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[54] **TRANSFER LINE BURNER USING GAS OF LOW OXYGEN CONTENT**

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[51] **Int. Cl.** **C10j 3/12**

[58] **Field of Search** .... 48/202, 204, 206, 210, 48/197 R, DIG. 4; 201/38, 31; 252/373

[56] **References Cited**

UNITED STATES PATENTS

2,588,075	3/1952	Barr et al.	48/206
2,654,665	10/1953	Phinney	48/206
2,694,623	11/1954	Welty Jr. et al.	48/197
2,729,552	1/1956	Nelson et al.	48/197
2,884,303	4/1959	Metrailer	23/1
2,912,315	11/1959	Haney	48/214

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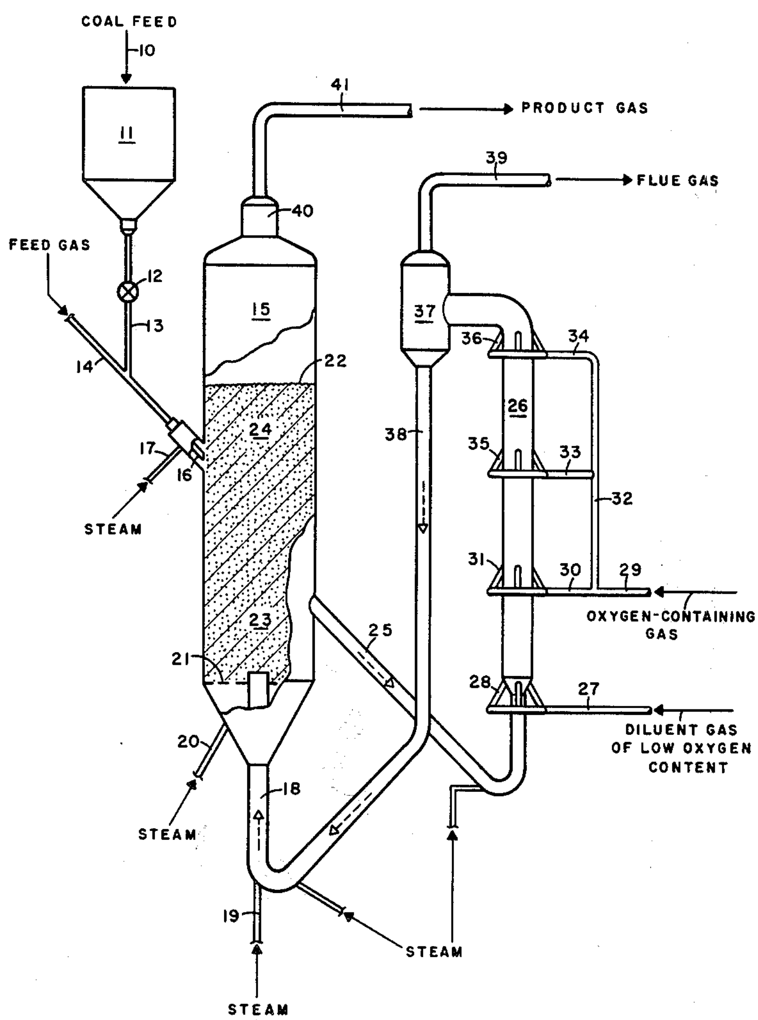
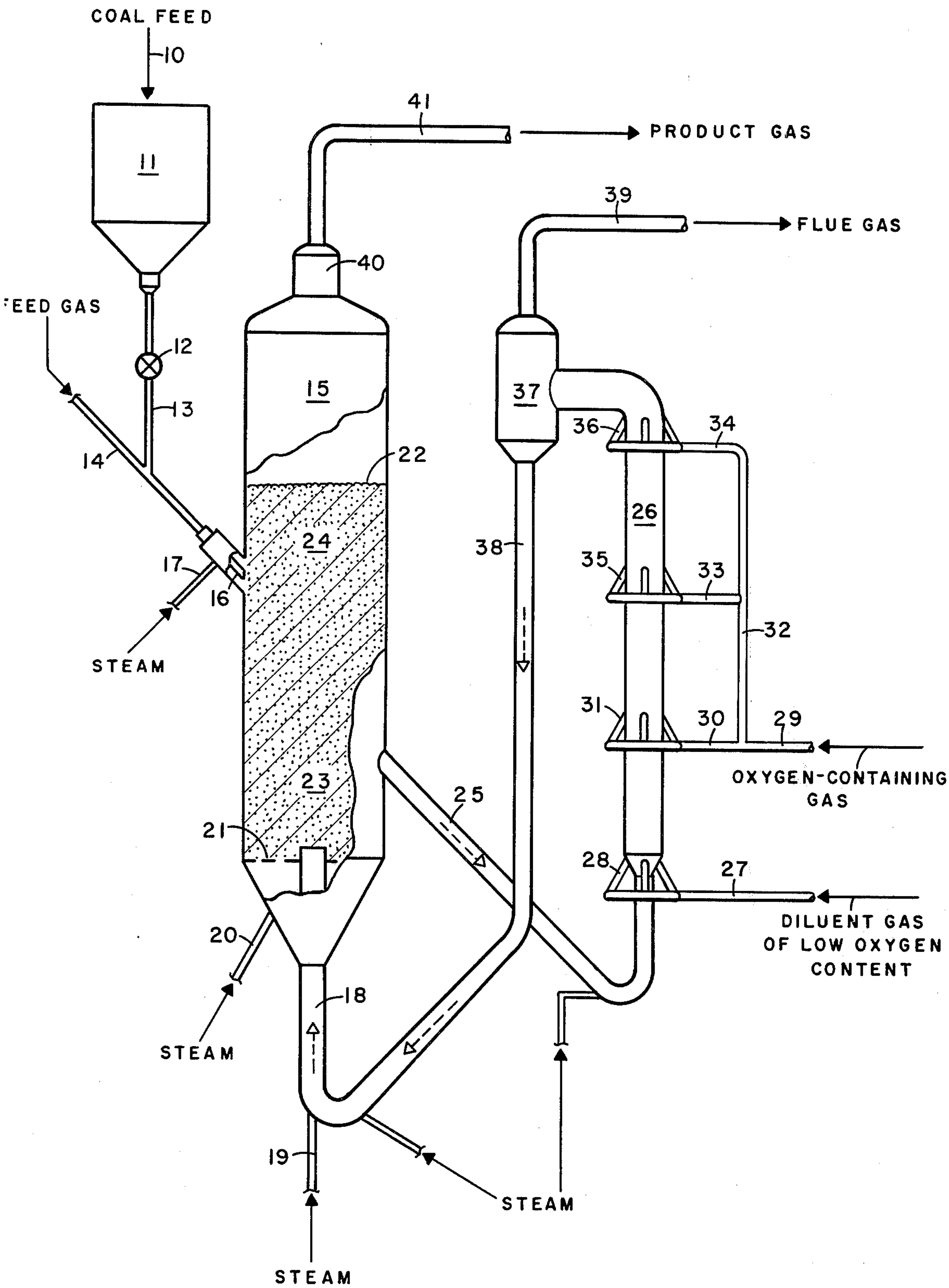
[57] **ABSTRACT**

