	2,783	3.42	7.31	22.2	55.9	27.1	2.0	-8.3	2.06	41.0	34.0	93.5	15.7
8	200.4	498	1,220	3,195	2.49	6.09	18.1	50.3	27.6	2.3	- .1	1.82	40.7	31.7	86.8	21.4
9A	207.0	507	1,066	3,096	2.45	5.25	19.0	41.2	26.3	2.5	11.6	1.57	48.9	33.0	95.1	30.0
9B	193.3	506	954	3,043	2.63	4.94	17.3	43.0	25.9	2.5	10.3	1.66	51.8	35.7	96.0	32.1
9C	187.0	507	869	3,050	2.71	4.65	16.4	48.7	25.6	2.1	6.7	1.90	51.7	30.4	90.4	32.1
10	198.5	502	850	3,058	2.53	4.28	17.8	51.2	25.5	2.6	2.1	1.98	41.2	31.7	79.0	27.1
11	245.0	516	853	3,055	2.08	3.48	17.4	52.7	28.5	3.1	-2.0	1.85	36.0	29.3	77.7	29.8
12	247.0	507	853	3,303	2.05	3.45	16.9	49.0	26.5	3.0	3.9	1.85	37.2	26.6	76.5	26.2
13	311.6	505	1,150	2,777	1.62	3.69	15.2	49.1	30.5	2.9	1.8	1.61	28.7	22.9	63.2	21.6
14	305.4	500	1,153	2,706	1.64	3.78	16.5	53.1	30.6	3.2	-3.9	1.74	29.5	24.7	64.6	25.4
Ave.	249.5	607	1,288	2,904	2.48	5.20	18.6	48.2	27.1	2.3	2.8	1.78	40.7	30.5	85.0	22.8

[1] Carbon gasified is average between values based on measurement of residue and measurement of gas volume.

TABLE 17. - Principal results on pilot-plant runs, using low-temperature steam (runs 15A-25B)

Run No.	Raw-coal rate, lb./hr.	Process-steam rate, lb./hr.	Process-oxygen rate, S.C.F./hr.	Steam-inlet temp., °F.	Steam/lb. raw coal, lb.	Oxygen/lb. raw coal, S.C.F.	Synthesis-gas analysis, percent					H_2:CO ratio	Synthesis gas/lb. raw coal, S.C.F.	CO + H_2 made/lb. raw coal, S.C.F.	Total carbon gasified, percent	Steam decomposed, percent
							CO_2	H_2	CO	CH_4 + Ill.	N_2					
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
15A	242.3	384	2,792	245	1.58	11.5	33.7	33.3	28.3	2.5	1.4	1.18	34.7	21.4	91.5	21.9
15B	242.3	318	2,723	245	1.31	11.2	30.2	35.8	26.6	2.7	3.2	1.35	33.9	21.2	86.6	22.6
15C	242.3	245	2,644	245	1.01	10.9	29.9	36.3	29.2	2.1	1.6	1.24	34.1	22.3	88.0	31.4
16A	264.0	243	2,654	235	.92	10.1	24.6	36.3	35.6	1.8	.7	1.02	34.1	24.5	91.1	34.2
16B	264.0	297	2,659	237	1.13	10.1	24.4	37.4	35.1	2.0	.5	1.07	36.4	25.4	93.4	34.7
16C	264.0	345	2,659	240	1.31	10.1	27.0	36.4	31.8	1.2	3.1	1.15	36.0	24.4	91.6	28.1
17	245.8	146	2,550	233	.59	10.4	13.7	36.4	45.8	1.8	1.5	.80	35.2	28.9	90.0	40.7
18	230.4	98	2,370	234	.43	10.3	12.5	32.3	50.6	1.9	1.9	.64	31.8	25.4	89.0	33.0
19	245.7	60	2,171	230	.24	8.8	3.4	27.4	64.7	1.9	1.9	.42	29.8	27.5	86.9	42.0
20	477.5	298	4,830	241	.62	10.1	16.8	35.1	43.2	2.0	1.7	.80	31.5	24.8	87.0	27.5
21	406.9	315	3,965	242	.77	9.6	18.0	37.8	40.9	1.9	.9	.93	30.4	23.9	86.4	22.5
22A	487.4	260	4,061	235	.53	8.3	18.1	36.2	43.9	1.2	.3	.82	25.7	20.6	77.2	18.5
22B	487.4	278	4,069	237	.57	8.4	15.8	37.0	44.5	1.7	.7	.83	27.9	22.8	82.7	25.9
23A	458.0	182	3,725	235	.40	8.1	12.2	36.3	48.9	1.5	.8	.74	27.0	23.0	77.9	26.8
23B	458.0	172	3,526	235	.38	7.7	10.5	35.1	50.6	2.2	1.1	.69	26.6	22.8	78.3	29.2
24A	477.0	91	3,322	235	.19	7.0	6.5	31.3	54.4	2.9	.8	.63	23.8	21.1	76.4	28.6
24B	477.0	90	3,193	235	.19	6.7	6.5	35.9	54.1	2.5	.7	.66	23.9	21.5	72.3	34.8
25A	446.0	178	3,785	245	.40	8.5	12.4	35.5	50.6	.8	.3	.70	28.1	24.2	82.2	30.2
25B	446.0	178	4,420	245	.40	9.9	11.2	32.5	51.6	1.5	2.1	.63	30.4	25.7	90.1	25.7
Ave.	361.2	220	3,270	238	.68	9.4	17.2	35.1	43.7	1.9	1.3	.86	30.6	23.9	85.2	29.4

TABLE 17. - Principal results on pilot-plant runs, using low-temperature steam (runs 15A-25B)

Run No.	Raw-coal rate, lb./hr.	Process-steam rate, lb./hr.	Process-oxygen rate, S.C.F./hr.	Steam-inlet temp., °F.	Steam/lb. raw coal, lb.	Oxygen/lb. raw coal, S.C.F.	Synthesis-gas analysis, percent					H_2:CO ratio	Synthesis gas/lb. raw coal, S.C.F.	CO + H_2 made/lb. raw coal, S.C.F.	Total carbon gasified, percent	Steam decomposed, percent
							CO_2	H_2	CO	CH_4 + Ill.	N_2					
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
15A	242.3	384	2,792	245	1.58	11.5	33.7	33.3	28.3	2.5	1.4	1.18	34.7	21.4	91.5	21.9
15B	242.3	318	2,723	245	1.31	11.2	30.2	35.8	26.6	2.7	3.5	1.35	33.9	21.2	86.6	22.6
15C	242.3	245	2,644	245	1.01	10.9	29.9	36.3	29.2	2.1	1.6	1.24	34.1	22.3	88.0	31.4
16A	264.0	243	2,654	235	.92	10.1	24.6	36.3	35.6	1.8	.7	1.02	34.1	24.5	91.1	34.2
16B	264.0	297	2,659	237	1.13	10.1	24.4	37.4	35.1	2.0	.5	1.07	36.4	26.4	93.4	34.7
16C	264.0	345	2,659	240	1.31	10.1	27.0	36.4	31.8	1.2	3.1	1.15	36.0	24.4	91.6	28.1
17	245.8	146	2,550	233	.59	10.4	13.7	36.4	45.8	1.8	1.5	.80	35.2	28.9	90.0	40.1
18	230.4	98	2,370	234	.43	10.3	12.5	32.3	50.6	1.9	1.9	.64	31.8	26.4	89.0	33.0
19	245.7	60	2,171	230	.24	8.8	3.4	27.4	64.7	1.9	1.9	.42	29.8	27.5	86.9	42.0
20	477.5	298	4,830	241	.62	10.1	16.8	35.1	43.8	2.0	1.7	.80	31.5	24.8	97.0	27.5
21	406.9	315	3,968	242	.77	9.8	18.0	37.8	40.9	1.9	.9	.93	30.4	23.9	86.4	22.5
22A	487.4	260	4,061	235	.53	8.3	18.1	36.2	43.9	1.2	.3	.82	25.7	20.6	77.2	18.5
22B	487.4	278	4,069	237	.57	8.4	15.8	37.0	44.5	1.7	.7	.83	27.9	22.8	82.7	25.9
23A	458.0	182	3,725	235	.40	8.1	12.2	36.3	48.9	1.5	.8	.74	27.0	23.0	77.9	26.8
23B	458.0	172	3,526	235	.38	7.7	10.5	35.1	50.6	2.2	1.1	.69	26.6	22.8	78.3	29.2
24A	477.0	91	3,322	235	.19	7.0	6.5	34.3	54.4	2.9	.8	.63	23.8	21.1	76.4	28.6
24B	477.0	90	3,193	235	.19	6.7	6.5	35.9	54.1	2.5	.7	.66	23.9	21.5	72.3	34.8
25A	446.0	178	3,785	245	.40	8.5	12.4	35.5	50.6	.8	.3	.70	28.1	24.2	82.2	30.2
25B	446.0	178	4,420	245	.40	9.9	11.2	32.5	51.8	1.5	2.1	.63	30.4	25.7	90.1	25.7
Ave.	361.2	220	3,270	238	.68	9.4	17.2	35.1	43.7	1.9	1.3	.66	30.6	23.9	85.2	29.4

TABLE 16. - Principal results on pilot-plant runs, using high-temperature steam (runs 1-14)

Run No.	Raw-coal rate, lb./hr.	Process-steam rate, lb./hr.	Process oxygen rate, S.C.F./hr.	Steam-inlet temp., °F.	Steam/lb. raw coal, lb.	Oxygen/lb. raw coal, S.C.F.	Synthesis-gas analysis, percent					H_2:CO ratio	Synthesis gas/lb. raw coal, S.C.F.	CO + H_2 made/lb. raw coal, S.C.F.	Total carbon gasified, percent 1/	Steam decomposed, percent
							CO_2	H_2	CO	CH_4 + Ill.	N_2					
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
1	211.5	483	1,180	2,960	2.28	5.58	21.1	51.4	28.8	2.6	-4.3	1.78	35.9	28.8	83.9	19.5
2A	200.0	593	1,116	3,043	2.97	5.58	24.4	46.2	21.1	2.7	4.3	2.19	41.7	28.1	69.5	18.7
2B	257.7	609	1,469	3,022	2.36	5.70	19.3	49.2	27.3	3.0	- .6	1.80	40.5	31.0	89.6	23.2
3	293.7	737	1,634	2,816	2.51	5.56	17.7	43.8	27.9	2.3	6.6	1.57	39.4	28.2	83.9	18.4
4A	309.0	674	1,664	2,772	2.18	5.39	16.8	44.6	26.9	2.0	7.5	1.66	42.0	30.1	85.6	25.4
4B	314.0	820	1,692	2,385	2.61	5.39	19.6	46.6	25.5	2.5	5.4	1.83	40.1	28.9	85.6	22.4
4C	354.0	866	1,887	2,850	2.45	5.33	16.7	43.5	27.4	2.1	7.7	1.59	41.6	29.5	85.6	21.7
5	299.0	752	1,785	2,889	2.52	5.97	20.6	46.3	28.3	1.6	1.6	1.64	39.6	29.5	88.0	20.2
6A	186.0	581	1,170	2,530	3.12	6.29	19.1	49.4	26.0	1.5	2.9	1.84	45.2	34.4	94.4	10.5
6B	232.0	667	1,460	2,638	2.88	6.29	19.5	47.8	27.5	1.1	3.8	1.74	43.5	32.8	91.4	14.9
7	230.0	815	1,740	2,783	3.42	7.31	22.2	55.9	27.1	2.0	-8.3	2.06	41.0	34.0	93.5	15.7
8	200.4	498	1,220	3,195	2.49	6.09	18.1	50.3	27.6	2.3	- .1	1.82	40.7	31.7	86.8	21.4
9A	207.0	507	1,066	3,096	2.45	5.15	19.0	41.2	26.3	1.5	11.6	1.57	48.9	33.0	95.1	30.0
9B	193.3	508	954	3,043	2.63	4.94	17.3	43.0	25.9	2.5	10.3	1.66	51.8	35.7	96.0	32.1
9C	167.0	507	869	3,050	2.71	4.65	16.4	48.7	25.6	2.1	6.7	1.90	51.7	38.4	90.4	32.1
10	198.5	502	850	3,058	2.53	4.26	17.8	51.2	25.8	2.6	2.1	1.98	41.2	31.7	79.0	27.1
11	245.0	510	853	3,055	2.08	3.48	17.4	52.7	28.5	3.1	-2.0	1.85	36.0	29.3	77.7	29.8
12	247.0	507	853	3,303	2.05	3.45	16.9	49.0	26.5	3.0	3.9	1.85	35.2	26.6	76.5	26.2
13	311.6	505	1,150	2,777	1.62	3.69	15.2	49.1	30.5	2.9	1.8	1.61	28.7	27.9	63.2	21.6
14	305.4	500	1,153	2,706	1.64	3.78	16.5	53.1	30.6	3.2	-3.9	1.74	29.5	24.7	64.6	25.4
Ave.	249.5	607	1,288	2,904	2.48	5.20	18.6	48.2	27.1	2.3	2.8	1.78	40.7	30.5	85.0	22.8

1/ Carbon gasified is average between values based on measurement of residue and measurement of gas volume.

