

THE GASIFICATION OF COAL IN A SLAGGING PRESSURE GASIFIER

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INTRODUCTION

In 1955 The Gas Council, Midlands Research Station at Solihull, began to be actively concerned in the development of a fixed bed slagging gasifier as a possible means for providing the gas industry with large gasification units to gasify a wide range of solid fuels, at high efficiency and low cost. A lean gas with a low carbon monoxide content, to be enriched with hydrocarbons to a calorific value of 500 Btu/ft³, was required. At this time the Lurgi process was well established in many parts of the world, but it used an excess of steam in the steam oxygen mixture supplied to the gasifier to prevent clinker formation in the fuel bed. As a result it was desirable to use coal of a high ash fusion temperature, and the comparatively small steam decomposition in the short contact time in the hot gasification zone limited the throughput of the gasifier and produced large volumes of phenolic liquor with its attendant disposal problem. It appeared then that better performance could be achieved by increasing the hot zone temperature of the gasifier and operating under slagging conditions.

The essence of slagging gasification is that the steam supplied per unit volume of oxygen is only that required for gasification. Under these conditions temperatures some hundreds of degrees centigrade above the fusion point of the ash are generated at the steam oxygen inlet, and the ash fuses to a liquid slag. Certain advantages follow from this, including high thermal efficiency, high throughput, a choice of fuel unrestricted by low ash fusion temperature or reactivity, and the absence of a mechanical grate. Gasification under slagging conditions does, however, introduce the hazards of high temperatures at elevated pressures.

The first slagging gasifier erected at Solihull was operated on coke at 5 atm. pressure and was used to make an assessment of the material requirements for gasification, and to investigate methods of controlling slag flow and removing slag from a pressure system⁽¹⁾. A feature of this gasifier was a flat-bottomed hearth with a side slag outlet consisting of a ¼ inch silicon carbide tube 30 inch long. Frequent blockages of the slag outlet were experienced due to the high heat loss from the slag, until an intermittent system of tapping was developed. Slag was allowed to accumulate in the hearth by directing up the tap tube hot combustion gases from a tunnel burner, and was then run off at a high rate by deflecting the burner and applying a differential pressure across the hearth. It was considered that this system of operation justified further development, but the side offtake was not satisfactory. The gasifier was therefore redesigned to take a new hearth with a short water-cooled slag tap at the centre, and was rebuilt together with the necessary auxiliary plant to provide for operation on bituminous coal at 25 atm. pressure and a gas output of 5 mill. ft³/day⁽²⁾.

DESCRIPTION OF PLANT.

Details of the gasifier and the gas cooling plant are shown in Fig. I. The fuel bed, 3 ft. diameter by 10 ft. deep, was contained within a refractory lined pressure vessel, at the top of which was mounted a water-cooled stirrer to break up any agglomerations in the fuel bed. Coal premixed with a suitable flux entered the gasifier through the fuel lock hopper, flowed by gravity through the stirrer unit, and was distributed over the top of the fuel bed as the stirrer rotated. The fuel capacity within the stirrer unit was just sufficient, when operating at the designed load, to ensure that the feed was maintained to the fuel bed during the period that the lock hopper was being recharged. Coal moved down the gasifier as it was continuously gasified by the steam oxygen mixture that was injected into

