

TABLE VII  
PRODUCTION COST FOR FISCHER-TROPSCH PRIMARY PRODUCT

	COST PER UNIT			PER TONNE OF LIQUID PRODUCT		
	Unit	Germany RM	UK £	Units	Germany RM	UK £
Coke .. .. .	tonne	17.79	3.0	4.38	85.55	13.14
Catalysts .. .. .	—	—	—	—	20.83	2.08
Wages .. .. .	—	—	—	—	10.71	1.07
Salaries .. .. .	—	—	—	—	2.41	0.24
Water .. .. .	m <sup>3</sup>	0.006	0.002	447	2.86	0.89
Steam .. .. .	tonne	2.60	0.3	15.8	41.10	4.75
Electric power .. .. .	1,000kwh	17.60	2.50	0.596	10.45	1.49
Fuel gas .. .. .	—	—	—	—	3.99	0.40
Compressed air .. .. .	—	—	—	—	0.13	0.01
Maintenance .. .. .	—	—	—	—	17.55	1.75
Materials .. .. .	—	—	—	—	12.57	1.26
Taxes and various costs .. .. .	—	—	—	—	21.70	2.17
<b>TOTAL .. .. .</b>					<b>229.85</b>	<b>29.25</b>
<b>Less Credit:</b>						
Liquefied gas .. .. .	tonne	200	10	0.097	19.40	0.97
Steam .. .. .	—	—	—	—	26.63	2.66
Residual gas .. .. .	—	—	—	—	8.16	0.82
<b>TOTAL .. .. .</b>					<b>175.66</b>	<b>24.80</b>
<b>Add:</b>						
Capital charges .. .. .					91.10	9.11
<b>TOTAL COST .. .. .</b>					<b>266.76</b>	<b>33.91</b>
<b>Cost of primary product (pence/gallon) .. .. .</b>						<b>27.1</b>

## B—Carbonisation and Gasification of Coal

### 1. INTRODUCTION

The production of synthetic oils, as will be seen from the attached chart (Fig. 3), is fundamentally dependent on coal carbonisation and gasification processes. The hydrogenation process for production of low molecular weight hydrocarbons from more complex structures requires compressed hydrogen of a fair degree of purity, usually not less than 96%, whereas the starting material for the Fischer-Tropsch synthesis is a gas containing hydrogen and carbon monoxide in the volume ratio of about 2 : 1.

The first step in the production of either hydrogen or Fischer-Tropsch synthesis gas is the manufacture of water gas. The conventional water gas process consists of steaming high-temperature coke from selected bituminous coal in intermittent generators, the heat required for the steaming reaction being supplied and stored in the coke bed during periods of blowing with air. The gaseous products from the blow period, consisting of nitrogen, carbon dioxide and a little carbon monoxide, are purged to atmosphere. This process was still used extensively in Germany, but it required lump coke, preferably of high ash-melting point, and limited availability of suitable coking coal had necessitated development of a wide variety of alternative processes capable of using inferior coal cokes, brown coal cokes and brown coal itself.

Low-quality cokes containing a high proportion of low melting point ash constituents were dealt with in continuous slagging generators. Heat was supplied continuously by adding oxygen to the steam, and ash

withdrawn as liquid slag, thus eliminating the need for mechanical grates.

The main method developed for the direct gasification of brown coal was the Winkler generator, which operated continuously with a mixture of oxygen and steam reacting with powdered brown coal or brown coal coke in a fuel bed maintained in an ebullient state of suspension. Latterly, this type of generator has been used largely on brown coal coke. It has a comparatively poor carbon efficiency because a large amount of un-gasified dust is blown overhead.

The next most important method was the Pintsch-Hillebrand process, using brown coal briquettes which were gasified continuously by reaction with a current of preheated steam and recycled carbonisation and water gas. The Wintershall-Schmalfeldt process was similar to the Pintsch-Hillebrand process, but used brown coal dust which was wholly gas-borne during gasification. The less-developed Koppers complete gasification process used brown coal briquettes, and was also similar to the Pintsch-Hillebrand generator.

Special low-temperature carbonisation processes were developed in Germany for the production of the coke required for water gas production. These were applied mainly to brown coal and to poorly coking bituminous coals. The two most important methods were the Lurgi-Spülgas and the Krupp-Lurgi systems. These processes were important to the German oil production programme not only because of their contribution to coke and water gas manufacture, but also because they gave rise to large quantities of low-temperature tars

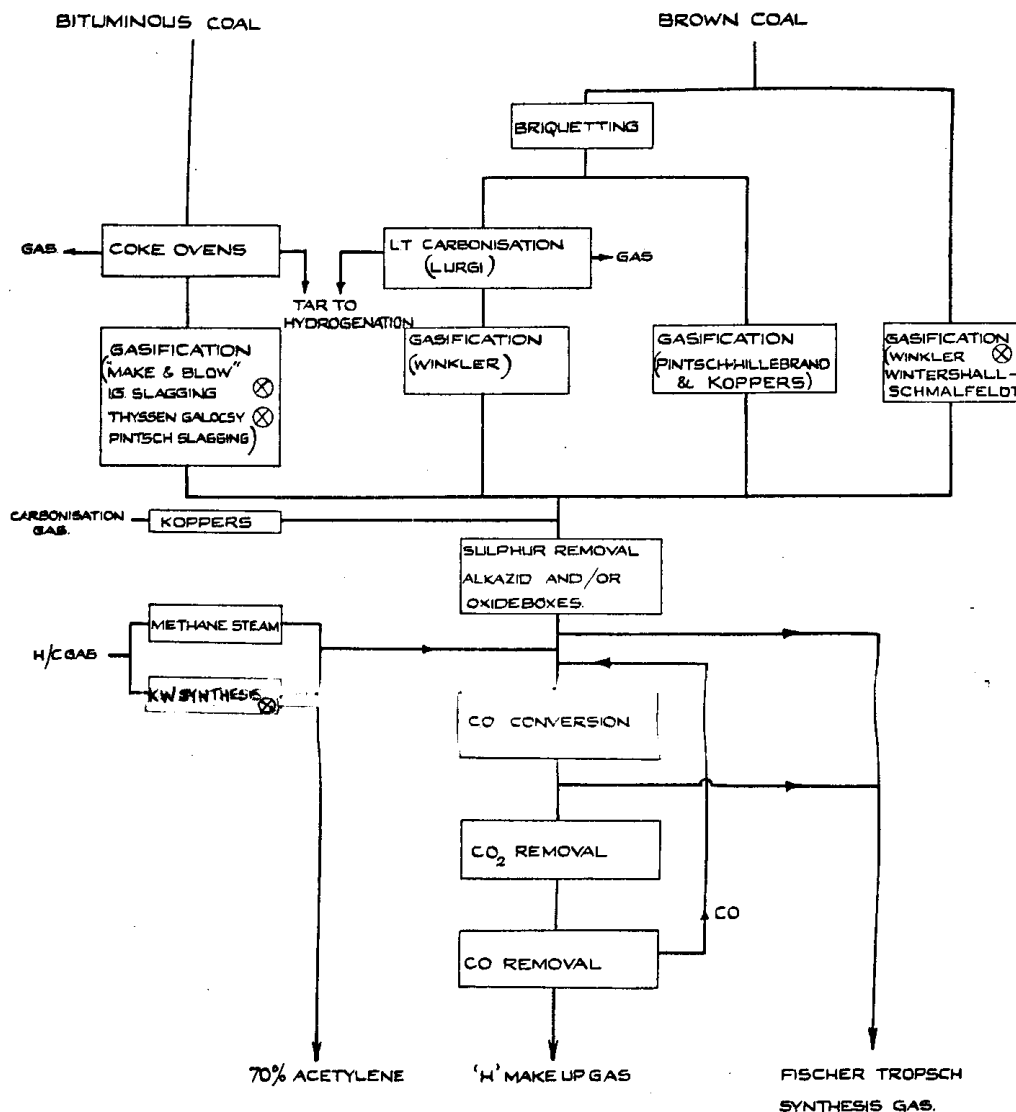
# 10 REPORT ON THE PETROLEUM AND SYNTHETIC OIL INDUSTRY OF GERMANY

which were used as raw materials for the hydrogenation processes.

Coal carbonisation is also the source of a gas which can be used to supplement coal and coke in the manufacture of water gas and hydrogen. Hydrogen can be

Three main methods for treatment of hydrocarbon gas were used in Germany, the methane-steam, the KW and the Koppers cracking processes. The methane-steam process has been used in both the UK and US, and is comparatively well known. Sulphur-free gas is

FIGURE 3

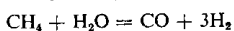


⊗ INDICATES THAT THE PROCESS REQUIRES OXYGEN

## DEPENDENCE OF GERMAN OIL SYNTHESIS ON CARBONISATION AND GASIFICATION PROCESSES

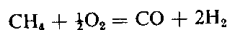
obtained direct by low-temperature fractional distillation of coke-oven gas, and various methods have been developed for the production of water gas from methane, also a constituent of carbonisation gases. These latter processes have also been used to convert to water gas, the methane, ethane, propane and butane obtained as by-products in the hydrogenation and Fischer-Tropsch processes.

reacted with steam at 700°C over a nickel catalyst contained in special steel tubes heated by flue gas. The reaction is shown by the equation:—



In the KW process the external heating, and therefore the use of an expensive tubular reactor, is avoided by generating the necessary heat for the reaction internally

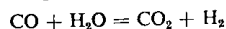
by burning part of the hydrocarbon gas feed with a controlled quantity of oxygen according to the equation:—



An interesting feature of this process is that conditions can be adjusted so that a synthesis gas containing 8–9% of acetylene can be produced.

Finally, the Koppers cracking process is a non-catalytic method of reacting steam and hydrocarbon gas in chambers filled with refractory chequerwork. The operation is intermittent, consisting of a gas-burning period followed by a reaction period. In the former, the chequer brickwork is heated up to 1,300°C.

The first stage in the conversion of water gas to hydrogen is the reaction of the carbon monoxide which it contains with steam to give carbon dioxide and hydrogen according to the equation:—



This same reaction is also frequently employed to adjust the  $\text{H}_2 : \text{CO}$  ratio of Fischer-Tropsch synthesis gas to the required figure.

For this reaction it is first desirable to remove  $\text{H}_2\text{S}$ . This was often effected by the well-established oxide box method but the Alkacid process in which the gas is scrubbed with a solution of an amino-acid salt was also in extensive use. An  $\text{H}_2\text{S}$  removal process using active carbon has been developed at Leuna.

The reaction of carbon monoxide with steam is effected by passing the purified gas and excess steam over chromium-promoted iron oxide catalyst. The process is normally operated at substantially atmospheric pressure, but in order to obtain maximum unit capacity and thereby conserve steel, the newer plants in Germany have adopted a modified process operating at 20–30ats. This process also saves compression costs but it has the disadvantage that parts of the plant become subject to corrosion.

Fischer-Tropsch synthesis gas was usually a mixture of water gas and gas from the water gas-steam reaction blended in the proportions necessary to give a  $\text{H}_2 : \text{CO}$  ratio of about 2 : 1. Before use, it had to be carefully purified from sulphur compounds.

For the manufacture of hydrogen, the product from the water gas-steam reaction has to be freed from carbon dioxide. This is usually effected by scrubbing with water at about 30ats. About 50% of the energy consumed in injecting fresh water is recovered by letting down the carbon dioxide solution through Pelton wheels. The carbon dioxide-free gas is next compressed to 200–300 ats, and scrubbed with ammoniacal copper solution to remove carbon monoxide. The carbon monoxide recovered in this way is recycled for further reaction with steam.

In the course of the investigations into the activities of the carbonisation and gasification industries, related to synthetic oil production, interesting detailed information was obtained on developments in the field of complete gasification of coal to a gas suitable for supplementing town gas. The main development was the Lurgi high-pressure process in which brown or non-coking bituminous coal is reacted continuously with a stream of steam and oxygen under a pressure of about 20ats. Under these conditions, the methane content of the final gas, and therefore its calorific value, remains high. The final gas after  $\text{CO}_2$  removal, contains approximately 20–25% methane, 45–50% hydrogen and 20–25%  $\text{CO}$ .

The detailed report which follows is sub-divided into the following sections:—

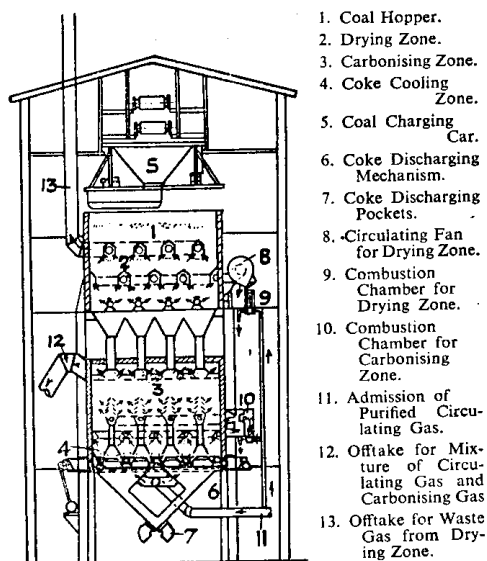
2. Low-temperature carbonisation.
3. Water gas production by coal or coke gasification.
4. Water gas production from hydrocarbon gases.
5. Conversion of water gas into hydrogen or Fischer-Tropsch synthesis gas.
6. Miscellaneous processes of interest to the gas industry.