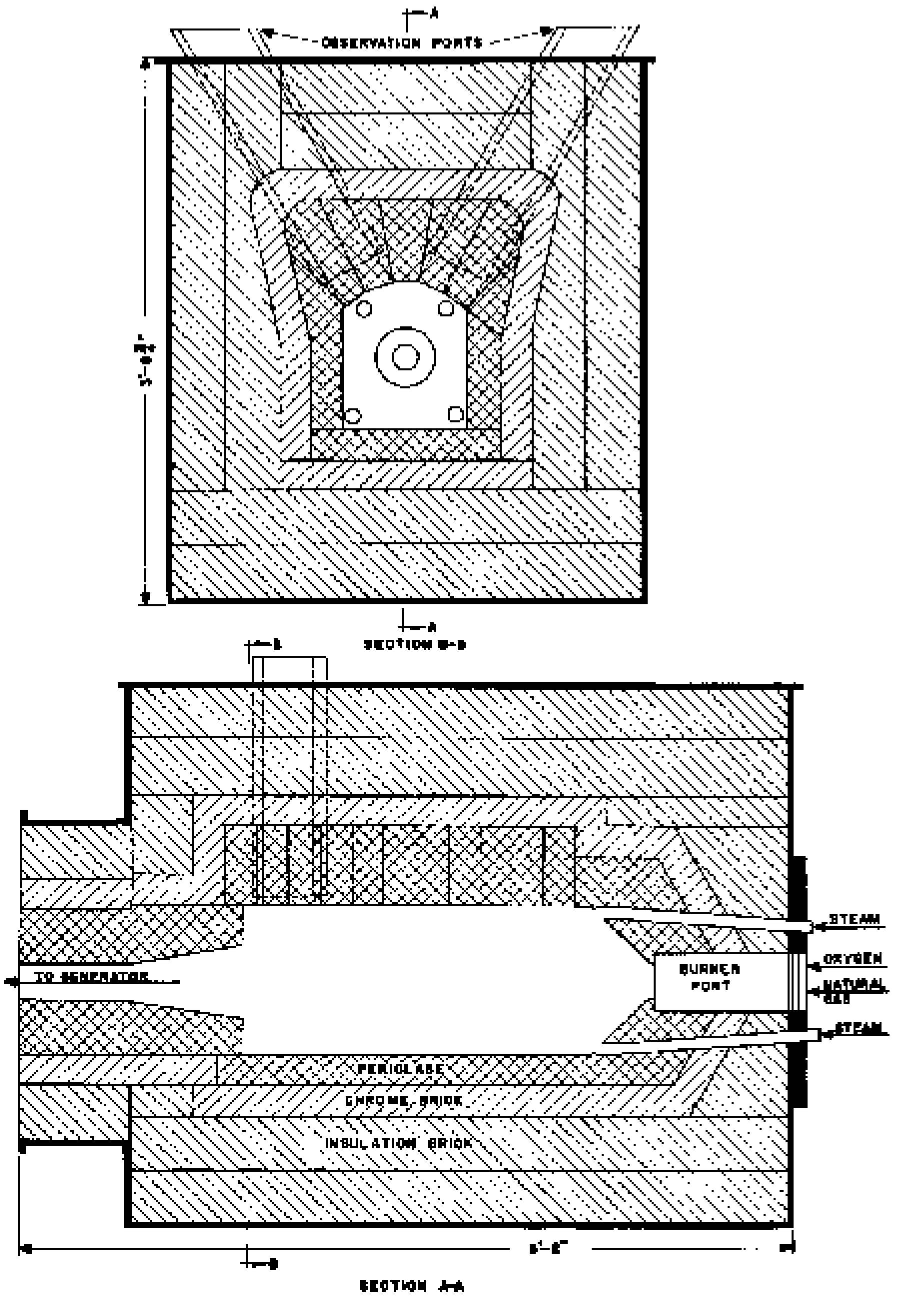

Figure 53. - Steam superheater in cross section.

TABLE 16. - Principal results on pilot-plant runs, using high-temperature steam (runs 1-14)

Run No.	Raw-coal rate, lb./hr.	Process-steam rate, lb./hr.	Process oxygen rate, S.C.F./hr.	Steam-inlet temp., °F.	Steam/lb. raw coal, lb.	Oxygen/lb. raw coal, S.C.F.	Synthesis-gas analysis, percent					H_2:CO ratio	Synthesis gas/lb. raw coal, S.C.F.	CO + H_2 made/lb. raw coal, S.C.F.	Total carbon gasified, percent[1]	Steam decomposed, percent
							CO_2	H_2	CO	CH_4 + Ill.	N_2					
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
1	211.5	483	1,180	2,960	2.28	5.58	21.1	51.4	28.8	2.6	-4.3	1.78	35.9	28.8	83.9	19.5
2A	200.0	593	1,116	3,043	2.97	5.58	24.4	46.2	21.1	2.7	4.3	2.19	41.7	28.1	89.5	18.7
2B	257.7	609	1,469	3,022	2.36	5.70	19.3	49.2	27.3	3.0	- .6	1.80	40.5	31.0	89.6	23.2
3	293.7	737	1,634	2,816	2.51	5.56	17.7	43.8	27.9	2.3	6.6	1.57	39.4	28.2	83.9	18.4
4A	309.0	674	1,664	2,772	2.18	5.39	16.8	44.6	26.9	2.0	7.5	1.66	42.0	30.1	85.6	25.4
4B	314.0	890	1,692	2,385	2.61	5.39	19.6	46.6	25.5	2.5	5.4	1.83	40.1	28.9	85.6	22.4
4C	354.0	866	1,587	2,850	2.45	5.33	16.7	43.5	27.4	2.1	7.7	1.59	41.6	29.5	85.6	21.7
5	299.0	752	1,785	2,889	2.52	5.97	20.6	46.3	28.3	1.6	1.6	1.64	39.6	29.5	88.0	20.2
6A	186.0	581	1,170	2,630	3.12	6.29	19.1	49.4	26.8	1.5	2.9	1.84	45.2	34.4	94.4	10.5
6B	232.0	667	1,460	2,638	2.88	6.29	19.5	47.8	27.5	1.1	3.8	1.74	43.5	32.8	91.4	14.9
7	238.0	815	1,740	2,783	3.42	7.31	22.2	55.9	27.1	2.0	-8.3	2.06	41.0	34.0	93.5	15.7
8	200.4	498	1,220	3,195	2.49	6.09	18.1	50.3	27.6	2.3	- .1	1.82	40.7	31.7	86.8	21.4
9A	207.0	507	1,066	3,096	2.45	5.15	19.0	42.2	26.3	1.5	11.6	1.57	48.9	33.0	95.1	30.0
9B	193.3	508	954	3,043	2.63	4.94	17.3	43.0	25.9	2.5	10.3	1.66	51.8	35.7	96.0	32.1
9C	187.0	507	869	3,050	2.71	4.65	16.4	48.7	25.6	2.1	6.7	1.90	51.7	38.4	90.4	32.1
10	198.5	502	890	3,058	2.53	4.28	17.8	51.2	25.8	2.6	2.1	1.98	41.2	31.7	79.0	27.1
11	215.0	510	853	3,055	2.08	3.48	17.4	52.7	28.5	3.1	-2.0	1.85	36.0	29.3	77.7	29.8
12	217.0	507	853	3,303	2.05	3.45	16.9	49.0	26.5	3.0	3.9	1.85	35.2	25.6	76.5	26.2
13	331.6	505	1,150	2,777	1.62	3.69	15.2	49.1	30.5	2.9	1.8	1.61	28.7	22.9	63.2	21.6
14	305.4	500	1,153	2,706	1.64	3.78	16.5	53.1	30.6	3.2	-3.9	1.74	29.5	24.7	64.6	25.4
Ave.	249.5	607	1,288	2,904	2.48	5.20	18.6	48.2	27.1	2.3	2.8	1.78	40.7	30.5	85.0	22.8

[1] Carbon gasified is average between values based on measurement of residue and measurement of gas volume.

TABLE 17. - Principal results on pilot-plant runs, using low-temperature steam (runs 15A-25B)

Run No.	Raw-coal rate, lb./hr.	Process-steam rate, lb./hr.	Process-oxygen rate, S.C.F./hr.	Steam-inlet temp., °F.	Steam/lb. raw coal, lb.	Oxygen/lb. raw coal, S.C.F.	Synthesis-gas analysis, percent					H_2:CO ratio	Synthesis gas/lb. raw coal, S.C.F.	CO + H_2 made/lb. raw coal, S.C.F.	Total carbon gasified, percent	Steam decomposed, percent
							CO_2	H_2	CO	CH_4 + Ill.	N_2					
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
15A	242.3	384	2,792	245	1.58	11.5	33.7	33.3	28.3	2.5	1.4	1.18	34.7	21.4	91.5	21.9
15B	242.3	318	2,723	245	1.31	11.2	30.2	35.8	26.6	2.7	3.5	1.35	33.9	21.2	86.6	22.6
15C	242.3	245	2,644	245	1.01	10.9	29.9	36.3	29.2	2.1	1.6	1.24	34.1	22.3	88.0	31.4
16A	264.0	243	2,654	235	.92	10.1	24.6	36.3	35.6	1.8	.7	1.02	34.1	24.5	91.1	34.2
16B	264.0	297	2,659	237	1.13	10.1	24.4	37.4	35.1	2.0	.5	1.07	36.4	26.4	93.4	34.7
16C	264.0	345	2,659	240	1.31	10.1	27.0	36.4	31.8	1.2	3.1	1.15	35.0	24.4	91.6	20.1
17	245.8	146	2,550	233	.59	10.4	13.7	36.4	45.8	1.8	1.5	.80	35.2	28.9	90.0	40.1
18	230.4	98	2,370	234	.43	10.3	12.5	32.3	50.6	1.5	1.9	.64	31.8	26.4	89.0	33.0
19	245.7	60	2,171	230	.24	8.8	3.4	27.4	64.7	1.9	1.9	.42	29.8	27.5	86.9	42.0
20	477.5	298	4,830	241	.62	10.1	16.8	35.1	43.8	2.0	1.7	.80	31.5	24.8	87.0	27.5
21	406.9	315	3,968	242	.77	9.8	18.0	37.8	40.9	1.9	.9	.93	30.4	23.9	86.4	22.5
22A	487.4	260	4,061	235	.53	8.3	18.1	36.2	43.9	1.2	.3	.82	25.7	20.6	77.2	18.5
22B	487.4	278	4,069	237	.57	8.4	15.8	37.0	44.5	1.7	.7	.83	27.9	22.8	82.7	25.9
23A	458.0	182	3,725	235	.40	8.1	12.2	36.3	48.9	1.5	.8	.74	27.0	23.0	77.9	26.8
23B	458.0	172	3,526	235	.38	7.7	10.5	35.1	50.6	2.2	1.1	.69	26.6	22.8	78.3	29.2
24A	477.0	91	3,322	235	.19	7.0	6.5	34.3	54.4	2.9	.8	.63	23.8	21.1	76.4	28.6
24B	477.0	90	3,193	235	.19	6.7	6.5	35.9	54.1	2.5	.7	.66	23.9	21.5	72.3	34.8
25A	446.0	178	3,785	245	.40	8.5	12.4	35.5	50.6	.8	.3	.70	28.1	21.2	82.2	30.2
25B	446.0	178	4,420	245	.40	9.9	11.2	32.5	51.6	1.5	2.1	.63	30.4	25.7	90.1	25.7
Ave.	361.2	220	3,270	238	.68	9.4	17.2	35.1	43.7	1.9	1.3	.56	30.6	23.9	85.2	29.4

TABLE 18. - Principal results on pilot-plant runs, using intermediate-temperature steam (runs 26A-31B)

Run No.	Raw-coal rate, lb./hr.	Process-steam rate, lb./hr.	Process-oxygen rate, S.C.F./hr.	Steam-inlet temp., °F.	Steam/lb. raw coal, lb.	Oxygen/lb. raw coal, S.C.F.	Synthesis-gas analysis, percent					H_2:CO ratio	Synthesis gas/lb. raw coal, S.C.F.	CO + H_2 made/lb. raw coal, S.C.F.	Total carbon gasified, percent	Steam decomposed, percent
							CO_2	H_2	CO	CH_4 + Ill.	N_2					
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
26A	399	493	3,605	1,670	1.24	9.04	20.6	40.6	35.1	1.4	1.9	1.16	33.6	25.4	78.0	16.3
26B	399	489	3,657	1,655	1.23	9.17	22.0	39.7	35.2	1.1	1.7	1.13	34.6	25.7	84.5	19.4
27	417	500	3,520	2,024	1.20	8.41	19.5	39.7	38.1	1.0	1.3	1.04	35.5	27.6	84.6	22.3
28 1/	428	484	3,571	1,600	1.13	8.31	20.8	38.4	38.2	1.2	1.0	1.01	33.5	25.7	90.6	16.8
29 1/	443	464	3,310	2,015	1.05	7.47	20.2	38.2	37.3	.9	2.7	1.02	31.6	23.8	85.9	14.2
30A 1/	421	470	3,775	2,086	1.12	8.97	20.7	38.0	39.1	1.0	.9	.97	36.2	27.9	95.9	20.2
30B 1/	421	464	3,761	2,042	1.10	8.93	18.6	39.1	40.4	.9	.7	.97	37.9	30.1	97.6	23.4
31A 1/	425	461	3,967	2,002	1.08	9.33	19.9	37.7	39.5	1.1	1.5	.95	36.1	27.9	95.3	18.7
31B 1/	425	468	3,882	1,997	1.10	9.13	19.6	37.1	39.9	1.0	2.1	.93	37.0	28.5	96.5	20.4
Ave.	420	477	3,672	1,899	1.14	8.75	20.2	38.7	38.1	1.1	1.5	1.02	35.1	27.0	89.9	19.1

1/ Wyoming Rock Springs coal.

The first attempts to tap slag continuously were not successful, because the slag tube was too long to be kept hot throughout its length. The generator base construction was modified as shown in figure 54. A chamber to hold about 6 to 7 cubic feet of slag was built into the side of the generator. A slag tube of Carbofrax refractory was placed to drain the slag from the generator floor into this space. A vent or gas outlet was arranged so that a small portion of the hot gases could be taken out with the slag to keep the slag tube and adjacent wall warm.

Provision for preheating the slag space was made by inserting a burner through the slag-chamber door. The coal and oxygen inlet enters the generator at an angle of 15 degrees upward from horizontal, the coal being fed with the pneumatic feeder described later in this report. The main generator "warm-up" burner is inserted in the tube as indicated. To obtain additional heat release near the generator base and entrance to the slag tube, an auxiliary oxygen nozzle was inserted as shown. Part of the oxygen for the reaction is introduced through this tube and the balance with the coal. Using this arrangement, runs have been made with steam at 300° F. Coal rates up to 700 pounds per hour have been reached with the percent carbon gasified about 90 percent. The gas made has contained about 11 to 15 percent carbon dioxide.

The indications are that the Carbofrax lining in the generator will give a reasonable service life.

The work to date has demonstrated that suitable refractories for heating steam to 3,500° F. are available; that the slag formed in the operation can be tapped continuously; and that a satisfactory dust-removal system can be made from standard equipment.

The study of aqueous triethanolamine solutions for the selective removal of hydrogen sulfide in the presence of carbon dioxide is being continued. Runs were made at 300 p.s.i.g., with inert gas containing about 14 percent carbon dioxide. Hydrogen sulfide in concentrations of 300 to 400 grains per 100 cubic feet was added to the gas.

Experimental results thus far indicate that a 50-percent solution is no more efficient than a 30-percent, and, owing to foaming difficulties, the maximum throughput is decreased. Employing a 30-percent solution and a liquid:gas ratio (gallon of circulating solution per 1,000 SCF gas) of 17.4, the hydrogen sulfide was reduced from 300 grains per 100 cubic feet to 20 grains, and the carbon dioxide from 11 mol percent to 7 mol percent. Steam consumption for reactification averaged about 1.3 lb. per gallon of circulating solution.

The use of 2-ethyl hexanol has been effective in reducing foaming of amine solutions. However, it was found that an appreciable quantity of the antifoam agent was stripped from the solution in the actifier, and that frequent additions were necessary. Other antifoam agents with lower vapor pressures are being investigated.

Substantially complete removal of both hydrogen sulfide and carbon dioxide from synthesis gas with diethanolamine solutions has been confirmed. Employing 30-percent diethanolamine and a liquid:gas ratio of 48.7, the hydrogen sulfide concentration was reduced from 350 grains per 100 cubic feet to 1 grain, and the carbon dioxide from 26 mol percent to 2 percent. Steam consumption averaged 1.1 pound per gallon of circulating solution.