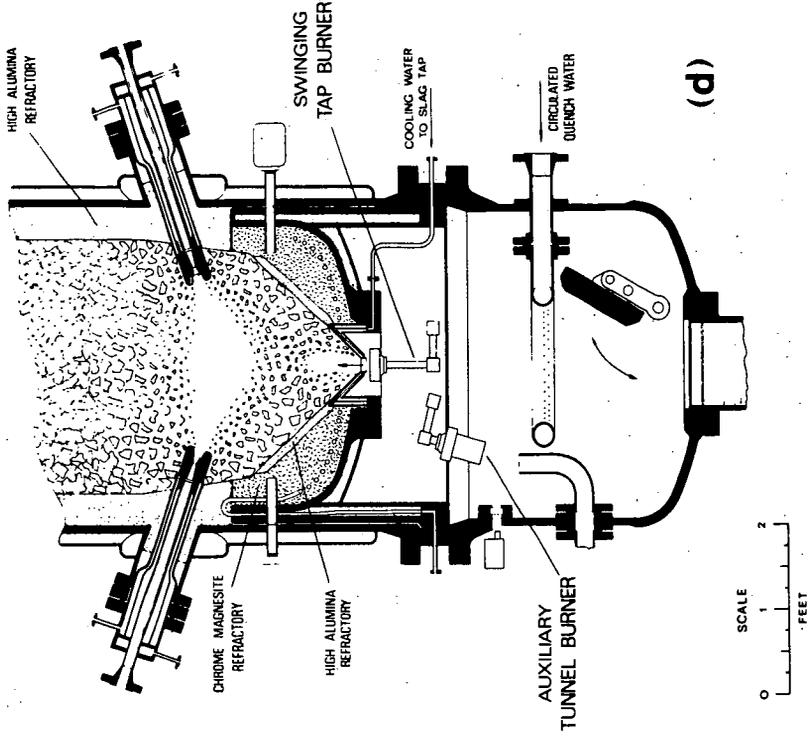


FIG. 2
Cross-Section of the Hearth and Quench Chamber

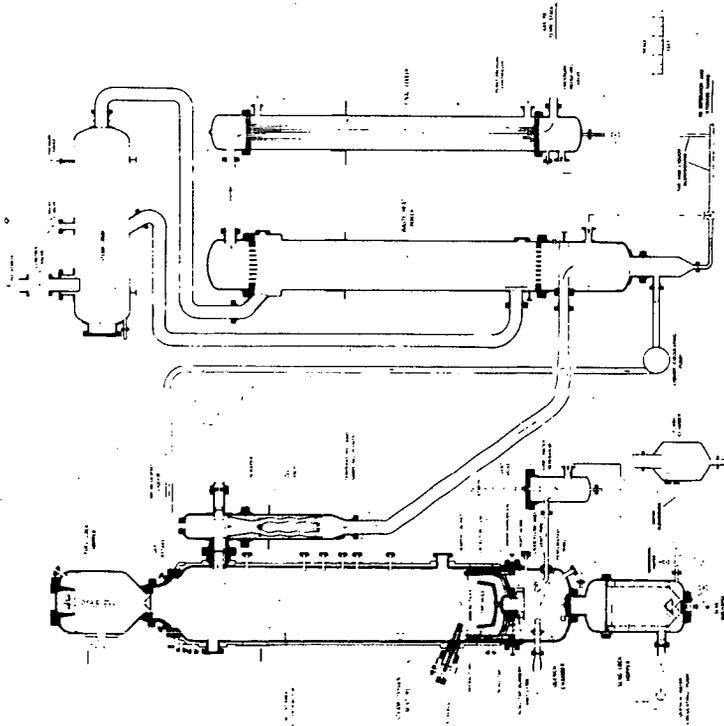


FIG. 1
The Slagging Gasifier and Gas Cooling Plant

the bottom of the fuel bed through four water-cooled tuyeres. In front of the tuyeres the temperature generated by the combustion of the fuel with the oxygen melted the ash which drained into the hearth below. The hot zone was confined to the centre of the gasifier away from the refractory walls by projecting the tuyeres 6 inch into the fuel bed, and using a blast velocity of about 200 ft/sec.

The hearth of the gasifier was contained within a water jacket supported from a carrier ring fitted between the main flanges of the gasifier and the quench chamber. Slag was discharged vertically downwards into water in the quench chamber where it formed a glassy black frit that settled in the slag lock hopper. Water was circulated at a high rate, from the slag hopper through a cooler and back into the quench chamber via a perforated ring, to create highly turbulent conditions that avoided stratification and helped to quench and break up the slag.

The product gases left the gasifier through an offtake fitted with a scraper that kept it free from deposits of tar and dust, and were quenched by liquor recirculated from the base of the waste heat boiler. After leaving the waste heat boiler the gases passed through a final cooler to a reducing valve that controlled the plant pressure. Tar and liquor condensate from the gas was blown down through valves controlled by the level of liquid in the sumps of the waste heat boiler and final cooler, and passed into a large storage tank where separation of the two phases occurred.

The cooled gases from the plant were burnt at ground level within an acoustically lined enclosure at the base of a 120 ft. high chimney stack.

DEVELOPMENT OF THE SLAG TAP.

In small slagging gasifiers the heat capacity of the slag stream even when the ash content of the fuel is increased artificially by fluxing is low in proportion to the high rate of heat loss, and problems with viscous slag, solidifying slag, and stalaicite formation are common when continuous tapping is attempted. To avoid these difficulties the tapping system for the experimental gasifier was based on intermittent tapping from a reservoir of slag contained in the hearth. This gave a homogeneous slag with consistent flow properties and enabled gasifying media to be introduced through the tap hole in an attempt to control the slag temperature independently of the main gasification reactions, but the increased slag residence time in the hearth encouraged the separation of liquid iron.

The fuels used in the slagging gasifier had a high iron content. In fact it was the presence of the iron compounds that gave the ash the low melting point and good flow properties for which the fuels were originally chosen. The strongly reducing conditions in the hearth inevitably led to the formation of iron, although the degree of reduction was greatly reduced by introducing oxygen through the tap hole. In the absence of a suitable refractory it was necessary to develop a water cooled metal slag tap that was compatible with the slag under the conditions existing in the hearth and that would also withstand liquid iron, which at temperatures above 1500°C has a very high heat capacity. To preserve the slag tap it was necessary to have a very high heat transfer rate to the cooling water to form a protective layer of solidified slag and to freeze any iron before it damaged the slag tap. It was therefore desirable to construct the slag tap with metal having a high thermal conductivity, use high cooling water rates, and avoid any accumulations of iron in the hearth. Originally austenitic stainless steels were used for the slag tap because of the type of slag tap burner employed, but later carbon steel was found to be satisfactory.