## 2. LOW-TEMPERATURE CARBONISATION

### i. GENERAL

The majority of the low-temperature coal carbonisation plants erected in Germany immediately prior to or during the war were of the Lurgi-Spülgas or of the Krupp-Lurgi type.

FIGURE 4



ELEVATION OF A LURGI-SPÜLGAS DIRECT-HEATING CARBONISATION PLANT

Under the four-year plan "Spülgas" plants, similar to those erected before the war in Canada and New Zealand, were built:—

- a. At Ofleben, Znachterstedt, Deuben, Profen, Regis, Deutzen, Böhlen, Hirschfelde, Rositz, Webau, Koepsen, Leuna, Magdeburg, Zeitz, Lützkendorf and Espenhain—for soft brown coal.
- b. At Brück—for hard brown coal.
- c. At Blechhammer, Karsten-Zentrum Grube and Auschwitz for non-coking or slightly coking bituminous coal.

Krupp-Lurgi plants were erected at Wanne-Eickel and Velsen. Similar plants with a modified coke discharge system (Brennstoffe Technik design) were built at Marienau (Saar) and at Mizluwitz Grube (Silesia).

### ii. LURGI-SPÜLGAS PROCESS

This process was considered in Germany to be best for the carbonisation of coal briquettes, although it had been applied to weakly coking coals.

The Spülgas oven, illustrated in Fig. 4, consists essentially of a deep shaft of rectangular cross-section

divided into three zones through which the briquettes pass successively. In the upper zone, the briquettes are pre-dried down to a moisture-content of less than 0.5%; carbonisation occurs in the middle zone, and the coke is cooled in the bottom zone which is followed by the coke discharging mechanism.

The shafts consist of a steel shell, 5mm thick, brick lined to a thickness of up to 327mm in the carbonising zone, which is about 6m high and of cross-section approximately 3m.

The plant is internally heated in both drying and carbonisation zones by the recirculation of gas produced in the carbonisation. This gas after washing, etc., is heated by the combustion of a part of it in chambers adjacent to the ovens. Usually, unpurified gas is used in both sections, but at the plants inspected purified gas was being used in the drying stage to minimise troubles caused by the condensation of sulphurous gases in cooler sections of the plant. The heating gases are circulated by means of blowers and enter the charge through a series of louvres across the section of the retort, those in the drying zone being made of metal, and those in the carbonisation zone of brick.

Hot gases enter the drying zone at a temperature of about 220°C, mix with the steam evolved from the briquettes, and pass off to atmosphere saturated at about 60°–80°C. A temperature of 600–700°C is maintained in the carbonisation zone, although this can be raised to 900°C for the production of coke of only 2% volatile matter. The temperature of the gas leaving the carbonising unit is 250°C.

In the lowest section of the shaft the coke is cooled by recirculated gas which is thus preheated before combustion. The coke is then discharged through a lock chamber fitted with a movable bell arrangement operated by a lever.

The carbonisation gas is passed successively through a dust separator, a washer-cooler, an electrostatic tar precipitator, a tubular cooler, a booster and a light oil absorption plant, before being recirculated.

The plants examined were at Böhlen and Regis, both of which were carbonising briquetted brown coal. As delivered from the mines, this coal contained over 50% water which was reduced to 10–25% before briquetting. Brown coal is particularly reactive and liable to form explosive mixtures with air; hence precautions to exclude air as much as possible were taken throughout the drying and briquetting processes.

At Regis the conveyors were all of the enclosed type (Redler) and cool waste gases were kept in continuous circulation during the operation of the plant. The wet coal was dried by contact with combustion gases at a temperature of 900–1,000°C. During the drying process the rapid evolution of steam caused a breakdown in size of the coal particles; approximately 50% of the dried coal will pass through a screen of 9,000 meshes per cm<sup>2</sup>. The dried brown coal was then briquetted, without binder, in a normal plunger type or in a Krupp ring roll press.

The Böhlen plant consists of 24 ovens with a throughput of up to 7,000te/day of briquettes. The plant produces 820te/day of tar and 1.5 million m<sup>3</sup> of gas per day, of which one-third is needed to heat the ovens. The Regis plant consists of 10 double-shaft retorts with a nominal throughput of 3,000te/day. The plant was producing 1,200te/day of coke immediately prior to shutting down.

The coke from brown coal is particularly reactive and at Böhlen the coke discharged from the ovens is further cooled in a rotary drier using nitrogen from the

air separation plant. At Regis, water sprays for use when necessary are fitted above the vibrating screens on to which the coke is discharged.

The coke discharged at Böhlen contains 20% breeze. The yields per tonne of dried brown coal are:—

Tar	..	..	130kg
Coke	..	..	450 "
Surplus gas	..	..	300m <sup>3</sup>
Gas C.V.	..	..	2,000k.cal/m <sup>3</sup>

The coke discharged at Regis was said to contain 5–8% of dust below 6mm size. The following were quoted as typical yields of products:—

Coke	..	50% of briquettes charged	
Tar	..	13% " " " "	(94% of Fischer Assay yield)
Surplus gas	..	160m <sup>3</sup> /te (plunger briquettes)	
		280m <sup>3</sup> /te (ring press briquettes)	
Gas C.V.	..	1,800k.cal/m <sup>3</sup>	

The coke normally contained 2.5% of sulphur but was said to be free from tar.

At Blechhammer the Spülgas process has been applied to the carbonisation of weakly coking coals, 5 ovens (300te/day per oven) making hard coke for hydrogen production and transport producers.

### iii. KRUPP-LURGI PROCESS

The Krupp-Lurgi oven is illustrated in Fig. 5. Each oven has 6 cells, with an average width of 84mm, and the 6 cells are charged simultaneously. The correct predetermined volume of coal is fed in a stream into each cell by passing through rolls. As the coal is fed in, a stamping bar, the length of the oven and of somewhat smaller width, is raised and lowered. The density of the charge is thus increased by about 8% to 0.835 te/m<sup>3</sup>. The time required is about 1.5mins. The top surfaces then have to be cleaned before the cover is replaced on the oven; this cleaning takes about 2min and is a disagreeable operation since within about 2min after the coal first enters, the tar fog begins to be liberated.

A new charging machine designed to overcome this tar fog nuisance has been built and a full-scale model tested. With this machine, charging can be made in 15sec and the increase in density above that of the unstamped coal is 15%. This is obtained by a "free" fall of the coal charge of 5m from a hopper through guiding slots into the individual cells. It is claimed that a uniform density from top to bottom of the cell is achieved by this method and stamping is not required. A "free" fall of 7.5m is said to give a 25% increase in density but the charging machine becomes too unwieldy.

Each of the 6 cells is 3,200mm long and 2,100mm high. The width at the top is 76mm and at the bottom 100mm. The capacity per cell is 0.585m<sup>3</sup> and per oven 3.51m<sup>3</sup>. The cells are separated by heating flues, the walls being 24mm thick "II" boiler-steel plate. An all-welded construction is used. The heating flues are of a cellular construction where the maximum dimension of a cell is not more than about three times the thickness of the webbing. Due to the taper of the coal cells, the flues taper in the reverse direction and are narrower at the bottom than at the top. Since the heat requirements at the bottom are greater than at the top, the hot gas enters at the bottom and the average temperature at the bottom is about 10°C higher than at the top.

The steel used has a strength at room temperature of about 50kg/mm<sup>2</sup> and this rapidly decreases above about 300°C to about 0.5kg/mm<sup>2</sup> at 600°C. Some difficulty has been encountered with distortion of the

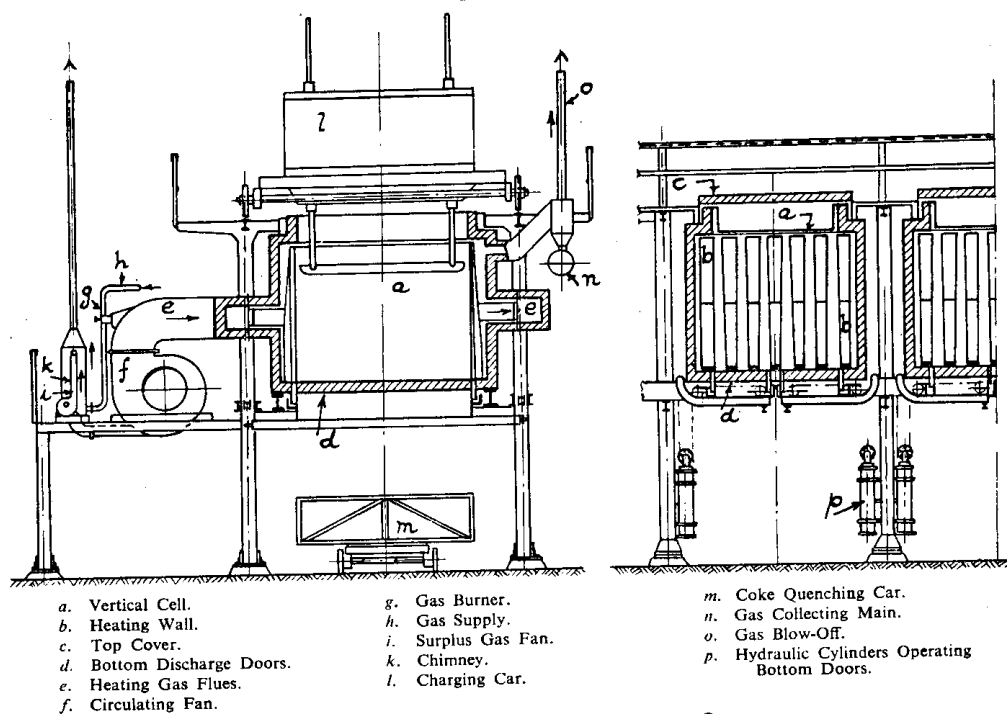
walls of the end flues but this was remedied by adding special supports within the refractory housing that surrounds the metal chambers for heat insulation. It is stated that no difficulties due to carburisation, decarburisation or oxidation are encountered. Some of the ovens had not been cooled down for 5 to 6 years. The wall thickness decreases about 0.5mm/yr and experience indicates that a life of 10 to 12 years can be anticipated.

Sometimes local buckling of the wall occurs, and two types of equipment, one mechanical and the other hydraulic, to flatten the metal walls have been developed. These are first used when the ovens are 2 to 3 years old.

On the average the gas is recirculated twelve times, i.e. make-up gas for maintaining temperature is about 8.5%. Both the heating gas and the air for combustion are preheated by the waste gases. There is a separate burner for each block of four ovens. The heat requirement for 1kg of coal with 10% water content is stated to be 540–580k.cal. When starting up a plant, one week is allowed to heat up the ovens.

The total time for carbonisation is 5.34 hours, including 1.5 min for charging (the new method of charging will reduce this to 0.25min), 2min for cleaning the top surface of the ovens and closing the top

FIGURE 5



#### KRUPP-LURGI LOW-TEMPERATURE CARBONISATION OVEN

The cover plate is water-sealed and does not make a tight seal against the top of the metal structure, thus providing a plenum chamber connecting to the gas offtake. The bottom door is about 3m above the floor level below, permitting a railway car to be moved under the oven to receive the discharged coke. The bottom doors are also water-sealed. In the newer type of ovens (at Wanne-Eickel), there is only one bottom door, inside which is a movable metal grid or grating which completely closes off the bottom openings of the cell. The grid can be moved to one side by a chain and pulley so that the coke can be pushed.

The heating of the ovens at Wanne-Eickel is by recirculation of flue gas obtained by burning the tail gas (Rest gas) from the Fischer-Tropsch plant. This tail gas has a net calorific value of 1,800–2,000k.cal/m<sup>3</sup>. Coal gas produced in the carbonisation can, however, be used for heating and this is done in the Velsen installation. It is important to avoid excess air in the combustion so as to maintain a reducing atmosphere in the heating flues. The gas entering the bottom of the flues is at about 620°C and leaves the top of the flue at about 570–580°C.

cover, and 3–5min for discharging the coke. With this time of carbonisation, the capacity per oven per day, using stamped coal with a bulk density of 835kg/m<sup>3</sup>, is 11.2te/day. With the new charging device the capacity is expected to be increased to about 12.0te/day owing to the greater density of the coal charge in the oven.

