

May 22, 1951

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2,554,263

GASIFICATION OF CARBONACEOUS SOLIDS

Filed Dec. 18, 1946

3 Sheets-Sheet 1

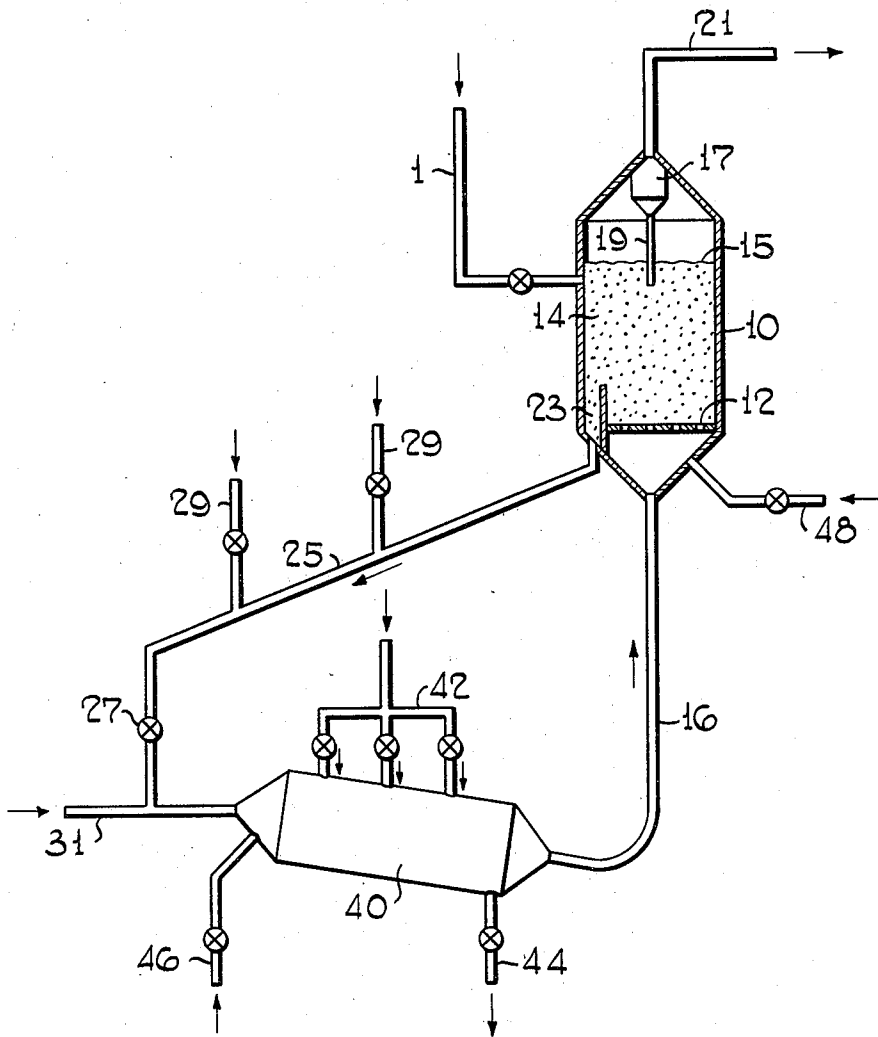


FIG.-1

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3 Sheets-Sheet 2

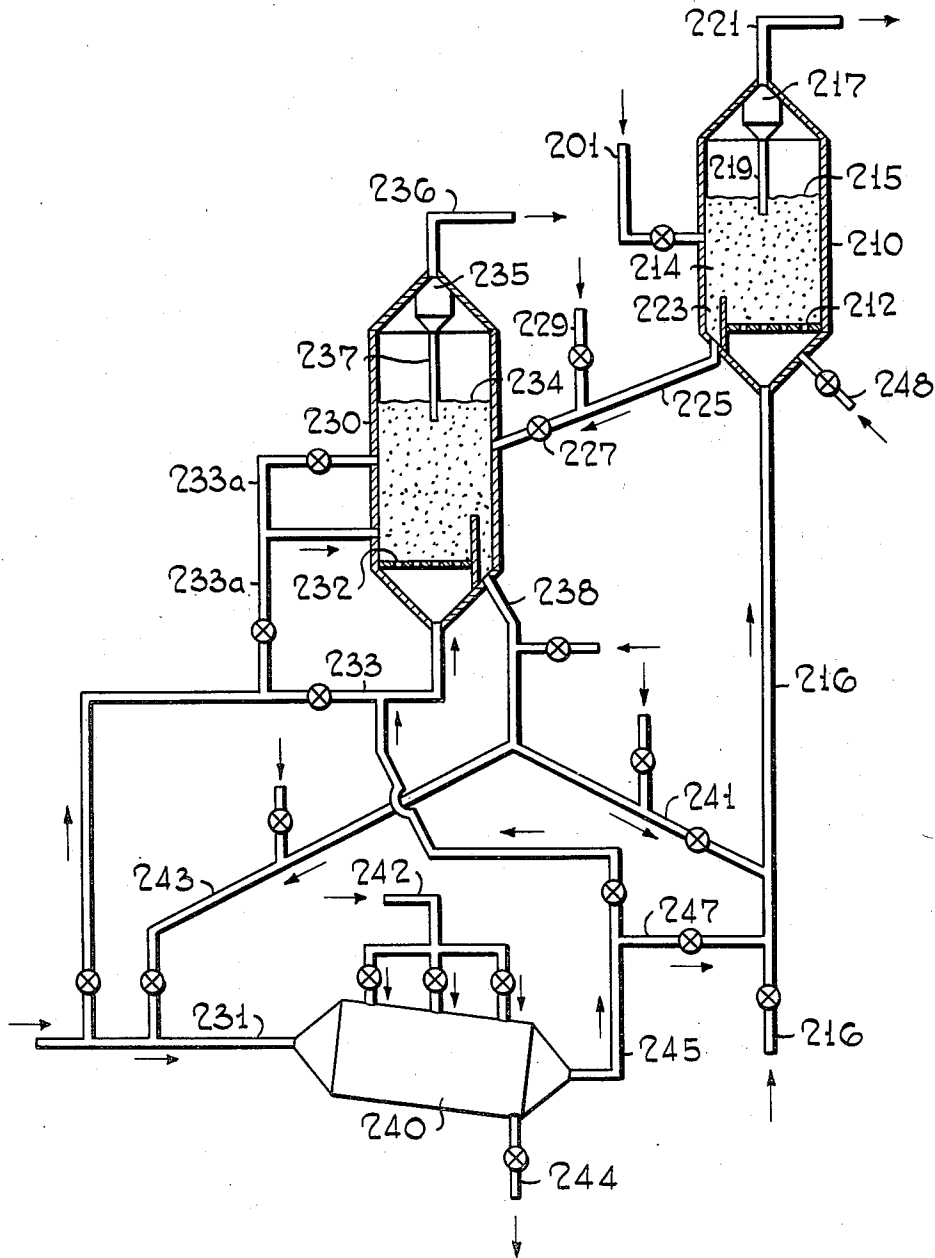


FIG.-2

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3 Sheets--Sheet 3

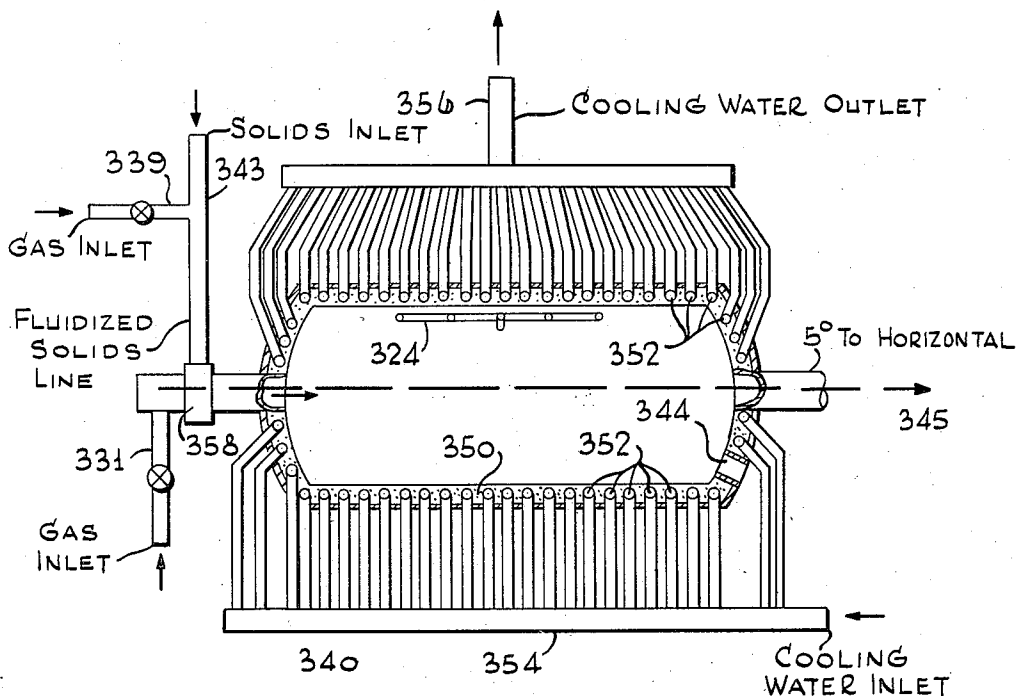


FIG. - 3

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UNITED STATES PATENT OFFICE

2,554,263

GASIFICATION OF CARBONACEOUS SOLIDS

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Application December 18, 1946, Serial No. 717,064

6 Claims. (Cl. 48—206)

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The present invention relates to the conversion of carbonaceous solids such as all types of coal, lignite, peat, oil shale, tar sands, coke, oil coke, cellulosic materials, including lignin, etc. into gases containing carbon monoxide such as water gas, producer gas and the like.

Prior to the present invention, it has been suggested to gasify carbonaceous solids with a gasifying medium such as steam and/or air to produce water or producer gas, in the form of a dense turbulent bed of finely divided solids having a particle size of about $\frac{1}{4}$ to $\frac{1}{2}$ in. down to about 400 mesh, fluidized by an upwardly flowing gas and maintained at gasification temperatures of about 1500°–2500° F. This technique is greatly superior to conventional fixed-bed operation. It provides larger solid reaction surfaces, better mixing and greatly improved temperature control, and it affords higher gas yields in fully continuous operation within shorter reaction times.