As a consequence of iron attack the hearth, slag tap and slag tap burners went through several stages of development, and these have been described in details by Hebden⁽²⁾ et al. The hearth depicted in Fig. 1 was abandoned at an early stage. It was unsatisfactory because of the erosion of the carbon tiles in the vicinity of the tap hole by the slag tap gases, and its shape permitted iron to collect. The slag tap withstood small quantities of iron, but it was found that masses of

10 to 30 lb. iron accumulated and destroyed the slag tap as they flowed uncontrollably from the hearth. Experience indicated that the hearth should slope steeply towards the tap hole with the minimum opportunity for iron to collect at any point. Thus, the hearth and slag tap were made funnel shaped, and the slag tap was tapped as frequently as possible to limit the quantity of iron that could collect.

The hearth and slag tap that proved reliable for more than 100 hours operation, and which were used for the performance tests, are shown in Fig. II. The floor of the refractory hearth sloped downwards at 45° to a carbon steel slag tap assembly, at the centre of which was the slag tap tube $1\frac{1}{2}$ inch diameter by 2 inches long. Very little erosion of the refractory hearth occurred for it was found that the high temperature gases and slag were confined to a zone about 20 inches diameter in the centre of the gasifier, and the refractory walls at the side were protected by unburnt fuel. Below the slag tap two swinging burners were mounted. The tap burner controlled the slag tapping and was used to introduce air, oxygen and town gas into the slag tap tube, where combustion occurred, giving a linear gas velocity sufficiently high to hold back the slag in the hearth. The auxiliary burner which had a refractory tunnel was normally in the retracted position and was installed to clear the tap hole in the event of a blockage a purpose for which the tap burner was unsuited. It was necessary to use this burner only on very rare occasions.

The slag tap shown in Fig. II was not completely immune from iron attack, which when it did occur, was invariably confined to an area at the outer edge of the tap cone entrance that was most difficult to cool effectively. To give this vulnerable area added protection water-cooled coils of copper tubing were installed in the hearth. The high heat transfer rates possible with the copper tubing enabled it to withstand large amounts of iron without suffering any damage. It is considered that copper is a very promising material for the construction of the slag tap, and the slag tap has been redesigned to enable it to be fabricated from this material.

SLAG TAPPING.

The intermittent slag tapping system proved to be so easy to operate and trouble free that it was possible to adopt an automatic tapping system. Initially the tapping was controlled on a time basis, but later it was controlled by the slag level in the hearth. With the slag tap burner directing the hot gases up through the tap hole the slag level in the hearth increased. On reaching the level of a collimated beam of gamma rays from the level detector, about 12 inches above the tap hole, a sequence of operations was started to tap the slag. The tap burner was swung away from the tap hole, and on reaching its fully retracted position a control valve on a vent line from the quench chamber was opened to reduce the pressure in the quench chamber below that of the gasifier. A controlled differential pressure across the hearth was maintained for a pre-set time period to force the slag to flow from the hearth, and the control valve was then closed. As the pressure in the slag quench chamber built up, due to the gases supplied to the burner, the slag flow stopped, and the tap burner was swung back to the tap hole. In a typical tapping cycle the slag was tapped for 20 seconds every 4 to 6 minutes.

During the development work on coke the slag was tapped at differential pressures of 5 to 15 lb/in.². When coal was tested it was found that at these differential pressures fuel was entrained in the slag stream. The differential tapping pressures were gradually reduced and it was possible to drain slag from the hearth at a rate of more than 10,000 lb/hr. with a pull of 1 lb/in.². At these low differentials carbon could not be detected in the slag, and the temperature of the slag stream was more uniform than at the higher differential pressures.

While the arrangement of swinging burners worked well on the pilot plant mechanical devices of this type would not be a practical proposition in a large plant. They could be replaced by a fixed burner arrangement mounted below the slag tap which would allow iron to pass out through the tap hole at any time while the

slag was retained and intermittently tapped to maintain a constant level, or reservoir, in the hearth.

GASIFIER PERFORMANCE

The first phase of the experimental programme was to develop the hearth design and the slag tapping system, and for this purpose a free flowing slag was required. Avenue coke was selected as a suitable fuel on the basis of its silica ratio, a parameter of chemical composition that gives a good correlation between the viscosity and temperature of slags⁽³⁾. The coke was mixed with blast furnace slag in the proportion 2 parts slag to 1 part ash to increase the volume of slag to be tapped. This simplified plant operation and avoided any complications that might arise with coal, so that attention could be concentrated on the hearth and slag tap. When the design of the hearth and slag tap had progressed sufficiently to enable the plant to be operated continuously for 4 days the changeover was made from coke to coal.

The first experimental runs with coal were made with Donisthorpe washed doubles, a weakly caking coal that was known to behave well in slagging boilers. It was mixed with blast furnace slag to increase its ash content and was successfully gasified at 300 p.s.i. and an oxygen rate of 40,000 ft³/hr. Using the same fuel the gasifier was next operated at its designed gas output of 5 mill. ft³/day for a test period of 81 hours to obtain data from which the performance of the gasifier could be accurately assessed.