The coke is pushed mechanically and received in a car for quenching with liquor and delivery to the coke wharf. The coke is collected as slabs that have no centre parting line. It is fed by conveyor belt to a breaker specially designed to produce a minimum of fines under 10mm in size. The slabs are carried over a set of eccentric rolls and as they pass, pusher bars are thrust by a system of cams against the slabs cracking them into large lumps. It is stated that this procedure produced only 7% material of less than 10mm as compared with 20% when a roll crusher is used.

The tar and gas formed in the carbonisation leave the oven and enter a primary cooler at about 190°C, where the heavy tar is condensed out. They then pass to a Lurgi-Cottrell precipitator at 150°C to recover tar-fog, and to an indirect cooler where the light tar is removed. After tar removal the gas is washed free from ammonia

and the resulting dilute ammonia liquor used for quenching the coke. The gas may be sent directly to a gas holder or it may be compressed to 10ats, cooled, freed from sulphur, and the benzene content lowered from 70 to 3gm/m<sup>3</sup> in a Feld washer. The wash oil used in the latter is part of the light tar condensate from the indirect cooler, after the benzene has been removed.

A plant of the Krupp-Lurgi type at Wanne-Eickel was inspected, and the following data relate to its performance.

The blended coal used contains about 5.4% ash, 8-10% water and 30% volatile matter. Fine grading is not desired since high density of charge increases throughput and gives a better product. Normally the size analysis of the charge is:—

0-1mm	..	30-40%
1-3	..	30-40%
3-6	..	20-30%
6-10	..	about 5%

Using this coal, the yield of dry coke based on dry coal is 84%, of approximately the following size distribution:—

40-90mm	..	70%
20-40	..	11%
10-20	..	7%
Below 10mm	..	12%

The last item includes the fines from discharging and handling as well as those produced in the crushing.

The large coke contains about 6.5% ash, 3-5% water and 7-8% volatile matter. Its porosity is approximately 50% and it has a high reactivity and is easily ignited. In the Micum drum test for coke strength, the percentage of coke greater than 40mm in size is about 65%. This coke is regarded as being of special value in gas production. Its high reactivity permits water gas formation at lower temperatures than those needed by high-temperature coke. It is claimed that the H<sub>2</sub>:CO ratio in the gas from the LT coke is 1.5:1, compared with 1.25:1 using HT coke. Thus the coke provides a more favourable gas for Fischer-Tropsch synthesis.

The three tars, heavy, tar-fog and light, are combined to form the low-temperature tar or Heizöl 5, and, after benzene removal, amount to 6.4% of the dry coal used. The raw heavy tar, after drying to less than 1% water, is used as a fuel by the Navy without further refining. The benzene recovered amounts to 1.25% of the coal charged and, on the plant inspected, has the following properties:—

Specific gravity at 20°C	..	0.820
Octane number	..	90
Phenol content	..	0.8% by volume
Olefin content	..	15%
Distillation begins at	..	50°C
25% distils up to	..	100°C
96% distils up to	..	195°C

The gas yield amounts to 95m<sup>3</sup>/te of dry coal and a typical analysis is:—

CO <sub>2</sub>	..	2.6%
Unsaturated hydrocarbons	..	2.4%
O <sub>2</sub>	..	nil
CO	..	3.8%
H <sub>2</sub>	..	25.7%
CH <sub>4</sub>	..	59.5%
N <sub>2</sub>	..	6.0%

The gross heating value is about 7,300k.cal/m<sup>3</sup>, and the net about 6,600k.cal/m<sup>3</sup>.

At Wanne-Eickel, 32 ovens were built in 1937 and a further 24 ovens in 1943. The rated monthly capacity of the plant was 17,000te semi-coke, 1,000te fuel oil (Heizöl) and 160te motor fuel (Schweibenzin), of which 7,200te semi-coke, 500te fuel oil and 97te motor fuel were obtained from the 1943 ovens. The total cost of the plant was RM5,000,000, the newer part costing RM2,000,000. About 100 men were required for operation; 8 men per shift for the 24 new ovens, an equal number for the 32 old ovens, one man per shift in the bunker house and the balance in the by-products plant.

The commercial possibilities of the process were considered to be good. In conjunction with a Fischer-Tropsch plant, it was claimed that this process gave 30-35% more total oil than would be obtained from a HT plant plus a Fischer-Tropsch plant. Apart from the operation of the plant in this way, it was claimed also that the process would be economically feasible for the production of a smokeless domestic fuel. The LT coke could also be blended advantageously with HT coke for central heating. The smaller sizes were also considered to be an excellent fuel for mobile gas producers, and another suggested use was in the production of ferro-silicon.

#### iv. LOW-TEMPERATURE CARBONISATION OF SHALE

Several plants for the LT carbonisation of shale, based on a tunnel furnace design, were under construction by the Lurgi Company in Estonia. There was also a similar plant at Heide in Holstein. Shale was also carbonised at a plant at Schomberg (Württemberg) using the Lurgi-Schweitzer system in which the retort is heated sufficiently internally by a mixture of air and steam to distil off the oil.

Other installations for shale carbonisation in Württemberg included:—

- Ten plants in which shale was carbonised in Meilers or heaps containing 1,500te shale.
- Twenty-eight Schweitzer retorts—capacity 1,000 te/day shale at Frommern.  
This installation was incomplete; it was to be heated by combustion of part of the shale.
- Eight Otto retorts each of 60te/day capacity at Dotternhausen. The retorts were heated internally by superheated steam.

### 3. WATER GAS PRODUCTION BY COAL OR COKE GASIFICATION

#### i. SLAGGING GENERATORS

##### (a) General

It had been known for some time that generators or producers were being operated in Germany with liquid slag removal, but information contained in the literature had always been scanty or had optimistically referred to results of tests of short duration in small installations. A good deal of information about this development has been obtained by the missions, which enables a much better assessment to be made of the economics of those processes found in large-scale operation, i.e. the IG, the Thyssen-Galocsy and the Pintsch methods.

##### (b) Lists of Installations

The installations shown in Table VIII of slagging generators are known to exist in Germany and elsewhere.

##### (c) IG Process

It is believed that prior to 1931 there were slagging generators in operation at both Leuna and Oppau, making producer gas for NH<sub>3</sub> synthesis, using a continuous

TABLE VIII  
LOCATION OF SLAGGING GENERATORS

Plant	Operating company	Installed	Units	Output in m <sup>3</sup> /hr H <sub>2</sub> + CO		Gas used for
				Unit	Total	
<i>a. IG type:</i> Leuna .. .. .	IG	before 1929	6	14,000	84,000	H <sub>2</sub> and methanol (formerly NH <sub>3</sub> )
Oppau .. .. .	IG	before 1929	?	?	?	NH <sub>3</sub>
<i>b. Thyssen-Galocsy type:</i> Wanne-Eickel .. .. .	Thyssen'sche	1943	1	4,000	4,000	Fischer-Tropsch
Duisburg-Hamborn .. .. .	Gas u. Wasserwerke	1938	1	200	200	Experimental
Pecs, Hungary .. .. .	—	1939?	1	1,000	1,000	NH <sub>3</sub>
<i>c. Pintsch type:</i> Scholven .. .. .	—	1939	8	8,000	64,000	H <sub>2</sub>

blast of air, sometimes with added steam. Tests were begun at Leuna in 1931 in a small slagging generator, using oxygen and hard coke, and since 1935 20,000m<sup>3</sup>/hr nitrogen-free water gas has been made continuously in slagging generators. To-day there are six generators installed at Leuna, mainly operated with a blast of oxygen and steam and mainly using "semi-ash"\* from ordinary coke water gas generators as fuel.

Fig. 6 shows a typical arrangement of a slagging generator, and the simplicity of design is very striking. The lowest and hottest part is brick-lined; the central portion consists merely of a water-jacketed mild steel shell, whilst the top, is again brick-lined. The whole generator stands on the ground. The blast is introduced through the sides of the base through eight water-cooled tuyeres or nozzles. The method of the slag removal is very simple; two tap holes are fitted at the extreme bottom, resembling the means for tapping blast furnaces. The fuel should not be less than 3cm but could be as large as 30-60cm. When using metallurgical coke as fuel it is usual to add 3% of limestone and 20% of solid slag from previous working to assist in forming liquid slag, but these additions are unnecessary when using "semi-ash" containing 50% carbon and 50% ash.

One of the six generators differs in having no water jacket; instead, cold water merely trickles down over the outside of the metal wall. This modification is recent and was carried out as the result of a serious accident in 1941 or 1942 which arose through a piece of clinker lodging just above a tuyere. As a result the incoming oxygen caused a very high local temperature, which burnt a hole in the jacket (see Fig. 6); the producer was flooded and quenched, and oxygen broke through into the gas mains and coolers, causing an explosion.

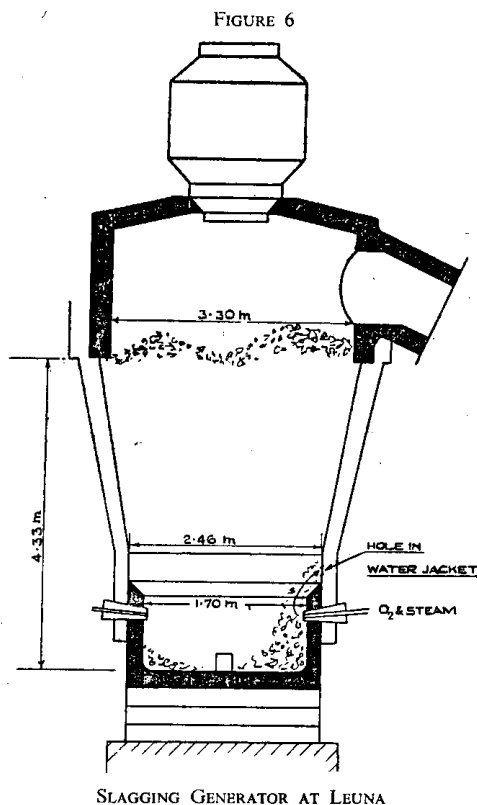
After leaving the generator the gas passes through a dust-catcher, hydraulic seal and coolers.

The output of each generator is 15,000m<sup>3</sup>/hr of nitrogen-free water gas, equivalent to an output of 1,650m<sup>3</sup> H<sub>2</sub> + CO per m<sup>2</sup> shaft area, measured at the generator top. This output is probably achieved only when using coke alone and is much less when using "semi-ash." Although the slag reaches a temperature of 1,700°C the exit gas temperature is only about 400°C. When tapping slag, molten iron can be obtained as the first runnings or by tapping at a lower hole, and is actually sold by Leuna.

As well as using hard coke and "semi-ash" as fuel, Leuna have also carried out tests with other fuels, such

\* "Rohschlacke": the grates of the coke generators are run fast to give ashes still reasonably high in carbon.

as low-temperature coke (made from bituminous coal), and crude coke (*i.e.* brown coal coke) and have also used CO<sub>2</sub> instead of steam.

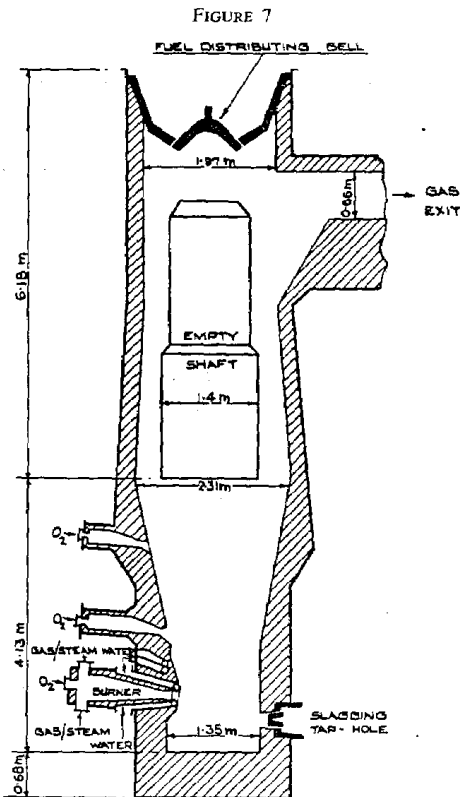


(d) Thyssen-Galocsy Process

The process was originally developed in Hungary, but was developed on a somewhat larger scale by the Thyssen'sche Gas und Wasserwerke. In 1938 they erected a small plant at Duisburg-Hamborn to gasify 2te/day coal and in 1943 a larger plant at Wanne-Eickel to gasify 40te/day coal, making about 4,000m<sup>3</sup>/hr H<sub>2</sub> + CO. Meanwhile just before the war a small unit was erected at Pecs, Hungary, gasifying 10te/day of coal. So far as is known there are no other units

running of any size, and since the running of the Wanne-Eickel unit was much interrupted by war causes it cannot be said that the process has been developed to anything like the same degree as the IG process.