Preliminary data indicate that diethanolamine solutions remove appreciable amounts of organic sulfur from the raw synthesis gas. At least a portion of this organic sulfur is not released during the regeneration of the fouled solution.

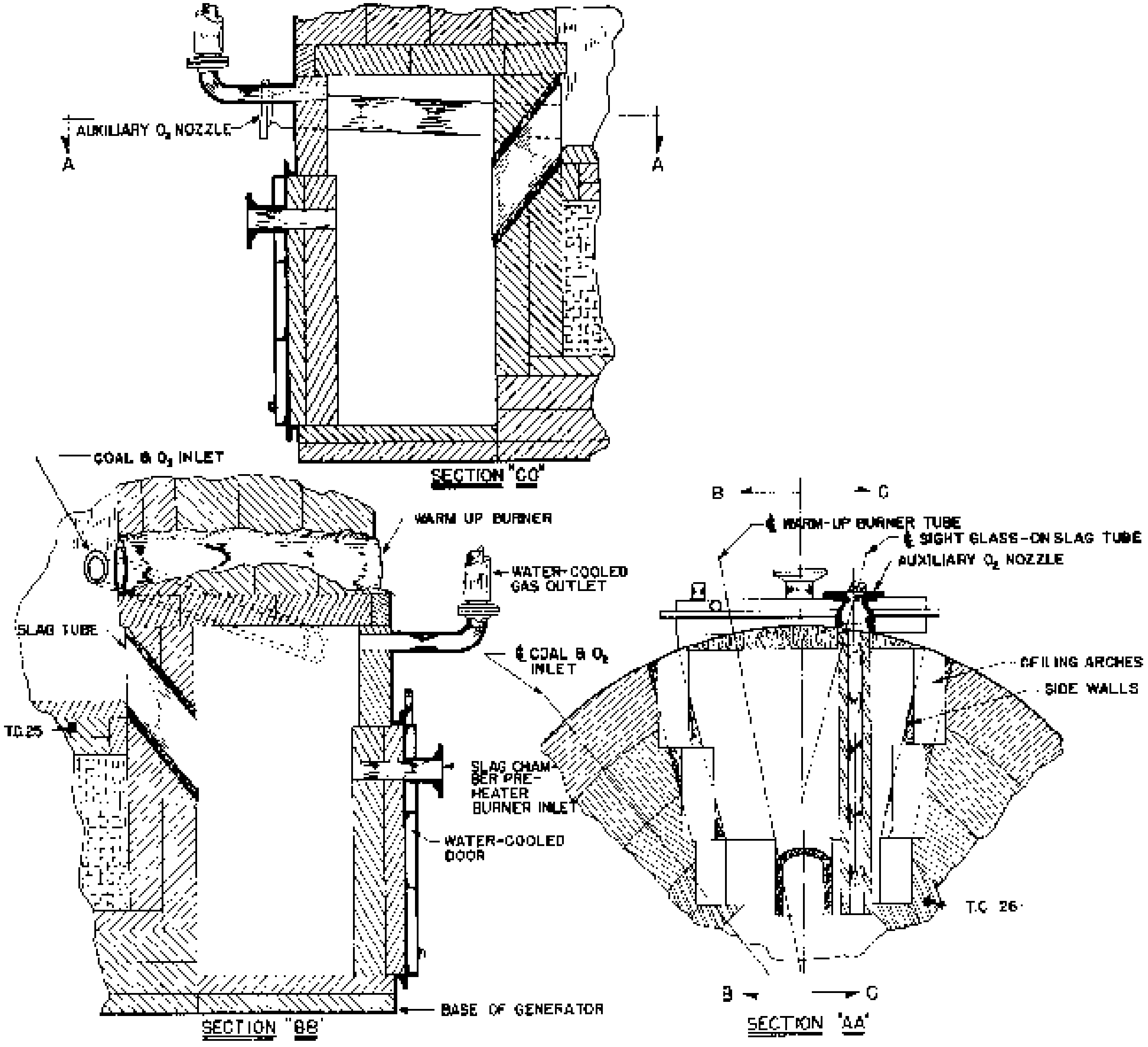

Figure 54. - Slag tap for atmospheric gasifier.

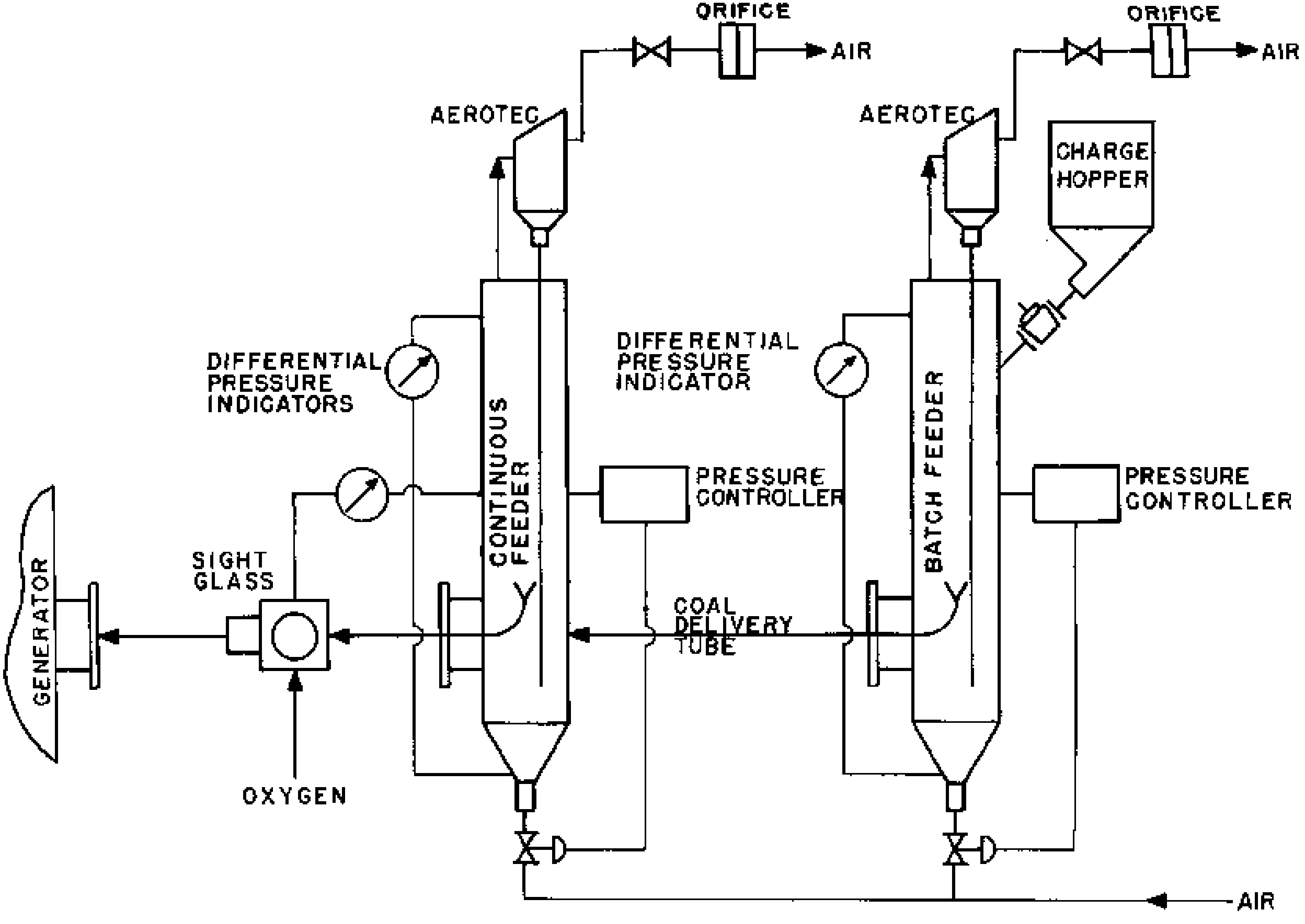

Figure 55. - Coal-feeding system for Morgantown pilot plant.

Fluidized Coal-Feeding Unit

The fluidized feeder[3/][4/] developed at Morgantown to supply pulverized coal at a steady rate to the generator has now been in operation for about 2 years. Extensive tests have demonstrated its reliability. Experience with the feeder in operation has shown that the average hourly rate of coal flow deviates an average of about 1 percent. The materials that have been fed successfully include 90-percent-through-200-mesh coal, 100-percent-through-10-mesh coal, sand,[5/] pulverized limestone, and grains.[6/] The feeder should be applicable to any broken solid that can be fluidized. The volumetric percentage of solid in the mixture of solid and gas flowing through the delivery tube is about 20 and is essentially independent of pressure, at least between 1 and 10 atmospheres.

A diagram of the feeder now in use at the Morgantown pilot plant is shown in figure 55. The feeder consists essentially of a fluidized bed of coal and a coal delivery tube, the entrance to which is near the bottom of the bed. A funnel that forms the entrance permits partial settling of the mixture, producing a uniform ratio of coal to air. The mixture flows through the delivery tube at about one volume of coal to four of gas. As the mixture flows as a fluid, the rate of coal delivery to the gasifier will be constant if the pressure drop causing flow is constant and if the ratio of coal to gas is constant. Constancy of pressure drop can be maintained by suitable instrumentation. Measurements of the coal:air ratio with an instrument developed for that purpose[3/][4/][7/] have shown that the standard deviation from the average per second rate of coal flow is 1 percent. The effect of pressure variations in the gasifier on the rate of coal delivery is minimized by having a relatively high pressure drop in the coal-delivery line as compared to the pressure variation in the gasifier. As the flowing mixture has a high weight ratio of coal to gas, the dilution of the make gas by the conveying gas is negligible; and as there are no moving parts to wear, constancy of calibration is maintained.

Tests on the feeder, using 90-percent-through-200-mesh coal and a delivery tube 60 feet long, have shown that the optimum feed rate for a 0.25-inch I.D. tube is 100 to 300 pounds of coal per hour; for a 0.33-inch I.D. tube, 200 to 600; and for a 0.44-inch I.D. tube, 500 to 1,500. The lower limit of usefulness depends on a minimum velocity in the feed line, and the upper limit on the point at which a large increase in pressure differential makes a small increase in feed rate. The superficial linear velocity through the fluidized bed ranges between 0.05 foot per second and 0.3 foot per second. Below the lower value, the flow through the coal-delivery line becomes unsteady, and above the upper value the quantity of entrained dust becomes

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[3/] Bureau of Mines, Synthetic Liquid Fuels. Annual Report of the Secretary of the Interior for 1949. Part I - Oil from Coal: Rept. of Investigations 4651, 1950, pp. 51-54.

[4/] Holden, J. H., and Albright, C. W., The Morgantown Fluidized Feeder: American Gas Association Tech. Paper PC-50-10, May 1950, 8 pp.

[5/] See p. 45 of reference cited in footnote 3.

[6/] Longhouse, A. D., Brown, B. P., Simons, H. P., Albright, C. W., The Application of Fluidization to Conveying Grain. Scientific Paper No. 420, West Virginia Agriculture Experimental Station, Morgantown, W. Va., July 1950, p. 349.

[7/] Dotson, J. M., Holden, J. H., Seibert, C. B., Simons, H. P., and Schmidt, L. D., New Method Measures the Solid:Gas Ratio in High-Solid Flow: Chem. Eng., vol. 56, 1949, pp. 128-130.

excessive. The velocity that appears to give the smoothest operation is in the lower part of the range. In the tests the mixture flowing through the coal-delivery tube had a coal:air ratio that ranged between 100 and 250 pounds of coal per pound of air. Using a 10-inch I.D. column, the ratio of fluidizing air to conveying air ranged from 10/1 to 30/1.

Results of pressure-drop measurements on the coal-delivery tube are shown in figure 56. Figure 57 shows a possible method of correlating these results. The expression $\Delta p / \Delta p_a$ is the ratio of the measured pressure drop to that of air flowing at the same average pressure level and at the same velocity as that of the mixture. The effect of increasing particle size is to cause a greater pressure drop at the same rate of coal flow. Runs made with coal crushed to pass a 10-mesh screen showed the pressure drop to be about three times that using 90-percent-through-200-mesh coal. Some investigations have been made at higher pressures, showing that the fluid flow properties are the same as those at the lower pressures; that is, the same ratio of 1 volume of coal to 4 of conveying gas flows through the delivery tube, and approximately the same fluidizing velocity is required.

Figure 58 shows a batch feeder designed for use at 30 atmospheres; it is to be used at the Morgantown station for feeding a gasifier operating at this pressure. The major differences between this feeder and the low-pressure feeder are the design to withstand high pressures and the recycle compressor. The recycle compressor is necessary to keep the power consumption of the feeder at a reasonable level.

Uses of the feeder other than those enumerated have not been investigated extensively, but the following appear feasible: To deliver solids to reactors for heterogeneous reactions, particularly where uniform addition of the solid with respect to time is desired; to proportion and mix one solid with another solid; to mix a solid with a gas in the desired proportions; in transferring heat to a flowing stream of solid and gas inside of a tube; and in the pneumatic conveying of a wide variety of solids. When pressure is near atmospheric, the power consumption is low.

Gas Treating and Testing

Occurrence of Organic Sulfur Compounds in Synthesis Gas

Sulfur analyses on the gas made in the Morgantown pulverized-coal gasifier indicated that the organic sulfur concentration in the gas is not necessarily proportional to the sulfur content of the coal gasified. Table 19 gives the sulfur contents of gas produced during some high-temperature steam runs.

It can be noted that, with a given coal, generally lower organic sulfur concentrations result from higher steam:coal ratios. This is undoubtedly due to hydrolysis and destructive hydrogenation of the organic sulfur compounds.

An experiment in which hydrated lime was added to the coal feed of the small laboratory gasifier showed that the addition of 0.038 pound of lime per pound of coal lowered the organic sulfur content of the gas but had no effect on the hydrogen sulfide concentration. Doubling this ratio of lime to coal further reduced the concentration of organic sulfur and reduced the concentration of hydrogen sulfide in the gas.

