TABLE 5. - Summary of data from runs made with various steam-coal ratios

Run No.	Number of burners	Raw-coal rate, lb. per hr.	Oxygen-coal ratio	Steam-coal ratio	Steam temp., °F.	Carbon conversion, percent
21A	4	1,340	9.8	1.43	1,695	81.5
21B	4	1,340	9.7	1.42	1,695	84.1
24	4	1,310	10.4	1.84	2,117	87.2
33A	4	1,500	8.9	.83	2,045	84.9
33B	4	1,500	8.9	.83	2,000	78.5

Run No.	Number of burners	Coal residence, sec.	Lb. coal per M std.c.f. CO + H_2	Std.c.f. O_2 per M std.c.f. CO + H_2	Economic factor	Average cone temp., °F.
21A	4	3.9	39.6	388	78.4	2,350
21B	4	3.8	38.2	372	75.4	2,450
24	4	3.2	36.3	378	74.1	2,540
33A	4	4.0	37.6	382	75.8	2,460
33B	4	4.2	39.6	351	74.7	2,350

The results obtained from these runs are somewhat poorer than the base date, although the differences are not enough to be truly significant. An analysis of the effect of the higher steam-coal ratios would be quite difficult, since not only the mass of steam, as affecting gasification, but also the residence time and thermal picture are changed. These effects are treated theoretically and in considerable detail in the papers on coal-gasification kinetics, by Busche, Batchelder, and Armstrong.8/

The principal conclusion to be drawn from these data is that the variations in operating conditions as noted have little effect on the gasification results. Thus, it was concluded that within the limits of the equipment available the prime variable affecting gasification performance of the Koppers gasifier was the oxygen-coal ratio.

Data and Calculations for a Typical Run

Method of operation and data and sample collection during the runs have been described. Data from run 18 have been selected for the summary below. In many respects this run was considered typical for this gasifier. The conditions of operation are well within the limits for the base runs, and the results check closely those from several other runs (17, 23B, and others).

All gas flows were measured by orifice differential meters and corrected to standard conditions (60° F., 30 in. Hg, dry).

Each hopper of coal used was weighed and sampled for moisture and mesh analysis. The analysis of the coal from the car used in the test was taken. If these overlapped during the run, a sample from the run was composited and analyzed separately.

8/ Busche, R. M., Batchelder, H. R., and Armstrong, W. P., Engineering Aspects of the Kinetics of Coal Gasification: Ind. Eng. Chem. Vol. 45, No. 9, Sept. 1953, pp. 1856-1878.

Typical Data for Koppers Gasifier

Date: August 29, 1949
Run 18

Hours duration		3.0
Number burners operated		6
Raw-coal rate	lb./hr.	1,976
Process O_2 rate	std.c.f./hr.	19,690
Steam rate	lb./hr.	1,800
Oxygen-dry coal ratio	std.c.f./lb.	9.8
Steam inlet temperature	°F.	1/1,700
Synthesis gas outlet temperature	°F.	2,295
Synthesis gas analyses, percent:		
CO_2		15.6
H_2		37.8
CO		42.9
C_1+		0.1
Synthesis gas	std.c.f./hr.	63,800
H_2 + CO	std.c.f./hr.	51,490
Total carbon gasified	percent	83.6
B.t.u. efficiency (heating value of gas/heating value of coal)	percent	63.8
Ratios per lb. raw coal:		
100% O_2	std.c.f.	9.5
Steam introduced	lb.	0.9
Steam consumed	lb.	0.1
Synthesis gas	std.c.f.	32.3
H_2 + CO	std.c.f.	26.0
Ratios per std.c.f. 100% O_2		
Synthesis gas	std.c.f.	3.4
H_2 + CO	std.c.f.	2.7
Ratios per M std.c.f. H_2 + CO		
Dry coal	lb.	37.2
100% O_2,	std.c.f.	364.0
Steam introduced	lb.	35.0
Steam consumed	lb.	2.3
H_2-CO ratio		0.9
Gasifier cone temperatures, °F.:		
North cone		2,620
South cone		2,630

1/ Estimated from steam temperature from pebble heater and normal temperature drop in line to gasifier. Thermocouple in steam line at gasifier was out of order in this run.

Typical Data for Koppers Gasifier (Cont.)