The next run was to attempt to gasify a smaller and dirtier coal with a high proportion of adventitious ash, i.e., untreated singles with an ash content of 11%. With the higher ash content it was no longer necessary to ballast the fuel, but to preserve approximately the same silica ratio a flux of dolomite was added in the ratio 0.3 lb/lb. ash. The gasifier worked well on the dirtier coal without any difficulty with the fuel bed, or with tapping the slag.

It was then considered of interest to establish the limit of the gasifier output, and a run was made at an output of 7½ mill. ft³/day, i.e., 50% above the designed output. The gasifier operated smoothly and slag tapping was consistent, but as the run progressed difficulties were experienced with the gas cooling system and run was terminated after 10 hours. The factor that limited the output was the lock hopping of the fuel. At this output the distributor was empty before recharging and repressurising of the lock hopper could be completed. This resulted in a widely fluctuating gas outlet temperature which was normally fairly steady at 350 to 400°C, it now fluctuated between 200 and 800°C. A greater capacity in the distributor and another lock hopper would be required before any material increase in output could be attempted.

The final run made with the slagging gasifier was to test its ability to handle a coal with a high fusion point, and a coal in the highest range of silica ratios was selected from Newstead Colliery in the East Midlands coalfield. The ash content of 6% was fluxed by the addition of dolomite to reduce the silica ratio to 65 and, in a run of 10 hours, no difficulty was experienced in tapping the slag which was in fact remarkable for its free flowing character. The indications were that a lower flux:ash ratio or steam:oxygen ratio could have been employed.

a) Fuel.

The details of the fuel and the chemical composition of the fuel ash and flux used in the performance tests are given in Tables 1 and 2.

TABLE 1. Fuel Analysis

Test No.	54	67	70	71	72
<u>Fuel</u>	Avenue No.2 Coke	Donisthorpe D.S. Nuts	Donisthorpe Untreated Singles	Donisthorpe D.S. Nuts	Newstead Doubles
Nominal size (in.)	1½ - 1	1½ - 1	1 - ½	1½ - 1	2 - 1
Rank	-	902	902	902	802
<u>Ultimate Analysis</u>					
Carbon wt. %	88.0	71.3	68.7	74.6	73.8
Hydrogen "	0.75	5.0	4.75	5.0	4.7
Nitrogen "	1.05	1.55	1.5	1.55	1.55
Sulphur "	1.15	1.45	1.8	1.3	0.7
Chlorine "	0.05	0.15	0.05	0.2	0.3
Oxygen, errors, etc.	1.45	13.2	11.75	11.7	11.15
Ash "	7.55	7.35	11.45	5.65	7.6
	<u>100.0</u>	<u>100.0</u>	<u>100.0</u>	<u>100.0</u>	<u>100.0</u>
Moisture as charged	9.45	12.7	15.1	13.8	12.6

TABLE 2. Ash and Flux Compositions

Composition wt. %	Avenue No. 2 Coke	Donisthorpe D.S. Nuts	Donisthorpe Untreated Singles	Newstead Doubles	Blast Furnace Slag	Dolomite
Al ₂ O ₃	26.81	28.50	27.52	32.40	20.6	2.18
SiO ₂	44.18	38.60	44.53	47.0	33.40	0.86
Fe ₂ O ₃	15.24	15.60	15.49	4.52	1.63	1.07
TiO ₂	1.03	1.22	1.17	0.95	0.73	0.04
Mn ₃ O ₄	0.17	0.22	0.23	0.20	1.26	0.18
P ₂ O ₅	0.45	0.49	0.32	1.25	0.12	0.17
Na ₂ O	2.36	1.09	0.55	2.14	0.53	0.08
K ₂ O	2.11	1.38	2.17	0.80	1.26	0.12
CaO	5.14	5.71	4.75	5.95	33.10	30.2
MgO	1.05	2.06	2.17	1.70	6.15	21.6
SO ₃	1.46	6.21	2.59	3.49	1.0	0.01
Ash Properties						
Silica Ratio*	67	62	69	80	46	-
Fusion Temps., °C						
Initial Reformation	1110	-	1165	1235	1230	-
Hemisphere Point	1215	1250	1350	1355	1250	-
Flow Point	1330	1345	1470	1545	1320	-

* Silica Ratio =
$$\frac{\text{SiO}_2 \times 100}{\text{SiO}_2 + \text{Equiv. Fe}_2\text{O}_3 + \text{CaO} + \text{MgO}}$$

b) Performance Data.

The performance data obtained during one test on coke and four tests on coal are summarised in Table 3 and cover changes in fuel, fuel size, gasifier output and ash fusion properties. The fuel quantities are expressed as dry and ash free and all product gas volumes are given on the nitrogen free basis to simplify comparison. The greater part of the nitrogen content of the gas produced originated from the air supplied to the slag tap burner and varied between 5 and 7 per cent depending on the operation of the slag tap and the gasifier output. To obtain a product gas with a low nitrogen content suitable for synthesis gas the slag tap burner could be run on a mixture of make gas, oxygen and steam.