While these great advantages make the application of the fluid solids technique to coal gasification appear highly attractive it has not as yet found the broad commercial application it would seem to deserve. One of the more important reasons of the slowness of this development lies in difficulties encountered in the substantially complete conversion of the carbon feed with the carbonaceous charge into product gas and heat required for the process at reasonably constant conversion conditions, satisfactory steam conversion rates, reasonable temperature levels and economic equipment design.

Such substantially complete utilization of the carbonaceous charge is an essential condition for the economic operation of the coal gasification process. On the other hand, the rate of conversion of the gasifying medium in the water gas as well as in the producer gas reaction decreases rapidly as the carbon concentration in the conversion zone decreases so that relatively high carbon concentrations are necessary for the production of satisfactory gas yields at a given temperature per unit of time and reactor space.

In conventional fluid solids operation the entire reacting mass of fluidized carbonaceous solids has a substantially uniform carbon concentration. As a result, the requirements of complete carbon utilization and highest possible carbon concentration can not be reconciled in an economical manner when conventional means of heat generation are applied.

For example, when heat is generated by a par-

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tial combustion of the carbonaceous charge within the fluid generator bed at high-carbon concentrations, a gasification residue of the same high carbon concentration is withdrawn from the gas generator in continuous operation. This high-carbon residue must be reprocessed in order to avoid carbon losses in the system. Gasification to low carbon concentrations at substantially constant conditions would require an excessively large reaction space without completely avoiding the carbon losses in the form of carbonaceous gasification residue.

Generation of heat by combustion of high-carbon gasification residue in a conventional external burner and heat supply to the generator in the form of sensible heat of combustion gases so produced require excessive amounts of heating gases in view of the high gasification temperatures to be maintained in the gas generator and the temperature limitations imposed by the relatively low heat resistance of economical construction materials for conventional burners.

It has also been suggested to generate heat by the combustion of solid carbonaceous gasification residue in an external heater and to supply heat to the gas generator in the form of sensible heat of solid combustion residue circulated from the heater to the gas generator. Efficient combustion at high temperatures in an external heater of this type requires low carbon concentrations to avoid excessive air requirements and/or carbon losses in the form of CO formed by the reduction of CO₂ with excess carbon. This requirement is incompatible with a high carbon concentration in the gas generator.

It will be appreciated from the above that the reconciliation of high carbon concentrations in a "fluid" gas generator with a complete conversion of the available carbon into gas and heat required for the process presents an important and difficult problem. The present invention is concerned with means for solving this problem.

It is, therefore, an important object of this invention to provide an improved process for producing combustible gases from carbonaceous solids employing the fluid solids technique.

Another object of my invention is to provide an improved process for the gasification of carbonaceous solids in the form of a dense, turbulent, fluidized bed of finely divided solids at optimum rates of conversion and with full utilization of available carbon.

Another object of this invention is to provide

improved means for supplying heat to a fluidized bed of finely divided carbonaceous solids undergoing gasification at optimum conversion rates by burning gasification residue in an external burner.

A more specific object of my invention is to provide improved means for supplying heat to a fluidized bed of finely divided carbonaceous solids undergoing gasification at optimum conversion rates, full utilization of available carbon in the process and convenient disposal of ash.

Other and more specific objects and advantages will appear hereinafter.

In accordance with the present invention, finely divided carbonaceous solids are subjected in a gas generation zone to a gasification reaction with a gasifying medium at gasification conditions of temperature and pressure, in the form of a dense turbulent bed of finely divided solids of substantially uniform and high carbon concentration conducive to highest rates of conversion of the gasifying medium used. Solid finely divided gasification residue having the average carbon concentration of the gasification bed is burned in a separate combustion zone at temperatures above the melting point of the ash of the carbonaceous charge, that is, about 2000°-3500° F., preferably 2700°-3300° F. and at an oxygen supply at least sufficient to permit complete combustion of the carbon introduced into the combustion zone. Liquid ash may be withdrawn from the combustion zone while flue gases are supplied to the fluidized solids bed of the gasification zone substantially at the extremely high temperatures of the combustion zone to supply at least a portion of the heat required by the gasification reaction. Low-carbon solids carry-over from the combustion zone to the gas generation zone is substantially avoided by the fusion and liquid drawoff of the ash in the combustion zone.

Complete combustion of the carbon supplied as gasification residue to the combustion zone normally requires a considerable excess of oxygen to be supplied to the combustion zone. This excess oxygen enters the gas generator as a constituent of the high temperature flue gas so that the heat balance required for the conversion of carbon and gasifying medium into fuel gas is established by the fast and extremely exothermic reaction between oxygen and carbon, for example, in accordance with the following equations:

- (1) $C + O_2 \rightarrow CO_2 + 170,000 \text{ B. t. u.}$
- (2) $C + CO_2 \rightarrow 2CO - 74,500 \text{ B. t. u.}$
- (3) $CO + H_2O \rightarrow CO_2 + H_2 + 13,100 \text{ B. t. u.}$

Since reaction (1) is very fast, reaction (2) relatively slow and reaction (3) of intermediate speed the desired gas composition may be established by adapting the contact time of solids and gases at any given gasification conditions of temperature, pressure, carbon concentration, reactivity of the solid charge and steam to oxygen ratio to the requirements of either reaction (2) or reaction (3) either of which requires contact times sufficiently long to allow for the completion of reaction (1) and for optimum utilization of the heat obtained from the latter. Adaptation of the contact time to the requirements of the slowest reaction (2) will permit the formation of sufficient CO_2 to generate the heat required while permitting a reduction of CO_2 formed to establish the desired minimum CO_2 content of the final product gas.

