(19) Europäisches Patentamt
European Patent Office
Office européen des brevets

(11) Publication number: 0 044 123
A1

(12) EUROPEAN PATENT APPLICATION

(21) Application number: 81300493.4

(22) Date of filing: 05.02.81

(51) Int. Cl.³: C 10 J 3/54

(30) Priority: 16.07.80 US 170311

(43) Date of publication of application: 20.01.82 Bulletin 82/3

(84) Designated Contracting States: BE DE FR GB NL

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(54) Coal conversion process.

(57) This concerns a gasification process for solid carbonaceous materials in a fluidised bed in the presence of steam, added hydrogen, carbon monoxide and a catalyst. The reactor volume is reduced by elevating either the point or points at which the solid carbonaceous feed is introduced, the point or points wherein hydrogen and carbon monoxide are introduced or the point or points at which both the solid carbonaceous feed and the carbon monoxide and hydrogen are introduced. When the points of feed introduction 31'-31" and the points of carbon monoxide and hydrogen introduction 35'-35" are elevated, the reduction in volume is greater than would have been predicted from a relocation of either of these points separately.

EP 0 044 123 A1

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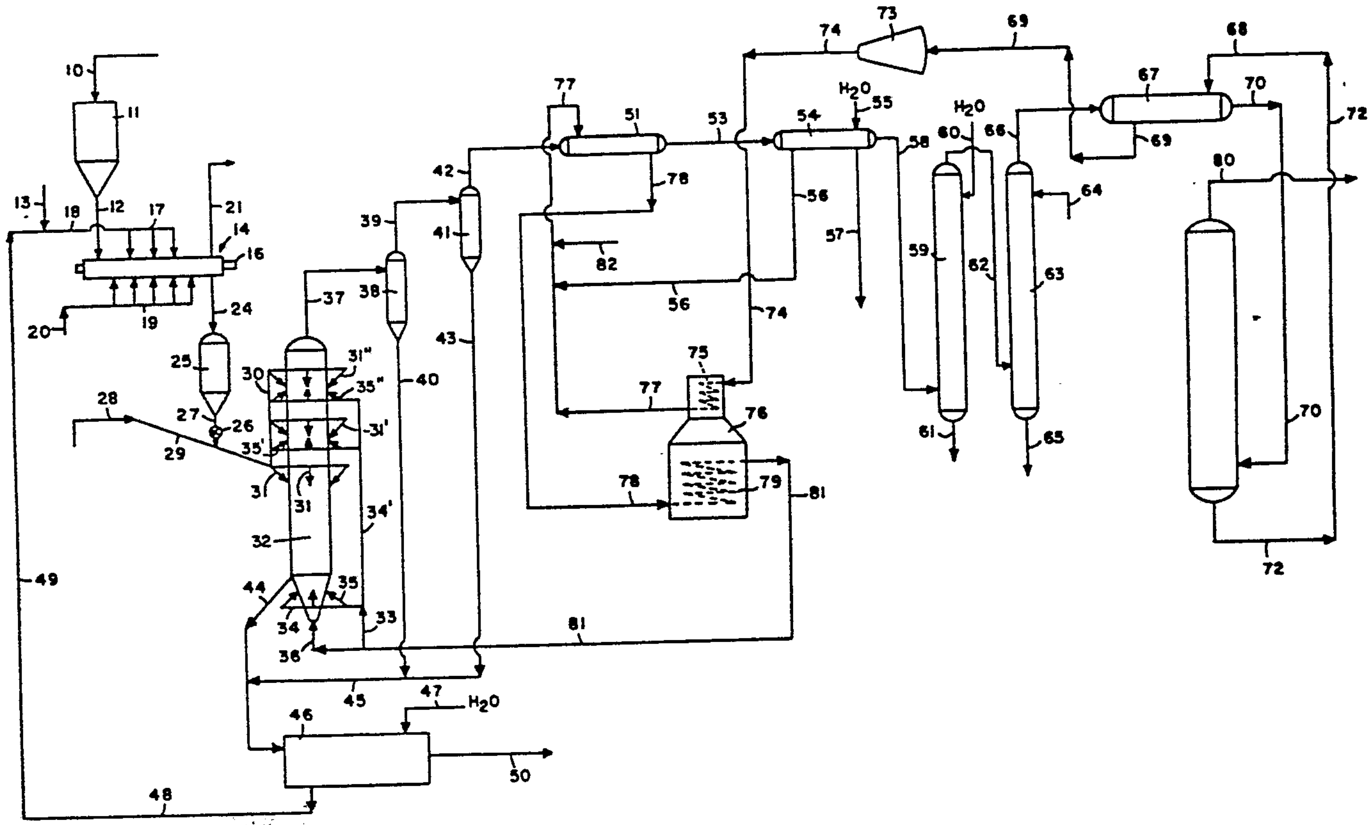

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"Coal Conversion Process"

This invention relates to an improved process for converting solid carbonaceous materials. More particularly, this invention relates to a process for gasifying solid carbonaceous materials.

