

REDUCTION OF IRON ORES USING NON-COKING COAL AS THE REDUCTION AGENT

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Introduction

Coke made from metallurgical coal using conventional coking processes is still the main reagent for producing pig iron from iron ore in the blast furnace.

However, the required coal grades are available only at certain points on the earth, and the resources of some deposits are limited. Metallurgical coke prices exhibit a rising trend.

Substitute materials such as natural gas, fuel oil, naphtha, etc. are employed in the blast furnace process to reduce the specific coke rate in relation to the ton of pig iron produced. Again, reduction gases made from these substitute materials are also being used on an increasing scale as the basis of modern direct reduction processes (sponge iron production).

The increasing use of these materials has led to progressive scarcity and a rising price trend.

In contrast, almost unlimited quantities of non-coking coal are available all over the world; these cannot be treated using conventional coking processes.

It is therefore an unavoidable fact that metallurgists must bring their knowledge to bear on the hitherto scarcely utilized possibilities of employing non-coking coal as heating and reduction materials for the production of pig iron and steel.

This paper is intended to constitute a contribution towards the solving of the attendant problems.

Non-coking coal as reduction agent for direct reduction processes

Commercially employed processes for removing oxygen from iron ore require solid carbon or the gas components CO and H₂ as reduction agent; the latter can also naturally be won by gasification of the carbon.

It is therefore expedient to consider the use of non-coking coal, firstly on a direct basis, secondly as a gasification agent for the production of reduction components CO and H₂.

Direct reduction processes using solid carbon bearers.

Retort processes

Naturally, the simplest method is to mix the iron ores with carbon bearers and sulphur-binding substances such as limestone or dolomite in vessels and apply heat to the latter from the outside to attain reduction temperature. Given sufficient time, very high degrees of reduction can be attained. After discharge of the vessels, magnetic separation is applied to collect the iron. This indirect heating system and the required process time involve a considerable calorie consumption rate, which limits the application of the processes both in scale and profitability; thus, for all practical intents and purposes, they are suitable only under special conditions. Two of these processes are briefly discussed by way of example, i.e. the Hogenas and the Echeverria processes.

With both processes, iron ores are mixed with solid carbon bearers and reduced by heating.

Rotating cylindrical furnace processes

The disadvantages of the discontinuous retort processes as discussed earlier, particularly as regards energy consumption, are in high degree eliminated by the rotating cylindrical furnace processes.

SL/RN and Krupp sponge iron processes

The principle of the rotating cylindrical furnace process is that lump, pelletized, or agglomerated iron ore is mixed with lump or fine carbon bearers; heating is normally effected on the counter-flow principle to generate hot gases. The reduced burden is discharged hot from the cylindrical furnace with a degree of pre-reduction of up to 95–98%, and then cooled in a separate, smaller rotating cylindrical kiln.

The gases leave the cylindrical furnace with a relatively high temperature. Under favourable conditions, some of the waste-gas volume can be used for pre-heating, sintering, pelletizing, or the production of hot air.

Further details would exceed the scope of this paper.

GHH-Freeman process

This is a rotating cylindrical furnace process based specifically on the use of the uniflow principle (Fig 1).

As shown by Fig. 1, the fine coal, e.g. lignite, is pneumatically conveyed on the ore charging side. The volume of air used is metered to the point where a reducing atmosphere is created. Temperatures are controlled and a reducing atmosphere created over the entire length of the furnace by burning coal fines with metered air volumes over the full length of the furnace.

As compared with other rotary cylindrical furnace processes, the furnace has a shorter length, which results in high waste gas temperatures and waste gas losses (waste gas temperatures approx. 1100 °C).

The principle of the GHH—Freeman process, and also the explanation for the short furnace length, is the preheating of the burden to about 1000 °C on a travelling grate of the appropriate size.

As illustrated in Fig. 1, the furnace gases travel in the same direction as the charge of pellets heated to approx. 1000 °C, the product being cooled to below 200 °C in a cooling drum to prevent reoxydation, in the same way as with the other rotary cylindrical furnace processes.

From the very outset, the temperature in the reduction chamber is 1000 to 1100 °C, which explains the

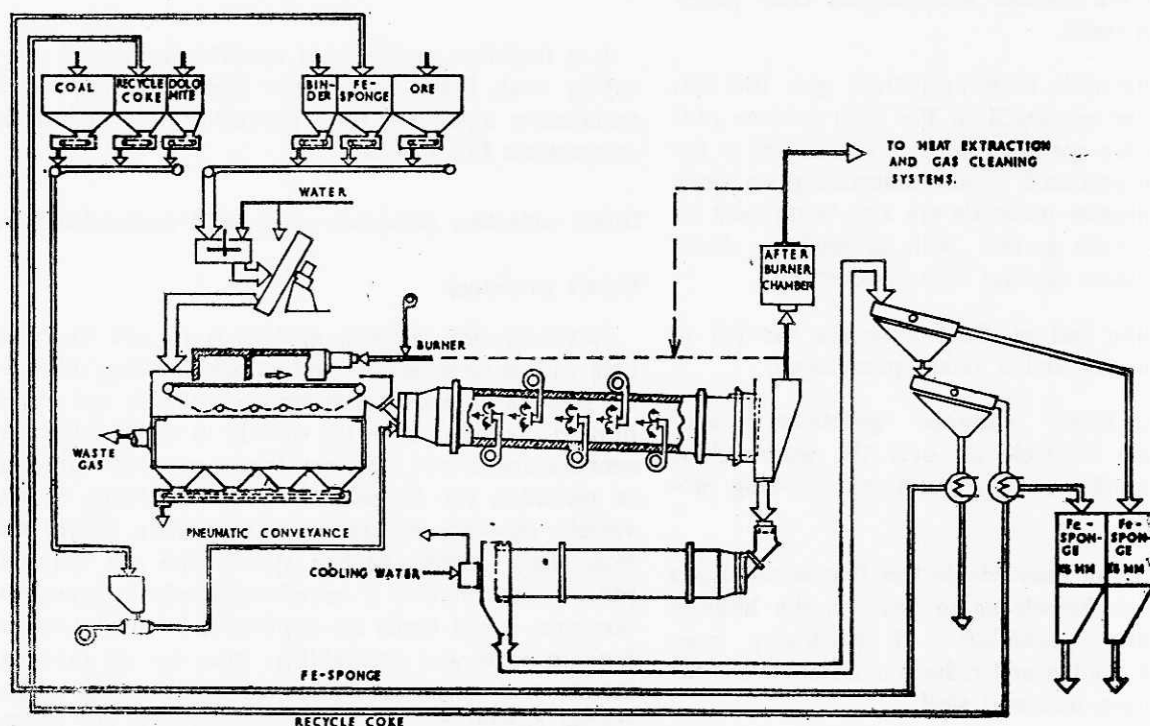


FIG. 1 THE GHH-FREEMAN ROTATING CYLINDRICAL FURNACE PROCESS FOR THE DIRECT REDUCTION OF IRON ORES.

shortness of the cylindrical furnace. For this reason, the furnace also has a higher degree of filling than is true of the other processes of this type, i.e. almost 40%.