In this manner, the gasification reaction may be conducted at optimum carbon concentrations and conversion rates at relatively low conversion temperatures while all available carbon not con-

verted in the gas generator is utilized for heat generation rather than lost as it would be if the gasification were carried out at similar gasification conditions but without the use of my external combustion zone. An additional advantage of my invention results from the fact that gas-generating equipment for high-carbon, low-temperature operation is more economical with respect to size and cost of construction materials than equipment for low-carbon and/or high-temperature operation.

The extremely high combustion temperatures required for the purposes of my invention may be produced, for example as has been recently shown, by passing a suspension of finely divided carbonaceous solids in a combustion-supporting gas such as air and/or oxygen tangentially along the axis of a cylindrical combustion zone thereby imparting a rotating motion to the suspension, if desired, aided by a gas, preferably secondary combustion-supporting gas, introduced tangentially into the combustion zone at a high velocity of about 200-800 ft. per second to bring the total oxygen supply to about 105-600% of the theoretical. However, it is only necessary to supply that amount of oxygen to the combustion zone, which is required to burn the carbon therein substantially completely while any excess oxygen thereover may be by-passed and admitted directly to the gas generator. The secondary combustion-supporting gas may amount to as much as about 98% of the total combustion-supporting gas supplied to the combustion zone. The combustion zone may be either substantially vertical or substantially horizontal with a tilt downward toward the discharge end to facilitate the flow and tapping of liquid ash. Instead of secondary combustion-supporting gas steam, CO_2 , or the like may be used as the gas introduced tangentially if the amount of primary combustion-supporting gas is high enough to satisfy the requirements of the process.

In operation, a violent swirling action takes place within the combustion zone due mainly to the tangential velocity of the secondary gas and to some extent to the rotary motion of the solids-in-gas suspension fed axially. As a consequence of centrifugal force, the internal walls of the combustion zone are covered with a film of molten ash which travels in a spiral manner and eventually discharges at the lower end of the combustion zone. The larger carbonaceous particles are caught in the slag film and burned by the combustion-supporting gas passing by at higher velocity, while the smaller particles burn in the gas zone. The fines burn almost instantaneously in the vortex to which a small volume of tertiary combustion-supporting gas, amounting to about 1-5% of the total combustion-supporting gas, may be fed to accelerate this combustion.

The combustion chamber may be a steel cylinder lined with refractory such as chrome ore or the like. Since the sintering or melting points of even the most heat resistant refractory linings usually lie substantially below the prevailing combustion temperatures, cooling tubes are preferably imbedded in the lining.

The temperatures reached in a combustion zone of this type when pure or concentrated oxygen is used as the combustion-supporting gas are substantially higher than those required to establish an efficient temperature differential between the combustion zone and the gas generation zone. In actual operation I prefer, there-

fore, to use air or oxygen diluted with steam or flue gases as the combustion-supporting gas the proper choice of which depends on the type of gasification reaction desired. When producer gas is to be manufactured in the gas generation zone, air is a suitable diluted combustion supporting gas while for the production of water gas oxygen diluted with steam is preferred.

As mentioned above, the external combustion zone is normally operated with an excess of oxygen so that unconverted oxygen becomes available for conversion and/or heat generation within the gas generation zone. For the generation of producer gas this excess of oxygen may be chosen high enough to supply, in combination with the CO₂ content of the flue gas, all the oxygen required in the gas generation zone to convert the desired amount of carbon into carbon monoxide. When water gas is produced, part or all of the steam required for the water gas reaction may serve as a diluent in the combustion zone and the excess oxygen from the combustion zone may be used in the gas generation zone to generate additional heat therein substantially as outlined above. Some steam dissociation and/or conversion may take place at the conditions of the combustion zone and any hydrogen and/or carbon monoxide produced in this manner will be recovered from the gas generator.

In accordance with a more specific embodiment of the invention, a combustion zone of the type described may be combined with a conventional "fluid" heater to supply heat to the gas generation zone in the form of sensible heat of solid heater residue. For this purpose solid gasification residue of relatively high carbon concentration may be circulated from the fluidized gas generator bed to an external heater wherein it is subjected to a combustion with air in the form of a dense, turbulent, fluidized bed at relatively high carbon concentration. A major portion of the solid, relatively high-carbon combustion residue from the heater may be returned to the gas generator for heat supply therein. A minor portion may be passed to a high temperature combustion zone of the type described above to completely burn the carbon, dispose of the ash and produce flue gas having a temperature far in excess of the temperature in the heater. This flue gas is returned to the heater.

The CO₂ introduced into the heater with the flue gases from the combustion zone will, at the high carbon concentration of the heater, react with carbon to form CO, causing losses of heat and effective carbon unless sufficient oxygen is present in the heater to favor the formation of CO₂ over that of CO. It is advisable, therefore, to maintain, for example, at least a slight excess of oxygen in the heater over that consumed by the desired combustion of solid carbon in the heater. This may be accomplished by splitting the air feed to the system in a suitable manner and feeding part of the air to the high temperature combustion zone and part to the heater, if desired, in a plurality of streams.