Carbonaceous solids are heated by passing a stream of such solids through a transfer line burner, introducing a stream of gas of low oxygen content into the burner near the lower end thereof in a quantity sufficient to produce a transition from dense phase to dilute phase flow, and introducing into the upper portion of the burner at one or more points a gas of higher oxygen content in a quantity sufficient to promote the conversion of carbon to carbon dioxide and the generation of sufficient heat to raise the temperature of the solids to the desired level.

15 Claims, 1 Drawing Figure

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PRODUCT GAS
FLUE GAS
OXYGEN-CONTAINING GAS
DILUENT GAS OF LOW OXYGEN CONTENT
STEAM
STEAM

TRANSFER LINE BURNER USING GAS OF LOW OXYGEN CONTENT

BACKGROUND OF THE INVENTION

1. Field of the Invention

This invention relates to the heating of fluidized beds containing coal particles or other carbonaceous solids and is particularly directed to coal gasification and related processes in which heat is generated by burning a portion of the carbonaceous solids in a transfer line burner.

2. Background of the Invention

Fluidized bed processes for the conversion of coal and similar carbonaceous materials into gases useful as fuels normally require the continuous input into the system of large quantities of heat. One of the methods by which this can be done is through the use of a transfer line burner. Such a burner normally consists of a large vertical line into the lower end of which finely divided coal char or similar carbonaceous material is introduced from the fluidized bed reaction zone. Air or other oxygen-containing gas is introduced into the burner near the lower end thereof in a sufficient quantity to carry the carbonaceous solids upwardly through the burner in dilute phase flow. The oxygen present in the gas stream burns a portion of the carbon from the solids, thus generating sufficient heat to raise the unburned particles to the desired temperature level. The hot solids carried overhead by the combustion gases are removed by means of cyclone separators or similar equipment and recycled to the fluidized bed. The gases, which will normally contain ash and fines not removed by the separation equipment, are then generally scrubbed and handled in the conventional manner.

The use of a transfer line burner for heating solids withdrawn from a fluidized bed as described above has important advantages over other heat-generating systems but also has certain disadvantages. Experience has shown that the combustion taking place in the burner is difficult to control. The transient temperature rise which takes place directly above the air injection point is high and may result in localized overheating in areas where the relative ratio of air to char or other carbonaceous solids is higher than average. Maldistribution is particularly apt to occur as the char entering the burner moves from the region of dense phase flow into the region of dilute phase flow near the point of air injection. The char solids employed may contain ash having a fusion temperature only slightly higher than the desired transfer line exit temperature and hence fused ash deposits may form in such areas of localized overheating. Such deposits may produce plugging problems and other difficulties. In addition, the efficiency with which the combustion air is utilized may be reduced considerably due to the reduction of carbon dioxide to carbon monoxide as the combustion gases move upwardly in contact with the hot char. This increases the cost of operating the transfer line burner and may make it more difficult to handle the flue gases.

SUMMARY OF THE INVENTION

The present invention provides an improved method for the operation of transfer line burners used in coal gasification and similar operations which at least in part eliminates the difficulties outlined above. In accordance with the invention, it has now been found that the operation of such burners can be significantly improved by injecting sufficient flue gas or other gas of low oxygen content into the lower end of the burner to convey the carbonaceous particles upwardly in the burner and produce a transition from dense phase to dilute phase flow and then injecting air or other gas of higher oxygen content into the burner at one or more higher points in a quantity sufficient to convert carbon to carbon dioxide and generate the heat required to elevate the carbonaceous particles to the desired final temperature. The gas introduced near the lower end of the burner to effect the transition from dense phase to dilute phase flow may contain up to about **10** percent oxygen by volume but will preferably contain less than about **6** percent by volume. This results in relatively low temperatures in the lower part of the burner and thus avoids localized overheating and ash fusion problems in the area where the transition from dense phase flow to dilute phase flow takes place. The introduction of air or other gas of higher oxygen content at one or more points in the upper part of the burner permits generation of the heat required to raise the temperature of the solids to the desired level and reduces the time during which carbon dioxide contacts the hot carbonaceous solids. This in turn decreases the production of carbon monoxide by the reaction: $CO_2 + C \rightarrow 2CO$, and improves the thermal efficiency of the process.

Studies have shown that the rate at which carbon dioxide reacts with the solid carbonaceous particles tends to be proportional to the total surface area of the particles, including the area of the internal pores; whereas the primary combustion reaction wherein carbon is burned to carbon dioxide tends to proceed at a rate proportional to only the external surface area of the solid particles. Since the internal surface area is normally much greater than the external area, the reaction of carbon dioxide with carbon to form carbon monoxide tends to reduce the size of large and small particles alike. Because the fine particles have a much greater external surface-to-internal surface ratio than do the larger particles, the introduction of the combustion air into the upper part of the burner tends to minimize the overall reduction in particle size and instead promotes the preferential burning of fines carried over from the fluidized bed reaction zone to generate heat and carbon dioxide. This results in more efficient utilization of the carbonaceous solids by reducing the amount of unburned fines carried overhead with the flue gases.