In view of the special features of the process, which employs inexpensive lignitiferous coal, a material which appears to be particularly suitable for this process, it is thought that a furnace having a length of 30 to 40 m and an inside diameter of approx. 6.5 m is sufficient for the production of 1 million tonnes of sponge iron; this could yield appropriately low investment costs.

The calorie consumption rate is above 4 Gcal/t Fe.

Gas reduction processes on the basis of non-coking coal

It is rather interesting to examine coal gasification as the basis of gas reduction processes. Of course, not all types of coal gasification processes can be discussed in this paper. Only such possibilities are discussed as have been the subject of publications made recently in connection with direct reduction processes.

Table 1 shows the material and heat transformation values applicable when using various gasification media.

Gasification of carbon bearers using carbonic acid gas is seldom applied, since the heat input of the heavily endothermal gasification reaction including heating of the carbon bearer and the recycle gas as carbonic acid gas bearer to reduction temperature incurs difficulties.

Water vapour as a gasification agent is in the main used only for cooling oxygen gasification or creating and enriching the H_2 content in these processes, disregarding the old principle of water gas production.

The gasification of carbon bearers using oxygen or air is a feature of interesting new developments. The increasing shortage of fuel oil and natural gas, and the resulting tendency towards rising prices in this sector make the gasification of solid carbon bearers for the production of reduction gases for direct reduction processes an interesting subject. The modification of the existing process principles as involved by conversion to carbon gasification is discussed later.

Production of reduction gases from non-coking coal Flame gasification processes

One problem attending the gasification of solid carbon bearers is the handling of the ash components, especially at temperatures exceeding the softening point of the ash. Work aimed at solving the difficulties of gasifying baking coal has led to the "uniflow gasification processes". These processes are virtually a transition to the flame gasification processes in which crack gas is produced from gaseous or liquid hydrocarbons and in which similar facilities are employed.

The basic principle of these processes is that the medium to be gasified and the gasification medium always move in one and the same direction, i.e. coal dust and gasification medium flow through the gasifier in the same direction.

High reaction temperatures are a prerequisite. The degasification products are almost completely cracked. Consequently, these processes can also be used for cracking heavy fuel oil and natural gas. Usually, these processes employ oxygen as the gasification medium to attain the high temperatures needed.

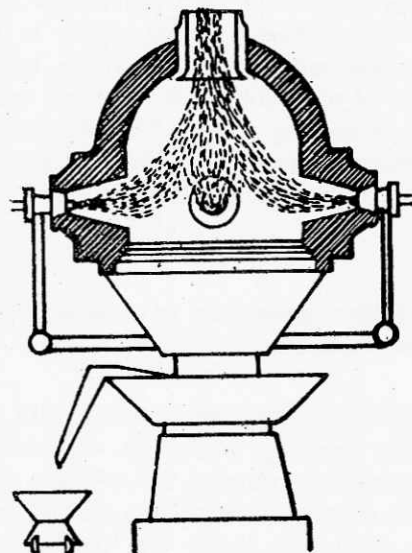
Table 1
Stoichiometric gas compositions and reaction enthalpies in the gasification of carbon bearers using various gasification media

Gasification medium	Reaction formula	Stoichiometric gas composition			Volume in Nm ³ kg. C	Reaction enthalpy ΔH_{293} in kcal/kg
		% CO	% H ₂	% N ₂		
Oxygen	$C + 1/2 O_2 = CO$	100	—	—	1.865	-2 450
Atmospheric oxygen	$C + 1/2 O_2 = CO$	35	—	65	5.371	-2 450
Water vapour	$C + H_2O = CO + H_2$	50	50	—	3.730	+2 511
Carbonic acid gas	$C + CO_2 = 2 CO$	100	—	—	3.730	+3 346

Some of these processes employ low pressures, some high pressures.

Fig. 2 shows a principle diagram of a Koppers Totzek gasifier.

Fig. 2



PRINCIPLE DIAGRAM OF A KOPPERS

In this process, the gasifier is of horizontal design; from both sides, mixtures of the material to be gasified are first fed in with a particle size of less than 0.1 mm through nozzles together with oxygen. After entering the gasification chamber, which is refractory

lined and features a high temperature, gasification immediately commences (flame gasification).

The danger of an explosion of the dangerous mixture of oxygen and coal dust is prevented by controlling the process in such a way that the flow velocity in the feed pipe is higher than that of ignition, which means that conversion does not take place until the gasification chamber proper is reached.

Coal dust explosions can be prevented by filling the bins with nitrogen, the latter then also being available for conveying purposes.

Water vapour with (atmospheric) oxygen, which is also used as an additional gasification agent, is blown through an annular nozzle in such a way that it envelops the reaction zone between coal dust and oxygen (protection of the refractories). (Endothermal gasification reaction).

Ash components which do not leave the generator in the form of dust, which is collected in a cleaning system, are granulated in liquid form and discharged (approx. 50%).

Attention is drawn to the need for a fully efficient control and measuring system, and in particular for adjusting the proportions of coal dust, oxygen, and water vapour.

Table 2 shows examples of gas analyses obtained using this process.

