

Jan. 3, 1956

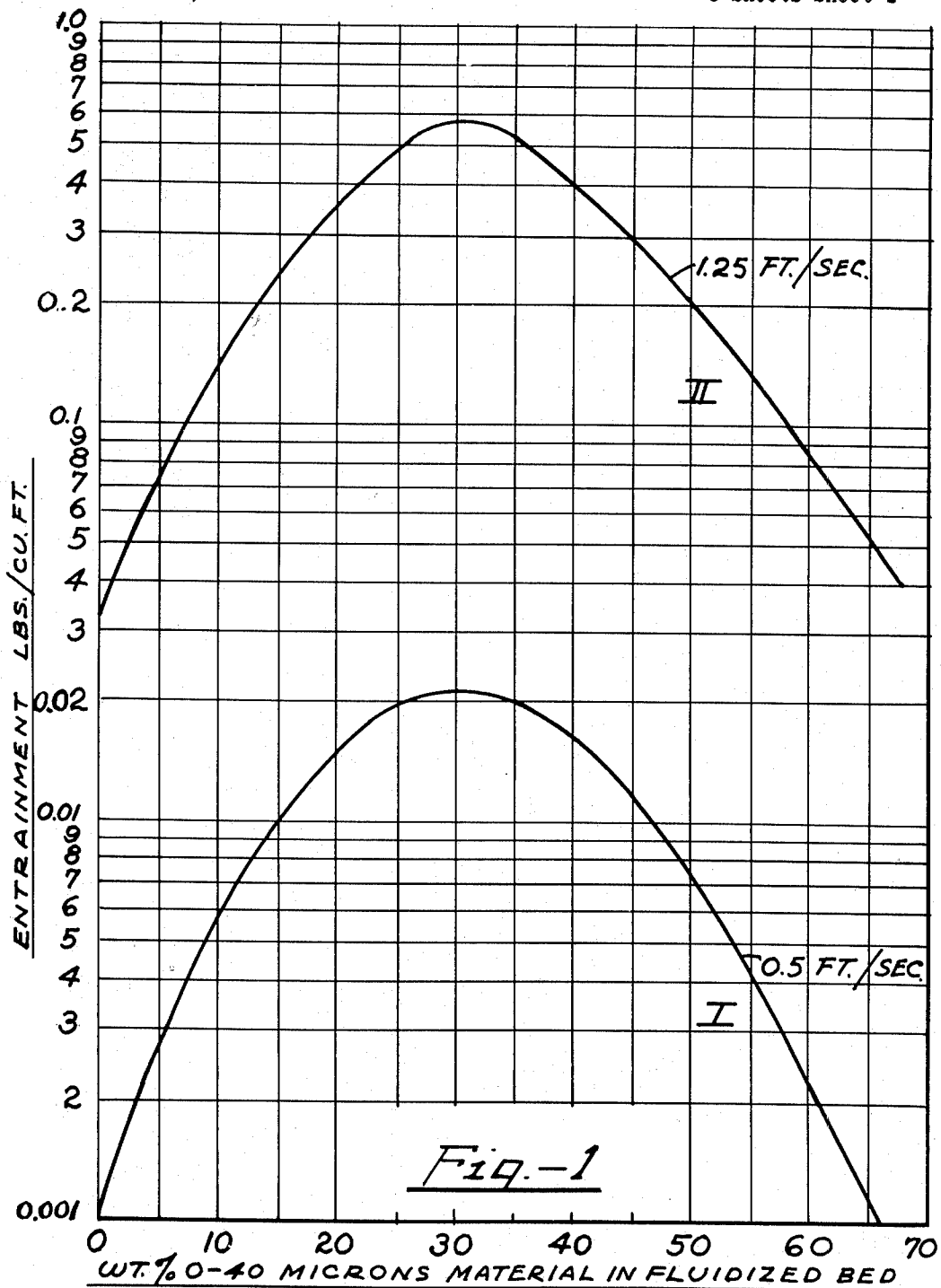
K. J. NELSON ET AL

2,729,552

PROCESS OF CONTACTING GASIFORM CARBONACEOUS SOLIDS

Filed Dec. 24, 1949

8 Sheets-Sheet 1



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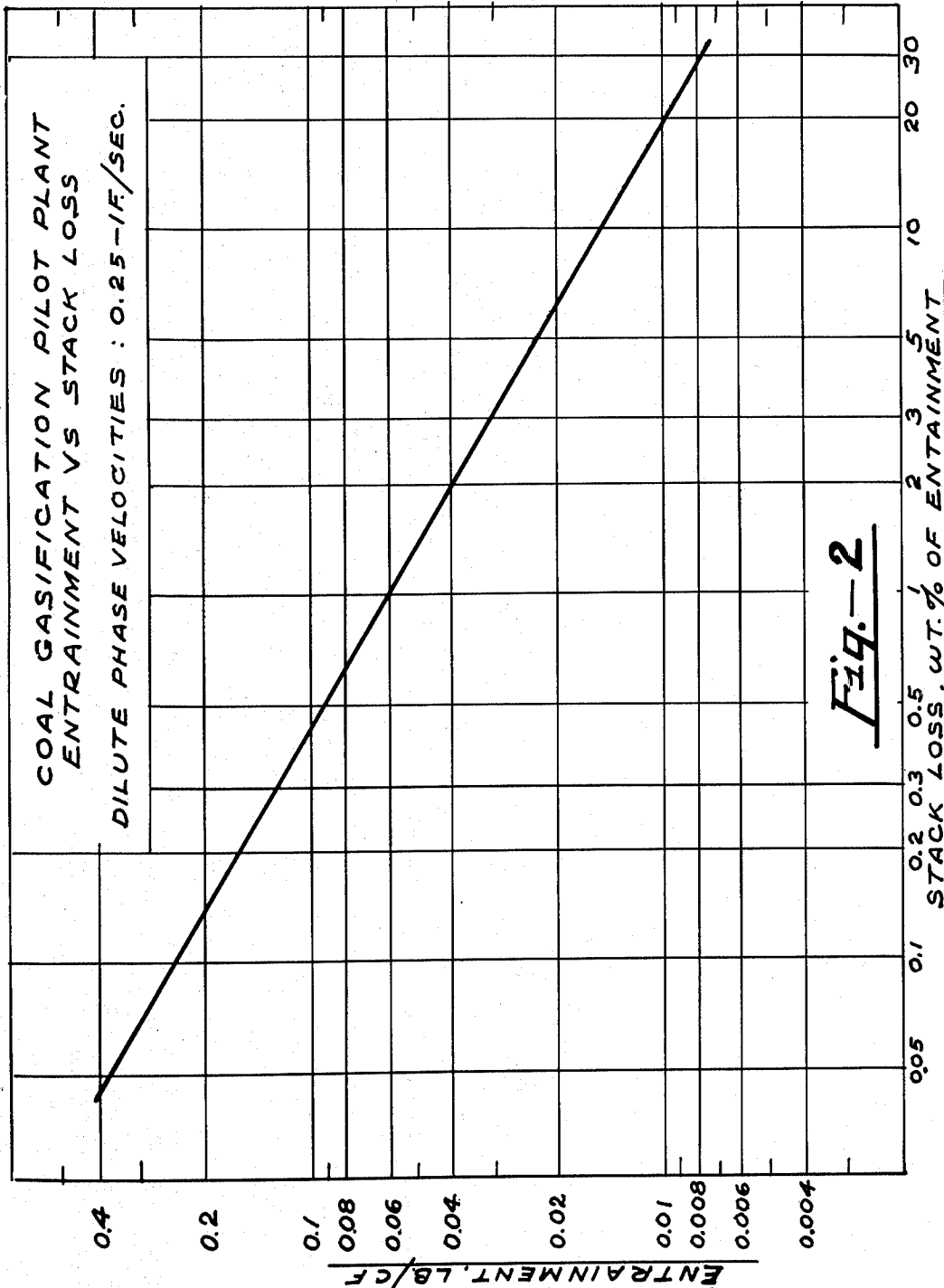
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PROCESS OF CONTACTING GASIFORM CARBONACEOUS SOLIDS