The steam oxygen ratio was maintained almost constant. A ratio of 1.1 vol./vol. gave satisfactory slagging conditions on coal, being slightly reduced when operating with the higher melting point ash (Test 72). Similarly, the rates of supply of gas, air and oxygen were not varied by significant amounts.

The gas compositions are characterised by low carbon dioxide and high carbon monoxide contents. It is interesting to note that in the high output run (Test 71) the calorific value of the gas was unaffected. The organic sulphur content was rather higher than expected; Lurgi gas usually contains less than 10 grains/100 ft.³.

During the tests on coal the rate of slag flow through the tap hole, calculated from the slag tapping times, was between 10,000 and 14,000 bl./hr. whereas the rates were as high as 30,000 lb./hr. with coke. This was not due to any difference in the rheological properties of the slags, but to the use of lower differential pressures across the hearth when tapping coal slag.

The absence of volatile matter and the lower reactivity result in coke having a higher oxygen consumption than coal. The variation in the oxygen consumption figures for coal are hardly significant, but they reflect the increase which can be expected with higher ash contents and the use of a flux. The specific throughput figures are noteworthy in that for these conditions a gasification rate of 1,000-1,500 lb. d.a.f. coal/hr. ft.² would appear to be readily attainable without any undue sacrifice in performance. Gas output from the fuel bed is extremely high and is equivalent to more than 30,000 to 42,000 ft.³/hr. per ft.² of fuel bed.

c) Mass and Heat Balances.

Due to the scale of operations it was difficult to accurately measure the make gas, and a carbon balance was used to obtain this major item. A typical mass balance is given in Table 4. The hydrogen and oxygen items show differences which are relatively small and can be attributed to the loss of flash steam during the blowing down of liquor from the waste heat boiler.

A heat balance for the same test is given in Table 5. It shows that 1.5 per cent of the total heat supplied is lost as high grade heat from the hearth 0.3 per cent being in the slag stream and 0.7 per cent being in the cooling water from the tuyeres and slag tap. The heat lost from the jacket and stirrer amounts to a similar quantity but in a commercial gasifier this would be recovered as gasification steam. Of the total heat supplied 80.3 per cent appears as potential heat in gas and 90.8 per cent as potential heat in gas plus by-products.

TABLE 3. Performance data obtained on the slagging gasifier operated at 300 lb/in.²
(Fuel figures are on dry and ash-free basis)

Test No.	54	67	70	71	72
Duration, (hr.)	54	81	30	12	14
<u>Supplied</u>					
Fuel	Avenue Coke	Donisthorpe D.S. Nuts	Donisthorpe Untreated Singles	Donisthorpe D.S. Nuts	Newstead Singles
Size Feed rate (in.) (lb./hr.)	1½ - 1 3,640	1½ - 1 6,930	1 - ½ 6,520	1½ - 1 10,150	2 - 1 6,560
Flux	Blast Furnace	Blast Furnace	Dolomite	-	Dolomite
Flux:ash ratio (lb./lb.)	Slag 0.83	Slag 0.75	0.3	0.0	0.32
Steam:oxygen ratio (vol./vol.)	1.17	1.10	1.10	1.10	1.06
Steam rate (lb./hr.)	1,650	2,038	2,040	2,965	2,045
Oxygen rate (ft. ³ /hr.)	30,100	39,650	39,640	57,615	41,240
Slag tap burner					
Gas rate (ft. ³ /hr.)	3,000	3,000	2,735	3,000	3,000
Air rate (ft. ³ /hr.)	16,050	16,050	12,590	16,050	16,050
Oxygen rate (ft. ³ /hr.)	1,200	900	1,120	1,200	1,200
<u>Produced</u>					
Gas (N ₂ free basis)					
CO ₂	1.85	2.55	3.5	2.7	2.35
C _n H _m	0.0	0.45	0.7	0.55	1.05
H ₂	27.25	28.05	29.25	28.1	28.65
CO	70.45	61.3	60.05	60.85	60.55
C _n H _{2n} + 2	0.45	7.65	6.5	7.7	7.25
	100.0	100.0	100.0	100.0	100.0

TABLE 3. (Contd.)