TABLE 2
Analyses of gases produced from various fuels using the Koppers-Totzek process

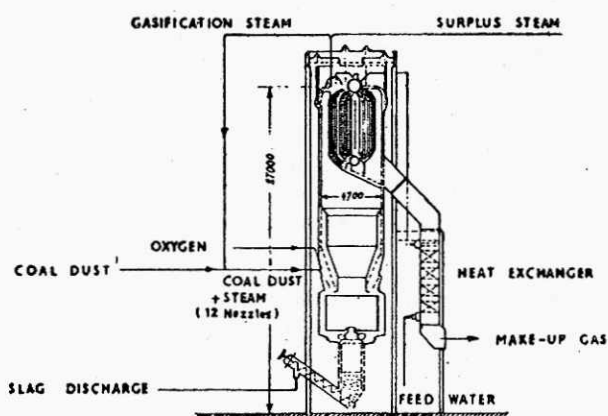
TYPE OF FUEL	Powdered Coal High Ash (Medium Coking)	Powdered Coal Low Ash (Non-Coking)	Powdered Coke	Fuel Oil	Naphta	Natural Gas.
FUEL ANALYSIS-DRY						
C-%	55.18	74.30	79.9	84.60	84.11	CO ₂ -1.0%
H-%	3.18	4.75	1.0	11.30	14.34	H ₂ -1.0%
O-%	3.82	13.30	1.5	0.13	-	CH ₄ -98%
N-%	0.96	1.05	0.3	0.40	1.42	-
S-%	0.76	0.95	0.8	3.50	0.13	-
Ash%	35.10	4.64	15.5	0.07	-	-
Gross C.V.-Btu's/lb	9820	12990	12150	18770	20700	975 per scf
Gross C.V.-Therms/Ton	220.0	291.0	272.2	420.5	463.7	514.2
ANALYSIS OF TOTZEK GAS						
CO ₂ -%	11.9	11.8	11.8	4.5	4.7	3.6
CO-%	55.3	57.6	65.9	48.4	43.7	34.0
H ₂ -%	31.1	29.0	21.0	45.4	50.7	61.3
N ₂ -%	1.3	1.2	0.9	0.8	6.8	1.0
CH ₄ -%	0.1	0.1	0.1	0.1	0.1	0.1
H ₂ S-%	0.3	0.3	0.3	0.8	Trace	-
Total-%	100.0	100.0	100.0	100.0	100.0	100.0
Gross C.V.-Btu's/scf	276.8	278.0	278.5	303.6	301.0	304.6
Sp. Gravity (Air=1)	0.755	0.774	0.846	0.587	0.538	0.437
NO-p.p.m.	4.5	4.5	4.5	Less than 1	Less than 1	Less than 1
Organic S-grains/100 scf	21.9	21.9	21.9	43.7	-	-
Gas Yield per ton-scf	56600	75100	70450	109700	121300	2610/1000scf
Steam Generated:						
Tons per ton of fuel	1.06	1.42	1.40	2.0	2.20	2.46
Lb. per 1000 scf Totzek Gas	41.93	42.35	44.5	40.85	40.64	40.0
Consumptions per 1000 scf Totzek Gas						
Oxygen-scf	360	350	370	277	274	274
Power (excluding oxygen) kWh	1.48	1.32	1.58	0.4	0.4	0.37
Steam (low pressure) lb.	12.5	4.65	12.8	16.0	15.7	4.9
Water 'make-up'-gallons	17.5	17.5	17.5	11.7	11.7	11.7
*THERMAL EFFICIENCY-%	71.3	71.7	72.1	79.2	78.7	81.3
C conversion ratio-%	96.0	97.7	95.4	95.6	97.3	99.3

High-capacity plants of this type are at present being erected in India for the manufacture of fertilizers on the basis of coal (oxygen gasification).

Attention is also drawn to a process which operates on the same principle and which utilizes waste heat in a combined waste-heat boiler (dust gasifier Babcock and Wilcox Co.).

This process, which has similar gas analyses and consumption values as the coal-dust gasifier just described (flame gasifier) has also been utilized in big units (up to 7000000 Nm³ CO+H₂/day).

Fig. 3



PRINCIPLE DIAGRAM OF A BABCOCK & WILCOCK COAL-DUST GASIFICATION PLANT.

High-temperature Winkler process

With the object of reducing the considerable costs of grinding coal, fluidized bed processes have been developed for gasifying fine coal for the production of rich gases (CO+H₂). On an industrial scale, they generate crack gases with low methane contents as the basis for synthesis gases (particle size approx. 0—8 mm).

Originally, these processes were designed for gasifying large surplus volumes of fine coal unsuitable for solids generators.

The principle of these fluidized bed process is that the fine-grain materials are suspended in a similar way to a liquid by a rising flow of gas.

A very important factor with these processes is the intensive heat exchange between gas and fluidized bed, resulting in an almost uniform temperature throughout the latter.

One of the most important fluidized bed gasification processes is the Winkler process. The principle is shown in Fig. 4.

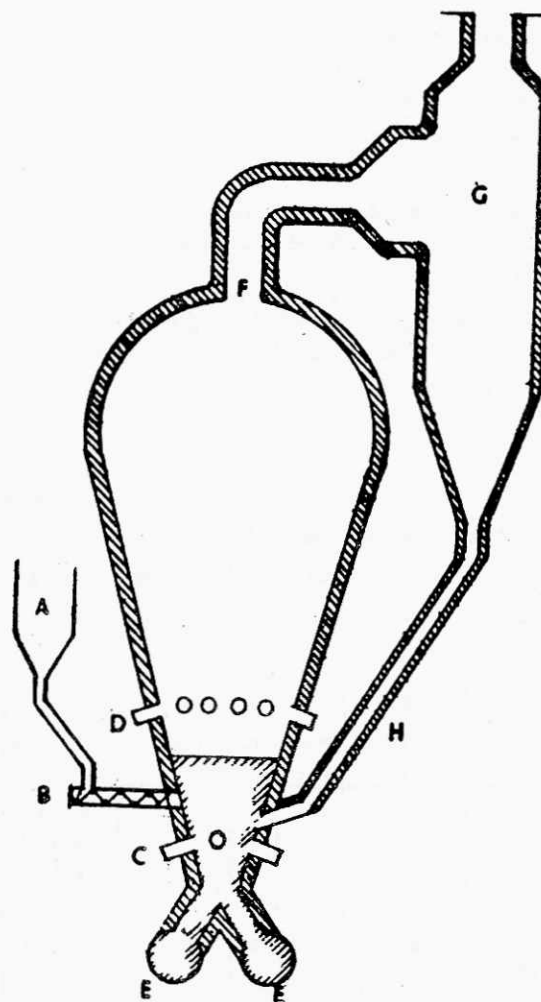
High-capacity generators have been built (up to 70,000 Nm³/h.) Normally, oxygen and steam are used as gasification media. The illustration shows that the actual carbon bearer layer has a very low height.

The Technical University of Aachen has refined the Winkler process in a particularly characteristic way for hard coal and ligniferous coal, which to date have proved difficult to process in the Winkler generator. The further-developed process is known as the high-temperature Winkler process.

One feature of this refined process is that atmospheric oxygen can also be used as a gasification medium instead of pure oxygen as required for the original Winkler process.