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8 Sheets-Sheet 2



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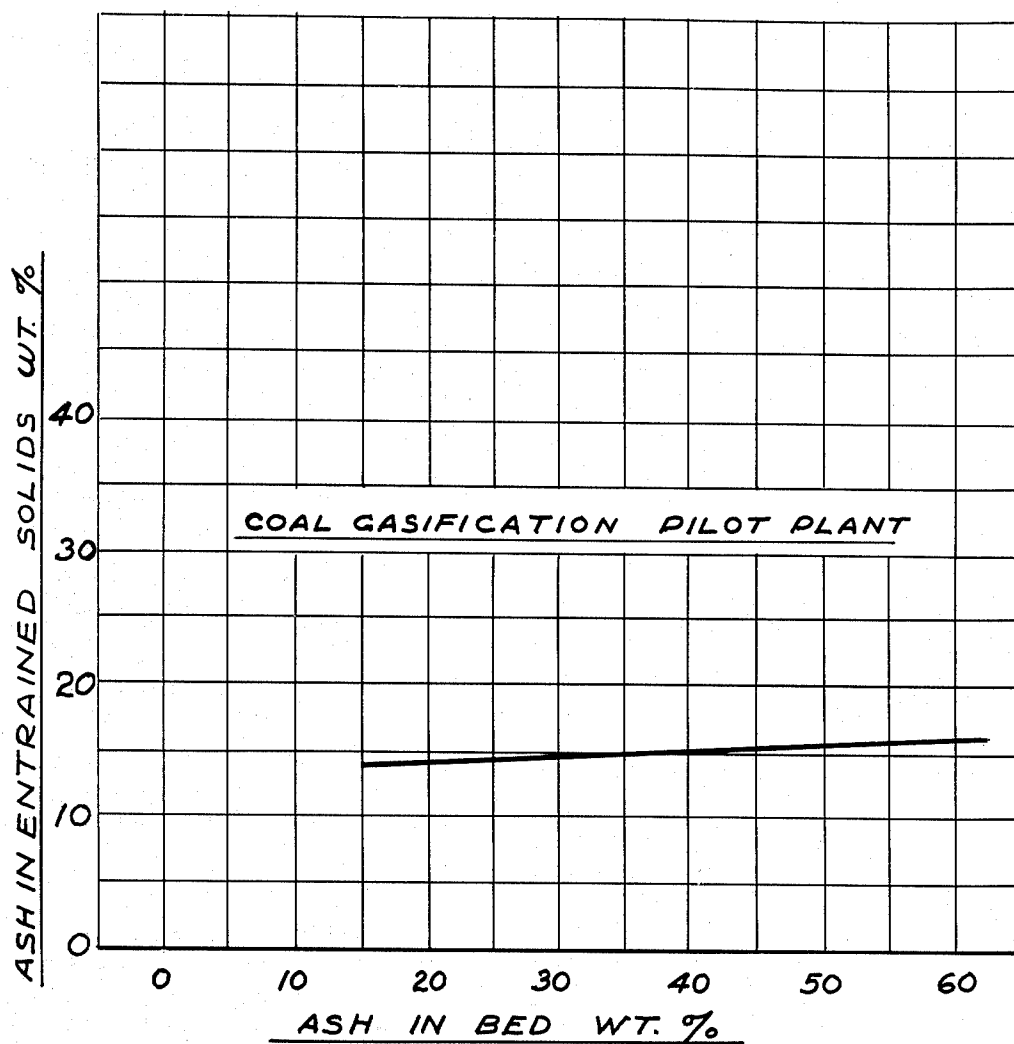
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PROCESS OF CONTACTING GASIFORM CARBONACEOUS SOLIDS