As indicated above, the method of the invention makes possible significant improvements in the operation of transfer line burners for the heating of coal char particles and similar finely-divided carbonaceous solids. The introduction of recycled flue gas or a similar gas stream of low oxygen content into the lower part of the burner to produce a transition from dense phase to dilute phase flow and the injection of air or other oxygen-containing gas for combustion of carbon to carbon dioxide into the burner at a higher point or points eliminates localized overheating in the transition zone and related problems due to ash fusion, results in more efficient combustion and the transfer of more heat to the suspended solids, and permits significant reductions in the amount of carbon monoxide and fines in the flue gases, thus reducing pollution problems and simplifying later treatment of the gas stream to comply with applicable pollution control regulations. These and other advantages provide economic incentives for use of the method.

BRIEF DESCRIPTION OF THE DRAWING

The single FIGURE in the drawing is a schematic flow sheet of a process for the production of a methane-rich gas from coal in which the improved transfer line burner of the invention is employed.

DESCRIPTION OF THE PREFERRED EMBODIMENTS

The drawing depicts a process for the production of product gases of relatively high methane content from bituminous coal, subbituminous coal, lignite, or similar carbonaceous solids. The solid feed material employed in the process, preferably a bituminous or lower rank coal, is introduced into the system through line **10** from a suitable feed preparation plant or storage facility which is not shown in the drawing. To facilitate handling of the solid feed material in a fluidized state, the coal or other carbonaceous material is introduced into the system in the form of finely-divided particles, preferably less than about 8 mesh on the Tyler screen scale. The process depicted is operated at elevated pressure and hence the coal or other feed material introduced through line **10** is fed into vessel **11**, from which it is discharged through a star wheel feeder or similar feeding device **12** into line **13** at the system operating pressure or at a slightly higher pressure. In lieu of or in addition to vessel **11** and star wheel feeder **12**, parallel lock hoppers, aerated stand pipes operated in series, or other conventional equipment may be employed to permit introduction of the input stream of coal or other solids into the system at the required pressure.

A feed gas stream is introduced into the system through line **14** to permit the entrainment of solid particles from line **13** and the introduction of the solids into gasifier **15**. High pressure steam, product gas, or an inert gas may be employed as the feed gas. The use of recycled product gas simplifies downstream processing of the product and is normally preferred. The feed gas stream is introduced into the system at a pressure between about 50 and about 1000 psig, depending upon the pressure at which gasifier **15** is operated and the solid feed material employed. This stream is injected into the gasifier through one or more shrouded nozzles **16** into which steam is admitted through line **17** to keep the nozzle temperature below about 600° F. and thus minimize difficulties due to fouling of the nozzle with agglomerating coal solids. If an agglomerating coal feed material is employed, an injection nozzle designed to promote intimate and extremely rapid mixing of the injected coal with the hot solids present in the gasifier will normally be used. Nozzles especially designed for this purpose have been described in the literature and will be familiar to those skilled in the art.

The gasifier vessel **15** which is employed in the system shown in the drawing contains a fluidized bed of char particles introduced into the lower part of the vessel through line **18**. Steam for maintaining the particles in the fluidized state and reacting with the char to produce a synthesis gas containing substantial quantities of hydrogen and carbon monoxide is introduced into line **18** by means of line **19**. Additional steam may be introduced through line **20**. The total steam rate will generally range between about 0.5 and about 2.0 pounds of steam per pound of coal feed. The entering steam and char particles form a fluidized bed which extends upwardly above distribution grid or similar device **21** to a level above the point at which the coal particles are introduced through nozzle **16**. The upper surface of this fluidized bed is indicated by reference numeral **22**.