The process air is preheated to approx. 650 °C (using heat from the process) before being fed into the fluidized bed (coal particle size 0—8 mm). Since

Fig. 4



PRINCIPLE DIAGRAM OF A WINKLER GAS GENERATOR

- | | |
|------------------|------------------------|
| A—FUEL BIN | E—ASH DISCHARGE SCREWS |
| B—FEED SCREW | F—GAS OUTLET |
| C—BOTTOM NOZZLES | G—DUST SEPARATOR |
| D—TOP NOZZLES | H—RECYCLE DUST. |

no other heat consumers such as steam or carbonic acid gas are present, the temperature rapidly rises with combustion of part of the coal to 1400 to 1500 °C.

In order that the ash components do not disturb the process, sufficient limestone is added by injection to increase the basicity of the ash to the point where it remains dry and can be easily discharged (reduction of the volume of friable components). At the same time, the high lime content of the fluidized bed ensures an adequate degrees of desulphurization.

In view of the high temperatures and the high reaction rate, the equilibrium conditions between H_2O and CO_2 are so favourable that their proportion is less than 1%.

(Results of a semi-industrial plant at the Rhenish-Westphalian Technical University Aachen — Department of Ferrous Metallurgy).

The possibility of marketing the friable slag — which as a factor of the ash content and chemical analysis is held to provide an excellent clinker for cement production — has a favourable effect on profitability.

Table 3 shows the analysis of the gases obtained from various carbon bearers in the high-temperature Winkler generator.

Fig. 5 shows the expediency of the high-temperature gasification method through the less than 1% of CO_2 at temperatures above 1300 °C.

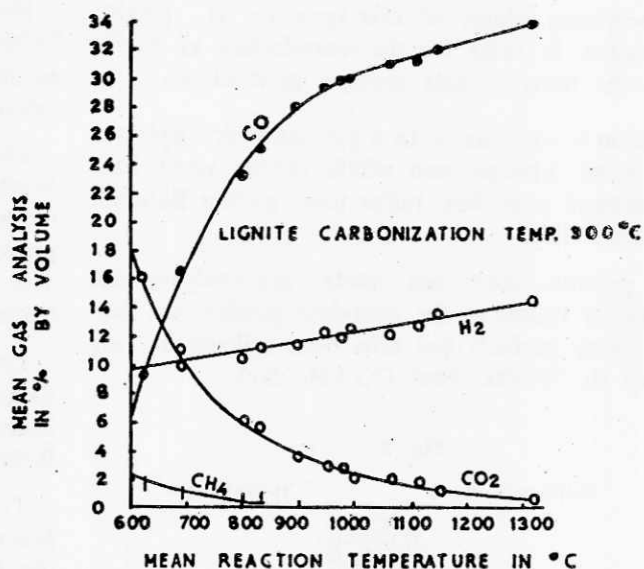


Fig. 5 Dry gas analysis relating to gasification of pre-carbonized Rhenish lignite with air using the Winkler process at various reaction temperatures (lignite carbonization temperature 300 °C)

Process possibilities based on coal reduction gas
Direct reduction process possibilities based on coal reduction gas

As shown in Table 1, oxygen gasification yields in theory a pure CO gas.

As compared with reduction gases having higher H_2 contents, this incurs the following problems:

Table 3

Composition and calorific value of the gases generated using various solid fuels at a gasification temperature of 1400—1500 °C and adding 200 kg limestone/kg fuel.

Vergasungs- brenn- stoff	Gasanalyse Gas analysis						Heizwert (kcal/Nm ³) Calorific value	
	CO ₂	CO	H ₂	CH ₄	N ₂	S	H ₀ net	H ₀ gross
	(Vol.-%) % by vol.	(Vol.-%) % by vol.	(Vol.-%) % by vol.	(Vol.-%) % by vol.	(Vol.-%) % by vol.	(g/100 Nm ³) x100		
Coke Koks	0,6	34,0	0,5	traces Spuren	64,8	traces Spuren	1039,7	1042,1
Anthrazit- kohle (85% fl. Bestand- teile)	Anthracitic coal (85% volatiles)	33,8	5,8	traces Spuren	59,5	traces Spuren	1170,4	1198,3
Bituminöse Steinkohle (16% fl. Bestand- teile)	Bituminous hard coal (16% volatiles)	31,0	8,4	traces Spuren	59,3	traces Spuren	1152,9	1193,2

(a) As indicated by Fig. 6, higher hydrogen contents in the reduction gas yield better utilization factors at higher temperatures in accordance with the equilibrium formulae (Bauer—Glaessner), i.e. lower surplus volumes are required or lower reduction temperatures can be employed.

(b) With recycling of the tail gas, the oxydation product of the reduction, H_2O (steam), can easily be removed by condensation, especially as the tail gas is normally cooled anyway in connection with dust collection.

In contrast, removal of the carbonic acid gas of the reduction — where it is not used as a gasifying medium, which is difficult in coal gasification — is a costly procedure. The carbonic acid gas scrubbing plant involves a considerable investment and operating cost outlay.

A further problem is briefly mentioned in this context. Naturally, the profitability of direct reduction plants, particularly when charging prerduced ores into conventional pig-iron producing facilities (blast furnace) is governed in substantial degree by the plant size; these should be in the order of more than 1 million tonnes per annum.

With shaft furnace and also retort processes, difficulties are to be expected with large inside diameters as regards uniform distribution of gas through to the furnace centre, since in the reactor of the direct reduction processes, as compared with the blast furnace shaft, the readily reacting coke is missing as a binding agent between the ore charge components.

New developments are concerned with positive guidance of the reduction gas in shaft and retort process to permit uniform gas flow also in big units with the correspondingly higher and more uniform degrees of reduction then possible.

The use of reduction gases obtained from non-coking coal in shaft furnace process,

Existing shaft furnace processes.

Some shaft furnace processes operate on the basis of the gasification of natural gas with top gas carbon dioxide as gasification medium. This produces a relatively hydrogen-rich gas. After condensation of the majority of the water vapour, sufficient top gas carbon dioxide is recycled as required as a gasification medium to supplement the consumed reduction gas (including recycling of the components $CO + H_2$ not consumed). The remainder of the top gas is used to help cover the heat requirements for the heavily endothermic gasification and heating of the components.

About 70 per cent of the top gas volume are recycled, after condensation.

The difficulty of applying coal gasification to these processes lies in the mentioned use of the top gas carbon dioxide as the gasification medium in coal gasification.

Thus, with both regenerative and recuperative gasification of coal with top gas carbon dioxide, it is apparent that the gasification processes based on natural gas must be entirely re-designed. Mention is also made of the fact that lower utilization coefficients are to be expected when there is a heavy percentage increase in the proportion of carbon monoxide. A certain control of the proportion of H_2 permits the use of steam and/or the moisture in the coal.