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8 Sheets-Sheet 3

Fig.-3



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8 Sheets-Sheet 4

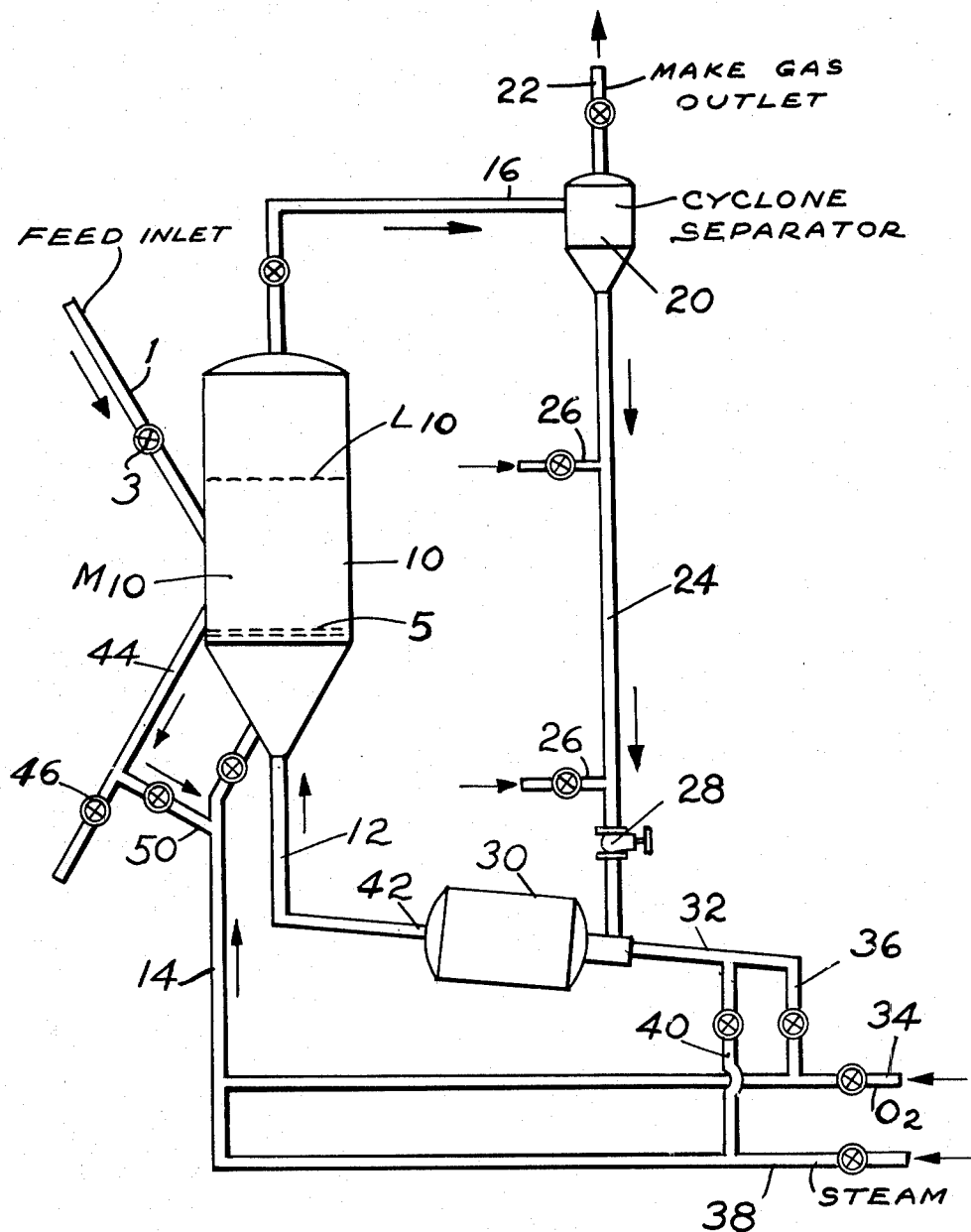


Fig. -4

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PROCESS OF CONTACTING GASIFORM CARBONACEOUS SOLIDS

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8 Sheets-Sheet 6

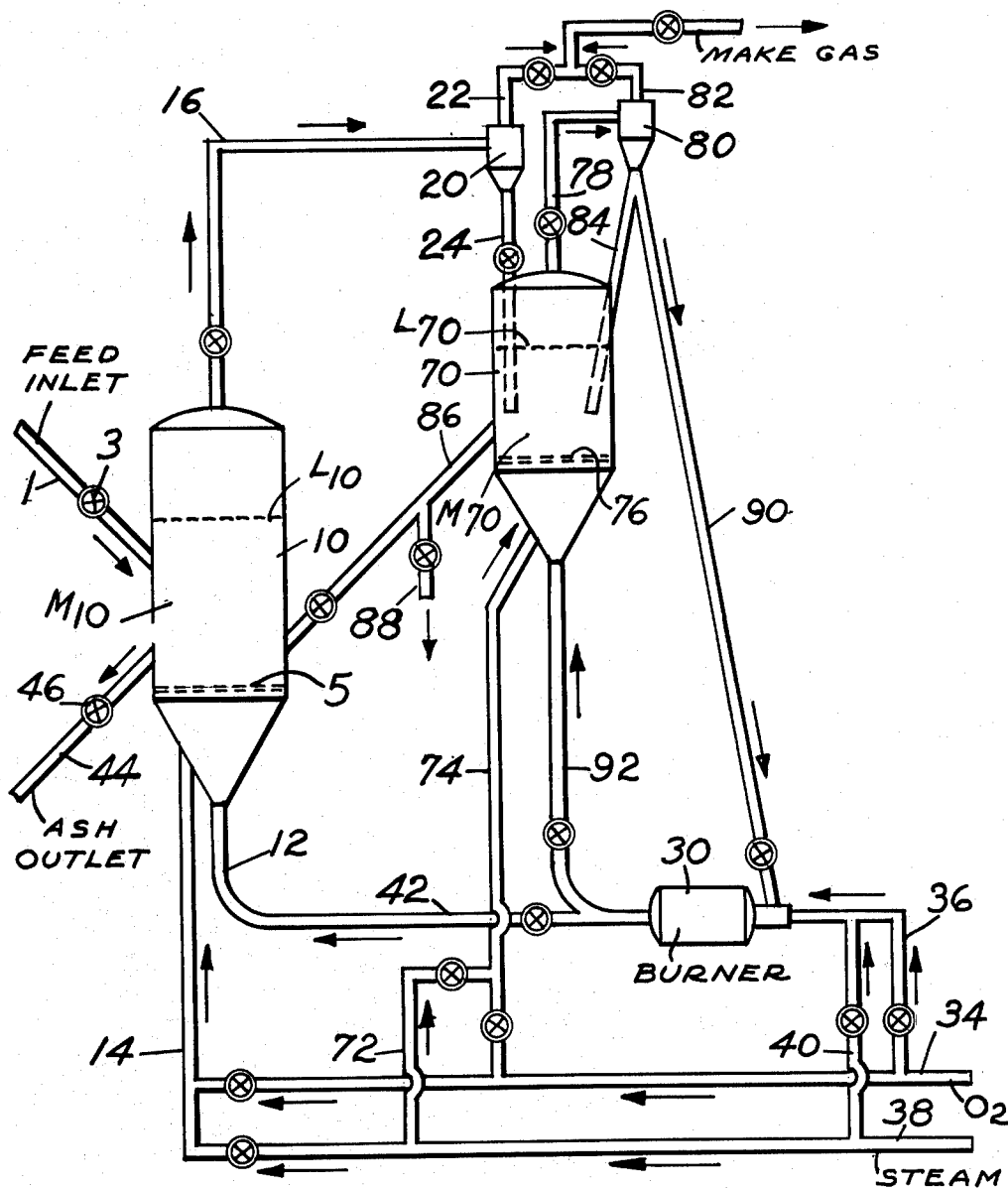


Fig. - 6

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PROCESS OF CONTACTING GASIFORM CARBONACEOUS SOLIDS

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8 Sheets-Sheet 7

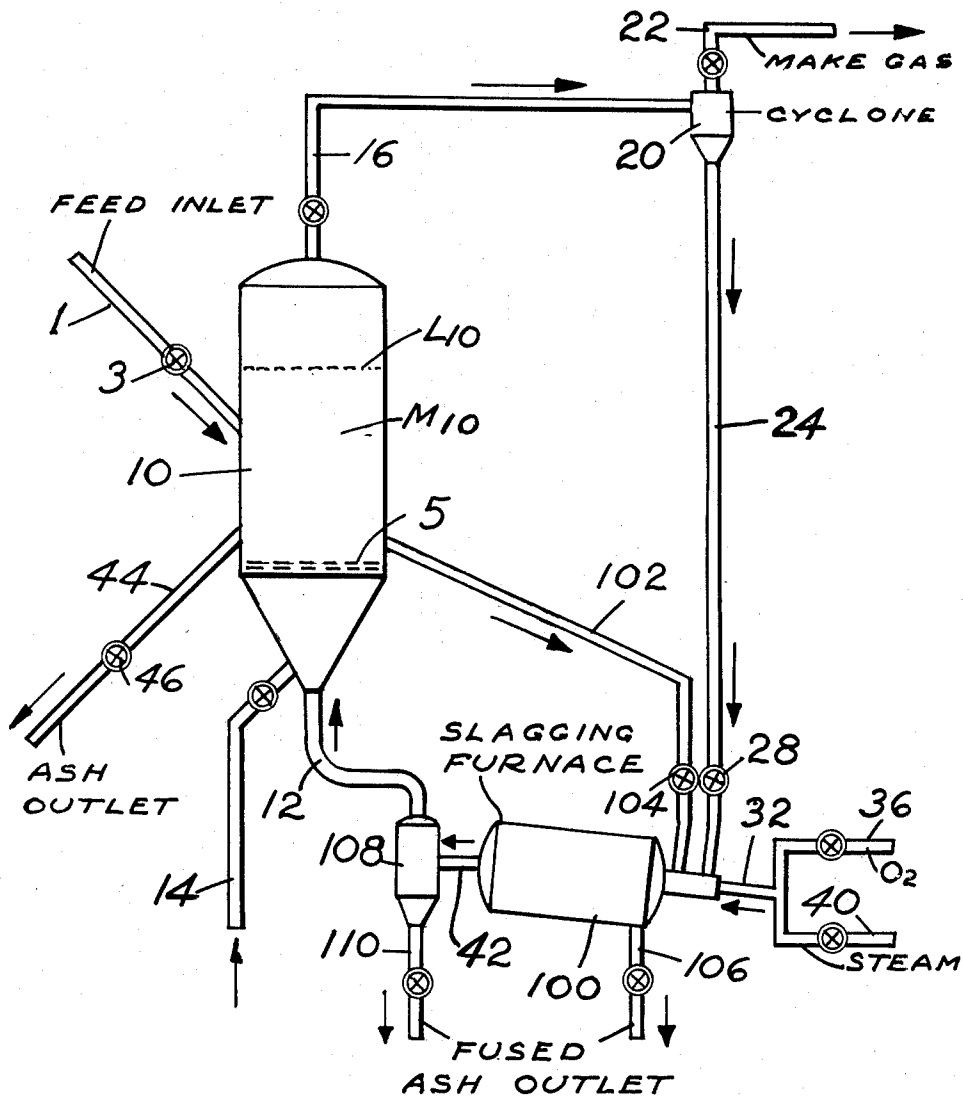


Fig. -7

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PROCESS OF CONTACTING GASIFORM CARBONACEOUS SOLIDS

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8 Sheets-Sheet 8

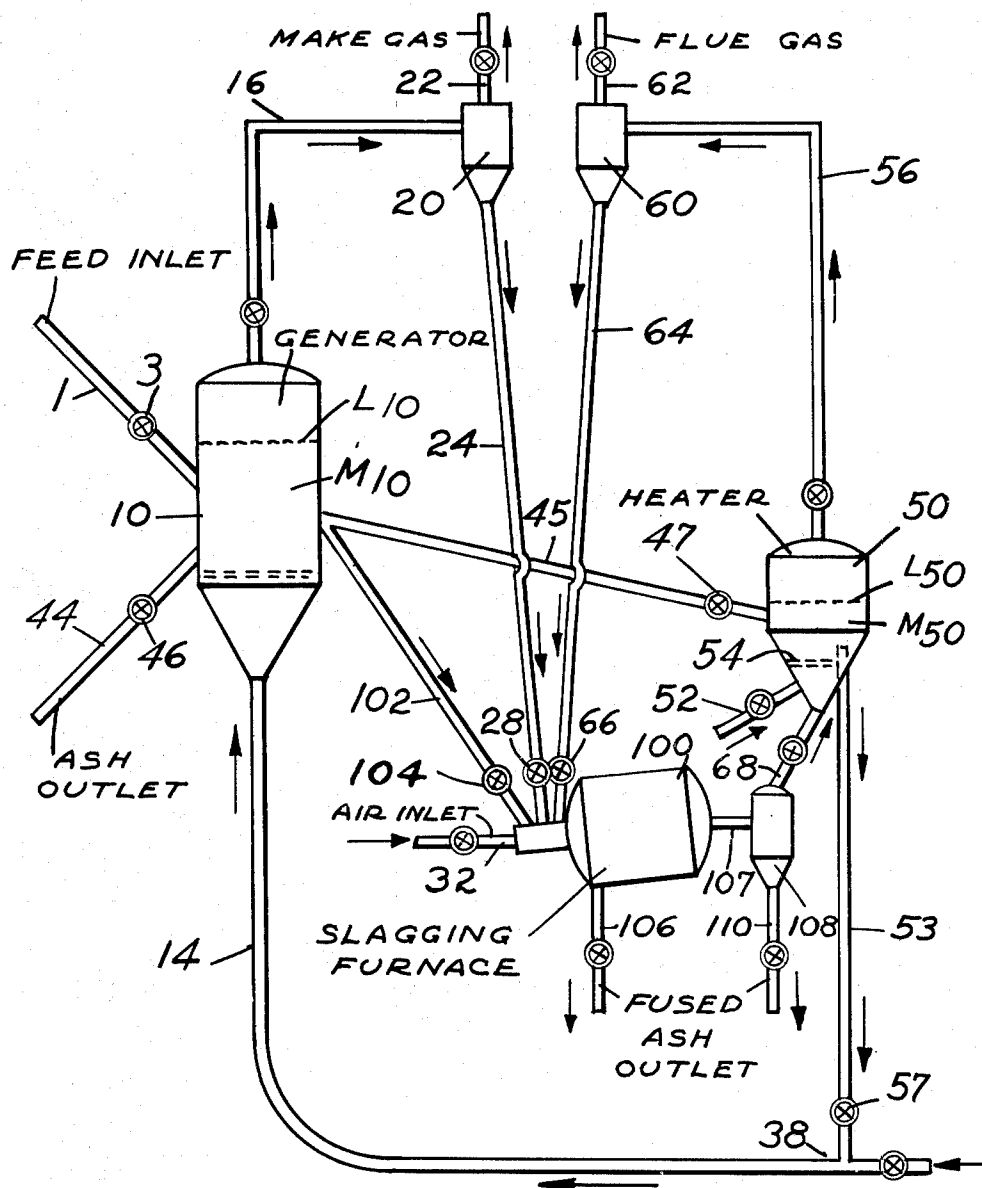


Fig.-8

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PROCESS OF CONTACTING GASIFORM
CARBONACEOUS SOLIDS

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Application December 24, 1949, Serial No. 134,976

9 Claims. (Cl. 48—197)

The present invention relates broadly to the handling of fluidized solids and finds specific application in the conversion of carbonaceous solids, such as all types of coal, lignite, peat, oil shale, tar sands, cokes, oil coke, cellulosic materials, including lignin, etc., into gases containing carbon monoxide, such as producer gas, water gas, synthesis gas, and the like. More particularly, the invention pertains to a novel method for controlling solids entrainment from fluidized solids beds. Specific embodiments of the invention involve a gasification process wherein the carbonaceous solids undergoing gasification are maintained in the form of a relatively dense turbulent mass of subdivided solids fluidized by an upwardly flowing gas to resemble a boiling liquid and wherein fines entrainment from the fluidized bed and particle fusion by overheating within the fluidized bed are minimized by withdrawing entrained fines and, if desired, dense phase solids from the gas generation zone and subjecting the same to combustion in a separate burner supplying at least a portion of the heat required for the gasification reaction, or to a secondary gasification reaction so as to control the concentration and/or the carbon concentration of the fines in the fluidized bed.

Prior to the present invention, it has been suggested to gasify carbonaceous solids with a gasifying medium, such as steam and/or oxygen 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 bed 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.

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This problem is seriously aggravated by the excessive fines entrainment in conventional fluid gasification systems. Experimental work on a pilot plant scale has demonstrated that entrainment from fluidized char gasification beds may run as high as 1.0 lb. of solids or higher per cu. ft. of product gas at gas velocities of about 0.2–2 ft. per second, in spite of the provision of as much as or more than 6 ft. of dilute phase settling space above