The lower portion of the fluidized bed in gasifier **15** between grid **21** and the level at which the coal solids are introduced, indicated generally by reference numeral **23**, serves as a steam gasification zone. Here the steam introduced through lines **19** and **20** reacts with carbon in the hot char to form synthesis gas in accordance with the reaction: $H_2O + C \rightarrow H_2 + CO$. At the point of steam injection near the bottom of the gasifier, the hydrogen concentration in the gaseous phase of the fluidized bed is essentially zero. As the steam moves upwardly through the fluidized char particles, it reacts with the hot carbon to produce synthesis gas and the hydrogen concentration in the gaseous phase therefore increases. The temperature in the steam gasification zone will generally range between about 1450° and about 1800° F. Depending upon the particular feed material and the particle sizes employed, the gas velocities in the fluidized bed will normally range between about 0.2 and about 3.0 feet per second.

The upper part of the fluidized bed in reactor vessel **15**, indicated generally by reference numeral **24**, serves as a hydrogasification zone where the feed coal is devolatilized and a portion of the rapidly convertible carbonaceous material in the coal reacts with hydrogen generated in the steam gasification zone to produce methane as one of the principal products. The point at which the coal feed stream is introduced into the gasifier through nozzle **16** and hence the location of the steam gasification and hydrogasification zones depends primarily upon the properties of the particular coal or other carbonaceous solid which is employed as the feedstock. It is generally preferred to select the nozzle location so that the methane yield from the gasifier will be maximized and the tar yield will be minimized. Generally speaking, the amount of methane produced increases as the coal feed injection point is moved nearer the top of the reaction vessel. The tar formed from some coals has a tendency to foul downstream processing equipment. The tar yield normally increases as the coal injection point is moved upwardly in the gasifier, and decreases as the coal input point is moved nearer the bottom of the reaction vessel, other operating conditions being the same. The coal feed stream should therefore generally be injected into gasifier **15** at a point where the hydrogen concentration in the gas phase is in excess of about 15 percent by volume, preferably between about 25 percent and about 50 percent by volume. To secure acceptable methane concentrations in the product gas stream, the upper surface **22** of the fluidized bed should normally be located at a level sufficiently above the nozzle **16** to provide at least about 4 seconds of residence time for the gas phase in contact with the fluidized solids in hydrogasification zone **24**. A residence time for the gas in contact with the solid phase above the point of coal feed injection between about 7 and about 20 sec. is generally preferred. It will be understood, of course, that the optimum coal injection point and the optimum gas residence time above the point of coal injection will vary with different types and grades of feed coal and will also change with variations in the gasifier temperature, pressure, steam rate, and other process conditions. Higher rank coals normally require somewhat more severe reaction conditions and longer gas residence times to obtain high methane yields than do coals of lower rank. Similarly, higher reactor temperatures and lower steam rates, for a given solids holdup below the coal injection point, generally tend to increase the hydrogen concentration in the gas phase and thus reduce the gas residence time needed for acceptable methane yields from a particular feed coal.

The heat required to sustain the overall endothermic reaction taking place within gasifier **15** and maintain the gasifier operating temperature within the range between about 1500° and about 1800° F. is provided by withdrawing a portion of the char solids from the fluidized bed through line **25** and passing this material into the lower end of transfer line burner **26.** Steam may be injected into line **25** in the vicinity of bends in the line in order to promote smooth flow of the solids and avoid any danger of clogging. The solid particles moving downwardly through line **25** will be in dense phase flow. Recycled flue gas or a mixture of flue gas and air containing up to about 10 percent, preferably less than 6 percent, oxygen by volume is introduced through line **27** and multiple injection nozzles **28** into the solids stream near the bottom of burner **26,** preferably in a volume sufficient to provide the particle and gas velocities necessary for the transition from dense phase to dilute phase flow. The recycled flue gas will normally be injected in an amount between about 0.1 and about 5 actual cubic feet of gas per pound of char solids, as measured at the temperature and pressure of the mixed gas and char. The superficial gas velocity in the transfer line burner above the flue gas injection nozzles **28** will normally range between about 15 and about 100 feet per second. The recycled flue gas will preferably contain less than 1 percent oxygen by volume and will consist primarily of nitrogen and carbon dioxide. Small amounts of water vapor, carbon monoxide, and other gases will, of course, also be present. As the recycled flue gas and entrained solids move upwardly within the transfer line burner, little combustion takes place. At a higher point in the burner, a gas of higher oxygen content, preferably air, is injected into the burner through lines **29** and **30** and multiple injection nozzles **31** spaced at intervals about the periphery of the burner to provide intimate contact between the upflowing solids and injected gas.