New shaft furnace processes.

ZEE process

Fig. 7 shows a flow sheet relating to the ZEE process.

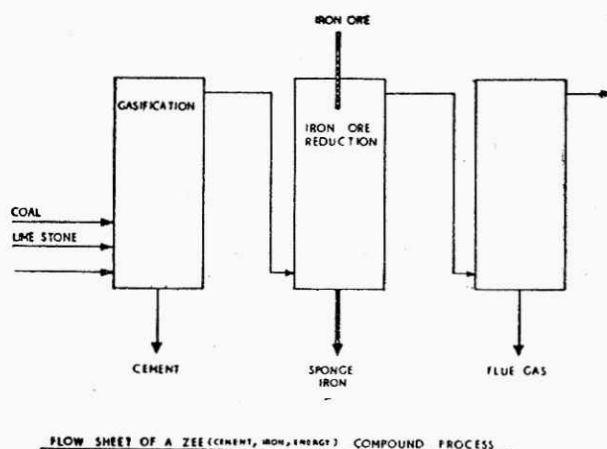


Fig. 7

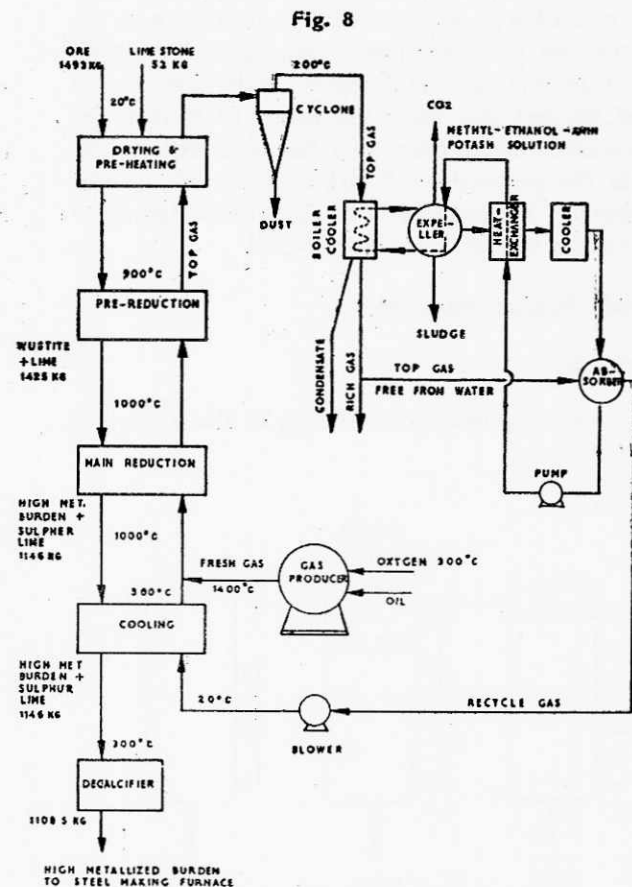
This process is based on the gasification of volatile coal in the fluidized bed of the Winkler high-temperature generator with atmospheric oxygen. The exothermic heat of reaction of the gasification process provides the required sensible heat for reduction work in the shaft. The top gas, which is not suitable for recycling as a gasification medium in view of the high nitrogen content stemming from air gasification (there is no economic method of removing nitrogen), can be employed for power generation. The high melting point of the coal ash in the Winkler generator, as obtained from the limestone addition, permits dry discharge of the same. The friable ash is a welcome component for mixing with cement production materials. These production components give the ZEE process its name, i.e., Z = Zement (cement)

E = Eisenschwamm (sponge iron)

E = elektrische Energie (electric energy)

DMK process

This process is in fact based on gasification using technically pure oxygen. A line diagram of the process is shown in Fig. 8.



LINE DIAGRAM OF THE DMK HIGH METALLIZED BURDEN PROCESS

Gasification is carried out in a flame gasifier, in the same way as with the Koppers-Totzek process. The gas is introduced into the shaft at the required tem-

perature and with the required surplus volume. The temperature is obtained from the exothermic gasification.

The top gas is not employed as gasification medium; after condensation separation of the water vapour of reduction, the reduction carbon dioxide is washed out using one of the known scrubbing processes. This yields a cold recycle gas with an analysis similar to that of the fresh gas, since it carries the unused CO and H₂ components on an almost complete basis.

The cold recycle gas is introduced into the sponge iron discharge, cooling this and protecting it against reoxydation. At the same time, it is itself warmed and mixed with the new charge of fresh gas.

The actual reduction gas consumption rate is only a little above the stoichiometrically required gas volume. Oxygen consumption and scrubbing costs are economic factors which must be included in calculations.

The use of reduction gases obtained from non-coking coal in the fixed-bed process

Existing fixed-bed processes

With the HyL process, the tail gas leaving the retort after final reduction (FeO—Fe) has an appropriately higher content of H₂O as compared with CO₂ (from reduction), in accordance with the higher hydrogen contents of methane gasification with water vapour. This fact permits use of the tail gas in reduction stages Fe₂O₃ — Fe₃O₄ — FeO after condensation of the water vapour and reheating (recuperative and partial combustion).

However, if the proportion of CO is very high through application of coal gasification, and if only small volumes of water vapour are condensed out, then the reduction capacity of this gas may under certain circumstances no longer be adequate for pre-reduction purposes.

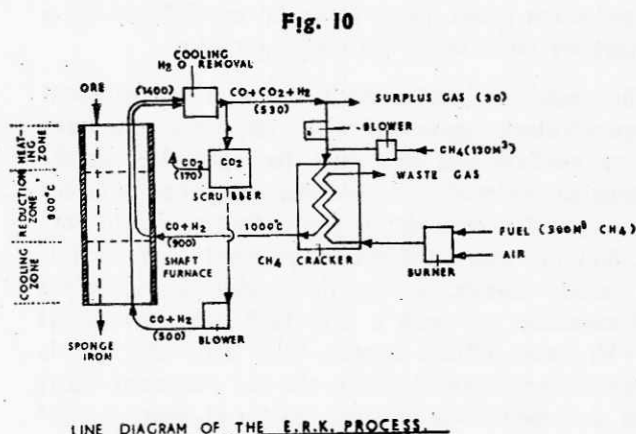
The following table demonstrates this fact using a simplified flow sheet.