the fluidized bed. In order to prevent the loss of these carbonaceous solids from the fluid bed, intricate and expensive gas-solids separation systems must be used from which separated solids may be returned to the bed. However, the solution of the entrainment problem by this means is complicated by the fact that the high temperatures and the corrosive atmosphere of the gasification zone make it desirable to locate the gas-solids separators, at least in part, outside the gasification zone, which results in substantial heat losses from the separated solids. When such vast amounts of separated entrainment solids are returned to the gasification zone at a reduced temperature, it becomes difficult, if not impossible, to maintain optimum gasification temperatures. In addition, more oxygen or air is needed to make up for these heat losses.

A further operating problem arises from the fact that partial fusion of fluidized solids takes place when concentrated oxygen is injected into the fluidized bed, particularly at relatively high bed temperatures and oxygen feed rates. This fusion results in an agglomeration of particles which may reach a point at which fluidization of the bed becomes difficult, if not impossible. In addition, such agglomerates will accumulate in generator sections adjacent to the oxygen feed point and will eventually plug the oxygen feed line and surrounding portions of the vessel.

The present invention overcomes the above mentioned difficulties and affords various additional advantages as will be apparent from the detailed description below wherein reference will be made to the accompanying drawing in which:

Figures 1–3 are graphical illustrations of entrainment conditions and entrainment composition encountered in various fluid-solids operations; and

Figures 4–8 are schematic illustrations of gasification systems adapted to carry out various specific embodiments of the invention.

In normal fluid operation, such as in fluid catalytic cracking, entrainment from the fluid bed is governed mainly by the velocity of the fluidizing gas and by the particle size of the fluidized solids. Thus it is well known that entrainment from the fluid bed increases at a rate which is a power function of the increase in gas velocity and that generally entrainment increases as the average particle size of the fluid bed is decreased at otherwise equal conditions. It is also known that quite generally a certain proportion of fines of, say, up to about 15–25% is desirable in fluidized solids beds to assure smooth fluidization and to avoid slug formation. If fines concentration is excessive, the fluidization characteristics of fluid beds of this type of solids are detrimentally affected. From the experience with cracking catalysts, it was concluded that in fluid operation quite generally no beneficial results may be derived from any increase in the fines concentrations beyond the level required for smooth fluidization. On the contrary, it was assumed that such further increase in the fines concentration would lead both to fluidization troubles and excessive fines entrainment from the fluidized bed.