In this manner, the heater temperature and thus the temperature differential between heater and gas generator may be maintained at least at similar and even at higher levels as if carbonaceous solids of low carbon concentration were burned in the heater. However, the carbon concentration of the heat-carrying solids returned to the gas generator is now high enough to establish a carbon concentration within the gas generator at levels desirable for high con-

version rates. In this, as in the embodiment operating without solid heat carrier, the total available carbon is eventually converted into product gas and heat required for the process and the ash is effectively disposed of.

Having set forth its objects and general nature, the invention will be best understood from the more detailed description hereinafter in which reference will be made to the accompanying drawing wherein

Figure 1 is a partly schematical partly diagrammatical illustration of a system utilizing the sensible heat of high temperature flue gases as a source of heat for the gas generation reaction,

Figure 2 is a similar illustration of a system using the sensible heat of high temperature solids for the same purpose, and

Figure 3 is a semi-diagrammatical illustration of a high temperature combustion chamber suitable for the purposes of the invention.

Referring now to Figure 1, the system shown therein essentially comprises a water gas generator 10 and a high temperature combustion zone or burner 40 whose functions and cooperation will be forthwith explained. While low temperature coke will be referred to hereinafter as the carbonaceous solid used any other solid carbonaceous material may serve as charge to my process.

In operation, a finely divided preferably highly reactive coke produced by the carbonization of a bituminous coal in a fluidized solids bed at temperatures not substantially exceeding 1000° F. is supplied through line 1 to gas generator 10. Line 1 may be part of any conventional means for conveying finely divided solids such as an aerated standpipe, a pressurized feed hopper, a mechanical conveyor, etc. The particle size of the coke may fall within the wide ranges of 1/2 in. to 400 mesh, preferred size ranges being about as follows:

- 1-70% through 200 mesh screen
- 20-40% through 50 mesh screen
- 40-60% through 30 mesh screen
- 80-100% through 1/4 in. screen

The finely divided coke forms in generator 10 above distribution grid 12 a dense turbulent mass 14 of solids fluidized by the gaseous reaction products and the gas and vapors supplied through line 16 and grid 12 as will appear more clearly hereinafter. Linear gas velocities of about 0.1-10 ft. per second, preferably 0.3-3 ft. per second, within mass 14 are generally suitable for this purpose at pressures ranging from about atmospheric to about 400 lbs. per sq. in. and for bed densities of about 10-50 lbs. per cu. ft.

Heat and gasifying medium such as air, oxygen and/or steam, depending on the product gas desired, are supplied through line 16 sufficient in amounts to maintain bed 14 at the desired gasification temperature of about 1400°-2500° F. and at a carbon concentration of about 15-60%, preferably about 30 to 50%. At gas temperatures in line 16 of about 2700°-3000° F. the amount of air required to produce producer gas at the carbon concentrations indicated is about 1.5 to 6.0 lbs. per lb. of coke charged and the amount of steam required to produce water gas is about 0.4 to 4.0 lbs. per lb. of coke charged.

Product gas is withdrawn overhead from level 15 of mass 14 and passed through a conventional gas solids separator 17 provided with

solids return line 19. Gases substantially free of solids leave through line 21 and flow to further processing equipment and/or any desired use such as a hydrocarbon synthesis reactor or the like (not shown), if desired, after heat exchange with solid and/or gaseous feed materials of the process.

Solid fluidized gasification residue of the average carbon concentration of bed 14 is withdrawn downwardly through well 23 and passed under the combined pseudo-hydrostatic and gas pressures of bed 14 through standpipe 25 to which a small amount of steam, air and/or oxygen is added through taps 29 to facilitate the flow and strip the solids in standpipe 25. The rate of solids flow through standpipe 25 is adjusted by means of valve 27. A solids withdrawal rate of about .05 to 0.3 lb. per lb. of coke charged through line 1 is generally adequate.

Carbonaceous solids discharging through valve 27 are suspended in a stream of combustion-supporting gas such as air and/or oxygen flowing through line 31 and the suspension formed is blown into burner 40, preferably tangentially, along its axis. The amount of air and/or oxygen entering burner 40 through line 31 including any combustion-supporting gas admitted through taps 29 may be about 1-20% of the total combustion-supporting gas supplied to burner 40. The remainder of the combustion-supporting gas required to bring the total amount of gas above 100%, say, to between about 105 and 600% of the amount theoretically required for complete combustion of carbon available in burner 40 is supplied through a manifold 42 or the like, tangentially to the combustion zone of burner 40 at a linear velocity of about 300-700 ft. per second, preferably about 500 ft. per second. Details of the design of burner 40 will be described below in connection with Figure 3 of the drawing. It should be noted, however, that burner 40 may also be arranged in a substantially vertical position with a downward flow of feed and ash.

The temperature of burner 40 is maintained above 2500° F., the desired temperatures depending on the fusion point of the ash and the amount of carbon to be burned. Liquid ash is tapped at 44 while hot flue gases enter line 16 and generator 10 substantially at the temperature of burner 40.