Before the turn of the century it was known that hydrocarbon gases and liquids, tars and chemicals could be obtained not only from petroleum, but from coal and other carbonaceous liquids solids. Very early processes employed destructive distillation, coal being transformed into gases and petroleum-like liquid products. Primary emphasis in many of these processes is on gasification of the coal with the objective of improving processes for the production of water gas, producer gas, or hydrogen, as opposed to the production of coal liquids. For the past several decades, due to disallocations of supplies, there have been reoccurring periods of interest in the gasification of coal to produce fuel gases, first primarily in Europe; and then, in this country. The art reflects the various periods of interest in terms of peaks defined by large numbers of patents, and literature. Presently existing and projected shortages of natural gas in this country have sparked a renewed and very keen interest in the gasification of coal, and it appears that this will be a long-range trend. Consequently, intensive research and development efforts are now underway to produce synthetic high-BTU, intermediate-BTU and synthesis fuel gases for commercial usages.

It was early recognized that some mineral and trace inorganic constituents naturally present in some coal could exert favorable catalytic influences in gasification reactions vis-a-vis thermal reactions, and a variety of catalytic materials have been added to coal to alter the natural chemistry inherent in various of the early coal gasification processes. The thrust of present research is to develop processes for the production of synthetic high-BTU gases with far higher efficiencies than was possible in the classical European, or early Euro-American processes. There are, however, inherent chemical kinetic limitations in coal gasification processes which have defied solution, and these problems are yet unsolved. Yet, solution is essential, and there remains a strong interest in providing better coal gasification processes, or catalysts for use in catalytic coal gasification processes.

In a coal gasification process, i.e., one whose object is to produce a high BTU gas, an intermediate BTU gas or a synthesis fuel gas; steam or a similar reagent and particulate coal are fed to a gasifier at elevated temperature and pressure and converted to a synthesis gas, or gaseous mixture of high methane content, which contains significant amounts of carbon monoxide and hydrogen. Recently, it has been proposed to separate the methane from the carbon monoxide and hydrogen in a catalytic process and to then recycle the carbon monoxide and hydrogen to improve thermal efficiency where a high BTU gas is desired or recycle a portion of the entire stream when an intermediate BTU gas is desired. Generally, the methane in the recycle stream may be reformed to carbon monoxide and hydrogen prior to the recycling step. Moreover, the entire hydrocarbon gas may be reformed when synthesis gas is the desired product. Processes of this type are described in U. S. Patents Nos. 4,094,650 and 4,118,204. Practical objectives, however, require thermal efficiencies coupled with a reduction of reactor size.

More recently, it has been discovered that the recycling of hydrogen to the reactor retards the steam gasification of coal and similar liquid and solid carbonaceous materials thereby increasing the size of the gasification reactor. The need for an improved gasification process exhibiting the same thermal efficiently but requiring a smaller gasification reactor is, therefore, believed to be readily apparent.

It has now, surprisingly, been discovered that the foregoing disadvantages of the prior art gasification processes may be overcome with the method of the present invention and an improved gasification process provided thereby. It is, therefore, an object of this invention to provide an improved method for gasifying carbonaceous materials. It is another object of this invention to provide such an improved process wherein the thermal efficiency normally associated with the introduction of carbon monoxide and hydrogen directly to the gasification reactor is realized. It is still another object of this invention to provide such an improved process wherein a smaller gasification reactor is required. These and other objects and advantages will become apparent from the description set forth hereinafter.

In accordance with this invention, the foregoing and other objects and advantages are accomplished by gasifying a carbonaceous material in a fluid bed at elevated temperatures and pressures such that either the carbonaceous feed, added hydrogen and carbon monoxide or all three are introduced into the gasification reactor at a point generally above the bottom of the bed but sufficiently below the top thereof to permit substantial equilibration of the gas phase and to avoid tar breakthrough from the bed. Surprisingly, thermal efficiency is maintained by introducing added carbon monoxide and hydrogen into the gasification reactor and into the fluid bed at an elevated point or points. As pointed out more fully hereinafter, raising the point or points of introduction of the carbonaceous feed and the carbon monoxide-hydrogen mixture results in a synergistic reduction in gasification reactor size.