It has now been found that fines entrainment from fluidized solids beds composed of particles of substantially all sizes within the fluidizable range of, say 1–1000

microns or larger, increases as the bed concentration of fines of —40 microns size increases, up to a certain maximum beyond which a further increase in the fines concentration of the fluid bed results in a corresponding decrease in the solids entrainment rate. These findings are illustrated in the diagram of Figure 1 in which entrainment rate as pounds of solids per cu. ft. of gas (ordinate) is plotted against fines concentration of the fluid bed in wt. percent of —40 microns particles. Curves I and II were obtained by varying the fines concentration (0–40 microns) between 0 and 70 wt. percent in fluidized beds of char withdrawn from a fluid-type gas generator, having otherwise the following ranges of particle sizes:

Microns :	Weight percent
40–60	0–57
60–80	0–24
80–150	0–62
150–1000	0–95

at linear superficial gas velocities of 0.5 and 1.25 ft. per second, respectively. Both curves show a well defined maximum of the entrainment rate, which lies at a fines concentration of about 30 wt. percent. On either side of this maximum the entrainment rate drops off sharply.

The experimental data summarized in Figure 1 demonstrate that, contrary to prior belief and expectation, solids entrainment from a fluidized solids bed may be reduced by maintaining in the fluidized bed a fines concentration substantially in excess of that required for smooth fluidization. The data also explain why this phenomenon could not be observed in conventional operations involving progressive solids disintegration and fines formation in the course of the process, as it is the case, for example, in fluid-type shale distillation or the fluid-type iron-catalyzed synthesis of hydrocarbons from H_2 and CO. These processes normally have been started at a fines concentration far to the left of the maximum of the entrainment curve of the fluid bed involved. As fines formation proceeded as a result of solids disintegration, the entrainment rate increased as a function of, but more rapidly than, the increase in fines concentration so that the maximum of the entrainment curve was approached from the left side, but never exceeded, and prohibitive entrainment rates were attained.

Based on the above considerations and the data reported, the present invention proposes to control the solids entrainment rate from fluidized solids masses by maintaining in the fluidized masses initially a concentration of particles of —40 microns size substantially different from that which will cause maximum solids entrainment from the fluidized bed at the fluidization and reaction conditions involved. Fines concentrations suitable for the purposes of the invention may lie substantially below or substantially above said maximum value and may vary within wide range depending mainly on the character of the solids involved, the particles size distribution of the fluid bed and the linear superficial velocity of the fluidizing gas. However, the determination of the fines concentration causing maximum solids entrainment and with it the determination of the fines concentrations suitable for a substantial reduction of the solids entrainment is a matter of a few preliminary routine tests which may be carried out by the expert without any difficulty in the manner indicated with reference to the curves of Figure 1. Since a certain proportion of fines is required for smooth fluidization and this proportion may frequently come close to that causing maximum entrainment, it is in many cases preferred to operate at fines concentrations substantially above the level causing maximum entrainment. For example, in the case of coal gasification this concentration should be in excess of 30 wt. percent and preferably within the range of 40–60 wt. percent.

Coal gasification experiments carried out on a pilot plant scale have further shown that an additional effect is superimposed on the phenomenon described above, under actual gasification conditions. In this case, entrainment appears to be strongly affected by the rate at which entrained solids are returned to the fluid bed. The graph of Figure 2 illustrates this phenomenon by a curve correlating solids entrainment from the fluid bed as pounds per cu. ft. of product gas to solids losses through the stack expressed as wt. percent of the solids entrainment from the fluid bed, on a log-logarithmic scale. Stack loss in this case is equal to 100 minus the per cent recovery efficiency of the cyclone separation system arranged outside the generator. The conditions at which this curve was determined were as follows:

Temperature, °F.	1600–1900
Pressure, p. s. i. g.	0–4
Gas feed composition, mol percent:	
Steam	0–71
O ₂	14–39
N ₂ +CO ₂	6–81
Gas velocities, ft./sec.:	
Fluid bed	1–2
Dilute phase	0.25–0.50
Initial particle size distribution, wt. percent:	
0–40 microns	0–50
40–100 microns	10–40
100–500 microns	15–75
+500 microns	0–25

The curve of Figure 2 shows that as stack loss decreases, i. e. as cyclone recovery efficiency increases, solids entrainment from the fluid bed increases very rapidly. The influence of stack loss on entrainment overshadowed the normal strong effects of velocity and particle size to the extent that a single line on the correlation could be drawn even though fluid beds varying widely in particle size and velocities were employed. Furthermore, entrainment rates in many instances were many-fold those which would be predicted from the velocity-particle size relationships for the condition employed.

In those pilot plant operations where the initial particle size distribution of the fluid bed indicates that the fines concentration was less than 30 wt. percent, the concentration at which fines entrainment reaches a maximum according to the curves of Figure 1, the phenomenon of increasing entrainment with increasing cyclone efficiency and fines return may be partly explained. However, it is indicated that the curve of Figure 2 substantially holds, i. e. that solid entrainment increases with increasing cyclone efficiency, even in the cases where the fluid gas generator was operating initially to the right of the maximum of the curves of Figure 1 with respect to fines concentration. This phenomenon, therefore, introduces a new complication into the problem of controlling fines entrainment at least in the case of coal gasification.

In the course of the experiments described with reference to Figure 2, the ash content of the entrained solids was compared with that of the bed from which the solids were entrained at different ash concentrations in the bed. The results of this investigation are summarized in the diagram of Figure 3 wherein wt. per cent of ash in the entrained solids is plotted against wt. per cent of ash in the fluid bed. It is evident from Figure 3 that for ash contents in the bed of 20–60 wt. per cent, the ash in the entrained solids remained substantially constant at about 15 wt. per cent. These fluid beds contained appreciable quantities of particles of the same diameter as comprised the entrainment. When comparing ash contents (carbon essentially equals 100% ash) on the basis of the same particle sizes, it was again found that the ash in the particle entrainment remained substantially constant at about 15%, for ash contents in the

bed of 20-60 wt. per cent. In other words, the entrained solids have a carbon content higher than the average carbon concentration of the bed and are similar in carbon content to coke fines entering the generator with the fresh feed.

Figures 2 and 3 demonstrate that if it is attempted to retain fines in the system by the conventional means of increasing cyclone efficiency and thus the return of entrained solids to the fluid bed, the rate of entrainment will be excessive while attempts at reducing the entrainment rate will result in increased stack losses of entrained solids, i. e. in the selective loss of solids of very high carbon concentration and thus in poor overall carbon utilization in the process. The present invention overcomes this difficulty.

In accordance with one embodiment of the present invention, the preferential entrainment of high carbon level solids is minimized by reducing the carbon level of those solids in the bed, the particle size of which is in the same range as that of the entrained solids. The efficiency of this procedure is demonstrated by the fact that fines retained in the bed which have the same particle size as the fines entrained from the bed have carbon concentrations approaching that of the bed, i. e. substantially lower than the carbon concentration of the entrained fines. In carrying out this embodiment of the invention, the solids entrained from the fluid bed are separated from the product gas in conventional gas-solids separation equipment, subjected to combustion in a separate combustion zone at conditions conducive to a substantial removal of their carbon content, and the low carbon ash particles so obtained are returned to the fluid bed. Such solids fines of reduced carbon content will not be entrained again in accordance with the relationship of Figure 2 at the prevailing fluidization conditions but entrainment from the fluid bed becomes again substantially a function of bed velocity and particle size. In this manner, the fines concentration may be maintained at desirably high levels, thus maintaining the entrainment rate at the right side of the maximum of the curves of Figure 1, if desired, and in all cases the fines concentrations required for proper fluidization may be retained.

The separate combustion zone is preferably a suspended-phase burner which may have the form of a transfer line burner, cyclone furnace or the like. The solids fines to be burned may be suspended therein in a mixture of oxygen and steam in such a manner that the temperature of the suspension is raised in the combustion zone to a level about 10°-50° F. below the ash fusion point. The hot suspension is preferably discharged as a whole into the fluid bed to supply at least a portion of the heat, oxygen and steam required for gasification. Additional oxygen and steam may be supplied directly to the fluid bed.