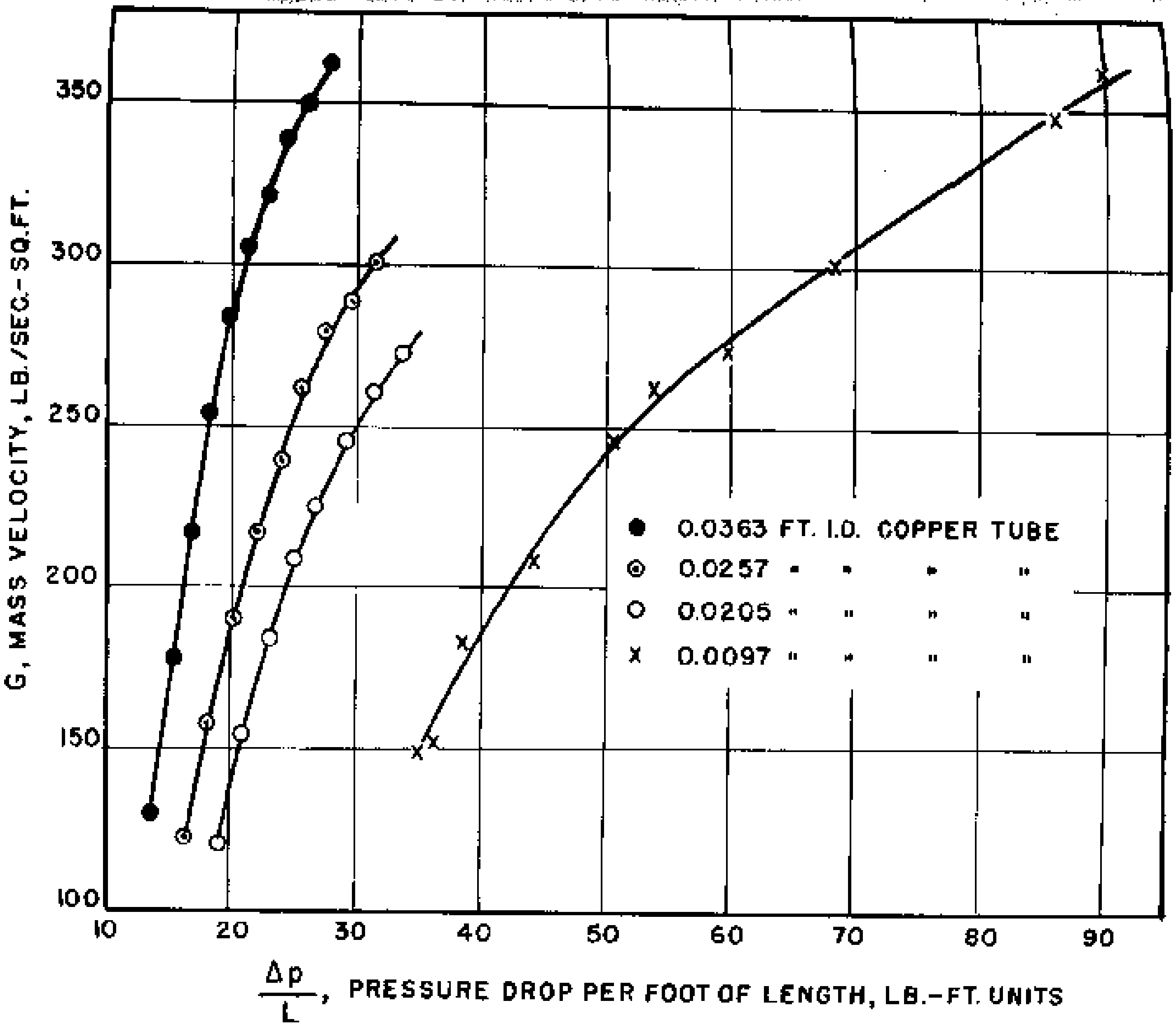

Figure 56. - Pressure-drop measurements for flow of dense coal-air measurements.

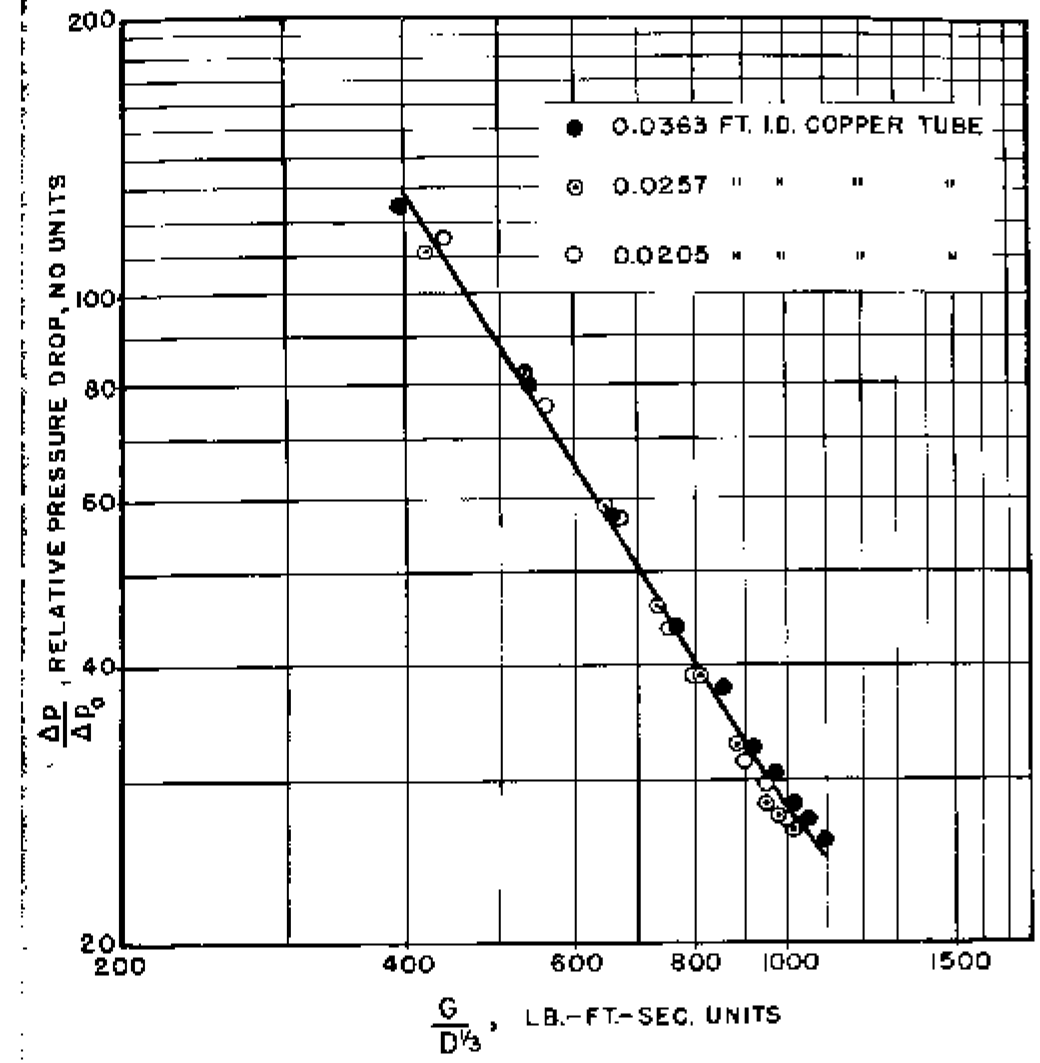

Figure 57. - Correlation of pressure-drop measurements shown in previous figure.

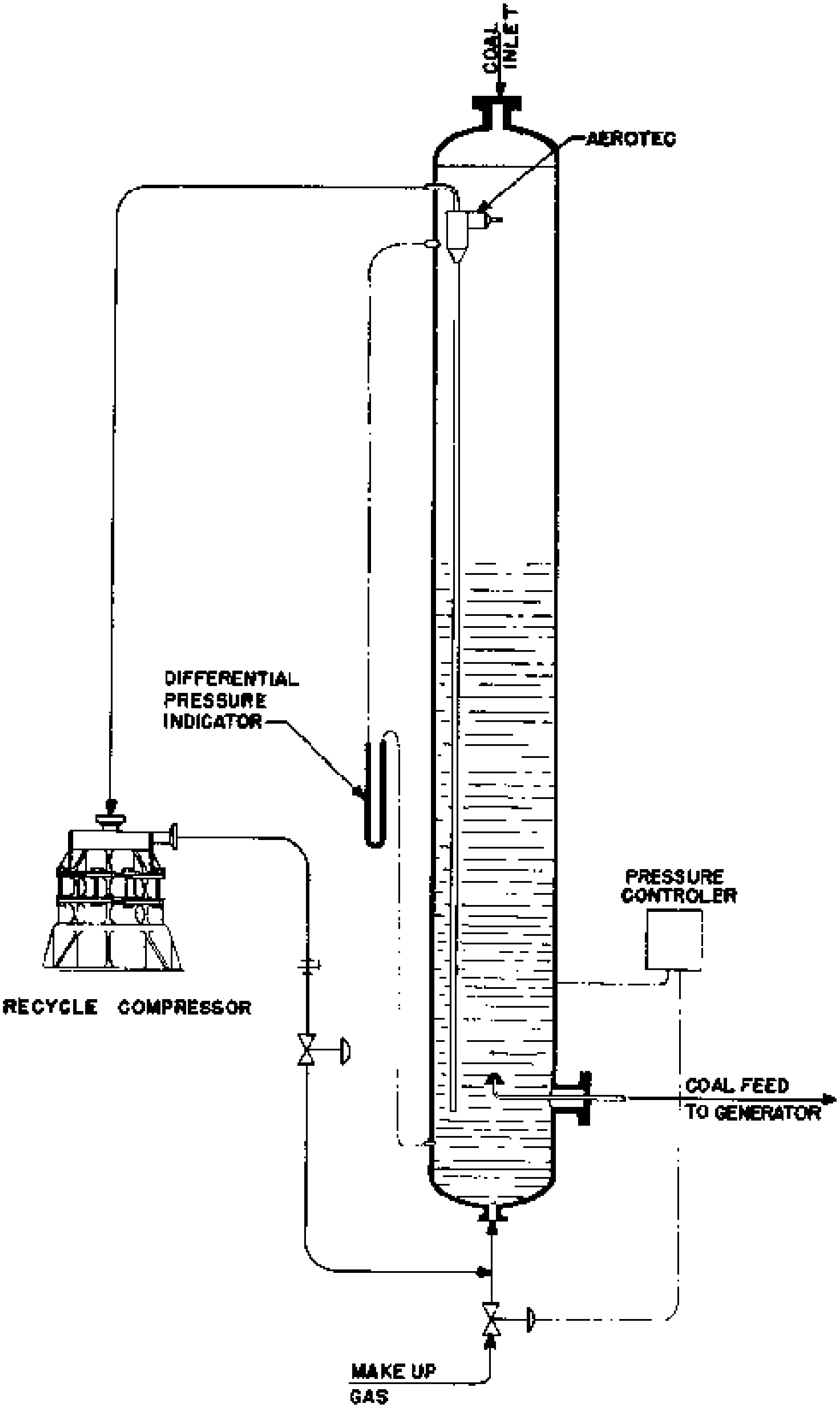

Figure 58. - Coal feeder (600 p.s.i.g.).

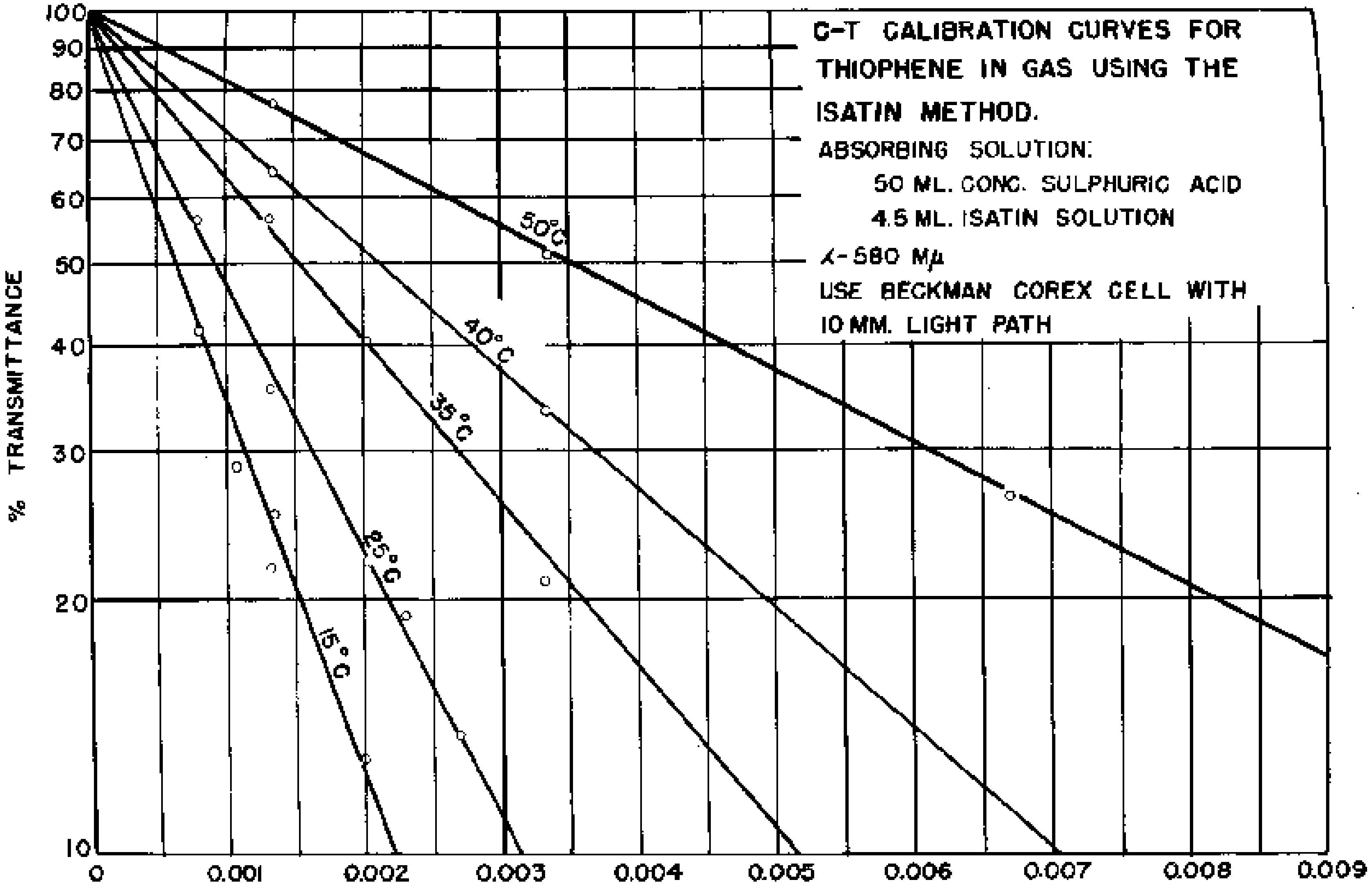

Figure 59. - Grains of sulfur as thiophene per 54.5 ml. of solution.