TABLE 4
Simplified flow sheet of the reduction components in the HyL Process for various gasification modes

	Natural gas gasification				total	Coal gasification				
	CO	H ₂	CO ₂	H ₂ O		CO	H ₂	CO ₂	H ₂ O	total
Assumed reduction gas Nm ³ /t Fe	226	1278	—	—	1504	1128	376	—	—	1504
Reduction FeO — Fe	-60	-340	+60	+340	+ 400	+300	-100	+300	+100	±400
Condensation				-340					-100	
Tail gas for prereluction	166	938		60	1164	828	276	300		1404
Proportion CO + H ₂ as %	= 95.85%					= 78.63 %				

The basic principle of the process is that ores charged into chamber retorts are brought up to reduction temperature by waste gases of combustion or by hot

Fig. 10 shows a line diagram of this type of process principle (basis CH_4 -CO gasification).



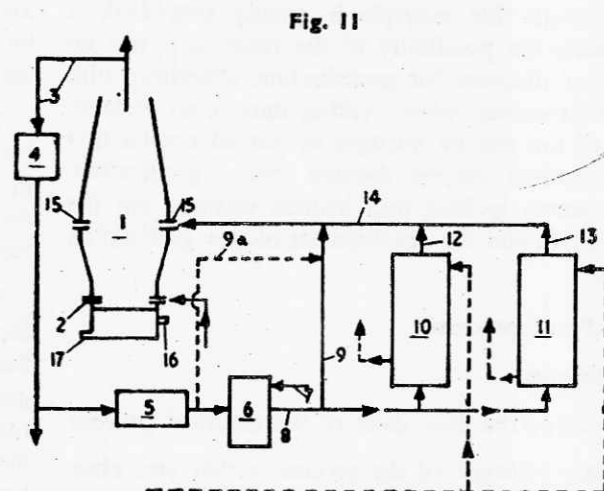
Possibilities of fluidized bed processes working on the basis of non-coking coal

The fluidized bed processes, which work with very rapid gas circulation, normally require very high hydrogen concentrations, and this for two reasons: firstly, high hydrogen contents permit operations with relatively low reduction temperatures, which is of importance with regard to preventing the caking of the ore fines in the charge; secondly, high hydrogen contents, which as oxydation products of reduction mainly have only water vapour, permit simple regeneration of the tail gas by condensation. Thus, the employment of carbon monoxide rich gases creates difficulties. Naturally, there is the possibility, though costly, of converting the carbon monoxide to hydrogen using water vapour and then remove the carbon dioxide so formed by scrubbing. $\text{CO} + \text{H}_2\text{O} = \text{CO}_2 + \text{H}_2$.

Introduction of reduction gas from non-coking coal into the blast furnace

In contrast to the problem of charging sponge iron into the blast furnace, mention is made of an equally competitive development of the blast furnace process, namely the injection of reducing gases above the tuyere level into the bosh of the blast furnace.

Fig. 11 shows a line diagram of such an arrangement.



LINE DIAGRAM FOR THE INTRODUCTION OF REDUCING GASES INTO A BLAST FURNACE ABOVE THE TUYERE LEVEL.

- 1 Blast furnace, standard design
- 2 Tuyeres
- 3 Top gas line
- 4 Gas scrubber
- 5 Clean gas pump
- 6 Recycle gas collecting chamber
- 7 Introduction of hydrocarbons
- 8 Gas mixing line
- 9 Cross-connexion to temp. control
- 10/11 Ceramic gas cracking furnaces, with pebble filling
- 12/13 Crack gas line
- 14 Gas line
- 15 Reduction gas feed
- 16 Slag taphole
- 17 Iron taphole

It is believed that the blast furnace coke rate can be cut to less than 300 kg coke/t pig iron, with an appropriate capacity increase.

In the extreme case, the coke charged into the blast furnace is employed only for providing coke carbon and thermal heat. The drop in carbon monoxide is balanced by the use of a reducing gas (CO and H_2).

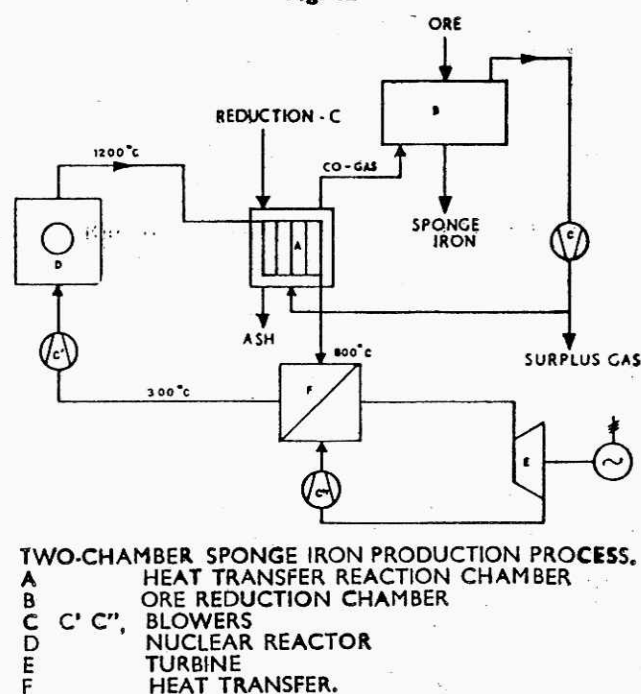
Proposals have already been made to exhaust low-nitrogen boundary gas and use this in part as a gasification medium on the recycle for gasifying fuel oil or natural gas.

The missing quantity of sensible heat of the tuyere gas is made up by feeding in the reduction gas with raised temperature. Thus, we virtually have a sponge iron process inside the blast furnace. Investigations using small blast furnaces with hearth diameters of 1.5 m have yielded coke rate cuts to about 320 kg/t pig iron. Part of the nitrogen-low boundary gas drawn off using special exhausting appliances, dedusted, and reformed to reduction gas by mixing in natural gas and top gas carbon dioxide. With coal-oxygen gasification, the carbon dioxide can be scrubbed out and

Possibilities of using nuclear heat with processes employing gasification of carbon

Fig. 12 shows the utilization of the sensible heat of the helium circuit for a coal gasification process using top gas carbon dioxide (two-chamber process).

Fig. 12



nuclear reactor in perfect condition. Plans are also held regarding the accelerated use of catalysts (e.g. Na_2CO_3).

The use of nuclear reactor heat can also be of interest for the generation of reduction gases on a coal basis for injection into the blast furnace shaft.

In countries with surplus non-coking coal or insufficient coking coal, prereduction is doubtlessly not the sole solution: as a result of extremely promising new developments, the possibility of manufacturing briquettes or formed coke to replace conventional blast-furnace coke must also be taken into consideration.