Test No.	54	67	70	71	72
Value of 'n'	1.0	1.123	1.128	1.129	1.114
Calorific value	315	374	366	375	379
H ₂	-	250	233	-	144
Organic sulphur	-	206	126	-	-
Rate	3.7 x 10 ⁶	5.1 x 10 ⁶	5.0 x 10 ⁶	7.23 x 10 ⁶	5.09 x 10 ⁶
(Btu.ft. ³)					
(gr./100 ft. ³)					
(ft. ³ /day)	542	938	1,026	534	602
(lb./hr.)	1.1	4.0	5.3	3.1	3.6
(min./hr.)	19.7	77.5	69	229	63
Dust Carry-over	-	508	501	-	453
(lb./hr.)	-	1,609	1,807	-	1,413
Ter	-	30,100	34,700	-	35,600
(lb./hr.)	-			-	
Permanganate value	-			-	
(p.p.m.)	-			-	
<u>Specific Data</u>					
Consumption					
Fuel	516	981	924	1,440	930
(lb./hr.ft. ²)					
(lb./therm)	3.40	2.56	2.66	2.62	2.54
(ft. ³ /therm)	71.4	55.2	56.2	54.7	56.6
Production					
Gas	94,800	68,800	71,900	66,500	72,400
(ft. ³ /ton)					
(therms/ton)	299	258	263	250	274
(therms/hr.ft. ²)	69	113	109	153	114
(per cent fuel charged)	0.45	0.9	0.8	1.84	0.8
Dust	-	15.4	16.2	-	15.5
(gal./ton)	-	2.0	2.35	-	1.76
Tar	-			-	
Liquor	-			-	
(lb./therm)	-			-	

TABLE 4. Mass Balance for Test 67 - lb./hr.

		Carbon	Hydrogen	Oxygen
IN	Coal	5,323	373	993
	Moisture	-	121	972
	Steam and Oxygen	-	226	5,106
	Slag Tap Burner	41	15	378
		<u>5,364</u>	<u>737</u>	<u>7,449</u>
OUT	Gas	4,896	509	5,925
	Dust	48	2	6
	Tar	406	38	45
	Liquor	14	176	1,376
	Difference	0	12	97
		<u>5,364</u>	<u>737</u>	<u>7,449</u>

TABLE 5. Heat Balance for Test 67.

		Therm/h.	%
IN	Coal (potential)	958	15.8
	Town gas (potential)	13.5	1.35
	Steam and Oxygen (latent and sensible)	28.5	2.85
		<u>1000.0</u>	<u>100</u>
OUT	Gas (potential)	803.0	80.3
	Gas and vapours (latent and sensible)	55.5	5.55
	Tar (potential)	84.4	8.44
	Dust (potential)	8.6	0.86
	H ₂ S ammonia, etc. (potential)	4.6	0.46
	Slag (sensible)	8.0	0.8
	Cooling water from tuyeres and hearth (sensible)	6.6	0.66
	Cooling water from gasifier jacket and stirrer (sensible)	15.2	1.52
	Unaccounted for losses, etc.	14.1	1.41
	<u>1000.0</u>	<u>100.0</u>	

d) Ash and Slag Balances.

By making a mass balance of constituents in the fuel ash, flux and slag and assuming a quantitative recovery of such refractory oxides as alumina lime, magnesia and titania the loss of the other constituents were evaluated. As was expected the losses increased as the temperature in the tuyere zone increased, and the percentage loss of each constituent would be correlated with the steam oxygen ratio. Actual temperatures measured by sighting an optical pyrometer through the tuyeres ranged from 1800°C at a steam oxygen ratio of 1.3 v/v to more than 1950°C at 1.10 v/v.

In addition, to the loss of the more volatile constituents such as sulphur phosphorus, chlorine and oxides of sodium and potassium there was a loss of iron oxides due to the formation of metallic iron, and also of silica. Loss of silica was due to its reduction to volatile silicon monoxide, which was subsequently reoxidised and carried away in the gas stream as silica fume. With coke the loss of silica was 5% at a steam oxygen ratio of 1.3 but this increased very rapidly at a ratio of about 1.10 v/v to more than 15%. Typical losses of sodium oxide, potassium oxide and sulphur were 5, 5 and 40% respectively at 1.3 v/v, increasing to 40, 40 and 80% at a ratio of 1.10 v/v. These materials were deposited in varying degrees in the gas offtake and in the sumps of the waste heat boiler and final cooler. The deposits in the gas offtake were enriched with alkali metal oxides and had a silica concentration of more than 50%. Most of the material was removed in the bottom of the waste heat boiler, but finely divided solids with an ash content greater than 90%, of which 75% was silica, separated from the gas in the final cooler and was removed in the blow down.

The loss of ash constituents from the coals tested were significantly lower than for coke, with the exception of chlorine and sulphur. Silica losses were negligible apart from the test with untreated coal that contained a high proportion of shale.

The loss of iron oxides is of considerable importance in the operation of a slagging gasifier due to the hazard presented by liquid iron, and to the effect upon the flow properties of the slag. Iron oxide acts as a flux and its loss raises the melting point, and by increasing the silica ratio produces an increase in the slag viscosity. A controlled series of experiments with coke showed that there was little effect on the reduction of iron oxides by decreasing the steam oxygen ratio from 1.40 to 1.20 v/v, but a further decrease to 1.10 v/v almost doubled the iron formation. However, it was found that this trend could be reversed by supplying extra oxygen to the gases passing through the tap hole, and reduction of the iron oxides was decreased to 15%. Although in the experiments with coal between 20 and 30% reduction occurred it is considered that this could have been much improved by supplying a larger proportion of the gasifying media through the tap hole.

e) Tar and Liquor.