Gasifier outlet temperature..... °F.	2,295
Pebble-heater temperatures, °F.:	
North combustion chamber.....	1,845
South combustion chamber.....	1,970
Steam outlet.....	1,795
Stack temperature, pebble heater..... °F.	430
Temperature of pebbles to feeder..... °F.	300
Synthesis gas analyses:	
Number of samples.....	6
Mass spectrometer analyses, percent:	
CO_2.....	15.6
H_2.....	37.8
CO.....	42.9
CH_4 and higher.....	0.1
N_2.....	3.3
A.....	0.1
O_2.....	0.1
H_2S.....	0.1
H_2 + CO.....	80.7
Gross heating value.....B.t.u./cu.ft.	262
Synthesis gas analyses:	
Number of samples.....	3
Orsat analyses, percent:	
CO_2.....	16.6
CO.....	41.5
O_2.....	0.0
Carbon in ash.....percent	66.5
Oxygen flow:	
Orifice meter.....c.f./hr.	19,560
Purity.....percent	95.2
Saturation temperature.....°F.	92
Flow temperature.....°F.	92
Pressure.....p.s.i.g.	7.2
Gas gravity.....	1.01
Moisture.....percent.	3.4
Flow correction to dry-process O_2.....	1.042
Process O_2 flow.....std.c.f./hr., dry	19,690
100% O_2 flow.....std.c.f./hr., dry	18,740
Synthesis gas flow:	
Orifice meter.....c.f./hr.	67,300
Saturation temperature.....°F.	87
Flow temperature.....°F.	87
Pressure.....in. H_2O	12.8
Gas gravity (referred to air).....	0.712
Moisture.....percent	4.19
Flow correction to dry gas.....	0.948
Flow of DRY synthesis gas.....std.c.f./hr.	63,800

Typical Data for Koppers Gasifier (Cont.)

Steam flow: Orifice meter	lb./hr.	1,800
Carbon conversion:		
Total carbon in coal	lb./hr.	1,416
Carbon in synthesis gas	lb./hr.	1,184
Unconsumed carbon	lb./hr.	232
Total carbon gasified	percent	83.6
Hydrogen from steam, lb./hr.:		
H_2 in dry synthesis gas		128
H_2 from coal		99
H_2 from reacted steam		29
Oxygen from steam, lb./hr.:		
O_2 in dry synthesis gas		2,010
O_2 from coal		241
O_2 in process O_2		1,580
O_2 from reacted steam		189
Steam consumption, lb./hr.:		
Reacted steam (hydrogen figures)		261
Reacted steam (oxygen figures)		214
Average reacted steam		238
Total steam introduced		1,900
Unreacted steam		1,662
Thermal data:		
Heat loss from gasifier	million B.t.u./hr.	1.00
B.t.u. efficiency (heating value of gas/heating value of coal net basis)	percent	63.8
Heat loss	percent of input	3.9
Thermal balance temperature	°F.	2,515
Pressures, in. H_2O:		
In gasifier		0
Out of washer-cooler		-1.0
In pulverized-coal bins		11.0
O_2 booster pressure	p.s.i.g.	7.2
O_2 north mixing header	p.s.i.g.	3.5
O_2 south mixing header	p.s.i.g.	3.0
In steam section, pebble heater	p.s.i.g.	1.4
In fired section, pebble heater	p.s.i.g.	1.4
Differential across neck, pebble heater	in. H_2O	+0.7
Coal analysis:		
Moisture	percent	3.4
Ultimate, dry basis, percent:		
H_2		5.2
C		74.2
N_2		1.7

Coal analysis (con.):

Ultimate, dry basis, percent (con.):	
O_2	12.6
S	0.9
Ash	5.4
Net heating value, B.t.u. per lb.	12,728
Mesh analysis, percent passing:	
60-mesh	99.5
100-mesh	97.9
200-mesh	87.5
300-mesh	73.0

The total carbon gasified is calculated from the coal weights and analyses and make gas and its analyses, thus:

	Lb./hr.
Carbon in coal (1,976) (0.966) (0.742)	1,416
Carbon in gas:	
CO (63,800) (0.429) $(\frac{12}{379})$	867
CO_2 (63,800) (0.156) $(\frac{12}{379})$	315
CH_4 (63,800) (0.001) $(\frac{12}{379})$	2
Total carbon in gas	1,184
Carbon conversion (1,184/1,416 x 100)	83.6 percent

This is checked roughly by determining the amount of carbon in the ash and weighing the slag from the run. For the 6-hour operating period of the run ash in the coal amounted to 620 pounds.