The basis is the agglomerating of non-caking coal unsuitable for normal coking procedures to make formed bodies which are then degassed. Accordingly coke-like bodies definitely suitable for the blast furnace can be made of almost all grades and sorts of coal.

All these new processes share a common feature: in contrast to the conventional coking process, which is discontinuous, they are continuous in operation. This is extremely suitable as regards adopting measures aimed at prevention of air contamination; the processes may thus be termed progressive as regards protection of the environment.

A condition for the use of such coke in the blast furnace is the availability of adequate abrasion strength and fracture strength, limitation of the residual content of volatiles, adequate reactivity, and adequate voidage.

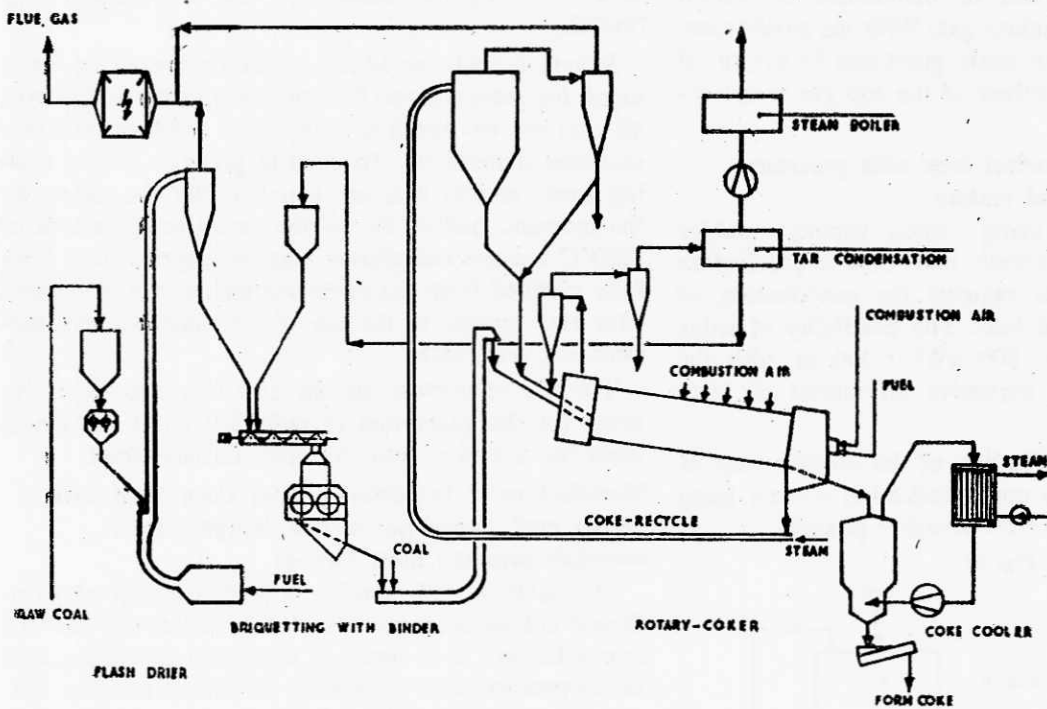
Methods of manufacturing formed coke

As already mentioned, the basis of formed coke is the agglomeration of coal fines to obtain a solid body, which is then degassed. These bodies can be made by briquetting or pelletizing. This can be effected cold, using some binder, but also by adding caking coal at temperatures between 400 and 500 °C.

Accordingly, the following processes are arrived at:

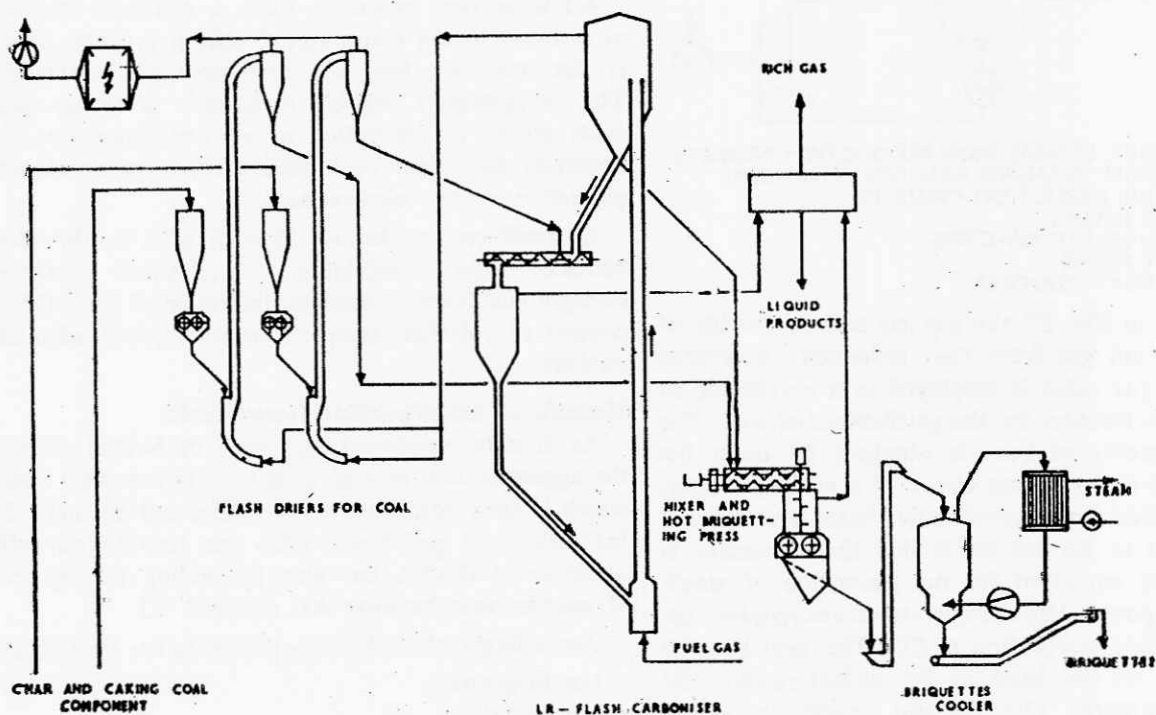
- hot briquetting
hot pelletizing
cold briquetting
cold pelletizing.

Fig. 13



PRODUCTION OF PITCH-BOUND BRIQUETTES AND COKING-OF THE BRIQUETTES IN THE LURGI-ROTARY-KILN

Fig. 14



FLOW SHEET OF THE BF-LURGI HOT BRIQUETTING PROCESS.

Hot briquetting process

This is described on the basis of the BF Lurgi hot briquetting process. Fig. 13 shows a line diagram of this process.