Although the tar and liquor condensates from the gas were separated in two stages they were discharged into a common separating tank for ease of handling. This, together with the high liquor circulation rate from the sump of the waste heat boiler to provide the gas quench, produced a tar water emulsion. However, the emulsion was not persistent and three phases that could be separated without much difficulty were formed; a lower layer of solids, a middle layer of tar and an upper layer of liquor.

The liquor originated almost entirely from the moisture in the coal and had a permanganate value of 30,000 p.p.m., i.e., more than twice that from a Lurgi gasifier. Its smaller volume and, hence its reduced capacity to take into solution oxygen absorbing agents means, however, that its total oxygen absorbing potential per therm of gas is only half that of Lurgi liquor.

The tar, even after prolonged standing, retained about 20 to 25% water and about 4% suspended solids. It has been analysed by the Coal Tar Research Association⁽⁴⁾ and found to have the general characteristics associated with a vertical retort tar, i.e., a similar pitch yield and contents of naphthalene and tar acids. The high ratio of 2 to 1 methyl naphthalene and the high content of light oil, which was greater than that normally found in a carburetted water gas tar, indicated a high temperature of formation.

f) Behaviour of the Fuel Bed.

There was no evidence of channelling as a result of blockage of part of the fuel bed with fines, or any other abnormal behaviour during the operation of the slagging gasifier for a total of 500 hours with coke and 300 hours with coal. Screen analysis of the fuel bed at the end of the experimental tests with coal indicated that degradation increased as the fuel moved down the bed, but there were two main zones where this occurred, in the caking zone at the top of the bed due to the action of the stirrer, and in the tuyere zone due to the highly turbulent motion of the fuel. Table 6 shows the size distribution in the fuel bed immediately below the stirrer, and at the tuyere level, when the fuel charged to the gasifier was $\frac{1}{2}$ inch to 1 inch and 1 inch to $\frac{3}{4}$ inch. The proportion of fuel below $\frac{1}{2}$ inch at the tuyere level showed little variation in the tests on coal and averaged 30%

TABLE 6.

Fuel Size Inch	Test 67			Test 70		
	As Charged	Below Stirrer	At Tuyere Level	As Charged	Below Stirrer	At Tuyere Level
$\frac{1}{2}$ to 1	24.6	-	-			
1 to $\frac{3}{4}$	40.0	-	-	21.0	-	-
$\frac{3}{4}$ to $\frac{1}{2}$	21.0	59.1	10.9	15.6	21.4	9.2
$\frac{1}{2}$ to $\frac{1}{4}$	9.7	29.9	49.3	53.0	57.0	29.2
$\frac{1}{4}$ to $\frac{1}{8}$	1.4	3.1	8.1	2.2	8.6	30.1
below $\frac{1}{8}$	3.3	3.7	29.7	8.2	14.0	30.6

The fuel below $\frac{1}{2}$ inch entering the gasifier in Test 70 was more than double that in Test 67 but there was no significant change in the amount of dust carryover. In each of these tests the gas output was equivalent to four times that of a typical Lurgi gasifier but the fuel carryover was less than 1% of the fuel charged. However, on raising the gas output by 50% (Test 71) the carryover was doubled. With coke the carryover was only half that associated with coal, due to the absence of any mechanical breakage by the stirrer and the lower gas velocity above the fuel bed. There is little doubt that the carryover could be reduced by giving attention to stirrer design and perhaps by adopting the recirculation of tar dust mixtures as practised on the commercial Lurgi gasifier.

h) Comparison with a Lurgi Gasifier.

The Donisthorpe coal used in the slagging gasifier for Test 67 and 71 is similar to the coal used at the Lurgi gasification plant at Westfield, so that it is possible to make a direct comparison between the performances of the two types of gasifier. In Table 7 the performance data from Test 67 is compared with data given by Ricketts⁽⁵⁾.

Recent developments in the Lurgi process have resulted in lower steam consumptions and corresponding increases in gasifier output, but the output is still much lower than a slagging gasifier. At Westfield it has been possible to obtain nearly 20% more crude gas than the guaranteed maximum of 12 mill. ft.³/day, although at the higher loads more exacting operating conditions are required to avoid clinker formation. This compares with an output from the slagging of 5.2 mill. ft.³/day of

gas having a higher calorific value, which can be increased by 50% to 7.3 mill. ft.³/day without any significant loss of performance. The output of the experimental slagging gasifier when operating at a lower pressure is therefore more than half that of the commercial Lurgi gasifier. When expressed as the weight of fuel gasified or the volume of synthesis gas ($\text{CO} + \text{H}_2$) per unit cross sectional area of shaft the output of the slagging gasifier is at least four times greater than the Lurgi gasifier.