The oxygen in the injected gas stream rapidly reacts with carbon on the surface of the entrained particles to form carbon dioxide. This reaction is accompanied by the reduction of carbon dioxide by hot carbon to form carbon monoxide which takes place more slowly. After essentially all the free oxygen in the gas stream has been exhausted, this second reaction tends to cause a reduction in the carbon dioxide content and a corresponding increase in the carbon monoxide content of the gas. The discharge of substantial quantities of carbon monoxide represents a loss in the thermal efficiency of the process. To maintain an acceptably low carbon monoxide concentration in the burner gases discharged from the system, all or part of the oxygen used to support combustion should therefore be injected into the burner at a point or points sufficiently near the upper end of the burner that the exposure of carbon dioxide to hot carbon in the absence of free oxygen is minimized. Studies of a typical transfer line burner used in coal gasification operations have shown that the injected oxygen is generally consumed by the time the gases have moved downstream from the injection point a distance of from about 30 to about 36 inches. This distance will vary, of course, with changes in the dimensions of the burner, the volume of gas injected, the oxygen content of the gas, the type and amount of char solids entrained in the gas, and other factors. The point at which essentially all the oxygen has been consumed for a particular burner and said operating conditions can be calculated or determined by taking gas samples from the burner and hence the optimum injection point or points can be located.

As indicated earlier, it may be preferred to introduce the combustion oxygen into the burner at two or more points along the length of the burner in order to further reduce the danger of localized overheating and fusion of the ash. If this is done, the points selected should be sufficiently far apart to permit the consumption of substantially all of the oxygen introduced at each point before the oxygen introduced at the next point contacts the upflowing carbonaceous solids in the gas stream. In the system shown in the drawing, a portion of the oxygen-containing gas passes through line **32** and is introduced into the burner through one or more downstream injection lines **33** and **34.** The associated injection nozzles **35** and **36** are spaced about the periphery of the burner to provide intimate contact between the upflowing solids and injected gas, thus resulting in more efficient utilization of the oxygen and further reducing the danger of localized overheating of the entrained solids. The amount of oxygen introduced through nozzles **31, 35,** and **36** should be sufficient to generate enough heat to raise the temperature of the unburned solids entrained in the gas to a temperature of from about 50° to about 300° higher than that in the fluidized bed in reaction vessel **15.** A temperature rise of about 200° F. is generally preferred. The quantity of oxygen-containing gas needed to produce this temperature rise will depend upon the oxygen content of the injected gas, the input temperature of the gas stream, the type and quantity of carbonaceous solids moving upwardly within the burner, the composition of the flue gas, and the heat losses from the burner and can be readily calculated for any particular burner and set of operating conditions.

The rapid combustion of carbon on the surface of the entrained carbonaceous solids in response to the introduction of oxygen-containing gas at one or more points in the upper portion of the burner results in the conversion of carbon to carbon dioxide and the generation of sufficient heat to raise the temperature of the unburned solids to the desired level. The hot gases and entrained solids are withdrawn from the burner and passed rapidly to separation zone **37** to minimize the conversion of the carbon dioxide to carbon monoxide by reaction with the hot carbon. The separation zone will normally comprise one or more cyclone separators designed to remove entrained solids greater than about 325 mesh on the Tyler screen scale from the gas stream. These hot carbonaceous solids separated from the gas are withdrawn to dipleg **38** and returned to the fluidized bed through inlet **18** at the bottom of reaction vessel **15.** Flue gas containing fines not removed from the gas stream is taken overhead from the separation zone through line **39** and sent to downstream equipment for removal of the fines and recovery of heat from the gas. The amount of fines in the flue gas will generally be somewhat lower than in a conventional system where substantially greater conversion of carbon dioxide to carbon monoxide takes place. Similarly, the carbon monoxide content of the flue gas will be relatively low and hence further processing of the gas to comply with applicable pollution control regulations will be less costly than in a conventional system. Conventional equipment may be employed for the downstream removal of fines and recovery of heat from the gas.