When water gas is to be produced in generator 10 by a gasification reaction with steam, oxygen may be supplied as the combustion-supporting gas through lines 31 and 42. In order to prevent the temperature within burner 40 from rising beyond desired levels, at least a substantial proportion of the steam required for the water gas reaction may be admitted through lines 46 and/or 42 to burner 40 to act as a diluent of the oxygen. Any additional steam required to bring the total up to 0.4 to 4.0 lbs. per lb. of coke charged as it is needed for the desired gasification may be added through line 48 directly to generator 10. Burner temperatures of about 2700°-3300° F. are generally suitable for most of the conventional carbonaceous charge materials. Excess oxygen entering generator 10 burns an equivalent amount of combustibles in bed 14 to generate additional heat therein. In general, a total of 0.2 to 2.0 lbs. of oxygen charged to the burner per lb. of carbon to be gasified is sufficient for the production of water gas.

When generator 10 is to be used for the manufacture of producer gas, air instead of oxygen is supplied through lines 31 and 42 and the tem-

perature in burner 40 is controlled by the excess of air, part of which may be admitted through line 46. Additional air required for gas generation in bed 14 may be supplied through line 48. The amount of total air required for the manufacture of producer gas may fall within the approximate limits of 1.5 to 6.0 lbs. per lb. of carbon to be gasified.

Instead of feeding the carbonaceous charge through line 1 directly to bed 14 it may be suspended in the gasifying medium flowing through line 16 or line 48 and passed upwardly through grid 12 in a manner known in the art of fluid solids handling.

Referring now to Figure 2, the system shown therein essentially comprises a gas generator 210, a fluid solids heater 230 and a high temperature burner 240.

Finely divided fresh coke is charged through line 201 to generator 210 to form therein above grid 212 a dense turbulent bed of solids fluidized by gas supplied through line 216 and forming an upper level 215 to undergo gasification substantially as described in connection with generator 10 of Figure 1. Product gas is withdrawn upwardly through gas-solids separator 217 provided with solids return line 219 and thence through line 221.

High-carbon gasification residue flows through solids withdrawal well 223 and standpipe 225 provided with control valve 227 and aerated and stripped through one or more taps 229 with air and/or oxygen to heater 230 to form therein above grid 232 a dense turbulent bed of carbonaceous solids fluidized by hot air and flue gas supplied through line 233 to form a well defined upper level 234. Linear gas velocities and bed densities in heater 230 may be substantially the same as those specified in connection with generator 10 of Figure 1.

The amount and distribution of air supplied to heater 230 should be sufficient to prevent substantial reduction to CO, of CO₂ present in heater 230 including CO₂ contained in the hot combustion gases issuing from burner 240 and entering 230 through lines 245 and 233 as will appear hereinafter. Proper distribution of the air fed to heater 230 may be accomplished by the use of one or more manifold branch lines 233a, depending on the height of the fluidized bed in heater 230.

The combined effect of combustion and high oxygen supply raises the temperature of the solids in heater 230 to temperatures of about 1500°-2500° F., preferably about 1700°-1900° F., that is, below the fusion point of the ash but substantially above the desired gasification temperature in generator bed 214. Flue gases are withdrawn overhead from level 234 through cyclone separator 235 and pipe 236 to be used for any desired purpose including heat exchange with process solids and/or gases. Solids separated in separator 235 may be returned through pipe 237 to heater 230.

Solid fluidized combustion residue from heater 230 is withdrawn through standpipe 238 and passes substantially at the temperature of heater 230 through branch standpipe 241 into line 216 where it is suspended in the gaseous gasifying medium and carried through grid 212 into generator 210 to supply the heat required for gasification. As a result of the high temperature and high carbon concentration of the solids so supplied, the carbon concentration in generator 210 remains as high as about 15 to 60%, prefer-

ably about 30 to 50%, at gasification temperatures of about 1400°-2500° F. Solids circulation rates through standpipe 241 of about 50 to 500 lbs. per lb. of carbon to be gasified are generally sufficient for this purpose at the conditions indicated.

A minor proportion of the solids entering standpipe 238 is branched off to standpipe 243 and fed to line 231 wherein it is suspended in air and thence supplied to burner 240 for complete combustion at temperatures above the ash fusion point substantially as outlined in connection with burner 40 of Figure 1. Secondary high velocity air is admitted tangentially through manifold 202 to bring the total combustion air above the amount theoretically required for complete combustion. Liquid ash is withdrawn through tap 244 and hot flue gases pass at a temperature of about 2500°-3300° F. through lines 245 and 233 to heater 230 as outlined above.

It will be understood that the amount of solids circulated through standpipe 225 will depend on the feed rate of carbonaceous charge and the rate of solids circulation through standpipe 241 and pipe 216 so that major fluctuations of level 215 will be avoided. For similar reasons the rate of solids withdrawal through standpipe 238 will depend on the rate of solids supply through standpipe 225. In general, solids circulation rates through standpipe 225 of about 50 to 500 lbs. per lb. of coke charged through line 201 and of about 49 to 499 lbs. through standpipe 238 per lb. of coke charged through line 201 are adequate for the purposes of the invention.

The exact reaction conditions in generator 210 depend on the kind of product gas desired. For the production of water gas, steam is supplied through lines 216 and temperatures of about 1400°-1900° F. and pressures of atmospheric to about 400 lbs. per sq. in. may be used, the higher pressure ranges being conducive to the formation of a high B. t. u. gas rich in gaseous hydrocarbons. For the manufacture of producer gas the same or higher temperatures, say 2000°-2400° F., and similar pressures may be applied while predominantly air, if desired, admixed with some steam is supplied through line 216.