This embodiment of the invention may also be applied to gasification systems of the two-vessel type in which a separate fluid heater-burner is used to generate the heat required for gasification by combustion of fluidized generator residue and no oxygen is supplied to the generator. When so operating, cyclone fines from both the heater and generator are burned with at least a portion of the total air requirement in a separate high temperature burner to remove carbon from the fines and all the ash fines so produced are returned to the fluid bed in the heater-burner. The high solids circulation rates between heater-burner and generator in systems of this type tends to equalize the carbon concentration on the fines in both vessels.

In accordance with another modification of this embodiment of the invention, the total entrainment from a fluid generator bed is passed to a secondary fluid-type generator to form therein a fluid bed consisting exclusively of solids entrained from the primary generator. Gasification is carried out in both generators. The secondary fluid bed, as a result of its extremely high per-

centage of -40 microns particles, has a very low entrainment rate. Its particle size distribution may be about as follows:

Microns:	Wt. per cent
0-40	25-95
40-100	5-75
100-150	0-25
+150	0-10

Char masses of this type may be readily and evenly fluidized at superficial linear gas velocities of about 0.1-2.0 ft. per second to form fluid phases having densities of about 1 to 25 lbs. per cu. ft. In addition, at least a substantial portion of the carbon content of the primary entrainment is removed in the form of product gas, whereby not only carbon losses are minimized but a further reduction in the entrainment rate is also accomplished. Solids entrained from this secondary bed and separated in suitable gas-solids separation means may be either directly returned to the secondary bed or first further decarbonized by a high temperature combustion and then returned to the secondary and/or the primary fluidized bed for an additional control of the entrainment rate. Similarly, relatively low carbon fines from the secondary fluid bed may be directly passed from this fluid bed to the primary fluid bed for fluidity and entrainment control therein. The heat required may also be generated in a system of the two-vessel type as indicated before.

In accordance with another embodiment of the invention, entrainment rate is reduced by discharging at least a substantial portion of the entrained solids from the system without returning the same in any form to the fluid bed from which they were entrained. When so operating, the fines concentration of the fluid bed may be maintained on the left side of the entrainment maximum of the curves of Figure 1. However, stack losses will be relatively high and provisions must be made for utilizing the carbon content of the entrained solids prior to their rejection from the system.

One modification of this embodiment of the invention involves the combustion of the solids entrained from the fluid generator bed in a separate slagging-type combustion zone from which the ash may be removed in the liquid state while the hot combustion gases may be passed to the generator bed to supply at least a portion and preferably all the heat required for gasification. When so operating, the entrainment rate is reduced for the reasons outlined above and, simultaneously, fusion within the fluid generator bed may be avoided by supplying all the oxygen required for heat generation to the separate slagging-type combustion zone in such a manner that the oxygen is substantially completely consumed therein and practically no free oxygen will enter the fluid generator bed. If indicated by the heat requirements of the system, coarse solids from the fluid generator bed may likewise be passed to the slagging-type combustion zone. Operating conditions of slagging-type burners of this kind, when used in combination with a fluid gasification system, are described in detail in the copending Nelson application, Serial No. 717,064, filed December 18, 1946, now Patent No. 2,554,263, issued May 22, 1951, and assigned to the assignee of the present application. The disclosures of this copending application are here expressly referred to. Carbon remaining unconverted in the generator may be utilized for steam generation or for other suitable purposes.

The modification last described may be likewise readily adapted to systems of the two-vessel type in a manner generally analogous to that described with reference to the return of decarbonized entrainment solids to the fluid bed from which they originated.

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 Figures 4-8 of the drawing which illustrate various

systems adapted to carry out the embodiments of the invention described above.

Referring now to Figure 4, the system shown therein essentially comprises a water gas generator 10, gas-solids separation means 20 and a high temperature combustion zone or burner 30, the functions and coaction of which will be forthwith explained using as an example the production of water gas from low temperature coke. It should be understood, however, that the system may be used for the manufacture of producer gas and for the gasification of other carbonaceous solids in a generally analogous manner.

In operation, a finely divided low temperature coke is supplied through line 1 to gas generator 10 at a rate controlled by a valve or other metering device 3. 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 feed may fall within the wide ranges of 1/2 in. to 400 mesh or smaller in diameter, preferred size ranges being about as follows:

Microns:	Wt. per cent
0-40 -----	40-50
40-100 -----	25-35
100-150 -----	10-15
150-500 -----	10-15
+500 -----	0-5

The finely divided coke forms in generator 10, above distributing grid 5, a dense turbulent mass M₁₀ of solids fluidized by the gaseous reaction products and the gas and vapors supplied through lines 12 and 14 and grid 5, as will appear more clearly hereinafter. Mass M₁₀ forms a more or less well defined upper level L₁₀ and it may have an apparent density of about 5-10 lbs. per cu. ft. Linear superficial gas velocities within mass M₁₀ of about 1.0 ft. per second are generally suitable for this purpose.

Gasifying media, such as steam and oxygen, and heat in the form of sensible heat of hot combustion gases and entrained solids from burner 30, are supplied through lines 12 and 14, sufficient in amounts to maintain bed M₁₀ at the desired gasification temperature of about 1800°-1900° F., a carbon concentration of about 30-40 wt. per cent, and an operating pressure of about 400-500 p. s. i. g. At gas and solids temperatures in line 12 of about 2200°-2400° F., the amount of steam required to produce water gas is about 0.4-0.8 lb. per lb. of coke charged.

Product gas containing, at the conditions specified, about 0.05-0.15 lb. of entrained solids per cu. ft. is withdrawn overhead from level L₁₀ and passed through line 16 to separator 20 which is preferably arranged outside generator 10 and which may be a single or multiple cyclone system having an efficiency of about 85.0-99.99%. Make gas containing very small amounts of solids fines, mostly of -10 microns size, may be recovered through line 22 and further processed in any conventional manner (not shown).

The solids separated in cyclone 20 contain about 70-90 wt. per cent of carbon and may have a particle size distribution about as follows:

Microns:	Wt. per cent
0-40 -----	90
40-100 -----	8
100-150 -----	2

These solids flow down a standpipe 24 aerated through one or more gas taps 26 with small amounts of an aerating gas, such as steam, air, oxygen, flue gas, etc. The rate of solids flow through standpipe 24 may be controlled by slide valve or other metering device 28. The solids passing valve 28 enter suspended phase-type burner 30 wherein they are suspended in a mixture of steam and oxygen supplied from line 32 which receives oxygen from feed line 34 via line 36 and steam from feed line

38 via line 40. The absolute and relative amounts of steam and oxygen supplied to burner 30 and the solids feed rate through valve 28 should be so controlled and correlated that by virtue of the combustion the temperature of the suspension is increased to a level which is at least 10°-50° F. below the fusion point of the ash, which lies between about 2500° and 3000° F. for most conventional carbonaceous solids. About 0.4-0.7 lb. of steam and about 0.3-0.5 lb. of O₂ per lb. of fresh char charged are usually suitable for this purpose. At the conditions specified, the carbon concentration of the solids burned in burner 30 may be reduced to about 20-50 wt. per cent.

The suspension of burned solids-in-gases leaves burner 30 through line 42 and passes through line 12 and grid 5 into mass M₁₀ to supply heat thereto and adjust the carbon content and overall concentration of the -40 microns fines which determine the solids entrainment rate from mass M₁₀. At the conditions specified, the average carbon concentration of the -40 microns particles in bed M₁₀ may be about 20-60 wt. percent, that is substantially below that of the fines entrained if burner 30 were not used. With burner 30, however, total entrainment and carbon losses from the system are maintained at a minimum. The remainder of the oxygen and steam requirement of generator 10 may be supplied from lines 34 and 38, respectively, to line 14. At the conditions of the present example, about 0.3-0.5 lb. of additional oxygen and about .1-1.0 lb. of additional steam per lb. of char charged may be supplied through line 14. Solid gasification residue of average particle size and a carbon concentration of about 30-40 wt. percent may be withdrawn from generator 10 via standpipe or similar withdrawal means 44, provided with slide valve 46. The char so recovered may be fed to the boiler house or used for other suitable purposes.