TABLE 19. - Sulfur in gas from coal gasifier

Coal used	S in coal, percent	Steam:coal ratio	H_2 in gas, percent	Generator temp., °F.	Grains/100 cu. ft.	
					H_2S	Org. S
West Virginia bituminous	2.9	2.71	48.7	2,296	327	11.5
Do.	2.9	2.05	49.0	2,150	317	13.4
Do.	2.4	.92	36.3	2,115	335	28.3
Do.	2.4	1.12	37.4	2,238	290	19.6
Do.	2.4	.43	32.6	2,358	358	30.1
Do.	2.0	1.28	39.7	2,375	300	21.6
Wyoming bituminous......	1.1	1.10	39.1	2,536	150	19.0
Do.	1.1	1.08	37.7	2,468	157	21.4

No attempt was made to identify the types of organic sulfur compounds, other than thiophene sulfur, in the gas made in the Morgantown gasifiers. Results from tests in which the isatin method was used showed that there was no thiophene in the synthesis gas. Figure 59 gives the calibration curves used with the isatin test as developed at this station.[8]

Bench-Scale Purification Experiments

Bench-scale experiments on the removal of hydrogen sulfide and organic sulfur from gas were conducted. These studies included catalytic conversion experiments and activated carbon adsorption tests.[9]

Although cost estimates indicate that an alkalized iron process for removing organic sulfur from gas is more expensive than one employing a low-cost activated carbon,[10] this process was investigated because it is included in the purification train of the Bureau's demonstration plant at Louisiana, Mo.

An alkalized-iron catalyst containing 26 percent sodium carbonate successfully lowered the organic-sulfur content of a simulated synthesis gas to less than 0.1 grain of sulfur per 100 cubic feet until 65 percent of the carbonate had been converted to sodium sulfate. Somewhat higher conversion of the carbonate to the sulfate (78 percent) was obtained whenever 100 percent soda ash was used. Experiments with these alkali catalysts were run at 250° to 300° C. and 300 p.s.i.g., employing linear velocities up to 760 feet per hour at 20 atmospheres and space velocities up to 7,000 std.cu.ft./cu.ft./hr.

Investigation of a copper-chromium-vanadium catalyst for removing both hydrogen sulfide and organic sulfur from gas at atmospheric pressure was continued. The catalyst, after absorbing 10 percent of its weight in sulfur, is revivified with air and steam. Table 20 gives data for several successive runs.

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[8] Wainwright, H. W., and Lambert, G. I., A Colorimetric Method for the Determination of Thiophene in Synthesis Gas: Bureau of Mines Rept. of Investigations 4753, Nov. 1951, 11 pp.

[9] Sands, A. E., Wainwright, H. W., and Egleson, G. C., Organic Sulfur in Synthesis Gas: Occurrence, Determination, and Removal: Bureau of Mines Rept. of Investigations 4699, July 1950, 51 pp.

[10] Sands, A. E., and Schmidt, L. D., The Recovery of Sulfur From Synthesis Gas: Ind. Eng. Chem., November 1950, pp. 2277-2287.

TABLE 20. - Sulfur removal with a Cu-Cr-V catalyst

Run No.	Sp. vel., hr.-1	H_2S, gr./100 cu. ft. In	Out	Removal, percent	Org. S, gr. S/100 cu. ft. In	Out	Removal, percent	S absorbed per gram cat., gr.
27.............	2,000	350	1.5	99.6	56	0.01	99.99	1.2
28.............	1,000	350	2.5	99.3	35	.05	99.86	1.7
29.............	2,000	530	2.5	99.5	60	.10	99.83	1.6
30.............	2,000	400	2.5	99.5	40	.05	99.88	1.6
31.............	2,250	250	2.5	99.0	30	.05	99.83	1.7
32.............	2,200	280	2.5	99.1	30	.03	99.90	1.6
33.............	2,200	450	3.0	99.3	25	.07	99.72	1.8

The above data were obtained from runs made at 450° C. At a lower temperature of 300° C., the purified gas contained less than 0.01 grain of sulfur as hydrogen sulfide per 100 cubic feet, but the organic sulfur conversion was somewhat less than that at the higher temperature, averaging 98.5 percent. This indicates that, to meet purification specifications for synthesis gas, two separate catalyst beds would be required, operated at different temperatures.

Numerous bench-scale tests were made to study the practicability of removing carbonyl sulfide from synthesis gas by means of activated carbon at 300 p.s.i.g. Although no determination of individual organic sulfur compounds (other than thiophene) has been made on the Bureau's synthesis gas, it is quite likely that carbonyl sulfide predominates. In the removal of organic sulfur from synthesis gas by activated carbon, the removal of carbonyl sulfide would be controlling, and therefore tests were made with this gas.

The results of these tests, if confirmed in the pilot plant and demonstration plant, indicate a very great saving in the cost of removing organic sulfur compounds when compared with the hot alkalized iron process that was standard in the German synthetic liquid fuel plants. Compared with the cost of purification by alkalized iron, the use of the relatively cheap grade of activated carbon tested (15 cents per pound) should effect a saving of $17,000 a year in the Bureau's demonstration plant and more than $1,000,000 a year in a 10,000-barrel-per-day commercial synthetic liquid fuel plant. Alkalized iron, which was included in the design of the demonstration plant, will be retained primarily as insurance with a very low (if any) consumption of the alkalized iron. It is anticipated that the demonstration plant will be able to prove that alkalized iron need not be included in commercial plants.

Figure 60 graphically presents data obtained from tests in which an inexpensive activated carbon was used for removing carbonyl sulfide from a simulated synthesis gas.

The presence of carbon dioxide in the gas greatly decreases the capacity of the activated carbon for carbonyl sulfide adsorption.

It is interesting to note that the service time of the carbon, that is, the length of the adsorption period before regeneration is necessary, is the same within limits of organic sulfur content between 10 and 30 grains per 100 cubic feet when the linear velocity is between 100 and 2,000 feet per hour. For concentrations above

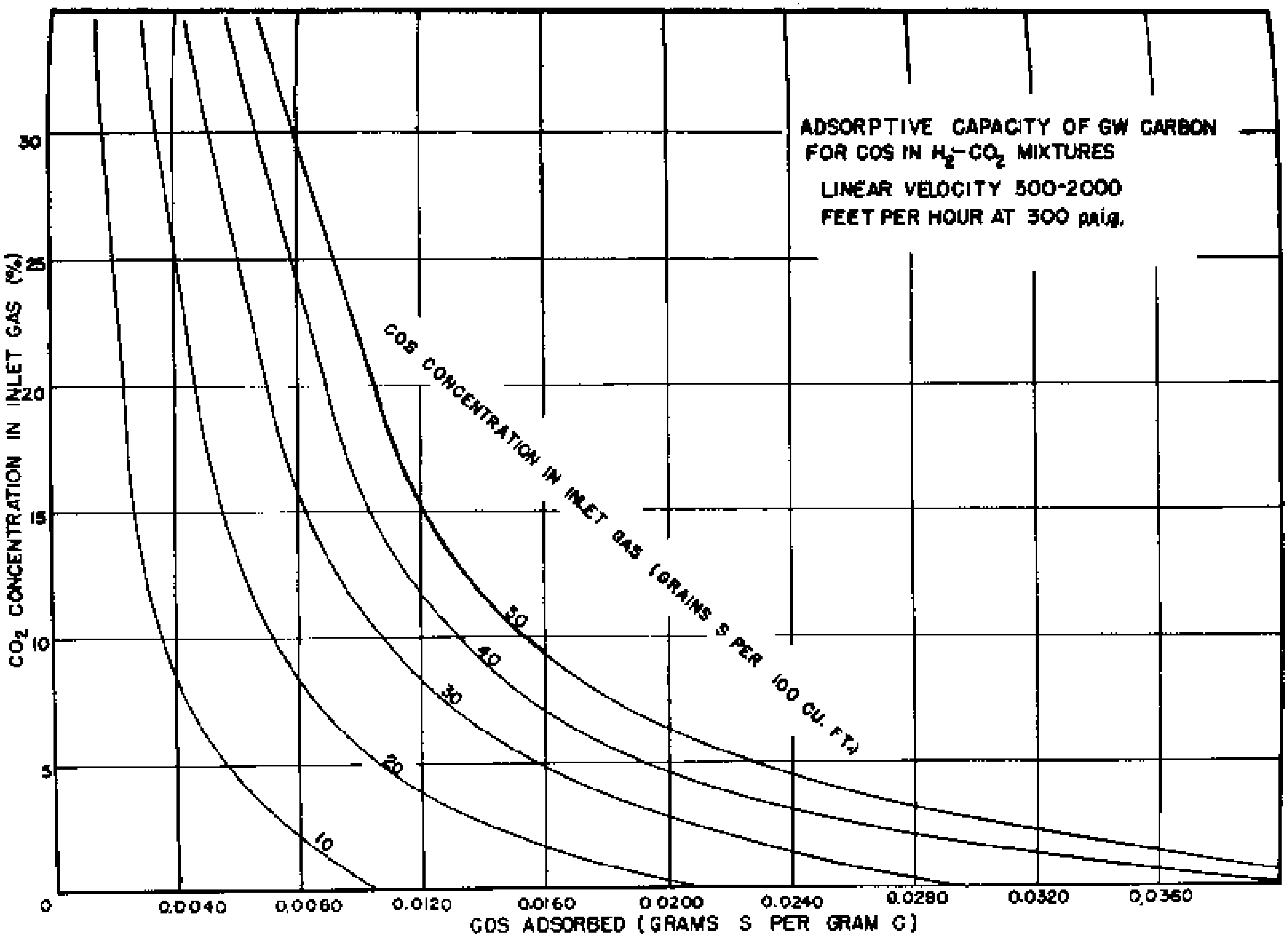

Figure 60. - Adsorptive capacity of inexpensive activated carbon for carbonyl sulfide in simulated synthesis gas.

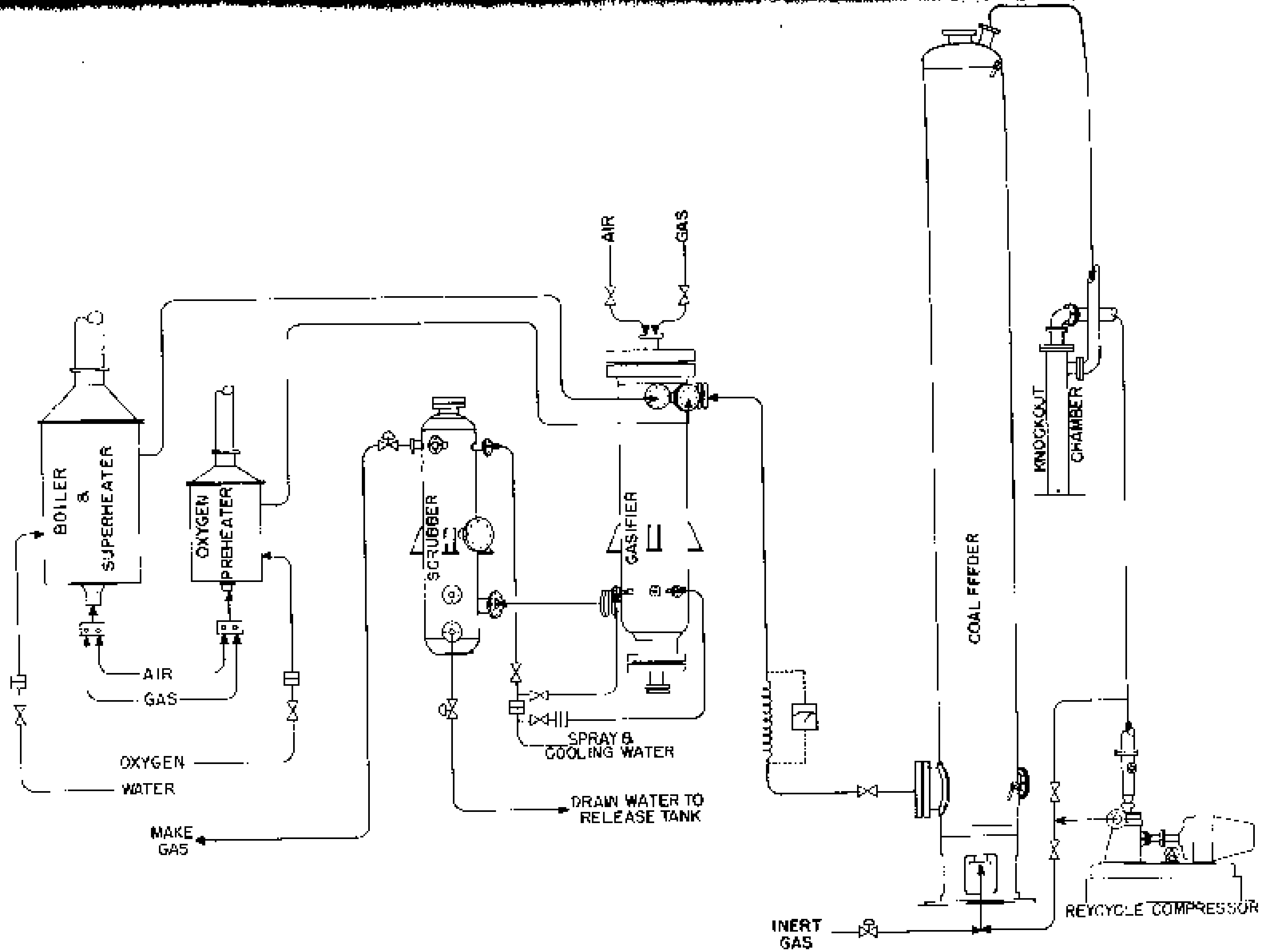

Figure 61. - Simplified flow sheet for pressure gasification.

30 grains per 100 cubic feet, and also when the linear velocity is below 100 feet per hour, the service time is affected by changes in concentration and velocity.

Bench-scale tests are in progress for studying the removal of both hydrogen sulfide and organic sulfur by means of activated carbon at elevated pressures. The hydrogen sulfide concentration in the gas being tested is approximately the same as that in the gas leaving the hydrogen sulfide-removal system (about 25 grains per 100 cubic feet). Data from a few preliminary tests indicate that activated carbon will remove at least 25 grains of hydrogen sulfide without reducing the capacity of the carbon for carbonyl sulfide adsorption. The effect on this operation of small amounts of oxygen in the synthesis gas is being investigated.

Pressure Gasification

Gasification at a pressure of 450 p.s.i.g. is believed to offer great possibilities in reducing the cost of coal gasification when the gas produced is to be used at this or higher pressures.

For example, consider a commercial plant generating 18,000,000 std.cu.ft. per hour of raw gas, which would produce approximately 10,000 barrels of product per stream day. We can assume that this gas would be 85 percent $CO + H_2$ and that the oxygen requirements would be 320 cubic feet of oxygen per M.c.f. of $CO + H_2$.

As commercial centrifugal compressors with an efficiency of 75 percent are now available, the estimate is based on this type of machine. If the 18,000,000 std.cu.ft. per hour of synthesis gas is compressed to 450 p.s.i.g. with 0.74 efficiency, the energy of compression becomes 107,000 horsepower.

If the synthesis gas is generated at 450 p.s.i.g. pressure, only 4,900,000 std.cu.ft. per hour of oxygen need be compressed to 500 p.s.i.g. The energy required is 30,000 horsepower. The net saving in energy is 107,000 minus 30,000 which amounts to 77,000 horsepower.