If desired, additional heat may be supplied to generator 210 in the form of hot, normally oxygen-containing burner flue gases branched off line 245 to lines 247 and 216, or a combustion-supporting gas such as air and/or oxygen may be added to generator 210 through line 216 or 248 to support a limited combustion within bed 214.

It will be understood that in place of any one or all of standpipes 225, 238, 241 and 243 other conventional means for conveying fluidized solids such as mechanical conveyors of various types may be used. It should also be noted that heater 230 may be arranged at higher level than generator 210 so that solids flow is by gravity from heater 230 to generator 210 and by way of combined gas and pseudo-hydrostatic pressure in the opposite direction. As stated before, burner 240 may also be arranged in a vertical position. Other modifications within the scope of my invention will appear to those skilled in the art.

Referring now to Figure 3, I have illustrated therein in greater detail a high temperature burner of the type schematically shown at 40 and 240 of Figures 1 and 2 respectively. While

the general principles of design and operation of burners of this type have been suggested by others prior to the present invention their specific adaptation to the gasification of carbonaceous solids employing the fluid solids technique forms an essential element of the present invention.

The burner, as illustrated, may consist of an outer steel shell 340 provided with an inner refractory lining 350 preferably of chrome ore having a fusion point not substantially below 2700° F. Cooling tubes 352 are imbedded in the refractory lining and supplied with cooling water through water header 354. Heated water and/or steam is withdrawn from the cooling tubes 352 through header 356. Manifold 324, preferably in the top of the combustion zone, serves the introduction of high velocity secondary combustion-supporting gas and/or diluent.

Carbonaceous solids having a particle size substantially as described above are supplied from the gas generator or heater through standpipe 343. About 1-15% of the total combustion-supporting gas required may be added through tap 339. The fluidized solids flow under the pressure of standpipe 343 to feed device 353 arranged on the horizontal axis of cylindrical shell 340. An additional amount of about 1-5% of the total combustion-supporting gas may be added through line 331.

The solids-gas mixture enters shell 340 tangentially along the axis, a more violently rotating motion being imparted to the combustion mixture by the tangential high velocity secondary combustion-supporting and/or diluent gas supplied through 324, which may amount to as much as 98% of the total combustion-supporting gas. The burner is operated above ash fusion temperature and liquid ash is withdrawn through a bottom orifice 344 close to the lower discharge end of the burner. Hot flue gases are withdrawn through port 345 for further use in the process, preferably after passing through an entrainment separator wherein entrained fluid ash may be removed in any conventional manner.

The violent swirling action of the combustion mixture causes an extremely intimate contact between fuel and combustion-supporting gas and makes possible the attainment of extremely high combustion temperatures in a relatively small combustion space. The cooling tubes 352 are operated so as to maintain the refractory lining in an effective operating condition, while withdrawing only a minimum amount of heat from the chamber.

The absolute dimensions of the burner depend, of course, on the output desired. It may be stated, however, that a combustion space of about 1 to 20 cu. ft. per 1,000 lbs. of coal per hour to be gasified in generators 10 or 210 is generally sufficient. Many modifications of the burner illustrated in Figure 3 may occur to those skilled in the art without deviating from the spirit of the invention.

While I have described above specific burner means suitable for the purposes of my process, the present invention is not limited to these specific means but is intended to include any means for completely burning finely divided carbonaceous solids of a high carbon concentration, say, above about 15% at temperatures above the fusion point of the ash.

It should also be understood that certain of the advantages of my invention will be realized when fixed or moving beds or dilute solids-in-gas sus-

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pensions are used in my process instead of the fluidized beds described above.

My invention will be further illustrated by the following specific examples.

EXAMPLE I

The continuous production of producer gas from bituminous coal using 95% oxygen and steam in a system of the type illustrated in Figure 1 of the drawing may be carried out at the conditions specified below.

Coal to fluid generator 10

Pounds per hour	100,000
Temperature, °F	60
Moisture, percent	1.5
Volatile matter, percent	33.0
Fixed carbon, percent	55.5
Ash, percent	10.0
Ash fusion temperature, °F	2500

Fluid generator 10

Temperature, °F	1800
Outlet superficial gas velocity, ft./sec	1.5
Fluidized solids, density, lbs./c. f	18
Fluidized solids, carbon concentration percent	40
H ₂ O conversion, percent	80

Burner 40

Temperature, °F	3000
Solids from generator, lb./hr	16,600
Carbon concentration, wt. percent	40
95% O ₂ to burner, lb./hr	77,500
95% O ₂ to burner, temperature, °F	800
H ₂ O to burner, lb./hr	74,000
H ₂ O to burner, temperature, °F	1700
Fused ash discharged, lb./hr	10,000
Gases and vapors to generator:	
CO ₂ , percent	8.4
O ₂ , percent	27.0
N ₂ , percent	1.9
H ₂ O, percent	62.7
	100.0

Dry gas from generator 10

M. M. s. c. f./hr	4.07
Composition:	
CO	56.9
H ₂	30.6
CO ₂	8.5
N ₂	1.1
CH ₄	2.9
	100.0
B. t. u./c. f., net	194

EXAMPLE II

For the continuous production of water gas from bituminous coal using a two vessel system with an auxiliary burner as illustrated in Figure 2, the following conditions may be employed.