Although the weight of the slag from this run was not obtained, an overall figure of 9 percent of the weight of the ash in the coal charged was obtained in runs 19 through 40. Using this, the weight of slag would be 620 x 0.09 = 56 pounds. Slags were analyzed and never found to contain more than a trace of carbon. Then the weight of ash that would end up in the settling basins would be 620 - 56 = 564 pounds. Dust analyses in the scrubbed gas showed only a few grains per 100 cubic feet; nor was there ever any evidence of tar or organic condensate from the "make gas." Thus, there was no carbon to account for in these materials. The carbon content of the ash was 66.5 percent, from which the total weight of ash (plus carbon content) was 564 ÷ 0.335 = 1,680 pounds, with a net carbon content of 1,116 pounds. This would be 186 pounds carbon per hour. In all, 1,416 pounds per hour of carbon was fed with the coal. Thus, the conversion was $\frac{1,416-186}{1,416}$ = 87 percent. This is about 3 percent higher than the figure obtained from gas production.

Steam reacted was calculated from the difference between the hydrogen in the gas and the hydrogen in the coal and also from the difference between oxygen in the gas and oxygen fed to the gasifier, both as process oxygen and oxygen in the coal. The agreement between the two steam-decomposition figures so calculated was taken as a measure of the reliability of the data, because this comparsion is quite sensitive to small errors.

Steam reacted (hydrogen figures):		Lb./hr.
Hydrogen in dry make gas (63,800) (0.378) $(\frac{2}{379})$	=	128
Hydrogen from coal (1,976) (0.966) (0.052)	=	99
Hydrogen from steam (by difference)	=	29
Reacted-steam equivalent (9 x 29)	=	261

Steam reacted (oxygen figures):

	Lb./hr.
Oxygen in dry make gas (63,800) (0.372) $(\frac{32}{379})$	2,010
Oxygen from coal (1,976) (0.966) (0.126)	241
Oxygen in process oxygen (18,740) $(\frac{32}{379})$	1,580
Oxygen from coal and process oxygen	1,821
Oxygen from steam (by difference)	189
Reacted-steam equivalent (9/8 x 189)	213
Discrepancy = $\frac{241 - 213}{241}$ x 100	11-1/2 percent

Thermal calculations were made for each run. In most instances a true "balance" was not made. Because of the uncertainties of measurement of gas temperatures, the procedure outlined below was felt to be more satisfactory.

The calculations for run 18 are:

Heat in: (all values on net basis, above 60° F. base temperature.):

	M B.t.u./hr.
Heating value of coal (1,976) (0.966) (12,728)	24,298
Less heat vaporization of H_2O in coal (1,976) (0.034) (1,060)	69
Heating value of coal as fed	24,229
Sensible heat in process steam at 1,700° F. $(\frac{1800}{18})$ (14,969)	1,497
Total heat in	25,726

Heat out:

	M. B.t.u./hr.
Heating value of make gas:	
H_2 (63,800) (0.378) (275)	6,630
CO (63,800) (0.429) (321)	8,780
CH_4 (63,800) (0.001) (914)	55
H_2S (63,800) (0.001) (596)	36
Total heat out	15,501
Heating value of unburned carbon (1,416 - 1,184) (14,100)	3,270
Sensible heat in products from gasifier at "thermal balance" temperature of 2,515° F.	5,955
Shell and miscellaneous heat losses from gasifier	1,000
Total heat out	25,726

A separate study was made to determine shell and miscellaneous heat losses from the gasifier at different internal temperatures. This is covered in detail in a following section, Miscellaneous Auxiliary Studies - Gasifier Heat Losses. The heat loss used in the calculations is based on this study.

The "thermal balance" temperature is obtained by trial and error. It is the value, which, when applied to obtain the sensible heat in the products from the gasifier and the shell and miscellaneous heat losses, will bring the heat output in balance with the heat input. This is considered to be the true equilibrium internal temperature of the gasifier in operation.

A true heat balance is not obtained in this calculation. However, a "thermal balance" temperature out of line with the observed gasifier temperatures indicated error in the data. After a number of runs had been calculated in this manner, the approximate values for each item were fairly well established, and any real discrepancy was also taken as an indication of possible error in the data.

In run 18 the thermal balance temperature of 2,515° F. lies between the average cone temperature of 2,625° F. and the exit gas temperature of 2,295° F. The cones were in a zone of radiant-heat reception from the oxygen-coal flame and undoubtedly were hotter than the average temperature in the gasifier. Numerous calculations indicated that the thermocouple for the exit gas temperature read at least 100° F. low. Thus, the 2,515° F. was a reasonable value for the equilibrium temperature in the gasifier.