The product gas formed in fluidized bed reaction vessel **15** passes through a separation zone **40** which will normally consist of one or more cyclone separators and is taken overhead through line **41.** The more effective utilization of relatively fine char particles in the burner system results in the return of fewer fines to the fluidized bed through dipleg **38** and injection nozzle **18** and thus reduces the range of particle sizes within the fluidized bed. This permits better control of the bed, makes possible higher gas velocities within the reactor, and reduces somewhat the volume of fines taken overhead with the product gas. This all tends to improve carbon utilization in the process and results in greater thermal efficiency than might otherwise be obtained.

The nature and objects of the invention can be further illustrated by considering the following specific example of a coal gasification process carried out in accordance therewith. A subbituminous Western coal dried to 4 weight percent moisture and ground and screened to a particle size less than about 8 Tyler mesh is fed into a gasifier system of the type shown in the drawing, at a feed rate of 100,000 pounds per hour. The ultimate analysis of the feed coal is:

	Pounds/Hour
Carbon	65,700
Hydrogen	4,600
Oxygen	16,400
Nitrogen	900
Sulfur	500
Ash	7,900
Water	4,000
Total	100,000

The feed coal is introduced into the gasifier with 16,700 pounds per hour of steam. The gasifier operates at a pressure of about 165 psig.

The coal flows into the upper hydrogasification zone where devolatilization takes place. Steam at a temperature of 400° F. and a pressure of 175 psig is introduced into the lower part of the gasifier at the rate of 47,300 pounds per hour. Synthesis gas produced by reaction of the steam with char in the steam gasification zone moves upwardly through the fluidized bed with a gas velocity of about 1.0 foot per second. In the hydrogasification zone, this synthesis gas reacts with the volatile products from the feed coal to produce the methane-rich product gas. This gas is taken overhead from the gasifier at the rate of 113,800 pounds per hour. The component mass rates for this gas are as follows:

	Pounds/Hour
CO	35,700
CO_2	26,600
H_2	4,400
H_2O	31,200
CH_4	9,000
C_2^+ Hydrocarbons	4,600
Other	2,300
Total	113,800

To maintain the fluidized bed temperature of about 1650° F., char particles are continuously withdrawn from the fluidized bed in the gasifier and passed to the transfer line burner at a rate of about 2,138,000 pounds per hour. To promote transition of the solid stream from dense phase to dilute phase flow, diluent gas is introduced into the lower end of the burner at the rate of about 22,600 standard cubic feet per minute. This is equivalent to about 0.21 actual cubic feet per pound of char solids. This diluent gas consists of flue gas recycled from the burner and has the following component mass rates:

	Pounds/Hour
CO	4,600
CO_2	27,300
H_2	100
H_2O	1,200
N_2	74,600
Others	1,400
Total	109,200

The diluent gas rate is sufficient to produce a superficial gas velocity of 30 feet per second in the lower section of the transfer line burner.

Air is introduced into the upper portion of the transfer line burner at the rate of 55,000 standard cubic feet per minute, having the following component mass rates:

	Pounds/Hour
O_2	58,400
N_2	190,200
Other	3,200
Total	251,800

The air stream will normally be preheated to a temperature of 450° F. The combustion gases and entrained solids are taken overhead from the burner at a temperature of about 1850° F. and separated to remove solid particles greater than about 325 Tyler mesh. The flue gas stream from the separator is treated downstream for the removal of fines. The flue gas stream has the following component mass rates:

	Pounds/Hour
CO	16,400
CO_2	96,400
H_2	200
H_2O	13,000
N_2	263,700
Others	4,800
Total	394,500

About 16,700 pounds per hour of fines having an average particle size of about 20 microns are recovered from the flue gas stream. The larger particles removed from the gas in the separation zone at the upper end of the transfer line burner are recycled to the gasifier at the rate of about 2,087,800 pounds per hour. A portion of the flue gas is cooled and recompressed for use as the diluent gas shown above.