In addition to conserving energy, there is the possibility that the gasification throughput per cubic foot of generator space will be greatly increased at elevated pressures, saving labor and material through greatly reducing the number of units.

To take advantage of these potential savings by pressure gasification, a pilot plant (fig. 61) has been designed and is under construction at Morgantown to gasify coal at 450 p.s.i.g. This plant is now 80 percent complete. The work on gasification under pressure is being done in cooperation with the Babcock & Wilcox Co., which maintains a resident engineer at Morgantown.

The pressure-gasification unit is designed for 500 pounds of coal per hour, with provision to vary the coal feed from 200 to 700 pounds per hour. The design figure on oxygen is 5,000 std.cu.ft. per hour, with provision to go to 7,000 std.cu.ft. per hour. The design figure on steam is 300 pounds per hour, with provision to go to 1,000 pounds per hour.

Description of Process

Coal is pulverized to 70 percent through 200-mesh in a standard ball mill. It is then conveyed to the fluidized feeder. The feeder holds 4,000 pounds of coal, or enough for an 8-hour run. The feeder is then pressurized, and the fluidizing gas is recycled. Through the draw-off tube, the pulverized coal can be fed to the generator. The coal enters the generator radially and cuts across the oxygen and steam streams.

The process steam is generated in a special high-pressure boiler at 600 p.s.i.g. Following the boiler is a superheater section, in which the temperature is raised to 1,400° F. The superheated steam is then conveyed to the gasifier, where it is introduced tangentially.

Process oxygen is stored in a Cascade unit at pressures up to 2,000 p.s.i.g. From the storage unit, the oxygen is taken through a preheater, where its temperature is raised to 1,000° F. The preheated oxygen is then introduced into the generator tangentially, causing the reaction products to swirl.

Coal, oxygen, and steam react in the gasifier to form H_2, CO, and CO_2. The anticipated reaction temperature is 2,500° F., although a higher temperature may be required to cause the slag to remain molten.

Generated gases and slag, together with unreacted coal, flow downward and out the slag hole at the bottom of the reaction section. Here the gases and slag encounter a water spray, and the slag stream is broken up into fragments that fall into a slag pot filled with water.

The gas is cooled by water spray and is then conducted to the scrubber, where it is washed and cooled. In the present stage of design, the scrubbed and cooled gas is flared.

Description of Special Apparatus

Fluidized Feeder

The fluidized feeder is a pressure vessel 34 feet high and 3 feet in diameter with a design pressure of 600 p.s.i.g. Fluidizing gas enters at the bottom and flows upward through a distributing plate composed of steel wool enclosed in a metal grid. The coal-gas mixture, in a fluidized state, is drawn off in a 1/2-inch tube and conveyed to the generator.

Recycle Compressor

To save compressor horsepower, the fluidizing gas is recycled. As the gas is contaminated with coal, a centrifugal compressor offered the most promise; however, the flows were much too low and the static pressure too high for any commercial type of equipment. Also, a range of static pressures from 60 p.s.i.g. to 500 p.s.i.g. had to be investigated.

As a joint effort with the Sawyer-Bailey Corp., a compressor was designed and built to utilize the centrifugal principle with a variable-speed electric drive for the service. Preliminary tests of this compressor show that it will operate at static pressures from 1 atmosphere absolute to 600 pounds per square-inch gage. The flow rate can be varied from 125 to 5,000 standard cubic feet per minute. A pressure rise of 25 pounds per square-inch gage is possible with 575 pounds per square inch-gage inlet pressure. A change in gravity of the recycle gas from 0.55 to 1.0 can be handled satisfactorily.

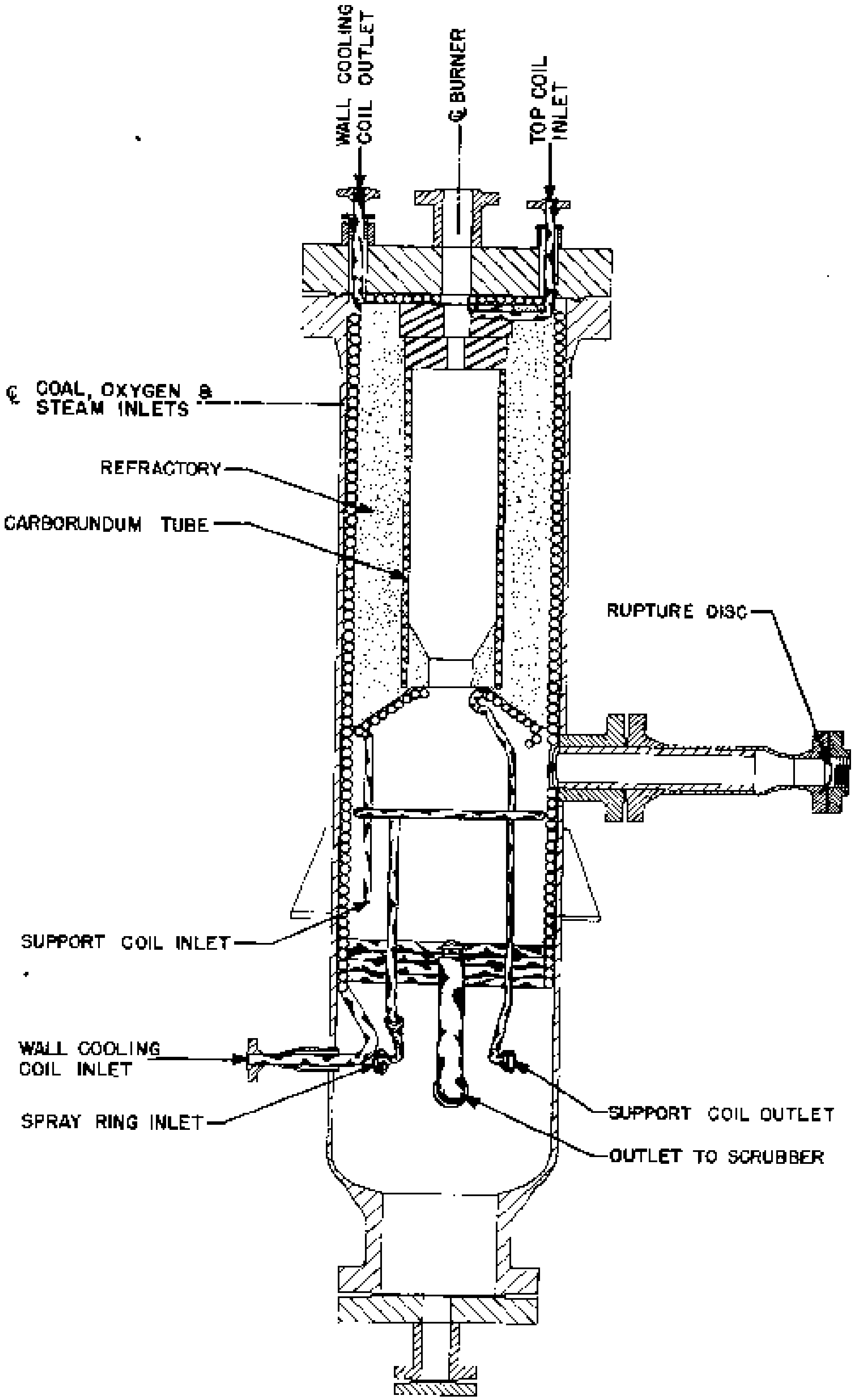

Figure 62. - Reactor for pressure gasification.

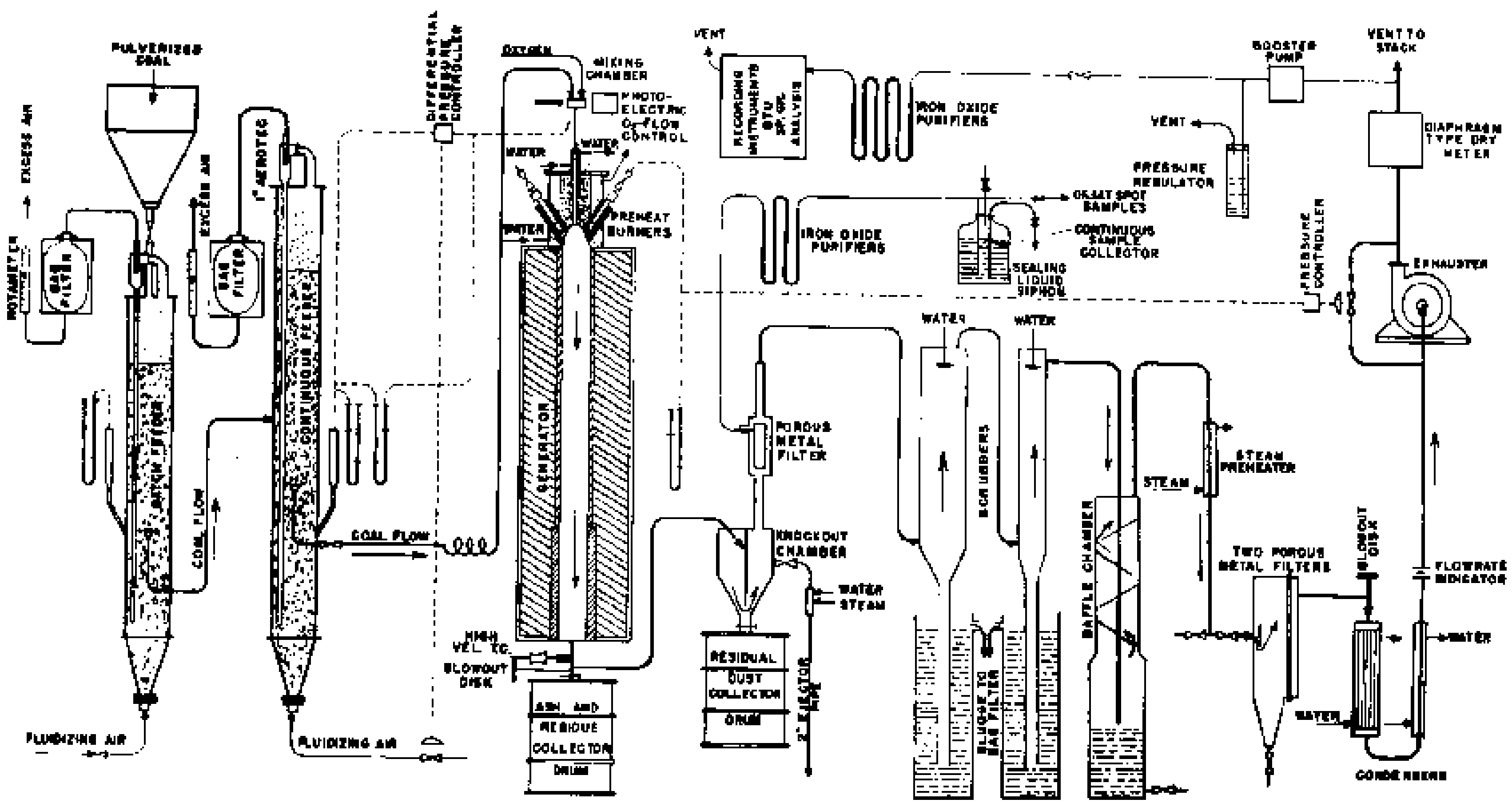

Figure 63. - Flow diagram of apparatus used for testing fuels for synthesis-gas production - 50-pound-per-hour gasifier, pneumatic feeding system, and purification train.

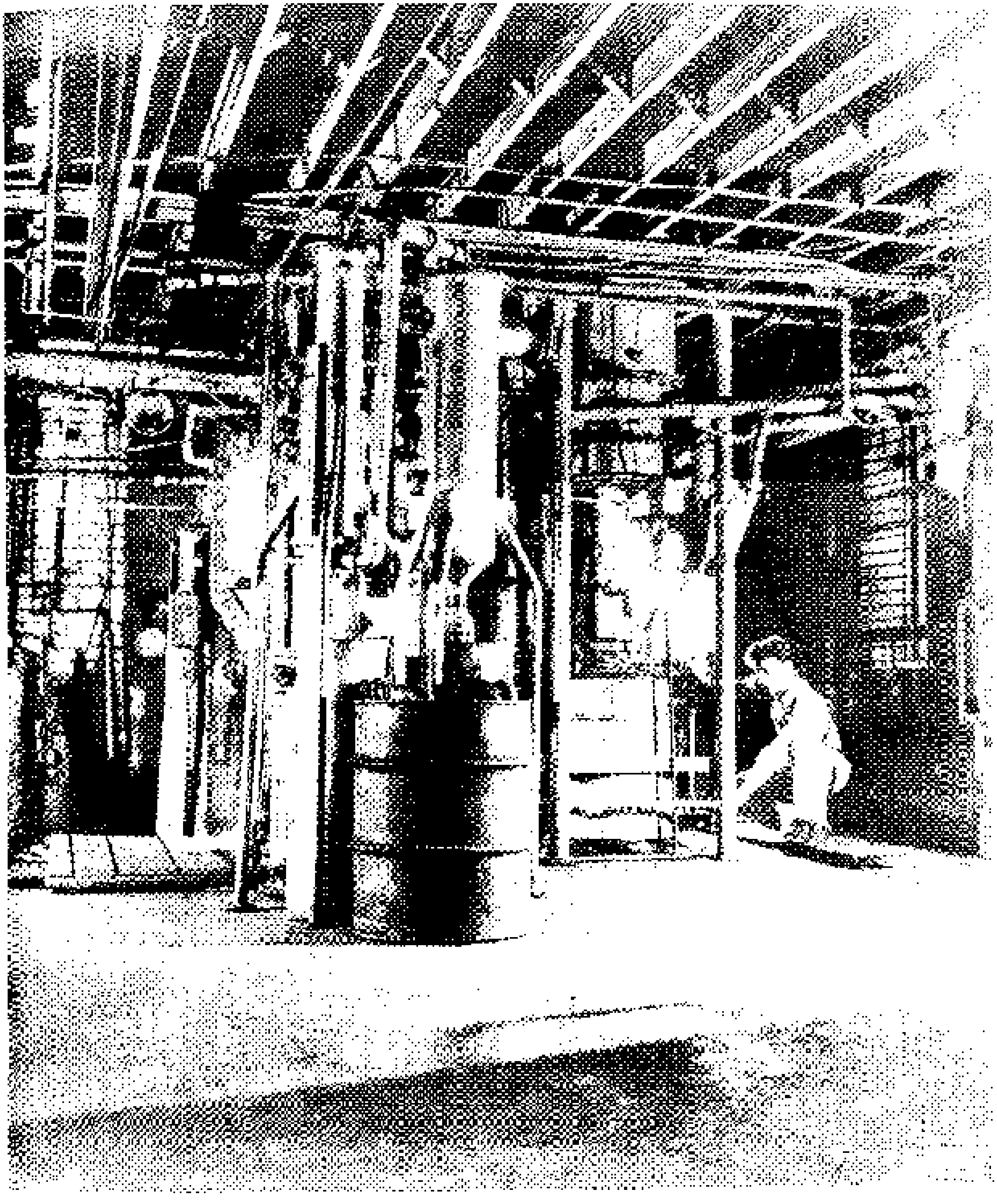

Figure 65. - Apparatus used for testing fuels for synthesis-gas production. Lower floor, showing lower part of 50-pound-per-hour gasifier and purification train.