The arrangement of the generator proper at Wanne-Eickel is shown in Fig. 7. The features showing the most important differences from the Leuna design are



THYSSSEN-GALOCSEY SLAGGING GENERATOR AT  
WANNE-EICKEL

the use of brick lining throughout; the much greater height (about double); the use of an inner steel shaft, kept empty by the method of charging, so as to even the flow of gas across the whole cross-section; and the three sets of tuyeres provided at the base as against one at Leuna. The five tuyeres in the lowest set are really water-cooled burners, where gas is burnt with oxygen and steam; the two upper sets of five each are not water-cooled but are used, one set at a time, to introduce more oxygen.

It is said that it is necessary to use the lowest set of tuyeres or burners in order to temper the reaction; in effect, highly preheated steam and gas passes over the coke, with sufficient additional oxygen introduced at tuyeres higher up to maintain temperatures. It is interesting that Leuna find it unnecessary to divide the gasification into two stages and claim that it is necessary only to ensure thorough mixing of steam and oxygen before feeding it to the tuyeres. Leuna state "... and at the only other place in Germany besides us where oxygen-gasification is used in slagging generators (Thyssen, Mülheim-Ruhr, after the patent of Galocsy, DRP.573,112), they have changed over to building pre-

combustion chambers before slagging generators, in which oxygen and gas are pre-combusted, giving a highly superheated mixture of steam and gas, which is then introduced into the fuel bed, so that in effect only endothermic reactions occur inside the generator. These elaborate precautions, which also reduce the output of the generator, are completely superfluous if there is early intimate mixing of oxygen and steam."

The gas used in the lowest tuyeres can be part of the gas made or some other suitable gas available. However, if part of the gas made is used, then the generator output is correspondingly reduced, although the thermal efficiency is not noticeably affected.

**Operating Results.**—Operating results for both the Leuna and the Wanne-Eickel generators when operating on hard coke (metallurgical coke) are set out in Table IX.

**Discussion.**—Comparing the efficiencies of the Thyssen-Galocsy and IG processes when working on hard coke, it can be seen that the carbon consumptions per 1,000m<sup>3</sup> H<sub>2</sub> + CO are very similar, with the TG process apparently achieving 100% gasification of carbon (calculated). The oxygen requirements of the TG process, however, are 20% higher, although less steam is used. It is difficult to give the exact reason for the higher oxygen requirements of the TG process; it may be an effect of relatively greater heat losses at the lower output, although the TG generator is better insulated. The lower steam rates and the recirculation of some gas to the burners would also tend to increase oxygen requirements, but the difference is much bigger than would be expected.

From the Leuna figures it appears that doubling the output has little effect on the efficiencies. Outputs per m<sup>2</sup> of shaft area are of the same order with each type of generator, and are higher than those obtained in "make and blow" generators.

When using "semi-ash" at Leuna the oxygen requirements are appreciably higher. A good deal of this is accounted for by the latent heat of fusion and sensible heat of the slag, which become important with the very high ash content of fuel.

**Comparison with Coke Water Gas Generators.**—An authoritative comparison between slagging generators using oxygen, coke water gas generators using air, and a "make and blow" process, is given by an IG gasification expert in BIOS Report 199.\* He concludes that for Leuna conditions, with oxygen costing 1.7p/gm<sup>3</sup>, the cost of pure hydrogen is about 8 to 15% higher using oxygen gasification than when using the "make and blow" process, the fuel in both cases being hard coke.

Figures used in the following comparison are based on the above, with some corrections.

Compared with "make and blow" generators, the slagging generator is thermally more efficient and appreciably less coke is used; the efficiency of the "make and blow" generator suffers from the CO content of blow gas, the sensible heat of blow gas and the greater quantity of undecomposed steam, whilst more carbon is lost as breeze and in the ashes.

Nevertheless, it will be seen that the coke savings of the slagging generator do not pay for the oxygen, although the coke consumption claimed above for the "make and blow" process is perhaps rather unusually low. Moreover, since any oxygen used must eventually appear as CO<sub>2</sub> in catalysed gas, the use of oxygen entails

\* "Ten Years of Oxygen Gasification at Leuna."—Translation by D. G. Fraser and R. J. Morley.



TABLE IX  
OPERATING RESULTS OF SLAGGING GENERATORS

Fuel		Thyssen- Galocsy, Wanne-Eickel	IG-Leuna		
		Hard coke	Hard coke	Hard coke	"Semi-ash"
Fuel analysis (as received)	C %	85	86.8	87.9	45-50
	S, etc. %	1.5	2.3	—	nil
	ash %	8.5	9.1	—	55-50
	H <sub>2</sub> O %	5.0	1.8	—	nil
C.V.	... k.cal/kg	6,865	—	—	—
Fuel size	... mm	40-60	—	—	—
Gas analysis	CO <sub>2</sub> %	2.5	6.8	—	9.7
	O <sub>2</sub> %	0.1	—	—	—
	CO %	70.7	61.4	64.6	66.5
	H <sub>2</sub> %	23.2	31.0	28.7	22.9
	CH <sub>4</sub> %	0.2	nil	0.1	nil
	N <sub>2</sub> %	3.3	0.8	0.9	0.9
	H <sub>2</sub> S %	—	0.6	—	0.4
Purity of O <sub>2</sub>	... %	90	98.5	98.5	98.5
100% O <sub>2</sub>	... m <sup>3</sup> /hr	1,053	—	—	—
Fuel	... kg/hr	1,680	—	—	—
Carbon	... kg/hr	1,430	—	—	—
Steam	... kg/hr	764	—	—	—
Auxiliary gas	... m <sup>3</sup> /hr	694	nil	nil	nil
Net gas made	... m <sup>3</sup> /hr	3,630	—	—	—
Net H <sub>2</sub> + CO made	... m <sup>3</sup> /hr	3,410	—	—	—
Gas exit temperature	... °C	—	400	390	200
O <sub>2</sub> in blast	... %	—	35.5	41.5	58
Cross-sectional area of shaft	... m <sup>2</sup> *	4.2	8.55	8.55	8.55
<i>Efficiencies (per 1,000m<sup>3</sup> H<sub>2</sub> + CO):</i>					
Fuel	... kg	493	490	496	—
Carbon	... kg	420	424	436	—
100% O <sub>2</sub> (measured)	... m <sup>3</sup>	309	258	256	342
Steam	... kg	224	346	294	207
Steam decomposition	... %	90	76.5	83	98
Carbon utilisation	... %	100	93	93	—
Output	... m <sup>3</sup> /hr/m <sup>2</sup> shaft area	810†	850	1,960	985‡

\* Shaft area for the Wanne-Eickel is calculated on widest diameter; the basis for the IG figures is unknown but is probably the same. † Limitation of oxygen supply.

‡ The output when using "semi-ash" as fuel was limited by charging and removal of the large quantity of ash.

TABLE X  
COMPARISON OF "MAKE AND BLOW" AND SLAGGING GENERATORS

			"Make and blow" Generator	Oxygen gasification		
				With slagging	Without slagging	
Gas analysis	CO <sub>2</sub>	%	5.0	6.8	22.6	
	CO	%	42.0	61.4	35.1	
	H <sub>2</sub>	%	49.0	31.0	40.5	
	N <sub>2</sub> and CH <sub>4</sub>	%	4.0	0.8	1.8	
<i>Consumptions (per m<sup>3</sup> of H<sub>2</sub> + CO):</i>						
Hard coke	..	..	kg	0.569*	0.490	0.514
Oxygen	..	..	m <sup>3</sup>	nil	0.262	0.260
Steam: to generator	..	..	kg	0.700	0.346	0.890
to CO conversion	..	..	kg	0.504	0.910	0.717
Total steam	..	..	kg	1.204	1.256	1.607
Volume dry catalysed gas	..	..	m <sup>3</sup>	1.50	1.69	1.72
CO <sub>2</sub> to be removed	..	..	m <sup>3</sup>	0.52	0.74	0.76
<i>Example of English costs (per 1,000m<sup>3</sup> of H<sub>2</sub> + CO)†:</i>						
Coke at 60sh/ton	..	..		34.0	29.4	30.8
Oxygen at 40sh/1,000m <sup>3</sup>	..	..		nil	10.5	10.4
LP steam at 4sh/ton	..	..		4.8	5.0	6.4
				38.8	44.9	47.6
Miscellaneous	..	..		nil	(Add cost of compressing extra CO <sub>2</sub> and removal by water wash.)	

\* This figure is probably derived after including a credit for carbon in ash, since the latter is recovered.

† No split of costs is given and those given here are merely intended as an example. Oxygen costs include all overheads and capital charges.

extra compression costs and extra cost of removing  $\text{CO}_2$  by pressure water washing, described later in Section B5(iii).

The above figures comparing the slagging with the non-slagging generator, both using oxygen, agree with theory that a slagging generator is more efficient and gives cheaper gas than a non-slagging generator; because of the much higher temperature of the fuel bed in the slagging generator, steam decomposition is much greater, which, coupled with the lower steam rates and lower  $\text{CO}_2$  content of the water gas, leads to appreciably lower sensible heat losses in the water gas and undecomposed steam. These additional heat losses in the non-slagging generator have to be covered in effect by burning CO to  $\text{CO}_2$ . Thus, although the oxygen consumptions in the two processes are the same, the slagging generator uses less coke.

Like most other oxygen gasification processes, the slagging generator is economic only when using cheap fuel which cannot be gasified in a "make and blow" process; e.g. small fuels, fuels with low ash mp, fuels with high ash content. Thus, in Table X, if fuel were charged to the slagging generators at, say, 40sh/ton, the slagging generator would be more economical than "make and blow" generators, using coke at 60sh/ton. Presumably the process found favour at Leuna because it could utilise the "semi-ash" obtained from the "make and blow" water gas generators, enabling the latter to be run at appreciably higher outputs at the same time.

#### (e) The Pintsch Type of Slagging Generator

Although the gas plant at Scholven is described as consisting only of ordinary coke water gas generators, later information suggests that some of the generators are operated with liquid slag removal. There are eight Pintsch generators, each with a capacity of 8,400m<sup>3</sup>/hr water gas, of which 7,500m<sup>3</sup>/hr is low in nitrogen. They are 3.6m inside diameter, and are of "the hot grate type, with liquid slag removal." The flat grate is made of special cast steel, and slag is removed at 1,000°C from the centre of the grate and also through two diametrically opposite doors on the sides of the grate. The generators, however, do not use oxygen but operate on a 3min cycle, using air. The coke consumption is given as 0.57kg/m<sup>3</sup> water gas, or, say, 0.66kg/m<sup>3</sup>  $\text{H}_2 + \text{CO}$ .

Operation in this manner infers that the ash has a very low melting point. Although more details would be interesting, the efficiencies quoted could be obtained

easily in non-slagging "make and blow" generators, and it does not appear that the Pintsch type of generator offers anything more than a way of using fuel of low ash melting point.

**Conclusions.**—Slagging generators, using oxygen, are well established at Leuna and appear to be superior to Thyssen-Galocsy generators, which, however, have not yet been fully developed. When using coke, slagging generators are not so economic as "make and blow" water gas generators, although they are somewhat more economic than non-slagging generators using oxygen. Slagging generators may become economic when gasifying a cheap fuel, which cannot be used in a "make and blow" generator.

#### ii. WINKLER GENERATORS

The Winkler is now a proven generator and is established in at least five large factories, for the manufacture of water gas and producer gas, used mainly for the production of hydrogen, ammonia and methanol synthesis gas and power gas. All known plants use dry brown coal or brown coal coke as fuel, and it appears that the process is economic, in comparison with alternative processes, only when it can utilise cheap fuels, which cannot be conveniently gasified by other means.

##### a. Winkler Generator Installations

Table XI gives a list of known Winkler generator installations.

In addition, there are small units at Oppau, near Ludwigshafen, as well as at Leuna, operated by IG to test various coals. It is also possible that there are three generators in Japan.