In Table 7 the material requirements of the two gasifiers are compared for the production of a therm of crude gas and for 1,000 ft.³ of synthesis gas. Steam requirements show the greatest difference; that of the slagging gasifier being about one fifth of the Lurgi gasifier. However, the composition of the crude gases should be considered when comparing steam consumption if the gas is to be detoxified or if synthesis gas is required. In the Lurgi crude gas there is sufficient undecomposed steam to carry out the required reduction in carbon monoxide content without added steam or liquor, but extra steam would be needed with the slagging gasifier because of its high carbon monoxide content, small volume of undecomposed steam and lower outlet temperature. Under these conditions the slagging gasifier loses some of its advantage in steam consumption.

The oxygen consumption per therm of crude gas from the slagging gasifier is about 10 to 12% higher than that of the Lurgi gasifier. This is due to the formation of a smaller proportion of the exothermic products carbon dioxide and methane, and to the loss of high grade heat to the hearth slag tap and tuyere cooling water. The oxygen consumption per unit volume of synthesis gas shows little difference between the two gasifiers.

TABLE 7. Comparison of the results from the slagging gasifier and a commercial Lurgi Plant.

	Lurgi Gasifier	Slagging Gasifier
Operating pressure (lb./in. ²)	355	300
Fuel:		
Rank	902	902
Size range (in.)	1¼ to ¾	1½ to ¼
Ash, including flux (%)	14.6	11.4
Moisture (%)	15.6	14.7
Steam/oxygen ratio (vol./vol.)	5.4	1.10
Crude Gas Composition, (% Vol.)		
CO ₂	24.6	2.5
C _n H _m	1.1	0.45
CO	24.6	60.5
H ₂	39.8	27.75
C _n H _{2n+2}	8.7	7.6
N ₂	1.2	1.0
	<u>100.0</u>	<u>100.0</u>
Calorific value (Btu/ft. ³)	309	371
Steam consumption (lb./therm crude gas)	11.6	2.56
Steam consumption (lb./1000 ft. ³ (CO + H ₂))	56.1	10.7
Oxygen consumption (ft. ³ /therm crude gas)	49.5	55.2
Oxygen consumption (ft. ³ /1000 ft. ³ (CO + H ₂))	238	236
Synthesis gas (CO + H ₂) output (ft. ³ /hr. per ft. ²)	4,930	26,700
Gasification rate (lb. d.a.f. coal/hr. per ft. ²)	210	981
Efficiency $\frac{\text{(Potential heat gas)}}{\text{(Potential heat coal)}}$	81	82.5

CONCLUSIONS.

A pilot scale slagging gasifier has been developed that will gasify coke and weakly caking coals, and performance data have been obtained at 20 atm. pressure. The gasifier had a low steam consumption, high output, high thermal efficiency and handled coal with a wide range of ash fusion temperatures provided that dolomite was used to flux the more refractory ashes. A crude gas output of $7\frac{1}{2}$ mill. ft.³/day equivalent to 42,000 ft.³/hr. per ft.² of fuel bed was obtained without any limitation by the fuel bed.

Comparison of the results with those of a Lurgi installation showed that the steam consumption and liquor volume were one fifth, and the gas output per unit area of the fuel bed was from four to seven times greater than the Lurgi gasifier. The slagging gasifier produced a gas with a high carbon monoxide content and there was little undecomposed steam to convert it. However, with the added steam necessary to convert the carbon monoxide to the level required for synthesis gas the overall steam consumption would be lower than for the Lurgi gasifier. For detoxification to the level required in the United Kingdom it would be logical to convert part of the carbon monoxide and use methane synthesis to remove the remainder, at the same time increasing its calorific value.

The system of intermittent tapping, in which a reservoir of slag was maintained in the hearth, and enabled the slag to be run off at a high rate by the application of a controlled differential pressure, proved to be very satisfactory. Coal with a low ash fusion temperature, uncleaned coal with a high proportion of adventitious ash and coal with a refractory ash when suitably fluxed, gave a homogeneous slag that was tapped without blocking the slag tap. This tapping system could be used for scale up to a commercial size, by making modifications to the slag tap burner design to allow the free drainage of iron from the hearth.

Experience with the slagging gasifier indicated that water cooled metal surfaces were essential for the slag tap and areas of the hearth exposed to hot slag, but were prone to attack by liquid iron. Iron attack was greatly reduced by maximising heat transfer through the metal surfaces, by preventing iron from accumulating in the hearth, and by introducing oxidising gases through the tap hole to retain the bulk of the iron in solution in the slag. The water cooled steel slag tap used in the gasifier proved capable of withstanding hot slag and iron for long periods, but occasionally suffered some damage from iron attack. Complete resistance to iron attack is an obvious design requirement, and recent experiments suggest that this can be achieved by the use of copper.

The development of the slagging gasifier has now reached a stage that requires the proving of designs and materials for longer periods of time, possibly on a prototype gasifier. However, in the United Kingdom the gasification of solid fuels, even with the improvements offered by operation under slagging conditions, cannot at present compete on economic grounds with the new oil gasification processes now available. An alternative route for producing gas from coal, by hydrogenation and gasification in a fluidised bed, is therefore being investigated on the pilot plant scale.

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