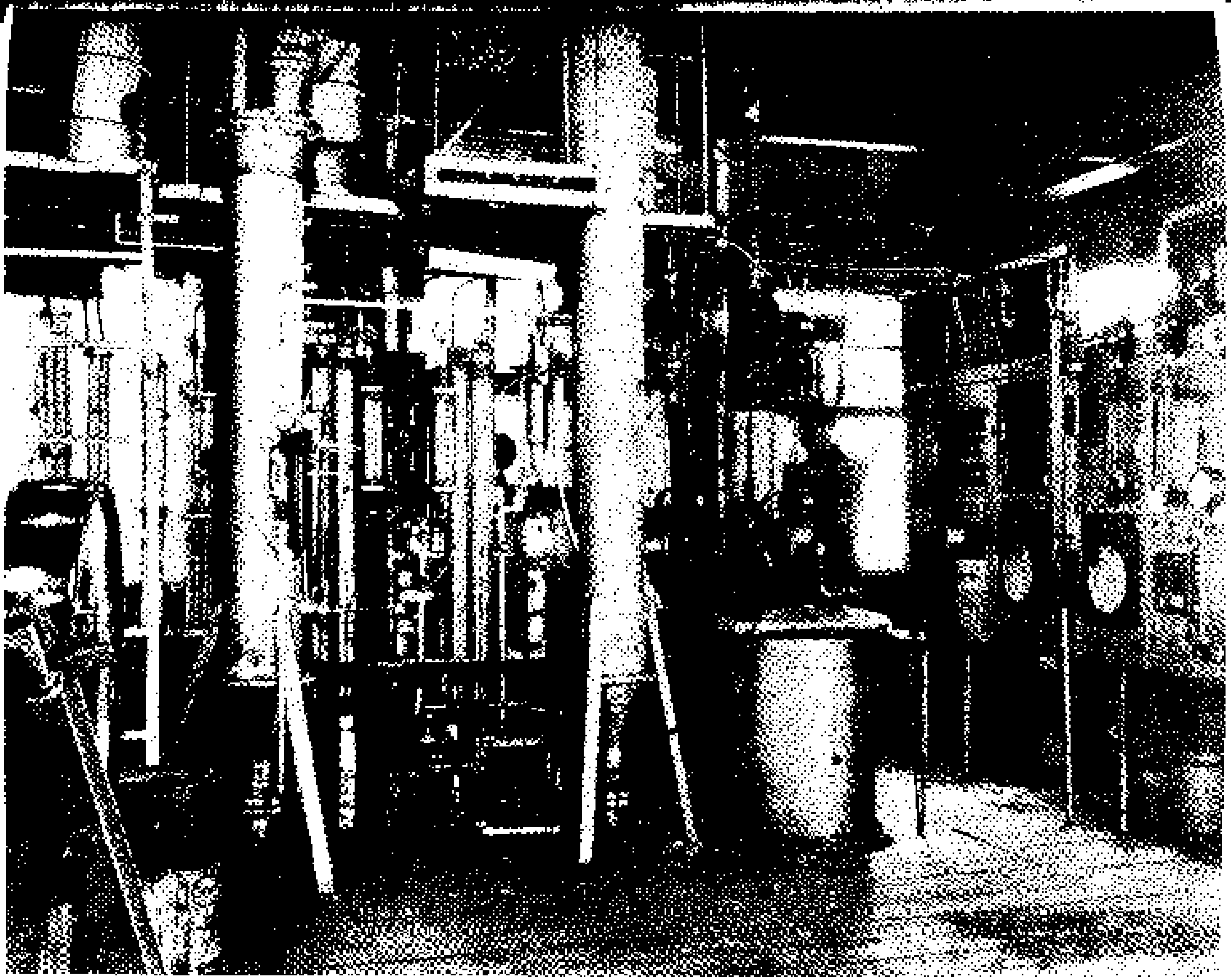

Figure 64. - Apparatus used for testing fuels for synthesis-gas production. Upper floor, showing the pneumatic feeding system, upper part of 50-pound-per-hour gasifier, and control board.

Gasifier

The gasifier is the most critical piece of equipment in the high-pressure pilot plant, for it must be operated at a pressure of 450 p.s.i.g. and an internal temperature of 2,500° F.

The gasifier, as designed, (fig. 62) consists of a steel shell 30 inches in diameter by approximately 12 feet in length. The design pressure is 900 p.s.i.g., and the working pressure 450 p.s.i.g.

The inside of the shell is lined with a continuous coil of 1-inch tubing, through which water is circulated. This affords protection to the shell in case of refractory failure. The design temperature on the shell is 650° F.

The reaction chamber proper is a refractory-lined cylinder 12 inches in diameter and 4 feet long. The reactants enter through nozzles at the top. Steam and oxygen enter tangentially and the coal axially. The bottom of the reaction chamber is necked down by cooling coils to provide a 6-inch diameter slag tap. The bottom of the reactor is a water-filled container for slag.

Future plans call for continuous slag removal. However, for the present the run will be terminated when the slag pot is filled.

Anticipated completion date for the high-pressure gasification plant is December 31, 1950.

Laboratory-Scale Powdered-Coal Gasification

The development of the small-scale pilot unit for testing fuels for their utility in synthesis-gas production on a commercial scale has been described in the synthetic fuel reports of the Secretary of the Interior for 1948 and 1949.

Modifications made since then include the substitution of a much larger continuous feeder (10-inch I.D. fluidizing tower) for the former 4-inch diameter vessel and the use of a single batch feeder of the same diameter, as shown in the flow diagram in figure 63. The dry method of removing dust from the synthesis gas was replaced by two efficient scrubbers and a baffle chamber capable of recovering the bulk of the carbonaceous residual dust, followed by an improved micrometallic filter for removal of the last traces of residue.

These improvements and installation of additional automatic controls resulted in a simple and very easily controlled down-draft entrainment process for the gasification of powdered fuels at feed rates up to 55 pounds per hour. The apparatus (see figs. 64 and 65) is capable of gasifying different types of fuels under variously chosen operating conditions at unusually high coal-throughput and corresponding gas-output rates. These and other operating results were found to be functions of the rank of the coal gasified and, probably, their oxygen contents and reactivities. Operating results obtained with a West Virginia bituminous coal, a Colorado subbituminous coal, and a Pennsylvania anthracite are compared in table 21.

TABLE 21. - Effect of type and rank of coal gasified

Operating data and results	Colorado sub-bituminous coal	West Virginia bituminous coal	Pennsylvania anthracite
Run number and date	71 2/50	70 1/50	67 12/49
Operating conditions:			
Size of coal, percent through 200 mesh	89.0	89.0	73.4
Duration of run, hours	4.57	4.55	4.00
Rate of coal feed, lb./hr.1/	54.3	52.2	49.3
Oxygen:coal ratio, lb./lb. dry, ash-free coal2/	0.887	0.866	0.658
Steam:coal ratio, lb./lb. dry, ash-free coal3/	0.208	0.159	0.199
Steam temperature, °F.	705	575	297
Operating results:			
Output of synthesis gas, std.cu.ft./hr.4/	1,406	1,358	841
Analysis of synthesis gas:			
Percent CO_2	12.1	10.3	18.2
Percent H_2	33.8	33.3	31.0
Percent CO	50.7	53.4	48.1
H_2/CO ratio	0.67	0.62	0.64
Temperature of gas made at outlet, °F.5/	1,635	1,955	2,100
Dry residue output, lb./hr.	10.3	12.4	28.2
Total ash in residue, percent	36.3	29.4	24.2
Coal input, lb./M.c.f. CO + H_2	45.7	44.3	74.2
Oxygen input, cu.ft./M.c.f. CO + H_2	310	402	477
Steam input, lb./M.c.f.; CO + H_2	8.39	8.50	12.68
Carbon gasified, lb./100 lb. C in coal	84.3	70.6	46.7
Thermal efficiency, heat of combustion of gas.. per 100 B.t.u. in coal charged	72.8	56.0	34.9
Yield of CO + H_2, cu.ft./lb. dry, ash-free coal	24.6	25.4	15.7
Coal throughput rate, lb./cu.ft. space/hr.	41.8	40.2	37.9
Gas output rate, cu.ft.CO+H_2/cu.ft. space/hr.	914	905	511
Total heat losses, B.t.u./lb. of coal charged6/	533	479	626

1/ On "as-charged" (moist) basis.
2/ Consists of oxygen injected plus that originally present in the dry coal.
3/ Consists of steam injected plus that originating from moisture in coal.
4/ All gas volumes refer to 60° F., 30 in. Hg, dry.
5/ Measured by high velocity thermocouple.
6/ Radiation and unaccounted for losses.

The operating results were found to be closely reproducible. As a rule, gasification in the 75 runs made to date was purposely discontinued after 4 to 5 hours, although some runs of 30 hours' duration have been made and could have been continued indefinitely. Very little slag formation was ever observed on the silicon carbide lining of the 6-inch I.D. and 7-foot long generator tube.

The last 15 runs made included investigations of the effects of a number of operating variables and such special experiments as the addition of lime to the coal

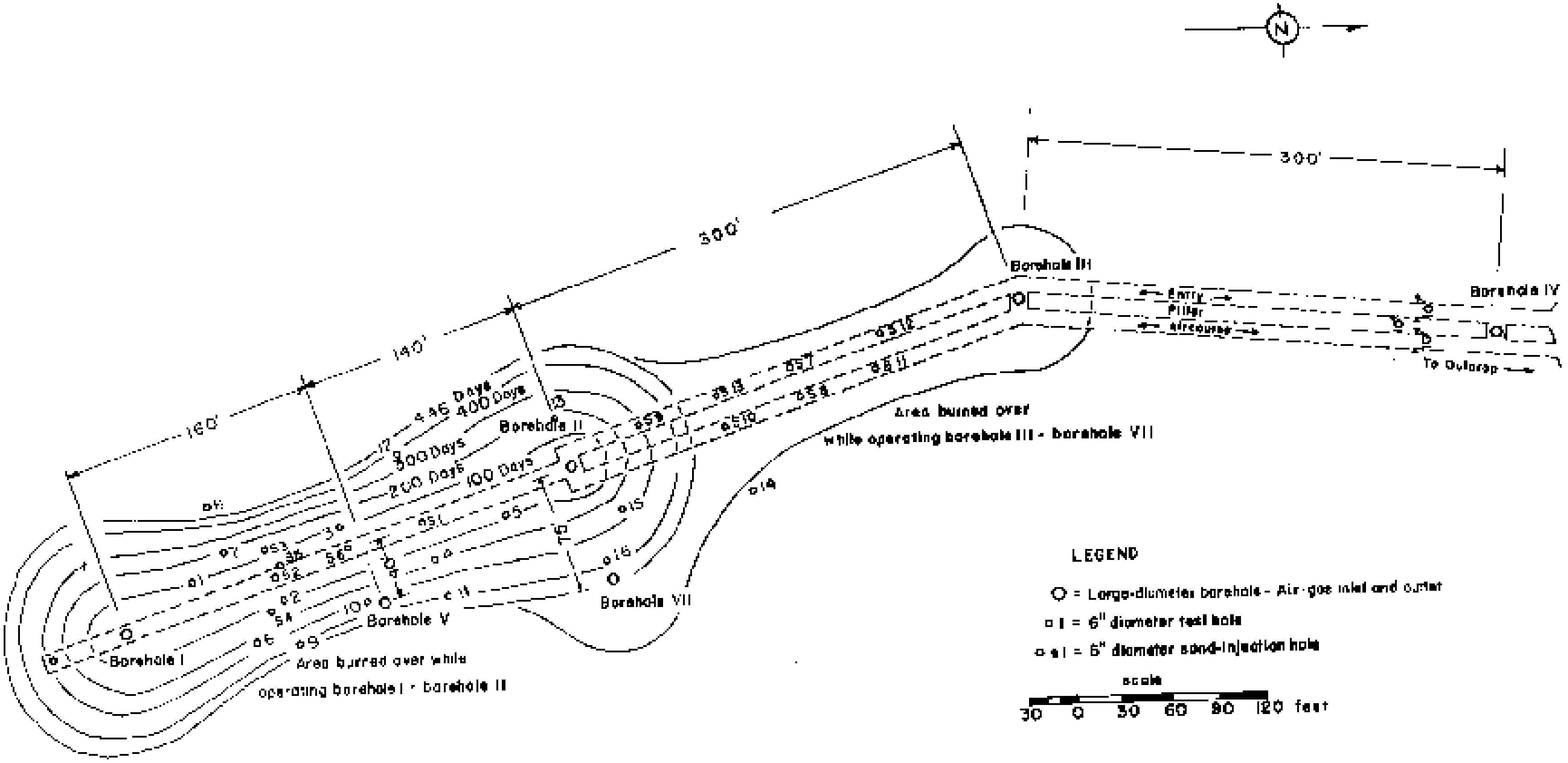

Figure 66. - Development of underground gasification system at Gorgas, Ala., and area burned over between March 15, 1949 and September 30, 1950.

gasified. These, together with a thermodynamic translation of the operating results to conditions prevailing in the gasification on commercial scale, will be described in a forthcoming publication.

Underground Gasification Project, Gorgas, Ala.

Successful development of a process for underground gasification of coal would make available a source of energy for electric-power generation and gaseous products suitable for the manufacture of synthetic liquid fuels. It may also be possible by this means to exploit coal deposits that are not now mined because of thinness of bed, presence of mineral impurities, or difficult mining conditions.

Construction and operation of the underground coal gasification project at Gorgas, Ala., was planned and supervised by Bureau of Mines engineers. Under a nonprofit contract with the Government, the Alabama Power Co. provided the personnel and equipment necessary for construction and operation. The Gorgas site, including the surface area and the underlying coal bed, was provided by the Alabama Power Co. without cost to the Government.

This experiment on underground gasification is being conducted in a 160-acre tract of the Pratt coal bed isolated from the main body of the bed by nature. The coal bed is 42 inches thick and dips slightly to the southeast. The cover above the experimental system ranges between 125 feet and 160 feet. The upper 25 feet of cover is surface soil, clay, and weathered sandstone; below this lie stratified, unweathered shales and sandstones.

The underground development consists of entries driven in the coal from the outcrop and connected at intervals with the surface by means of large-diameter vertical boreholes. An entry and aircourse, each 10 feet wide and separated by a 10-foot pillar of coal, were driven 1,240 feet from the northern outcrop. They were directed almost due south for 940 feet and then turned to the southeast in order to remain under the greatest possible cover. An additional 300 feet of single entry was constructed, as it was thought that operational characteristics would be more clearly defined in a single entry. Five large-diameter vertical boreholes were provided as air inlets and gas outlets and intersect the entries at 300-foot intervals. An entry seal wall was constructed 140 feet from the outcrop to contain the air and gas and to prevent fire from spreading to the outcrop.