##### b. Description of the Process

Winkler generators differ from all other existing gasification processes in utilising a "boiling bed" of fuel, i.e. using particles of fuel small enough to be almost gas-borne and hence comparatively mobile. Under such conditions the fuel gives an appearance as if the bed were boiling, the bed finds its own level, as does a liquid, and circulation of particles within the bed is such as to give substantially equal temperatures throughout the bed. When making water gas, for manufacture of hydrogen or methanol synthesis gas, the continuous blast is a mixture of oxygen and steam, and when making

TABLE XI  
WINKLER GENERATOR INSTALLATIONS

Plant	Started up	Operating company	Units	Approximate output/unit m <sup>3</sup> /hr water gas	Remarks
<i>Germany:</i>					
Leuna	1926	IG	4	60,000	75,000 on producer gas. Only one unit works on water gas and one on producer gas at one time.
(32km W Leipzig)	1930		1	30,000	
Böhlen— (16km S Leipzig)	1938	Brabag	3	20,000	
Zeitz— (32km SSW Leipzig)	1939	Brabag	3	20,000	
Magdeburg— (80km NW Leipzig)	1939?	Brabag	3	20,000	
<i>Czechoslovakia:</i>					
Brüx— (130km SE Leipzig)	1942	Sudeten-ländische Treibstoffwerke AG	5 or 6	20,000	

# CARBONISATION AND GASIFICATION OF COAL

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TABLE XII

PERFORMANCE OF WINKLER GENERATORS MAKING WATER GAS FROM GRUDE

PLANT	LEUNA	BÜHLEN		ZEITZ	(Figures for unspecified plant quoted by Thau, Oel u. Kohle, 8.6.42)
		Large Generators	Small Generators		
Units	3	3	3	3	—
Actual water gas output	50,000–60,000 m <sup>3</sup> /hr/unit	20,000–25,000	16,600	—	—
Fuel (grude)					
Analysis on dry basis					
C	68.0	—	71	63.7	70.7
H	2.0	—	2.6	2.2	2.7
O	2.2	—	—	—	4.0
N	—	—	1.1	5.5	0.0
volatile S	1.3	—	—	—	0.6
ash	26.5	—	25.1	28.5	22.0
H <sub>2</sub> O	(2.0)	—	(2.1)	(2.6)	(6.4)
Calorific value	te.cal/te nett as received	5,780	5,200–5,400	5,650–5,850	5,700
Final grading					
	Av. 3mm with <10% > 5mm	0.06–5mm	—	2–10mm	mm %
					0–0.2 20
					0.2–0.5 16
					0.5–1.0 20
					1–2 26
					2–3 8
					3–6 10
Gas analysis					
CO <sub>2</sub>	20.0	24.4	24.4	23.1	16.5
CO	37.5	27.6	28.8	29.6	42.6
H <sub>2</sub>	39.5	45.3	44.4	43.8	39.0
CH <sub>4</sub>	1.5	1.5	1.25	0.75	0.7
N <sub>2</sub>	0.5	0.7	0.5	1.5	0.7
H <sub>2</sub> S	1.0	0.5	0.6	1.25	0.5
Calorific value	k.cal/m <sup>3</sup> nett	2,150	1,990	1,982	1,985
O <sub>2</sub> in total blast	%	40–50	21	22	21.5
Blast temperature	°C	—	150	—	160–180
Secondary O <sub>2</sub>	%	33	12–20	—	10
Fuel bed temperature	°C	850–900	900	800	900–950
Gas exit temperature	°C	900–950	900	900	900–1,000
Gas temperature after W.H. recovery	°C	200	200–300	300	200–250
Dust content of gas					
(a) Before dust removal	gm/m <sup>3</sup> dry gas	100–200	200–250	225–250	300–360
(b) After primary dust removal	—	—	—	40–60	60
(c) After final dust removal	—	—	—	0.002–0.005	0.003–0.004
C in dust	%	50–55	40	43.1	54–56
C in ashes	%	—	40	54.1	30
Efficiencies (per 1,000m <sup>3</sup> H <sub>2</sub> + CO):—					
Grude	kg	640	(1) 757 (2) 782	765	1,023
Carbon	kg	427	—	534	635
Oxygen	m <sup>3</sup>	320–335	305	335	324
Steam used	kg	350	825	960	830
Steam raised	kg	580	750	—	850
Steam decomposition	%	—	—	—	925
Carbon utilisation	%	88	—	—	35
Dust blown over	kg	130–260	—	—	57
Dust recovered	kg	—	—	245–270	458
Grate ash	kg	—	21	—	60
Power (excluding oxygen production)	kwh	—	—	39	70
Cooling water	m <sup>3</sup>	—	21	—	34

(1) Average of two days, 11th and 12th August, 1938.

(2) Average of 31 days, January, 1940.

\* In addition, one report quotes about 230kg/1,000m<sup>3</sup> H<sub>2</sub> + CO as being used for "Apparateheizung," i.e. heating of items of plant, at least for the winter months January to March; this use of steam is obscure.

producer gas, for use as power gas, the blast is air alone, without steam. In every case, however, some of the blast is added above the fuel bed.

The bed temperature is maintained at 900° to 1,000°C and gas leaves the generator at about this temperature. The sensible heat of the gas is recovered as steam in waste heat boilers and the considerable amount of dust is removed by multicyclones, followed by direct water scrubbers and Theisen disintegrators.

To illustrate the size and output of a generator, the

units used by Brabag may be taken; the inside diameter of the brick-lined generator is about 4.5m, the overall height 20m and the depth of fuel 1.5m. The output can be varied from a minimum of about 9,000m<sup>3</sup>/hr to a maximum of about 25,000m<sup>3</sup>/hr water gas, although usually 20,000m<sup>3</sup>/hr is not exceeded.

The performance when making water gas from grude (brown coal coke) is summarised in Table XII and from dry brown coal in Table XIII. The performance when making power gas is summarised in Table XIV (p. 21).

TABLE XIII

PERFORMANCE OF WINKLER GENERATORS MAKING WATER GAS FROM DRY BROWN COAL

PLANT		LEUNA				(Unspecified plant, see Table XII)
		Small Generator	Large Generator	12 months' average	one day	
Actual water gas output	m <sup>3</sup> /hr/unit	—	—	27,000	42,000	—
Fuel (dry brown coal):—						
Analysis on dry basis	%					
C		61.1	—	59.3	57.3	61.1
H		4.7	—	4.5	4.5	4.7
O		17.0	—	15.2	14.0	16.3
N		0.1	—	0.7	0.6	0.8
S		3.3	—	3.6	3.1	3.3
ash		13.8	—	16.7	20.6	13.8
H <sub>2</sub> O		(6.0)	—	(8.7)	(8.1)	(8.7)
Calorific value	te.cal/te nett as received	—	—	5,170	5,150	5,270
Final grading	mm	0-0.6	9.6	—	—	mm
		0.6-0.88	1.5	—	—	0-0.2
		0.88-1.0	9.0	—	—	0.2-0.5
		1-2	23.3	—	—	0.5-1.0
		2-5	16.5	—	—	1-2
		>5	0.1	—	—	2-3
						3-6
Gas analysis	%					
CO <sub>2</sub>		19	19	21.8	15.7	17.5
CO		38	38	35.3	44.4	41.8
H <sub>2</sub>		40	40	38.5	36.0	37.2
CH <sub>4</sub>		2	2	1.8	1.6	0.9
N <sub>2</sub>		1	1	1.1	0.8	1.0
H <sub>2</sub> S		?	?	1.5	1.5	1.6
Calorific value	k.cal/m <sup>3</sup> nett	2,162	2,162	2,117	2,295	2,195
O <sub>2</sub> in blast	%	40	—	40	48	40
Blast above fuel bed	°C	33	—	—	—	—
Fuel bed temperature	°C	—	—	—	—	800-950
Gas exit temperature	°C	—	—	—	—	950-1,000
Dust content of gases						
(a) Before dust removal	gm/m <sup>3</sup> dry gas	—	—	110	170	—
(b) After primary dust removal	"	—	—	—	—	—
(c) After final dust removal	"	—	—	—	—	—
C in dust	%	—	—	29	35	—
C in ash	%	—	—	42	40	—
Efficiencies (per 1000m <sup>3</sup> H <sub>2</sub> + CO):						
Coal	kg	800	800	830	855	790
Carbon	kg	461	—	452	455	444
Oxygen	m <sup>3</sup>	320-335	330	366	316	342
Steam used	kg	370	—	407	250	384
Steam raised	kg	—	—	600	—	—
Steam decomposition	%	—	—	33	27	—
Carbon utilisation	%	80.5	—	86.5	80	84
Dust blown over	kg	—	—	148	211	—
Grate ashes	kg	—	—	41	44	—
Power	kwh	—	—	48	—	—

It will be seen that the carbon efficiencies at Zeitz are very poor, due to the large amount of ungasified dust blown overhead. The variable operating conditions make it difficult to summarise the service requirements and efficiencies of the process, but the following list of ranges encountered might be a useful summary of known achieved results:—

	Per 1000m <sup>3</sup> H <sub>2</sub> + CO
Crude	630-1,000kg
Carbon	420-630kg
Oxygen (98%)	305-335m <sup>3</sup>
Steam used	350-900kg
Steam decomposition	30-35%
Carbon utilisation	88-57%
Dry brown coal	800-860kg
Carbon	445-460kg
Oxygen (98%)	315-360m <sup>3</sup>
Steam used	300-400kg
Steam decomposition	27-33%
Carbon utilisation	86-80%

#### c. Suitable Fuels

All the known Winkler installations are using dry brown coal or grude as fuel; these fuels are highly

reactive and relatively cheap (say RM4 to RM9/te). As far as is known there are no large-scale installations using any other fuel. There are reports in the literature that bituminous coal can be used and it is known that trials have been carried out, at least on small generators. With fuels which are less reactive than dry brown coal or grude, higher operating temperatures are required and oxygen consumptions are greater.

#### d. Comparison with other Processes for making Synthesis Gas

As stated before, the Winkler process is able to use low-grade fuels, difficult to gasify in other ways, and this at once may give it a great local advantage. When, however, it is considered for coals which can be gasified in other ways or where it has to use fuels having enhanced value for other purposes, e.g. as a boiler fuel, then it becomes much less attractive.

The two great drawbacks to the Winkler process are its relatively poor thermal efficiency, which is no higher than that of a conventional coke water gas generator, and the cost of oxygen. If fuel is at all expensive, the poor carbon efficiency is a disadvantage which is only

TABLE XIV  
PERFORMANCE OF WINKLER GENERATORS MAKING POWER GAS

PLANT			LEUNA (12 months* average)	(Unspecified plant, figures by Thau, <i>Oel u. Kohle</i> , 8.6.42)	
			Dry Brown Coal	Dry Brown Coal	Grude
Actual output			50,000	—	—
Fuel					
Analysis on dry basis					
	C	%	57.8	61.1	70.7
	H		4.3	4.7	2.7
	O		16.5	16.3	4.0
	N		0.6	0.8	0.0
	S		3.3	3.3	0.6
	ash		17.5	13.8	22.0
	H <sub>2</sub> O		(8.5)	(8.7)	(6.4)
Calorific value			5,190	5,270	5,700
Gas analysis					
	CO <sub>2</sub>	%	9.8	7.7	4.3
	CO		21.7	22.5	32.7
	H <sub>2</sub>		11.7	12.6	7.6
	CH <sub>4</sub>		0.7	0.7	0.5
	N		55.3	55.7	54.6
	H <sub>2</sub> S		0.8	0.8	0.3
Calorific value			984	1,020	1,140
Blast above fuel bed			25	—	—
Fuel bed temperature			950	—	—
Gas exit temperature			1,000	—	—
Dust content of gases					
(a) Before dust removal			65	—	—
(b) After primary dust removal			—	—	—
(c) After final dust removal			—	—	—
C in dust			33-55	—	—
C in ash			34-44	—	—
Efficiencies (per 1,000m <sup>3</sup> gas):—					
	Fuel	kg	330	ca.330	ca.285
	Carbon	kg	176	ca.186	ca.190
	Air	m <sup>3</sup>	700	ca.720	ca.720
	CO <sub>2</sub>	m <sup>3</sup>	25	—	—
	Steam raised	kg	450	—	—
	Carbon utilisation	%	83	ca.82	ca.97*
	Dust blown over	kg	65	—	—
	Grate ashes	kg	18	—	—

\* This appears very high and is probably an error; the fuel consumptions should be greater than shown.

partly mitigated where the dust can be collected and used as a boiler fuel. Even when made in modern Linde-Fränkl units, oxygen is never cheap, and it is interesting that the IG consider that it is more expensive to use continuous oxygen gasification of any fuel than to use a "make and blow" process with air, provided the same fuel can be got into a suitable form in both processes.

All processes using oxygen gasification have an additional disadvantage in that any oxygen added must eventually appear as CO<sub>2</sub> in the synthesis gas; hence greater compression costs and water scrubbing costs are incurred to remove it, with correspondingly increased capital costs.

Although the amount of gas produced per unit shaft area is relatively great (say 900 to 1,500m<sup>3</sup>/hr H<sub>2</sub> + CO/m<sup>2</sup>, as compared with, say, 600m<sup>3</sup>/hr H<sub>2</sub> + CO/m<sup>2</sup> for "make and blow" coke water gas generators), and although the absence of distribution difficulties in a boiling bed enables very large units to be used (e.g. up to 50,000m<sup>3</sup>/hr H<sub>2</sub> + CO at Leuna, compared with, say, up to 8,000m<sup>3</sup>/hr H<sub>2</sub> + CO in the largest coke water gas generators), nevertheless all the auxiliary equipment required means that the output of Winkler generators per unit area of site is no greater than that of coke water gas generators. Thus at Böhlen and Zeitz the capacity of three generators was obtained for a site area of about 50m<sup>2</sup> per 1,000m<sup>3</sup>/hr H<sub>2</sub> + CO installed capacity, including fuel handling,

waste heat boilers, coolers, etc., but excluding the oxygen plant; the oxygen plant and its associated share of the boiler and power plant probably occupied an equal area. A coke water gas plant, with all auxiliaries, would probably make the same amount of gas for a site area of about 40 to 60m<sup>2</sup> per 1,000m<sup>3</sup>/hr H<sub>2</sub> + CO installed capacity. Moreover, the large reaction space above the fuel bed, in terms of capital cost, largely offsets the advantage of a high output per unit area of grate.