What is claimed is:

1. In an endothermic process carried out in a fluidized bed wherein a stream of carbonaceous solids is continuously withdrawn from the fluidized bed and introduced into the lower end of a transfer line burner for the generation of heat and wherein unburned solids are withdrawn from the said transfer line burner near the upper end thereof and returned to the said fluidized bed, the improvement which comprises introducing a gas of low oxygen content containing up to about **10**

volume percent of oxygen into said transfer line burner containing said carbonaceous solids near the lower end thereof in a quantity sufficient to produce dilute phase flow of said carbonaceous solids upwardly within said transfer line burner and introducing a gas having a higher oxygen content than said gas of low oxygen content into an upper portion of said transfer line burner in an amount sufficient to raise the temperature of said unburned solids to a level above the temperature in said fluidized bed.

2. A process as defined by claim **1** wherein said gas of low oxygen content is a flue gas.

3. A process as defined by claim **1** wherein said gas of higher oxygen content is air.

4. A process as defined by claim **1** wherein said gas of higher oxygen content is introduced into said transfer line burner at a plurality of points spaced about the periphery of said transfer line burner.

5. A process as defined by claim **1** wherein said gas of higher oxygen content is introduced into said transfer line burner at at least two points sufficiently far apart to permit the consumption of substantially all of the oxygen introduced at one of said points before the oxygen introduced at the other of said points contacts said carbonaceous solids.

6. A process as defined by claim **1** wherein said gas of low oxygen content is a mixture of flue gas and air containing less than about 6 percent oxygen by volume.

7. A process as defined by claim **1** wherein at least part of said gas of higher oxygen content is introduced into said transfer line burner at a point near the upper end thereof.

8. A process as defined by claim **1** wherein said gas of higher oxygen content is introduced into said transfer line burner in a quantity sufficient to raise the temperature of said solids about 50° to about 300° F. above the temperature of said fluidized bed.

9. A process as defined by claim **1** wherein said carbonaceous solids comprise coal char particles.

10. In a coal gasification process wherein coal char particles are continuously withdrawn from a fluidized bed gasifier, passed to a transfer line burner for the combustion of carbon and the generation of heat, and the heated particles are thereafter returned to said gasifier, the improvement which comprises introducing a gas of low oxygen content containing up to about 10 percent oxygen by volume and comprising flue gas into said transfer line burner containing said char particles near the lower end thereof in an amount sufficient to effect a transition in the flow of said particles from dense phase flow to dilute phase flow, introducing air into an upper portion of said burner in an amount sufficient to heat said char particles to a temperature about 50° to about 300° F. above the temperature of said fluidized bed, withdrawing a gas stream containing entrained particles from the upper end of said burner, separating entrained particles from said gas stream, and returning the separated particles to said gasifier.

11. A process as defined by claim **10** wherein said gas of low oxygen content is a flue gas containing less than 1 percent oxygen by volume.

12. A process as defined by claim **10** wherein said gas of low oxygen content is introduced into said burner in an amount between about 0.1 and about 5 cubic feet per pound of char solids, measured at the burner temperature and pressure conditions.

13. A process as defined by claim **10** wherein the superficial gas velocity in said burner above the point at which said gas of low oxygen content is introduced is maintained between about 15 and about 100 feet per second.

14. A process as defined by claim **10** wherein said air is introduced into said burner at at least two vertically-spaced points sufficiently far apart to permit the consumption of substantially all of the oxygen introduced at the lowermost of said points before oxygen introduced at a higher point contacts said char particles.

15. A process as defined by claim **10** wherein said gas of low oxygen content is a mixture of flue gas and air containing less than about 6% oxygen by volume.

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