Operation of the underground system was started in March 1949 and has continued without interruption for 20 months. For the first 14-1/2 months operation was maintained in the section between boreholes I and II (see fig. 66). The section between boreholes III and VII was operated for 4 months, after which time the area between boreholes IV and VII was utilized. A gas turbine installation was operated for a short time at borehole III, utilizing the energy of the hot effluent gases to provide additional compressed air for injection into the underground system. By December 31, 1949, approximately 3,900 tons of coal (moisture- and ash-free basis) had been consumed in the section between boreholes I and II, and the area burned was approximately 27,000 sq. ft. In January 1950, borehole VI, was drilled midway between boreholes I and II and 40 feet east of the center line of the original entry. Borehole VI penetrated a zone of low-temperature coke near the reaction face, and it was planned to set up a reaction zone 150 feet in length between it and borehole II. Operation was started by injecting 7,200 cubic feet of air per minute into borehole VI and drawing gas off at borehole II. No difficulty was encountered in initiating flow in this section, and during the first cycle the gas obtained at borehole II had an average analysis as follows:

	Average, percent
CO_2	25.2
Ill.	.1
O_2	.9
H_2	.6
CO	.8
CH_4	.2
N_2	72.2

After 8 hours the flow was reversed, and approximately 5,000 cubic feet of air per minute was injected at borehole II, gas being drawn off at borehole VI. The rate of flow was limited to this value as a pressure of approximately 22 p.s.i.g. was required to force the fluids through the system. The average effluent gas analysis was:

	Average, percent
CO_2	3.3
Ill.	.0
O_2	.4
H_2	11.0
CO	13.8
CH_4	1.0
N_2	70.5
Heating value, B.t.u. per cu. ft.	90

After 5 hours the flow was again reversed, and 6,300 cubic feet of air per minute was blown into borehole VI. The back pressure during this cycle was maintained at 21 p.s.i.g., in contrast to 13.3 p.s.i.g. during the first cycle described. A partial blockage had occurred near the base of borehole VI. During this cycle a composite gas sample had the following analysis:

	Average, percent
CO_2	18.1
Ill.	.0
O_2	6.9
H_2	.9
CO	.9
CH_4	.5
N_2	72.7

After 8 hours' operation, the direction of flow once more was reversed. An air blast of 7,600 cubic feet per minute was applied at borehole II, and gas issued at borehole VI. This gas ignited at the top of borehole VI soon after the cycle was started, but it was extinguished in about 15 minutes. After 2 hours temperature rose sharply in borehole VI. This was accompanied by flame, molten rock, and slag issuing about 50 feet into the air from the borehole. Air input was stopped, and after the outlet had cooled it was found that it had become filled with slagged material to within 60 feet of the surface. Attempts to penetrate this material by churn drilling were unsuccessful. Operation of this section showed that it was possible to produce a combustible gas in large volume when the outlet borehole intersected the reacting face. It was further demonstrated that the boreholes used as gas outlets would require additional protection against the effects of heat.

In order to develop further the possibilities indicated by operation of borehole VI, it was decided to construct borehole VII east of borehole II on a line normal to the entry and intersecting the reaction face. A 6-inch churn-drill test hole 65 feet east of borehole II found coke at its junction with the coal bed. Accordingly, borehole VII was located 10 feet farther from borehole II, or 75 feet from the center line of the entries. Coal unaffected by heat, at a temperature of about 200° F., was found when drilling reached the coal bed. This hole was drilled 18 inches in diameter and was stopped after penetrating a few inches into the coal bed. A 12-inch-diameter steel-pipe lining was placed in the hole to within 1 foot above the coal bed, and the space between the lining and the wall of the hole was filled with cement and sand grout. The hole was then extended through the coal bed with a 12-inch bit. This construction insured against air or gas entering the hole above the level of the coal bed. Boreholes VII and III were fitted with water sprays at coal-bed level to protect them from the effects of high temperatures.

During the time required for construction of the borehole, and while the combustion face advanced to the bottom of borehole VII, the section between boreholes I and II was operated on a regular 8-hour cycle with a normal air input of 7,200 cubic feet per minute. At this time the section had increased in width to 90 feet midway between the boreholes and 130 feet opposite the boreholes.

The gas composition as of January 1, 1950, was approximately as follows:

	Percent
CO_2	6
Ill.	0
O_2	12
H_2	0.4
CO	0.3
CH_4	0.3
N_2	81

At the end of May, about the time that operation was ended in the original section, the gas composition reflected a further loss of contact between air and coal, as indicated by the increased oxygen content of the effluent gas.

	Percent
CO_2	4.9
Ill.	0
O_2	14.1
H_2	0.1
CO	0.1
CH_4	0.3
N_2	80.2

The regular periodic reversal of flow did, however, tend to concentrate the combustion zone near the middle of the section and thus produce a cigar-shaped, burned-out area.

Table 22 summarizes the operation of the section between boreholes I and II. It may be noted that the heat of combustion of the effluent gas reached a maximum of 34.1 per cent of the heat of combustion of the coal consumed and averaged 24.6 per cent for the entire period.

TABLE 22. - Coal consumed and heat balance on operation of section between boreholes I and II, March 1949 to June 1950

	Tons of coal burned, moisture-and ash-free basis	Tons of coal burned per calendar day, moisture- and ash-free basis	Per cent of heat of combustion of coal consumed			
			Sensible heat of dry gas	Heat of combustion of product gas	Moisture in product gas	Stored underground, by difference
1949:						
March	120	8.9	21.3	9.3	1/10.0	59.4
April	201	6.7	31.9	11.6	1/10.0	46.5
May	312	10.1	36.0	23.5	1/10.0	30.5
June	337	11.2	29.7	34.2	1/10.0	26.1
July	452	14.6	24.4	24.3	1/10.0	41.3
August	441	14.5	22.9	26.2	11.2	39.7
September	483	16.2	19.1	32.0	11.7	37.2
October	583	18.8	21.4	34.1	13.7	30.8
November	454	15.9	17.9	31.9	11.5	38.7
December	534	19.7	18.4	24.2	8.9	48.5
1950:						
January	486	15.7	7.7	22.9	12.8	56.5
February	393	14.0	9.8	18.9	14.0	57.3
March	326	10.5	10.0	14.2	15.4	59.4
April	384	12.8	9.5	16.9	16.5	57.2
May	327	10.6	10.9	20.9	19.6	52.6
June	59	11.7	4.9	16.4	14.7	59.0
Total or average	5,900	13.2	18.7	24.6	11.2	45.5

1/ Estimated.

In preparation for establishing a combustion zone between boreholes III and VII, sand fluidized with compressed air was injected into the west heading of the parallel entries between boreholes II and III. The object was to concentrate the blast in one heading and to limit the reaction to a single face in so far as was possible. Fifty tons of sand was injected before operation in the area was begun.

The section between boreholes III and VII was put into operation early in June 1950. Air at 7,200 cubic feet per minute and 12 p.s.i.g. pressure was introduced at borehole VII, and the gases were taken off at borehole III. After 8 hours, the flow was reversed, and for approximately 1 hour 2,000 cubic feet of air per minute was admitted at borehole III, and the gases were removed at VII. For the first 8 days cycles of 8 hours duration and 7,200 cubic feet per minute were maintained when blowing in the direction VII to III. When the direction of flow was III to VII, both the air-blast rate and the cycle time were increased gradually until, at the end of 8 days, the rates of flow and cycle lengths were equal in each direction. These preliminary unbalanced cycles were used to eliminate excessive temperatures in or near No. VII borehole and also to distribute the combustion zone between III and VII.

Immediately upon starting combustion in this area, sand injection on a 3-shift basis was started, utilizing 6-inch churn-drill holes along the lines of the underground passages. The purpose of this sand injection was to minimize by-passing of the coal faces by gas-making fluids and thus to insure maximum contact of the reactants. As the amount of sand injected was increased, the pressure necessary to force the air blast underground was increased. During the next several weeks a total of 836 tons of dry sand was injected into the area, including the 50 tons injected before operation of this section was started.

In the first week of operation, the effluent gas at borehole VII varied in heating value between 151 and 59 B.t.u. per cubic foot. The effluent gas at borehole III varied in heating value from 45 to 21 B.t.u. per cubic foot. At each borehole the quality decreased somewhat with time, and the average analyses follow:

	Average, percent	
	Borehole III	Borehole VII
CO_2	16.4	13.6
Ill	.2	.2
O_2	1.0	.8
H_2	2.6	13.7
CO	2.6	11.2
CH_4	1.2	1.5
N_2	76.0	59.0
Heating value, B.t.u. per cu. ft.	33	100

During this period the oxygen content of the effluent gases at either borehole was low and indicated that by-passing of carbon faces had been minimized. The coal consumption for this first week's operation averaged 45.8 tons of moisture and ash-free coal per day with an air input of 7,200 cubic feet per minute, which is approximately double what had been achieved before.

The section between boreholes III and VII was operated with cycles of 5 to 8 hours for 4 months. The operation was characterized by gradual increasing oxygen content of the effluent gas, indicating deterioration in the efficiency of contact with time. This effect was greater when the direction of flow was from III to VII. During the entire 4 months of operation the rate of coal consumption, when using an air blast of 7,200 cubic foot per day, averaged 23.6 tons per day.

Two test runs, each of 10 days' duration, were made with reduced air input in order to determine the effect of varying flow rates on the operation of the underground system. During these tests, operation was continued on a cyclic basis. The first run was made at a nominal air-input rate of 4,000 cubic feet per minute with an average cycle length of 6 hours. The second test run was made with a nominal air input of 1,700 cubic feet per minute, and the cycle time was increased to approximately 24 hours.

During these 10-day periods of operation, there was little difference in operating conditions between the normal 7,200-cubic-foot-per-minute rate and the 4,000-cubic-foot-per-minute rate; the quality of the effluent gas was comparable and did not vary greatly with time. The decrease in rate of coal consumption was approximately proportional to the decrease in air rate. At the beginning of the 1,700-cubic-foot-per-minute tests, a definite increase in quality of the effluent gases was noted over what had been obtained when using 7,200 and 4,000 cubic feet per minute. The oxygen content of the effluent gases was lower and the heating value higher. At the end of the test period, the heating value decreased to approximately the same value as was obtained at the higher rates. The rate of coal consumption per unit volume of air input was initially higher during the 1,700-cubic-feet-per-minute test, but this value decreased with time and approached that obtained at the higher rates. A study of the analytical data shows that the increase in heating value during the initial part of the 1,700-cubic-foot-per-minute run was due to an increase in content of coal-distillation products and to the water-gas reaction.

Table 23 summarizes the operation of the section between boreholes VII and III. The average rate of coal consumption and the percentage of the heat of combustion of the coal as chemical heat in the product gas are 1.7 and 1.3 times as great, respectively, as for the operation between boreholes I and II.

TABLE 23. - Coal consumed and heat balance on operation of section between boreholes III and VII, June 1950 to October 1950

1950	Tons of coal burned, moisture- and ash-free basis	Tons of coal burned per calendar day, moisture- and ash-free basis	Per cent of heat of combustion of coal consumed	
			Sensible heat of dry gas	Heat of combustion of product gas
June..................	774	31.0	12.7	32.3
July..................	665	21.5	14.5	34.8
August................	551	17.8	12.2	29.8
September.............	566	18.9	10.6	27.2
Total or average.....	2,556	22.9	12.6	31.3

Operation of the section, borehole III to VII, demonstrated again that a lean producer gas could be obtained when very hot carbon was accessible to air near the outlet borehole. However, when the same borehole was used as the air inlet, less combustible gas was produced.

When the system was operated with relatively long reversal times, the calorific value of the outlet gas gradually increased during the cycle. This increase has been noted both in section I to II and III to VII. As the heating value approached 35 B.t.u. per std. cu. ft. and the outlet-gas temperature reached 800° to 900° F., the gas ignited underground. Material balance calculations show that when ignition has occurred underground, coal consumption increased, and pressure required to force air into the system at a given rate also increased.

During operation of section III to VII an aircraft-type turbo-supercharger was installed at borehole III. It was planned to operate the underground system under pressure and to expand the hot gas issuing from borehole III through the turbine, using the energy obtained to compress air, which would be added to the primary air supply. Several preliminary trials indicated that a single turbo-supercharger would not deliver air at the required pressure. Consequently, two turbo-superchargers were installed, with the turbines operating in parallel, and their attached compressors in series. To attain the gas temperatures (1,000° to 1,400° F.) necessary for operation of this installation, air was injected underground at borehole VII, and gas was removed from borehole III for about 10 hours. About this time the gas ignited, as described above, and the turbines were started. When the pressure had built up sufficiently, the discharge of the air compressor was added to that supplied to the system. Four runs were made with this installation. It was found that an equilibrium was established when the input air pressure was about 18.0 p.s.i.g., the gas outlet pressure about 12 p.s.i.g., and the total air input underground about 9,000 std.c.f.m. A certain amount of water was injected into the gas, and its conversion into steam helped to control the gas temperature in the piping and the turbines. With the increase in air input, the oxygen content of the effluent gas decreased to zero, and the rate of coal consumption increased further. In three later runs, combustible constituents remained in the gas after it was discharged from the turbines, and a flame was visible at the turbine discharge, where the hot gas encountered the atmosphere. The combustible content of the turbine discharge increased slowly with time, but very high gas temperatures prevented prolonging the runs until there was no further increase.

Operation of the turbo-supercharger installation demonstrated that it is possible to recover a part of the energy of coal burning in place underground as usable heat. An underground gasification system designed to deliver gaseous products of combustion to the surface in the temperature range 1,500° to 2,000° F. requires relatively little underground development, will operate under a wide range of air-input rates, and will permit operation at a magnitude that may be attractive from an industrial viewpoint. Solution of the problems of containing air at high pressures and the construction of outlets to handle high-temperature gas will require further experimentation.

The operation at Gorgas has demonstrated that:

1. The combustion of coal underground can be advanced over a relatively wide area from a single passageway in the coal bed.

2. A large percentage of the coal's heat of combustion can be obtained at the surface at a usable temperature. This has been done during the operation of a small gas-turbine installation for a short period.

3. Production of a combustible gas has been obtained for short intervals, but by-passing of the carbon faces has greatly decreased the calorific value of the

effluent gas. Where air has contacted hot carbon faces, good-quality gas has been produced underground. A large percentage of this gas has been burned by secondary combustion due to air by-passing the carbon faces.

4. The major problem in the successful application of underground gasification is the by-passing of carbon faces. By-passing is affected by the action of heat and pressure on the rock overlying the coal bed. A partial control of by-passing has been achieved by filling the openings underground with fluidized sand.

APPENDIX. - BIBLIOGRAPHY OF PAPERS AND REPORTS PRESENTED AND PUBLISHED DURING 1950

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