#### e. Capital Costs

Available figures are of doubtful significance, but it is probably true to say that the capital cost of a Winkler plant, including oxygen plant, waste heat boilers, etc., is somewhat greater than that of a corresponding coke water gas plant, including coke ovens. The oxygen plant probably costs more than the Winkler plant it supplies.

#### f. Process Labour and Maintenance Costs

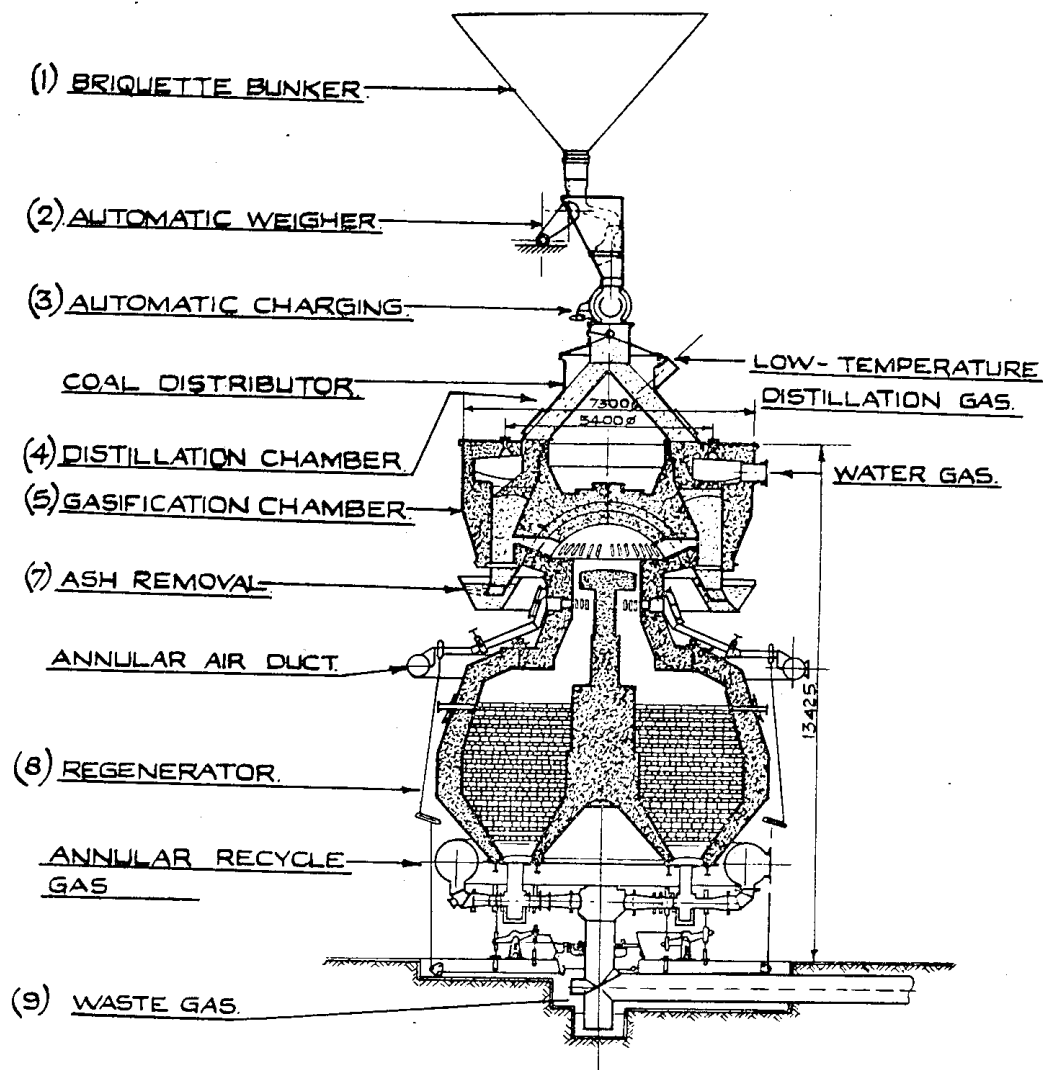
As a rough figure, process labour requirements can be taken as 0.4 man-hours/1,000m<sup>3</sup> H<sub>2</sub> + CO, when making water gas. At Leuna, during 1935 maintenance costs for water gas averaged RM1.75/1,000m<sup>3</sup> of H<sub>2</sub> + CO, for an average output of 20,000m<sup>3</sup>/hr H<sub>2</sub> + CO. Maintenance costs for power gas averaged RM0.46/1,000m<sup>3</sup> power gas, for an average output of 50,000m<sup>3</sup>/hr.

## g. Recent Technical Advances

The main result of the investigations was to obtain actual operating data, both as regards efficiencies and performance, thus supplementing the rather meagre information available in the literature. The only important technical advance to be uncovered was the

throughput of 50-60te/day of brown coal briquettes. This plant has been described in the literature, but operating data were lacking for the latest modification of the process. It was also known that erection of a large installation had been commenced in 1938 somewhere in Western Germany. This installation was found to

FIGURE 8



PINTSCH-HILLEBRAND GENERATOR

development of a *grateless* generator, in which the blast entered through tuyeres in the coned base of the generator, instead of through the normal stationary grate, made of cast iron bricks. There are two such generators at Leuna and at least one at Brück. Leuna claimed that these generators used 10% less oxygen and fuel. However, little additional information was obtained on this development.

## iii. PINTSCH-HILLEBRAND GENERATORS

## a. General

A single Pintsch-Hillebrand generator was installed at the Tiefstack Gas Works, Hamburg, in 1932, with a

be at the high-pressure hydrogenation plant at Wesseling, near Cologne. It consisted of 11 generators, each designed to produce 5,500m<sup>3</sup>/hr water gas for the hydrogenation process.

## b. The Plant and Process

The process consists of the continuous gasification of brown coal briquettes in a stream of recycled gas mixed with the required amount of steam, the gas and steam being preheated, by regenerators, sufficiently high to maintain the temperature of the fuel bed, and to supply the heat of the endothermic reaction of gasification. On its passage to the gasification chamber,

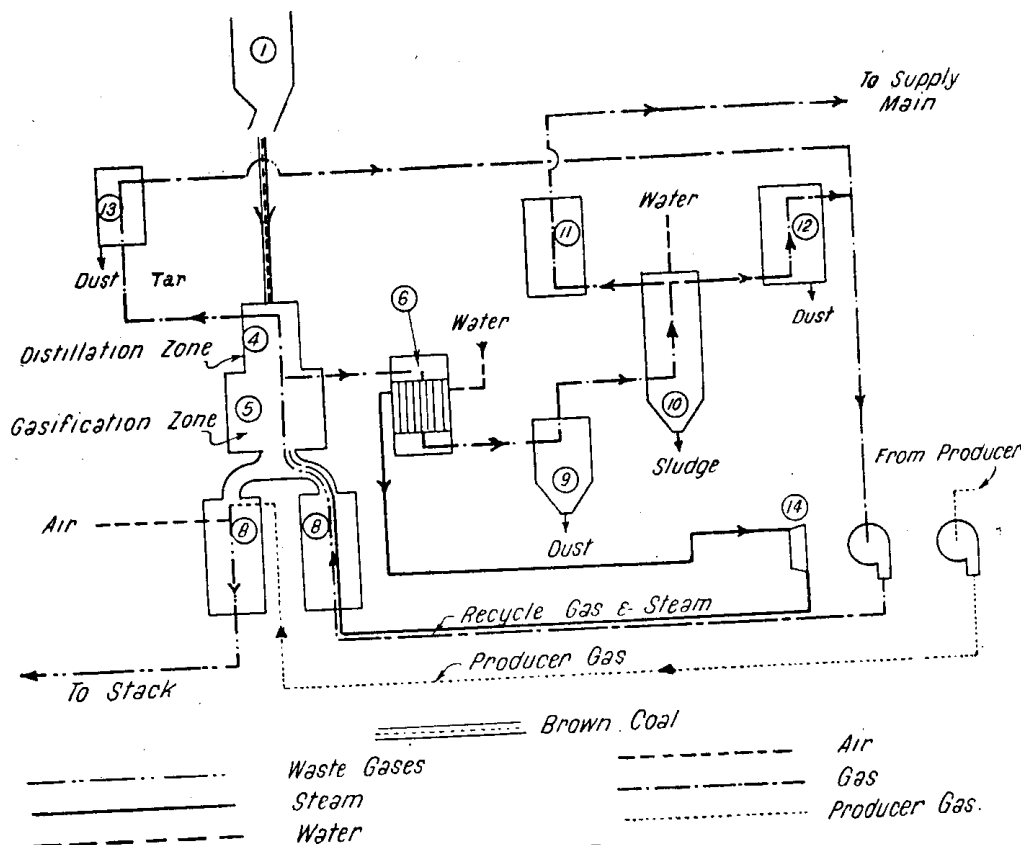
the fuel passes through a low-temperature carbonisation zone, the gas from which is added to the recycled part of the water gas.

The gas generators and regenerators are shown in section in Fig. 8. The brown coal briquettes pass from a bunker (1), through a weigher (2) and automatic charging valve (3) into the top of the conical low-temperature carbonisation zone (4). At the base, the fuel enters the gasification chamber (5), which is an annulus 55cm wide, with an internal diameter of 4.7m and a height of 2.5m. Ash is discharged through a water

disposed in cool positions in their pipe connections and are operated hydraulically from a central platform.

The preheated mixture of recycled gas and steam is distributed at 1,300°C to the base of the gasification zone through a slotted dome, and water gas is made by interaction with the fuel. The water gas leaves the gasification chamber at the top at 700°C. A small proportion of water gas is diverted to the carbonisation chamber, and after effecting low-temperature carbonisation of the fuel, it leaves the top of this zone, charged with the products involved. The temperature of the

FIGURE 9



PINTSCH-HILLEBRAND PROCESS

seal (7). It contains 40–50% carbon. The mixture of recycled gas and steam is preheated by upward passage in one of two regenerators (8). These two chambers each contain 110t refractory brickwork and are used alternately for 15min periods. The other regenerator is meanwhile heated to 1,500–1,600°C at the top brickwork by the combustion of producer gas (inlet not shown in Fig. 8) made in a separate installation and introduced at the top of the heated regenerator together with an air supply. The waste gas leaves the bottom of the heated regenerator and is discharged through a stack. Entry of producer gas into the waste gas is avoided by maintaining a slightly higher pressure in the latter. Thus a small amount of water gas and steam passes back into the regenerator which is being heated, but no valves are required between the regenerators and the gasification chamber. The regulating and reversing valves for the recycled gas, producer gas, air and waste gas, are all

gas leaving the carbonisation zone is maintained at 90°C, which is above the dewpoint, by controlling the amount of water gas passed into the zone.

The gas system is shown in the flow sheet, Fig. 9. The water gas leaves the gasification zone (5), and passes through a waste heat boiler (6), leaving at 230°C, then through a cyclone separator (9) and a water saturator (10), where it is cooled to 70°C. At this point the stream is divided. Part passes through a washer cooler (11) where it is cooled to 20°C, and this gas, containing 1–2mg/m<sup>3</sup> of dust, constitutes the water gas make.

The rest passes through an electrostatic dust precipitator (12) and joins the gas from the carbonisation zone, which has passed through an electrostatic tar precipitator (13). The combined stream is then boosted through a hot regenerator back to the gasification zone. Steam raised in the waste heat boiler is used to drive the booster turbine (14), and the exhaust steam (together with

additional steam from the boiler if required) is mixed with the recycled gas.

#### c. Performance and Efficiency

The Wesseling Pintsch-Hillebrand plant operated smoothly for 3 years before the factory was bombed out of action in 1944. It became evident with experience that satisfactory outputs could be achieved only if the briquettes used resisted mechanical disintegration. Such briquettes are readily manufactured if uniform briquetting pressures are used, and the moisture content of the brown coal is constant. The specific gravity of the briquettes is a criterion of correct pressing: it must be 1.23. The moisture content must lie between 13.5% and 14%. Using these briquettes, 9,400m<sup>3</sup>/hr of gas could be passed through the gasification chamber, the make being 5,400m<sup>3</sup>/hr, whilst 4,000m<sup>3</sup>/hr was recycled.

The operating data for a generator at full output are given in Table XV.