Whenever excessive solids fusion occurs in generator 10, this may be overcome by passing solid generator residue from line 44 via line 50 to line 14 in amounts sufficient to consume by combustion at least a substantial proportion of the oxygen present in line 14. In this manner, most or all of the combustion may be shifted from generator 10 to burner 30 and line 14 wherein the oxygen may be more quickly and uniformly contacted with larger amounts of solids and the heat generated by the combustion is so generated in direct contact with larger amounts of solids whereby localized overheating beyond the ash fusion point is substantially eliminated without loss in thermal efficiency.

It will be noted that in the system of Figure 4 the product gas is diluted with flue gases from the combustion taking place in burner 30 and generator 10. This may be avoided by employing a two-vessel type system as it is illustrated in Figure 5.

Referring now to Figure 5, the system shown is equipped with a heater 50 and a cyclone system 60 in addition to the essential elements of Figure 4. Like reference characters have been used to identify elements common to Figures 4 and 5.

In operation, generator 10 may be charged with coal or coke in a manner similar to that outlined with reference to Figure 4. Gasification and fluidization of the solids in generator 10 is accomplished by the supply of steam, preferably preheated to about 500°-1800° F., from line 38 through line 14 and grid 5, excess solids being withdrawn through standpipe 44. About 0.3-3.0 lbs. of steam per lb. of solids charged may be used. Heat is supplied to generator 10 by circulating about 10-100 lbs. of solid gasification residue per lb. of fresh solids charge from bed M₁₀ through standpipe or other conveying means 45 at a rate controlled by valve 47 to a separate fluid-type heater 50 wherein the char is heated by combustion to about 50°-500° F. above gasification temperature, as will appear more clearly hereinafter and from which the heated char is returned via standpipe

53 or the like provided with slide valve 57 to steam line 38 and to generator 10, in a manner generally known for two-vessel type gasification systems. Combustion in heater 50 may be maintained by air supplied through line 52 and grid 54 at a rate of about 1-3 lbs. of air per lb. of fresh carbonaceous charge to generator 10. The solids in heater 50 are fluidized to form a dense turbulent mass M_{50} having an upper interface L_{50} and an apparent density of about 5-30 lbs. per cu. ft.

Product gas which may contain anywhere between about .001 and 1.0 lb. per cu. ft. of entrained solids depending on operating conditions is passed from the top of generator 10 through cyclone 20, make gas being recovered through line 22 and separated solids being passed to suspended phase burner 30 via line 24 substantially as described with reference to Figure 4. However, burner 30 is now supplied through line 32 with air rather than oxygen and steam. Simultaneously, burner 30 receives entrainment solids carried overhead from heater 50 at a rate of about .001 to 1.0 lb. per cu. ft. of flue gas, depending on conditions, separated in cyclone system 60 and passed through standpipe 64 provided with valve 66 to burner 30. Flue gas may be vented through line 62, preferably after suitable heat recovery in any conventional manner. The heater entrainment will have a lower carbon content than the generator entrainment. The air supply to burner 30 is so controlled that the solids-in-gas suspension flowing therethrough is heated by combustion to about 10°-50° F. below the ash fusion temperature, which is substantially higher than the temperature to be maintained in heater 50. About .02 to 2.0 lbs. of air per lb. of fresh solid generator feed are normally sufficient for this purpose. The suspended low carbon solids in flue gas leaving burner 30 enters heater 50 through line 68 and grid 54 substantially at the temperature of burner 30 to supply heat to heater 50 and control the entrainment rate therein substantially as outlined with reference to Figure 4. Since decarbonized fines are returned from mass M_{50} to mass M_{10} at a high rate, the entrainment rate from generator 10 is similarly controlled.

A system suitable for controlling the entrainment rate by fines removal and gasification of the entrainment solids in a secondary gasification zone is illustrated in Figure 6. This system comprises a secondary fluid-type gasification zone 70 in addition to the essential elements of Figure 4, like reference characters identifying like system elements.

Referring now in detail to Figure 6, generator 10 may be operated essentially as outlined with reference to Figure 4 except for the handling of the entrained solids separated in separator 20 and withdrawn through line 24. These solids may have a particle size distribution about as follows:

Microns:	Wt. percent
0-40 -----	50-90
40-100 -----	8-40
100-150 -----	2-10

They are passed into secondary generator 70 which receives oxygen and steam for gasification from lines 34 and 38, respectively, via lines 72 and 74 and grid 76. About .04-.02 lb. of suitably preheated oxygen and about .05-.03 lb. of preheated steam per lb. of fresh charge through line 1 are normally adequate for gasification at a temperature of about 1700°-1900° F. Generator 70 is so designed that a linear superficial gas velocity of about 0.5-1.5 ft. per second may be maintained at these conditions so that the solids in generator 10 form a relatively dense turbulent mass M_{70} having an upper interface L_{70} and an apparent density of about 2-20 lbs. per cu. ft. As a result of the small average particle size and the reduction in the carbon content due to the secondary gasification of the solids in generator 70, the entrainment rate from generator 70 is at a relatively low

level of about .001-0.20 lb. per cu. ft. of secondary make gas. Secondary make gas of this solids concentration is withdrawn through line 78 into cyclone system 80. Make gas of low solids content may be withdrawn through line 82 to be combined with the primary make gas in line 22. Separated solids may be returned to mass M_{70} via dip pipe 84 and excess fines may be withdrawn from mass M_{70} through lines 86 and 88.

While the system of Figure 6 may be operated satisfactorily in the manner just described, further refinements in the control of the entrainment rates and in carbon utilization may be employed as follows. The solids separated in cyclone 80 may still have a relatively high carbon content. In order further to reduce this carbon content and thus the entrainment rate in generator 70, a portion or all of these solids may be passed through standpipe or the like 90 to a suspended phase burner 30 of the type described with reference to Figures 4 and 5. Oxygen and steam may be supplied to burner 30 in a manner previously described to heat the suspension therein to about 10°-50° F. below the ash fusion point and reduce the carbon content of these solids. The hot suspension may then be returned to mass M_{70} through line 92 and/or to mass M_{10} in generator 10 through line 42 to assist in particle size and entrainment control in these vessels as previously described. Additional improvements in the control of particle size and entrainment rate in generator 10 may be secured by circulating solids from mass M_{70} directly to mass M_{10} via line 86 at a rate adequate to maintain the concentration of low carbon fines of -40 microns size within mass M_{10} at a desirable level of about 5-15 wt. per cent.

A system wherein entrainment rate may be controlled by a complete elimination of the entrained solids without excessive carbon losses and which lends itself most readily to the prevention of particle fusion in the generator is illustrated in Figure 7. The general arrangement of the system shown therein is similar to that of Figure 4, like elements being identified by like reference characters. There are, however, significant operational differences, the most important of which reside in the use of a slagging-type furnace 100 in place of suspended-phase burner 30 and in the practical elimination of the feed of free oxygen to generator 10.

Referring now in detail to Figure 7, generator 10 may be charged with carbonaceous solids in the manner outlined with reference to Figure 4. Gasification and fluidization of the solids charge in generator 10 are accomplished with the aid of a hot mixture of steam and flue gas supplied from burner 100 via line 12 and grid 5, as will appear hereinafter. Gasification and fluidization conditions may be similar to those specified for generator 10 of Figure 4, although mass M_{10} contains a substantially smaller proportion of -40 microns size which may be readily maintained substantially below the entrainment maximum of the curves of Figure 1, by the elimination of entrained solids in the manner described below.