The prime feature of this process is the elimination of the following coking operation by mixing in a hot mixer the fine-particle carbonizing coke (0—3 mm) obtained in the fluidized bed or using the LR low-temperature carbonizer. Mixing is accompanied by the addition of dried caking coal. The coking coal is degassed by the effects of heat. The generated low-temperature carbonizer. Mixing is accompanied by the 5000 kcal/Nm³ (that from the fluidized bed approx. 1500 kcal/Nm³). Condensation components of the gas cooling phase are returned to the mixer.

A pilot plant with a capacity of approx. 100 tpd has been in operation for a number of years; an industrial plant rated at 300 tpd is under construction; a further plant for 650 tpd is to be built. Project work is being carried out for plants with capacities of up to 3 million tonnes of hot briquettes per annum. Elimination of post-coking proper keeps the investment costs low. The hot briquettes can also be coked, reducing their content of volatiles to approx. 1%.

Cold briquetting

Fig. 14 shows a line diagram of this process.

The fundamental consideration behind this process is the briquetting of non-caking coal, particularly weak and medium-caking coal, using tar pitch. The product is then degassed in a rotary kiln. The flow sheet illustrates the arrangement.

In contrast to the conventional coking process, rapid heating (to 20°/minute) is advantageous in the case of weak caking coal. Coals with higher contents of volatiles call for shorter heating times. The heating time is a yardstick for the size of the coking facility.

Apart from the selected example, the following coking facilities are also in use: shaft furnaces, chamber furnaces, sintering grates, rotary hearth units.

Hot pelletizing process

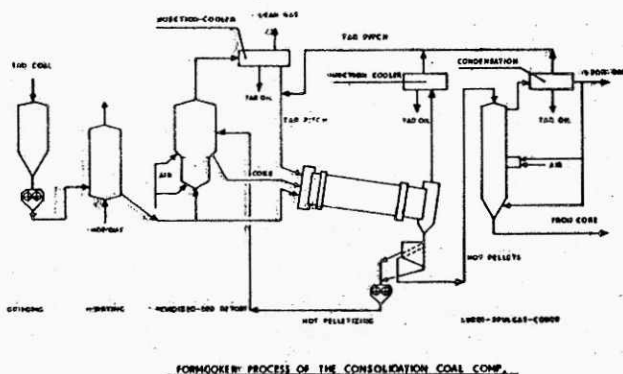
In this process, the caking coal is pelletized in a "rotary coker" at temperatures of approx. 450 °C on the basis of the plasticity of the excluded volatiles and small additions of binders such as tar. The pellets have a non-uniform shape. They are subsequently coked (process of the Consolidation Coal Company, USA).

Cold pelletizing process

Coal fines are conventionally pelletized using cold binding agents such as bentonite, sulphite liquor, etc. A good proportion of self-caking coal is expedient. In this case, it is expedient to dry the pellets prior to the actual coking phase. Coking can be effected in one of the ways described above.

Fig. 15 shows a flow sheet of one of these processes.

Fig. 15



Experience gained with formed coke in the blast furnace

A range of favourable results have been obtained with formed coke in the blast furnace. Table 5 shows a list of results obtained using formed coke relating to their charging into blast furnace, some of higher capacity, in the amount of more than 30000 t.

Most of the results show that formed coke is in certain respects a full substitute for normal metallurgical coke. Of special interest are the results obtained for hot briquettes, i.e. without subsequent coking; these can be processed without difficulty in spite of having a proportion of volatiles in the order of more than 7%. Lowering of the top gas temperature and a desirable increase in the calorific value of about 100 kcal/Nm³ top gas were the special points of hot briquette charging. No detriment to the gas system by the condensation of tarry components was noted.

There can be little doubt that the charging of certain grades of formed coke made from non-coking coal can attain significance in future, especially if there is any shortage of normal metallurgical coke.

Summary

A review is given of the possibilities of reducing iron ores on the basis of non-coking coal. Direct reduction on the basis of reducing gases obtained from such coal and the charging of formed coke or hot briquettes made from such coal are the available possibilities.

Brief indication is given to the possibilities of using nuclear energy in future times.

Decisions regarding the selection of the most suitable of the discussed methods should be made on the basis of a Feasibility Study and industrial tests. These should compare both the commercial and the technical aspects of each process.

Brief mention is made of the problems involved when converting to gas reduction processes.

TABLE 5

Known blast furnace trials with formed coke (according to Lurgi, 1972)

Test No.	Year	Quantity t	Type	Individual Size g or mm	Producer	Country	Hearth Ø m	Period of Input days	Company	Country
1	1962	2000	pitch briquettes	32x25x19	FMC	USA	1,32	30	US Steel	USA
2	1964	—	caked hot briquettes	12-44 Ø	Core Coal Co	USA	—	3	Bond of Mines	USA
3	1965	—	pitch briquettes	350	ICEM	Rumania	6,00	33	Brucetown ICEM	Rumania
4	1967	500	caked hot briquettes	21	BF Essen	Germany	1,40	—	Airbo Ougree	Belgium
5	1967	—	pitch briquettes	—	Carmaux	France	1,40	—	Airbo Ougree	Belgium
6	1967	—	—	32x25x19	FMC	USA	5,60	—	Armco Hamilton	USA
7	1969	—	—	21	HBNPC	France	1,40	—	Airbo Ougree	Belgium
8	1969	1400	caked hot briquettes	50	BF Essen	Germany	9,50	1	Hoesch Dortmund	Germany
9	1969	500	green hot briquettes	48	BF Essen	Germany	6,90	—	Rheinstahl Hattingen	Germany
10	1969	—	pitch briquettes	26,3	Inex	Belgium	1,40	—	Airbo Ougree	Belgium
11	1970	2700	green hot briquettes	48	BF Essen	Germany	6,80	3	Rheinstahl Hattingen	Germany
12	1970	7000	caked hot briquettes	44	BF Essen	Germany	6,80	7	Rheinstahl Hattingen	Germany
13	1970	200	pitch briquettes	—	FMC	USA	—	short	Inland Steel Chicago	USA
14	1970	—	green hot briquettes	17	EBV	Germany	4,00	3	Kupferhutte Duisburg	Germany
15	1971	2700	„	46	BF Essen	Germany	5,50	6	BSC Cardiff	Great Britain
16	1971	3500	green hot briquettes	51x48x25	EBV	Germany	5,00	9	Arbed Burbach	Luxemburg
17	1972	10000	pitch briquettes	45x45x32	—	Japan	8,40	57	—	Japan