TABLE XV

#### OPERATING DATA FOR PINTSCH-HILLEBRAND GENERATOR

Consumption of Rhineland brown coal briquettes	3,000kg/hr
Steam consumption (of which 60% is added as steam, 40% by saturation of the recycled gas)	4,200kg/hr
Producer gas consumption	3,000m <sup>3</sup>
Net calorific value of producer gas	1,460k.cal/m <sup>3</sup>
Water gas produced	5,400m <sup>3</sup> /hr
Recycled gas (of which 70-90% consisted of gas from the carbonisation zone, and up to 30% was bled off from the water gas line)	4,000m <sup>3</sup> /hr
Analysis of water gas: 14% CO <sub>2</sub> , 28% CO, 56% H <sub>2</sub> , 1% CH <sub>4</sub> , 1% N <sub>2</sub> .	
Low-temperature distillation gas: 17.5% CO <sub>2</sub> , 0.8% C <sub>2</sub> H <sub>6</sub> , 26% CO, 48% H <sub>2</sub> , 6.5% CH <sub>4</sub> and 1.2% N <sub>2</sub> .	

Table XVI gives the yields in 1943 and 1944 together with production costs.

TABLE XVI

#### PINTSCH-HILLEBRAND YIELDS AND COSTS

	1943	1944
Total water gas production/hr	48,600m <sup>3</sup>	52,000m <sup>3</sup>
Average production/generator/hr	4,950m <sup>3</sup>	5,100m <sup>3</sup>
Yield of water gas from briquettes	1,938m <sup>3</sup> /te	1,900m <sup>3</sup> /te
Producer gas used per unit volume of water gas made	828k.cal/m <sup>3</sup>	825k.cal/m <sup>3</sup>
Tar recovery, % by weight of briquettes	4.8	4.9
Steam requirement from external source, per unit volume of water gas made	0.126kg/m <sup>3</sup>	0.125kg/m <sup>3</sup>
H <sub>2</sub> + CO in water gas	83.5%	84%

Costs calculated on the H <sub>2</sub> + CO in gas	RM/1,000m <sup>3</sup>
Installation costs, including energy	16.15
Credit for by-products	-2.27
Operating costs	3.33
Capital charges	7.64
General expenses	3.11
	27.96

The output of the Pintsch-Hillebrand plant was, however, still the limiting factor in synthetic fuel production, and the capacity was increased by decreasing the recycled gas rate to 3,000m<sup>3</sup>/hr, thus increasing the make to 6,400m<sup>3</sup>/hr per generator. The heat lost by this reduction in recycled gas was made good by adding 400m<sup>3</sup>/hr tail gas from the hydrogenation plant.

#### iv. WINTERSHALL-SCHMALFELDT GENERATORS

##### a. General

Only one factory is known to work the Wintershall-Schmalfeldt process and this is at Lützkendorf, near

Mücheln. A full description of this plant has already been published in CIOS Report XXXII-90, Wintershall AG, Lützkendorf, by H. Hollings and R. Morley, so that little more than a summary is required here.

There are really two Schmalfeldt processes, one making producer gas and one synthesis gas, although at Lützkendorf they are complementary. Their special feature is the gasification of brown coal dust whilst wholly gas-borne in a highly preheated gasifying medium; there is no fixed boiling bed as in a Winkler generator. In the producers, brown coal dust is gasified with preheated air and steam, and the producer gas obtained is used as fuel gas, largely in the synthesis gas units. In the synthesis gas units, brown coal dust is gasified at about 1,000°C in a current of steam and recirculated synthesis gas, preheated to 1,300°C in chequer-filled regenerators, in turn heated with producer gas and air; oxygen can also be added to assist the gasification, although this is not essential; the sensible heat of the final synthesis gas is utilised in drying the raw brown coal needed for both the producers and synthesis gas units. After purification, synthesis gas is used both for Fischer-Tropsch synthesis and for the manufacture of hydrogen.

At Lützkendorf there were five producer units, each making at most 30,000 to 35,000m<sup>3</sup>/hr of producer gas, although normally making less. There were four synthesis gas units, each capable of making 30,000m<sup>3</sup>/hr synthesis gas when using oxygen, but only 15,000m<sup>3</sup>/hr when not doing so.

Typical operating figures are as follows:—

Raw coal analysis:—	% wt	
	As received	Dry basis
C	50-54	60
H	5-6	4
O	4-5	18-20
S	2-2.5	3
N	1-1.5	1
Ash	?	12
Synthesis gas analysis:—	% vol	
CO	25	
H <sub>2</sub>	50	
CO <sub>2</sub>	18-19	
CH <sub>4</sub>	3-5	
N <sub>2</sub>	3-5	

Efficiencies per 1,000m<sup>3</sup> H<sub>2</sub> + CO, including the requirements of producer gas allocated to synthesis gas manufacture:—

Carbon	1.0te	} including coal handling but excluding oxygen plant
Steam	0.9te	
Oxygen	133m <sup>3</sup>	
Nitrogen	210m <sup>3</sup>	
Electricity	90kwh	
Process labour	about 2.0 man-hours	}
Direct maintenance labour	about 0.6 man-hours	

The high fuel consumptions will be noticed; this arises mainly through the large loss of ungasified dust from both synthesis gas and producer gas units. The labour requirements are also high, mainly due to the several kinds of operation carried out and to the many points of control. The inert content (CH<sub>4</sub> + N<sub>2</sub>) is high, because drying and carbonisation products appear in the make gas, along with some flue gases purged from the regenerators.

##### b. Operating Difficulties

Since this installation was the first of its kind, a considerable number of operating difficulties were met, but these have now been largely overcome. Dust losses, however, remain high, and some trouble is still experienced with fusion of ash near the hottest points of the



generators. Serious difficulties in the Fischer-Tropsch plant were experienced, due to the presence of organic sulphur compounds in the gas, difficult to remove by ordinary means; this feature, however, caused no trouble in using the gas for making hydrogen for hydrogenation.

#### c. Comparison with Winkler Generators

The Schmalfeldt process allows run-of-mine raw brown coal to be charged directly to the generator; gasification is carried out in the entrained state and most of the heat of reaction is obtained by storing heat in regenerators, supplied by burning producer gas. The Winkler process must be supplied at least with dry brown coal; it uses a "boiling bed" and all the heat of reaction is supplied by burning fuel with oxygen simultaneously. Data for a complete comparison between Schmalfeldt and Winkler generators are not yet available, since this can really be done only on a cost basis. However, it is almost certain that the Schmalfeldt process is more expensive in running costs. The lower oxygen requirements do not offset the higher direct fuel and steam usage; process labour charges are greater and probably capital costs are at least as great. The high content of inerts ( $\text{CH}_4 + \text{N}_2 = 7\%$ , compared with 3% with Winkler generators) also represents a serious debit.

#### v. KOPPERS GENERATORS

The Koppers process for the production of synthesis gas from brown coal briquettes is in some ways similar to the Pintsch-Hillebrand process, but differs from it in two main features: (1) in the Koppers process only the gases leaving the carbonisation zone are recycled; (2) the Koppers process uses a pair of "Cowpers" and combustion chambers which are external to the generator.

Brown coal briquettes are charged into generators (or gasifiers), which are vertical chambers, the upper part of which forms a carbonisation zone, whilst in the lower part most of the coke produced is gasified in a stream of recycled gas heated to  $1,100^\circ\text{C}$ , using producer gas made from the residual coke discharged from the chambers in order to effect the heating of the recycled gas.

The system is illustrated in the flow diagram, Fig. 10. The gas drawn off from the top of the chamber, at about  $120^\circ\text{C}$ , passes through a tar separator to a blower. The gas is then mixed with steam, and passes up a hot "Cowper," followed by a combustion chamber, after which it enters the generator at  $1,100^\circ\text{C}$ . The synthesis gas is drawn off from the top of the lower zone at  $700^\circ\text{C}$ , and passes through a waste heat boiler, a cyclone separator, a washer, a Theisen washer and a washer cooler, to the blower.

The coke which escapes gasification by the heated mixture of steam and recycled gas is drawn off, and used to make producer gas in a separate installation. This producer gas passes into that combustion chamber which is not in use in the recycled gas stream; there it meets preheated air. The waste gases at  $1,500^\circ\text{C}$  pass down through the "Cowper" to be heated, and then away to a stack.

The yield of synthesis gas is  $1,200\text{m}^3/\text{te}$  of briquettes, with briquettes containing 14.5% water, 44.5% volatile matter and 54.5% carbon. The synthesis gas contains 84.9%  $\text{H}_2 + \text{CO}$ , the ratio of  $\text{H}_2 : \text{CO}$  being 2 : 1. Other quantities are given in the flow diagram.

Koppers had also developed a method of gasification of brown coal in the powdered form. During the missions to Germany, files were found in the Essen office of the Koppers Company, which revealed that a detailed proposal for a plant to produce  $100,000\text{m}^3/\text{hr}$  of synthesis

gas from powdered Upper Silesian coal or low-temperature coke had been submitted to Brabag in 1943. Very little information on the actual process was obtained.

#### vi. DIDIER-BUBIAG PROCESS

Before the Oil Mission visited Germany it was known (or claimed) that the Brabag Fischer-Tropsch plant at Ruhland-Schwarzheide obtained its synthesis gas from brown coal briquettes partly by the Didier-Bubiag and partly by the Koppers-Spülgas processes. German documents originating from the Ruhrchemie AG confirm this and indicate that four-fifths of the synthesis gas was provided by the Koppers plant and the remainder by the Didier plant.

The only information concerning the operation of the latter plant is contained in a Ruhrchemie document dated 1942. In this document it is stated that the Didier plant consisted of 80 chambers, 72 of which were in operation at that time and giving  $30,000\text{m}^3/\text{hr}$  synthesis gas. The chambers needed relining every two years and spraying with heat-resisting material at more frequent intervals to prevent disintegration. The costs of synthesis gas from the Didier and Koppers plant were given as follows:—

		Cost of synthesis gas, pfg/m <sup>3</sup>	
		1940	1942
Didier plant	..	3.0	2.8
Koppers "	..	1.8-1.9	2.2

The increased cost of the Koppers gas in 1942 appears to be due to the fact that at that time a proportion of coke from bituminous coal was being used as raw material.

Among documents obtained in Italy by the Ministry of Economic Warfare in 1944 was a brochure describing the production of synthesis gas by the Didier-Bubiag process based, apparently, on the plant erected at Ruhland.

#### vii. OXYGEN PRODUCTION

##### a. General

The economics of the oxygen-gasification processes already described depend largely upon the costs of production of oxygen. Information on this subject collected by the teams is therefore of considerable interest, and enables a fairly good assessment to be made of the present state of development of oxygen-producing plant.

All the information available deals with the well-known Linde-Fränkl process of air separation after partial liquefaction. There appear to be few fresh developments, beyond those already described in the literature, but actual operating results add to the state of knowledge of the process and especially confirm that power consumptions and running costs are appreciably greater in practice than those claimed by Linde, and in the technical press.

Practically all the large oxygen units are operating in connection with the previously discussed gasification processes.

##### b. Location and Details of Plants

The Linde's Eismachinen Gesellschaft has built about 900 air separation plants of all sizes since 1903, and about 76 Linde-Fränkl units; all but seven of the Linde-Fränkl units are located in Germany or in what was German-occupied territory, and the locations of most of these are known.

Table XVII (p. 27) summarises the information about the various plants.