Make gas containing about .05-.15 lb. per cu. ft. of entrained solids is passed through line 16 to cyclone 20 to be treated therein as described with reference to Figure 4. The separated solids containing about 70-90 wt. per cent of carbon are passed via standpipe 24 to slagging-furnace 100. Oxygen and steam are supplied to furnace 100 from lines 36 and 40, respectively, in relative and absolute amounts adequate to maintain the temperature in furnace 100 above the ash fusion point, to burn all the carbon present in furnace 100 and to generate all the heat required for the gasification reaction in generator 10. Since the carbon supplied via line 24 is normally insufficient to generate all the heat required by the system and to consume all the oxygen supplied to burner 100, coarse generator solids containing about 20-40 wt. per cent of carbon are passed directly from mass M_{10} via standpipe or the like 102 to furnace 100

at a rate controlled by valve 104. About 1-2 lbs. of coarse solids passing through line 102 per lb. of fresh solid generator feed are normally sufficient for this purpose. The oxygen and steam requirements of burner 100 are about 0.4-0.6 and about 0.1-0.3 lb., respectively, per lb. of fresh solid generator charge, at the conditions specified which include burner temperatures of about 2700°-3300° F.

Burner 100 may be designed and operated substantially as described in the said copending Nelson application, Serial No. 717,064 with reference to burners 40 and 340 of Figures 1 and 3 of said copending application which is here referred to for all details not specifically described herein. However, other conventional slagging-type furnaces may be used for this purpose.

Most of the fused ash is withdrawn in the liquid state directly from burner furnace 100 via tap 106. The hot mixture of unconverted steam and gases which is now substantially free of oxygen and entrained solids is withdrawn through line 42, passed to an entrainment separator 108 for the removal of entrained ash droplets and then supplied through line 12 to generator 10 substantially at the temperature of burner 100. Fused ash collecting in separator 108 may be discarded via tap 110. Excess coarse solids may be withdrawn from mass M₁₀ via line 44 to be sent to the boiler house or to any other suitable use. At the conditions specified, mass M₁₀ may have an average carbon concentration of about 20-40 wt. per cent and a particle size distribution about as follows:

Microns:	Weight per cent
0-40	5-15
40-100	25-35
100-150	10-15
150-500	30-50
+500	5-15

It may be readily fluidized at superficial linear gas velocities of about 1.0-2.0 ft. per second to assume an apparent density of about 15-25 lbs. per cu. ft.

An adaptation of the principles of the modification illustrated in Figure 7 to a two-vessel type gasification system wherein all the heat required in the generator is supplied in the form of sensible heat of hot solids highly heated in a separate heater is illustrated in Figure 8. The system shown in Figure 8 is generally similar to that of Figure 5 modified by the use of the slagging-type burner 100 of Figure 7, in place of burner 30. System elements common to Figures 5, 7 and 8 are identified by like reference characters.

Referring now in detail to Figure 8, generator 10 may be operated substantially as described with reference to Figure 5. The operation of slagging-type burner 100 is similar to that described with reference to Figure 7 with the exception that burner 100 receives in addition to entrainment solids and coarse solids from generator 10 via lines 24 and 102, respectively, also entrainment solids from heater 50 via lines 56, separator 60 and line 64 and that air rather than oxygen and steam, is fed through line 32 to burner 100. The solids flowing through line 24 will have a carbon concentration higher than those flowing through line 64, the latter approaching the carbon concentration in beds M₁₀ and M₅₀ which differ only slightly as a result of the high circulation rates between these two beds. The air requirement of burner 100 for the purpose of slagging the ash and complete carbon and oxygen consumption is about 0.1-3.0 lbs. of air per lb. of fresh solid generator feed.

The hot flue gases passing through line 107, entrainment separator 108 and line 68 to heater 50 substantially free of solids and liquid ash at a temperature of about 2000°-3300° F., may supply to heater 50 at least a substantial portion of the heat required by the gasification reaction, which heat is then transferred to generator 10 in the form of sensible heat of heater solids passed through lines 53, 38 and 14. Solids balance is main-

tained by circulating solids from mass M₁₀ to heater 50 via line 45 substantially as described with reference to Figure 5. This solids circulation rate may be about 10-200 lbs. per lb. of fresh solid generator feed. If desired, additional air may be supplied to heater 50 via line 52 to maintain the heater temperature about 50°-500° F. above the generator temperature.

Conversion and fluidization conditions in generator 10 and heater 50 of Figure 8 may be about as follows:

	Generator 10	Heater 50
Temperature, °F.	1,500-2,500	1,550-2,800
Entrainment, lbs. per cu. ft.	.001-0.2	.001-0.2
Particle Size Distribution, Wt. Percent:		
0-40 microns	0-20	0-20
40-100 microns	5-25	5-25
100-150 microns	15-30	15-30
150-500 microns	25-50	25-50
+500 microns	5-40	5-40
Carbon Concentration in Dense Phases, Wt. Percent	10-60	10-60

The systems of Figures 4-8 may be employed in the carbonization of carbonizable solids, such as coal, oil shale, etc. and for numerous other purposes in a generally analogous manner. Various other modifications of the systems shown in the drawing may appear to those skilled in the art without deviating from the spirit of the invention. For example, dense phase solids circulation as shown in Figure 4 with reference to line 50 for the purpose of preventing ash fusion in the fluid bed, may be employed in the systems of Figures 5-8 in a generally analogous manner. If desired, a portion of the fresh carbonaceous solids feed may be directly supplied to burners 30 or 100. Also, the fresh solids feed may be classified to remove fines and the fresh fines so separated may be fed directly to burners 30 or 100. While steam has been mainly referred to as the gasifying medium, it will be understood that carbon dioxide may replace a portion or all of the steam, if a gas relatively rich in CO is desired.

The above description and exemplary operations have served to illustrate specific embodiments of the invention but are not intended to be limiting in scope.

What is claimed is:

1. In the process of gasifying subdivided carbonaceous solids in the form of a dense turbulent mass of solids containing a substantial proportion of particles substantially larger than 40 microns in diameter fluidized by upwardly flowing gases in a gasification zone, the improvement which comprises reducing the entrainment of solids from said mass in said gases by maintaining the concentration of particles of -40 microns size within said mass at a level adequate for smooth fluidization but substantially above that at which said entrainment is at a maximum.

2. The process of claim 1 in which said carbonaceous solids are low temperature coke and said last mentioned level is about 30 wt. percent.

3. The process of claim 1 in which said first mentioned level is maintained by continuously removing entrainable solids from said mass, contacting said removed solids with additional gases reacting with carbon to form carbon oxides, at conditions conducive to said carbon oxide formation so as to reduce the carbon content of said removed solids, and returning solids of reduced carbon content so obtained to said mass.

4. The process of claim 3 in which said reaction comprises an endothermic gasification of carbon.

5. The process of claim 3 in which said reaction comprises a combustion carried out in a separate combustion zone at a temperature below the fusion point of the ash of said solids.

6. The process of claim 5 in which heat generated by said combustion is transferred to said gasification zone as sensible heat of effluent from said combustion zone.

7. The process of claim 6 in which a portion of the

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oxygen required to supply the heat required in said gasification zone is contacted in a separate contacting zone with solids withdrawn from said first-named mass independent of said upwardly flowing gases.

8. In the process of contacting gasiform material with carbonaceous solids in the form of a dense turbulent fluidized mass of solids containing a substantial proportion of particles substantially larger than 40 microns in diameter fluidized by upwardly flowing gasiform material, the improvement which comprises reducing the entrainment of solids from said mass in said gases by maintaining the concentration of particles of minus 40 micron size within said mass at a level adequate for smooth fluidization but substantially above that at which said entrainment is at a maximum.

9. The process of claim 8 in which said level is maintained by continuously removing entrainable solids from said mass and withholding such removed solids from the gasiform material.

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