Coal to fluid generator 210

Pounds per hour	100,000
Temperature, °F	60
Moisture, per cent	1.5
Volatile matter, per cent	33.0
Fixed carbon, per cent	55.5
Ash, per cent	10.0
Ash fusion temperature, °F	2500

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Fluid generator 210

Temperature, °F	1800
Outlet superficial gas velocity, ft./sec	1.5
Fluidized solids, density, lb./c. f	18
Fluidized solids, carbon concentration, per cent	50
H ₂ O, temperature, °F	1700
H ₂ O, conversion, per cent	80
Solids to fluid heater 230, lb./hr	10,730,000
Solids from fluid heater 230, lb./hr	10,700,000

Dry gas from generator 210

M. M. s. c. f./hr	3.61
Composition:	
CO	39.1
H ₂	50.8
CO ₂	5.8
N ₂	1.0
CH ₄	3.3
	100.0
B. t. u./c. f., net	296

Fluid heater 230

Temperature, °F	1900
Outlet superficial gas velocity, ft./sec	1.5
Fluidized solids, density, lb./c. f	18
Fluidized solids, carbon concentration, per cent	50
CO ₂ /CO ratio in outlet gas	4
Solids to burner, lb./hr	20,000

Burner 240

Temperature, °F	3000
Solids from fluid heater 230, lb./hr	20,000
Carbon concentration, weight per cent	50
Air to burner 240, lb./hr	336,000
Air to burner 240, temperature °F	1400
Fused ash discharged, lb./hr	10,000
Gases to fluid heater 230:	
CO ₂	7.3
O ₂	13.7
N ₂	79.0
	100.0

The foregoing description and exemplary operations have served to illustrate specific applications and results of my invention. However, other modifications obvious to those skilled in the art are within the scope of my invention. Only such limitations should be imposed on the invention as are indicated in the appended claims.

I claim:

1. The process of converting solid carbonaceous fuels into gases containing carbon monoxide by an endothermic reaction with a gaseous gasifying medium conducted at an elevated conversion temperature in a conversion zone, which comprises contacting carbonaceous solids with a sufficient amount of said gasifying medium at a conversion temperature to convert a substantial proportion of the carbon of said fuels into said gases in said conversion zone, withdrawing product gas from said conversion zone, withdrawing solid carbonaceous gasification residue from said conversion zone, passing said withdrawn residue to a heating zone, subjecting said residue in said heating zone to a heat-generating incomplete combustion at a temperature substantially above said conversion temperature and below the fusion point of its ash, withdrawing flue gas from said heating zone, withdrawing carbonaceous heating

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zone residue from said heating zone, passing a portion of said heating zone residue to a combustion zone, subjecting said portion to combustion with a combustion-supporting gas in said combustion zone at a combustion temperature above the fusion point of the ash of said carbonaceous solids and at a ratio of combustion-supporting gas at least sufficient to burn completely the carbonaceous constituents of said portion subjected to said combustion, withdrawing liquid ash from said combustion zone, passing hot flue gases from said combustion zone substantially at the temperature of said combustion to said heating zone and passing another portion of said heating zone residue substantially at the temperature of said heating zone to said conversion zone to supply at least a portion of the heat required by said reaction.

2. The process of claim 1 in which sufficient oxygen is supplied to said heating zone to favor the formation of CO₂ over that of CO.

3. The process of claim 1 in which a portion of the total oxygen supplied for said heating and combustion zones is fed to said heating zone and another portion to said combustion zone so as to supply sufficient oxygen to said heating zone to favor the formation of CO₂ over that of CO in said heating zone.

4. The process of converting solid carbonaceous fuels into gases containing carbon monoxide by an endothermic reaction with a gaseous gasifying medium conducted at an elevated conversion temperature in a conversion zone, which comprises contacting finely divided carbonaceous solids in the form of a dense turbulent mass fluidized by an upwardly flowing gas with a sufficient amount of said gasifying medium at a conversion temperature to convert a substantial proportion of the carbon of said fuels into said gases, withdrawing product gas upwardly from said mass, withdrawing finely divided solid carbonaceous gasification residue from said mass, passing said withdrawn residue to a heating zone, subjecting said residue in said heating zone to a heat-generating incomplete combustion in the form of a dense turbulent bed fluidized by an upwardly flowing

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gas at a temperature substantially above said conversion temperature and below the fusion point of its ash, withdrawing flue gas upwardly from said heating zone, withdrawing finely divided carbonaceous heating zone residue from said heating zone, passing a portion of said heating zone residue to a combustion zone, subjecting said portion to combustion with a combustion-supporting gas in a combustion zone at a combustion temperature above the fusion point of the ash of said carbonaceous solids and at a ratio of combustion-supporting gas at least sufficient to burn completely the carbonaceous constituents of the portion of said residue subjected to said combustion, withdrawing liquid ash from said combustion zone, passing hot flue gases from said combustion zone substantially at the temperature of said combustion to said heating zone to supply additional heat thereto and passing another portion of said heating zone residue substantially at the temperature of said heating zone to said conversion zone to supply at least a portion of the heat required by said reaction.

5. The process of claim 2 in which sufficient oxygen is supplied to said heating zone to favor the formation of CO₂ over that of CO.

6. The process of claim 2 in which a portion of the total oxygen supplied for said heating and combustion zones is fed to said heating zone and another portion to said combustion zone so as to supply sufficient oxygen to said heating zone to favor the formation of CO₂ over that of CO in said heating zone.

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