**CHARGE:**

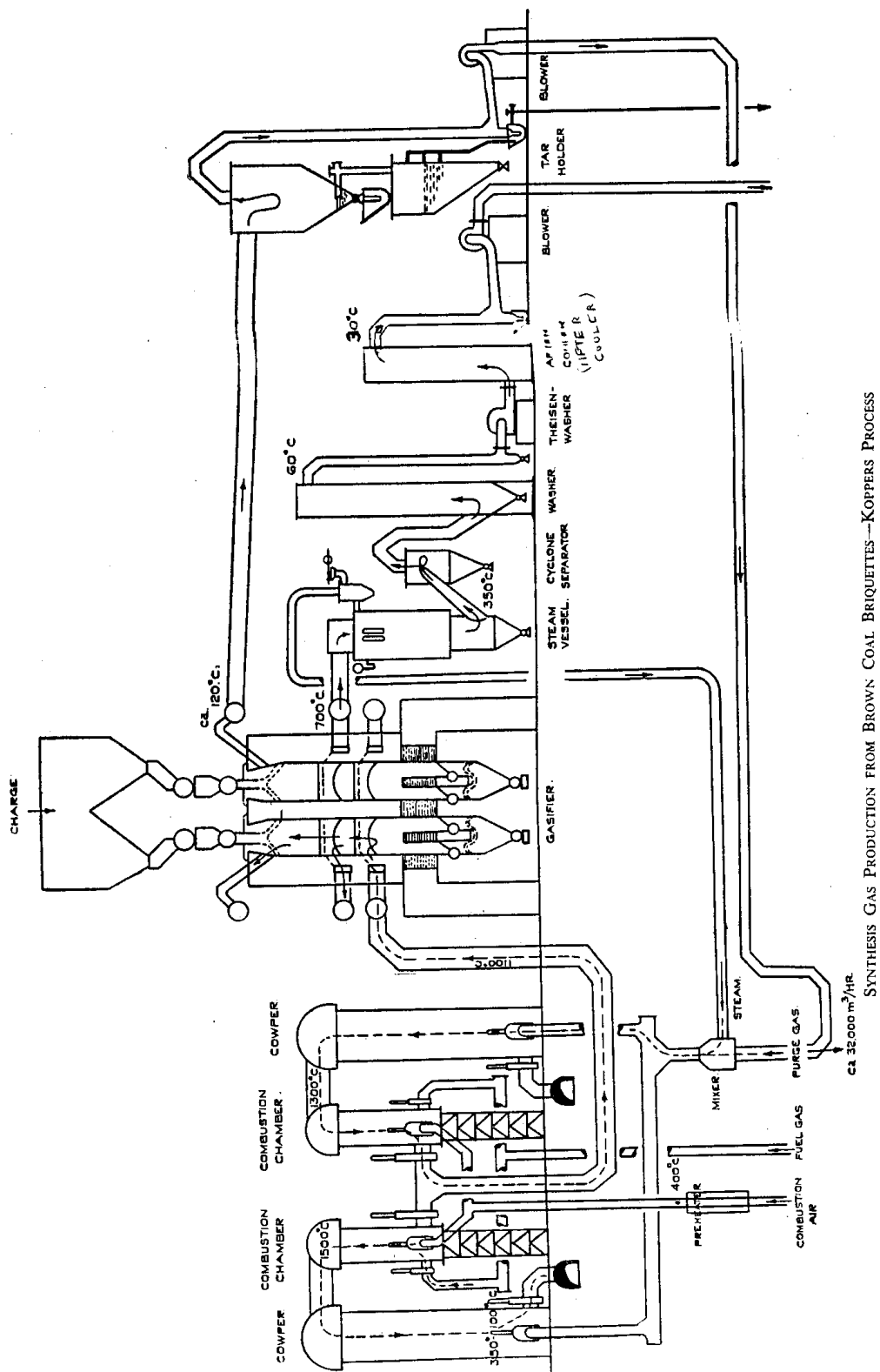


TABLE XVII  
DETAILS OF LINDE-FRÄNKL PLANTS

Location	Operating company	Installed	No. of units	Probable nominal O <sub>2</sub> capacity m <sup>3</sup> /hr		Use
				Unit	Total	
<i>Outside Germany:</i> Mosice, Poland	PFZA	1932	1	500	500	—
Nippon-Tar, Japan	—	1936	1	2,200	2,200	—
Leningrad, USSR	—	1936	2	3,500	7,000	—
Ajka, Hungary	V. Glühlampen u. Elektrizitäts AG, Ujpest	1936	—	(air processed 25,000)	—	Krypton producing plant
Irvine, Scotland	MOS	1937	2	930	1,860	Nitric acid
<i>Inside Germany and Leuna</i>	German-Occupied IG	Territory:— before 1928	2	2,000	27,250	Winkler generators and slagging generators; some N <sub>2</sub> used in NH <sub>3</sub> manufacture
		1932-5	2	2,800		
		1943	5	2,800		
Böhlen	Brabag	1938	1	3,650	11,500	Winkler generators
Böhlen	ASW	—	5	2,300		
		—	2	1,000		
		—	2	2,000	6,000	Lurgi pressure generators
Zeitz	Brabag	1938	6	3,000		
Magdeburg	Brabag	1939?	5 or 6	3,000		
Lützkendorf	Wintershall	1942-3	3	4,000	12,000	Schmalfeldt generators
Brüx, Czech.	Sudetendeutsche Triebstoffwerke A.G.	1942	10	3,300	33,000	Winkler generators
		Total	44	Average 2,850	125,750	
Location	Operating company		Use			
Oppau	IG		O <sub>2</sub> -CH <sub>4</sub> -H <sub>2</sub> O, etc.			
Heydebreck, Poland	IG		O <sub>2</sub> -CH <sub>4</sub> -H <sub>2</sub> O			
Auschwitz, Poland	Ostmärkische Stickstoff W.		O <sub>2</sub> -air-CH <sub>4</sub> -H <sub>2</sub> O for NH <sub>3</sub> synthesis			
Nr. Wanne-Eickel	Stickstoff-Hibernia		N <sub>2</sub> for NH <sub>3</sub> synthesis			

#### c. Description of Process

As the broad principles are well known only a very brief description will be given here; full details will be found in BIOS Final Report 591. Usually there are two air systems, one working at about 5 to 6ats absolute, and comprising the bulk of the air, and the other working at 160-200ats, comprising only 6-8% of the air; in some plants there is a third air system, working at about only 1.5ats absolute.

The high pressure air system, which uses reciprocating engines, forms the most important source of refrigeration, sometimes assisted by a liquid ammonia refrigeration system, and operates either by isentropic expansion (i.e. Joule-Thomson effect) or else by expansion in a turbine or reciprocating engine, with the performance of external work; this expansion process produces a certain yield of liquid oxygen and nitrogen.

The medium pressure system, which uses turbo-compressors, comprises the bulk of the air which passes in and out of the system as gas; heat is interchanged very efficiently by means of Fränkl regenerators, working on a cycle, and these incidentally remove the bulk of the water and CO<sub>2</sub> as solids from the incoming air, and give them up again to the outgoing nitrogen and oxygen fractions, so rendering it unnecessary to purify chemically anything but the high-pressure air.

The well-known Linde double-column enables both a reasonably pure nitrogen and a reasonably pure oxygen (97 to 99%) to be produced simultaneously. The lower column works at about 5 to 6ats and the upper column at about 1.5ats, and in some plants additional air at 1.5ats is introduced into the upper column, since theoretically this slightly reduces the energy requirements.

#### d. Energy Requirements

The cost of energy may account for up to one-half the cost of oxygen, so that the energy requirements are of great importance. The theoretical minimum requirements are about 0.07kwh/m<sup>3</sup>, but the practical approach to this is very poor. A figure of 0.45kwh/m<sup>3</sup> is often claimed for the Linde-Fränkl process. This figure probably refers to power at the compressor shaft, so this would be equivalent to about 0.50kwh to the motor, but even so no case is known where such efficiencies have been obtained. Table XVIII summarises the information obtained by the investigators from actual plant records. In some cases the figures are a little high because of the simultaneous production of pure nitrogen, sometimes because of the inclusion of nitrogen and oxygen fans, and sometimes because of the withdrawal of small amounts of liquid oxygen; these effects are small, however, and in no case is power for compression of oxygen included.

TABLE XVIII

#### ENERGY COSTS OF LINDE-FRÄNKL PROCESS

Plant	Period considered	Unit size m <sup>3</sup> /hr	Power to motor kwh/m <sup>3</sup> 100% O <sub>2</sub>
Leuna	4 years	{ 2 × 2,000 7 × 2,800 }	0.59-0.63
Böhlen (Brabag)	1 month	5 × 2,300	0.69
Böhlen (ASW)	—	{ 2 × 1,000 2 × 2,000 }	0.9-1.0
Zeitz	3 months	6 × 3,000	0.64-0.675
Lützkendorf	1 month	3 × 4,000	0.62

The high energy consumption at Böhlen (ASW) was attributed to the poor performance of the turbo-compressors, but apart from this it appears that achieved power consumptions are over 0.60kwh/m<sup>3</sup>. Lurgi were

TABLE XIX  
COSTS OF OXYGEN PRODUCTION

Plant	Power to motor kwh/m <sup>3</sup>	Power pfg/kwh	Cost of oxygen, pfg/m <sup>3</sup>			Total
			Power	Other costs	Capital charges	
Linde .. .. .	0.50	1.5	0.75	0.15	0.6	1.5
Böhlen-Brabag .. .. .	0.69	1.5	1.04	—	—	3.0
Leuna .. .. .	0.60	2.0	1.2	—	—	1.7-2.0
Heydebreck .. .. .	—	1.4	—	—	—	2.2
Böhlen, ASW .. .. .	0.9-1.0	1.1	0.9-1.0	0.4-0.5	0.9-1.0	2.2-2.5
Wanne-Eickel .. .. .	—	—	—	—	—	2.3
Estimated by Koppers .. .. .	—	2.0	—	—	—	2.5

using 0.67kwh/m<sup>3</sup> for purposes of estimating in 1942. From an analysis of the figures it appears that the main reason for the higher consumptions is the rather low efficiency of the turbo-compressors, and if this could be improved up to that obtained with larger machines, then the figure might well drop to 0.5kwh/m<sup>3</sup>, as claimed by Linde.

#### e. Capital Cost

Information is scanty and unreliable, but the capital cost of large units appears to be of the order of RM500 per m<sup>3</sup>/hr.

#### f. Operating Costs

Although several figures are available, it is difficult to assess their value. Table XIX summarises the information.

Thus the achieved operating cost for big units is of the order of 2.0 to 2.5pfg/m<sup>3</sup>, with power at 1.5pfg/kwh, but the official Linde estimate appears to be low.

To translate these costs into sterling a conversion figure of RM10 = £1 might be used, but costs derived in this way should be regarded with considerable reserve.

### 4. SYNTHESIS GAS PRODUCTION FROM HYDROCARBON GASES

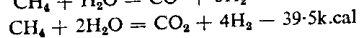
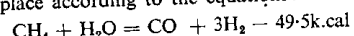
#### i. METHANE-STEAM PROCESS

##### a. General

The methane-steam process for the manufacture of hydrogen and CO from hydrocarbon gases has been widely described in pre-war technical literature, and the subject is therefore not treated in great detail in this section. Two plants of this type were installed in Germany, namely at Wesseling and at Stettin-Pölitz. The data given below refer to the achieved performance of the Wesseling plant.

##### b. Process

The feed gas to the methane-steam process at Wesseling consists of tail gas from the hydrogenation process. The methane-rich feed gas previously purified to a sulphur content of 5mg/m<sup>3</sup> is mixed with steam, and passed over a nickel catalyst at 600-700°C when reaction takes place according to the equations:—



##### c. Equipment

The Wesseling plant consists of four units each of 66 vertical reaction tubes arranged in three groups of 22 tubes and contained in producer gas heated furnaces. The incoming hydrocarbon gas and the steam are heated to 420°C in separate heat exchangers by the outgoing reaction gas at 750°C. The producer gas used as fuel is made from brown coal briquettes, has a calorific value

of 1,450k.cal/m<sup>3</sup>, and is also similarly preheated. The hydrocarbon gas to be reacted is passed at 420°C through four catalyst chambers in series. The first is filled with iron oxide, the second zinc oxide, the third with alternate layers of iron and zinc oxides, and the fourth with spent nickel catalyst from the methane-steam reactors. Iron oxide serves mainly to catalyse the conversion of organic sulphur compounds in the feed to H<sub>2</sub>S, zinc oxide mainly to absorb this by forming the sulphide, and the spent nickel catalyst, by promoting both these reactions, serves to bring about the final stages of purification. This treatment reduces the total sulphur content from 50mg/m<sup>3</sup> down to 5mg/m<sup>3</sup>. Purified gas, still at a temperature of about 420°C, is joined by preheated steam, the steam/hydrocarbon gas ratio being approximately 1.2kg steam/m<sup>3</sup> gas. The steam/hydrocarbon mixture then passes to the reaction tubes which are of Krupps' NCT6 (but NCT10 is preferred at Stettin-Pölitz). The catalyst is in ring form, 16mm × 8mm × 12.5mm.

The exit gas, which contains about 2% of residual methane, is cooled by passage through the heat exchangers, and passes to the main CO conversion plant. (See Section B5(ii).)

There are eight burners per furnace. The exit flue gases from the furnace pass first through a waste heat boiler and then through a preheater which heats the air used for combustion in the furnace. The flue gas then goes to stack. By the addition of an economiser to the waste heat boilers, the plant at Wesseling was made nearly self-supporting as regards steam. Dust in the incoming fuel gas led to trouble with the producer gas heat exchanger, necessitating a quarterly shut-down of one week for cleaning.

##### d. Operating Results

	1943	1944
Total cracked gas produced, m <sup>3</sup> /hr ..	33,600	40,000
Volume ratio of product gas to feed gas ..	2.717	2.400
Heating gas requirement, k.cal/m <sup>3</sup> of product gas ..	902	850
3.5ats steam required, kg/m <sup>3</sup> of product gas ..	0.554	0.550
18ats steam produced, kg/m <sup>3</sup> of product gas ..	0.320	0.450
% methane residue in product gas ..	2.7	1.5

##### e. References

The data on the methane-steam process collected during the course of the main missions were limited to the very brief account given in CIOS Reports XXVII-60 and XXVIII-40 on Wesseling. For the above summary, the information in the Wesseling reports has been supplemented by data supplied verbally by members of a Reparations Mission.

#### ii. KW PROCESS

##### a. General Process Description

The methane-oxygen or KW process for production of synthesis gas from hydrocarbon gases was developed