Extended Gasification Run 43

This run was made primarily to test the operability of the gasifier and its auxiliaries in an extended period of operation. It was intended to run with 21,000 to 22,000 standard cubic feet per hour of oxygen (98-1/2-percent purity) this being virtually the full capacity of the oxygen plant and adjust the coal rates to obtain an oxygen-coal ratio selected to obtain the maximum safe coal conversion without high-temperature damage to the lining.

The resultant coal rate was close to 2,650 pounds per hour (4-percent moisture content). The oxygen-coal ratio averaged somewhat above 8 cubic feet of oxygen per pound of coal. The steam rate was also adjusted within a limited range for control of the gasifier temperatures. Actually 1,750 to 2,000 pounds per hour was used, with the average ratio of about 3/4 pound of steam per pound of coal. As in previous runs, the coal and oxygen were fed at atmospheric temperatures. The oxygen was diluted to around 95 percent. The average steam inlet temperature was 1,600° F.

The run was started at 3 p.m. on April 3, 1950, and terminated, as planned, at 4 p.m. on April 14, following. Of the total elapsed time of 265 hours, there were 104 hours of operation with 6 burners, 97 hours with 5 burners, and about 24 hours with 4 burners or less. Down time totaled 40 hours.

Shutdowns for inspection were scheduled twice daily. These normally required about an hour. However, on several occasions this time was extended to make repairs. There were also a few occasions of enforced shutdowns for repairs. The nature of these repairs varied. None were major, and probably none were of a type that could not have been remedied while the gasifer was operating.

The 4-inch projection burner nozzles and the modified steam rings were used. Rock Springs, Wyo., coal, ground to an average of 88 percent through 200-mesh, was used throughout.

For the first time the waste-heat boiler, operating on hot gas from the gasifier, was put into service. Its performance was satisfactory. Soot was blown periodically, and no pressure buildup was noted. The tubes were slag free at the end of the run. Gas inlet temperatures averaged 1,500° F. During the extended operating period some 250 tons of coal, 4,100,000 cubic feet of oxygen, and 400,000 pounds of steam were used. Gas production amounted to 15,700,000 cubic feet, of which 12,700,000 cubic feet was CO + H_2. On this basis, the coal requirement per 1,000 cubic feet of CO + H_2 would be 39.4 pounds and oxygen would be 323 cubic feet. These figures give an economic factor of around 72.

Some difficulty in accounting for the coal usage developed in the latter part of the run. For this reason, the above economic factor may be a bit optimistic, and detailed data from the run are not considered pertinent. Data from a few selected portions of the run are presented in the Appendix. Additional average data from the run for 6-burner operation are presented below. In general, these data follow the pattern obtained from the base runs, as noted in previous sections.

Gas make, std.c.f./hr. = 80,000

Mass spectrometer analysis, percent:

H_2	39
CO	41.3
CO_2	14
N_2	4.9
CH_4	.3
A	.2
O_2	.2

H_2S (Tutweiler), gr./100 cu. ft.	125
Organic sulfur, gr./100 cu. ft.	8
Carbon conversion, percent	75
Lb. steam decomposed per lb. coal	0.12
Average cone temperature, °F.	2,450
"Thermal balance" temperature, °F.	2,375
Heat loss per lb. coal, B.t.u.	350
B.t.u. efficiency, 1/ percent	60
Pressure in gasifier, in. H_2O	-0.5

Ultimate coal analysis, percent, dry-weight basis:

C	74.6
H_2	5.2
O_2	12.2
N_2	1.6
S	1.2
Ash	5.2

1/ $\frac{\text{Net heating value of product gas}}{\text{Net heating value of coal}} \times 100.$

A brief analysis of the shutdowns reveals that the preponderance of trouble was due to the coal-feed system. These failures were mostly mechanical in nature and very well could have been remedied for routine operation.

Failure of the target tubes used for the optical pyrometers caused a few shutdowns. It was considered inadvisable to run without the temperature guide, but this would probably not be necessary on routine runs. Nevertheless, the short life of the target tubes is a problem that needs further attention.

Toward the end of the run difficulties developed with slag plugs in the ash legs. Most of these could be rodded out through sight ports in the top of the gasifier after shutting down the gasifier. However, on one occasion a piece had to be cut out of the side wall of the north ash leg. The leg was then rodded up through this to effect clearance. A more satisfactory slag-drainage system would be desirable for routine operation.