As indicated, supra, the present invention relates to an improved process for gasifying carbonaceous materials. Thermal efficiency is enhanced by introducing a mixture of carbon monoxide and hydrogen into a fluid bed comprising the carbonaceous material at varying degrees of gasification. The gasifier reactor size required is reduced by elevating the point or points at which either the carbon monoxide-hydrogen mixture or the carbonaceous feed is introduced to the fluid bed. Maximum reduction in the gasifier reactor size required is realized when both the carbon monoxide-hydrogen mixture and the carbonaceous feed are introduced into the fluid bed at a point or points above the bottom of the fluid bed and sufficiently below the top to permit substantial equilibration of the gas phase and to avoid tar break through from the bed.

In general, the process of this invention may be used to gasify any carbonaceous material that will fluidize in a gas stream. The process is, therefore, particularly suited to the gasification of solid carbonaceous materials such as coal, coal char, metallurgical coke, petroleum coke, charcoal, activated carbon and the like. In some cases, inert carriers having carbon deposited on the surfaces thereof may also be gasified in the process of this invention.

As indicated previously, the gasification of coal and similar carbonaceous materials normally produces a synthesis gas composed primarily of hydrogen and carbon monoxide. The principal reactions which take place in such a system include the following:

$$C + H_2O \longrightarrow CO + H_2 \text{ (Endothermic)} \quad (1)$$

$$C + 2H_2 \longrightarrow CH_4 \text{ (Exothermic)} \quad (2)$$

$$C + CO_2 \longrightarrow 2CO \text{ (Endothermic)} \quad (3)$$

$$CO + H_2O \rightleftarrows CO_2 + H_2 \text{ (Exothermic)} \quad (4)$$

The reaction kinetics during conventional gasification operations are such that the product gas normally contains varying amounts of methane. In steam gasification, the methane which is present occurs primarily as a result of devolatilization of the coal. The direct hydrogenation of carbon in accordance with equation (2) above is known to be very slow as compared to the endothermic reactions of steam and carbon dioxide with carbon as set forth in equations (1) and (3). The products of conventional steam gasification operations are thus primarily hydrogen and carbon monoxide and such operations are highly endothermic. As pointed out earlier, it has been proposed that this endothermicity be reduced by carrying out the operation in the presence of hydrogen to promote the exothermic carbon-hydrogen reaction of equation (2) but this normally requires a substantially higher reaction temperature than is needed for the steam-carbon reaction. Moreover, it is now known that hydrogen inhibits the reaction of steam with carbon to produce carbon monoxide and hydrogen. As a result, gasifier reactors of a larger size than would otherwise be required are required when hydrogen and/or carbon monoxide-hydrogen mixtures are introduced at or near the bottom of the gasifier.

In general, the gasifier reactor size is reduced when the feed point or points are raised above the bed since the inhibiting effect of the devolatilization gases is limited to that portion of the bed at and above the point of feed introduction. In this regard, it should be noted that the devolatilization will occur rapidly after feed introduction and the gases will flow upwardly. The solid particles, on the other hand, will be distributed throughout the entire fluid bed since such a bed approaches a perfectly mixed characterization. Ideally, the feed point or points will be located at the highest elevation possible without tar breakthrough from the bed.

Similarly, the gasifier reactor size is reduced when the point or points at which carbon monoxide and hydrogen are introduced is raised in the reactor. In this regard, it should be noted that the introduction of carbon monoxide and hydrogen into the gasifier will improve thermal efficiency in gasification processes of the type described in U. S. Patents No. 4,094,650 and No. 4,118,204.

In addition to the gas phase reaction illustrated by equation (4) above, the following gas phase reactions occur:

$$2CO + 2H_2 \longrightarrow CO_2 + CH_4 \qquad (5)$$

$$CO + 3H_2 \longrightarrow H_2O + CH_4 \qquad (6)$$

$$CO_2 + 4H_2 \longrightarrow 2H_2O + CH_4 \qquad (7)$$

All of these gas phase reactions are exothermic and, when an amount of carbon monoxide and hydrogen in excess of the equilibrium amount is introduced, heat is produced. Ideally, then, the point or points of introduction will be at the highest elevation which will permit the gas phase to substantially reach equilibrium at the top of the fluid bed. In this way, the heat generated will be distributed throughout the fluid bed by the well mixed solid particles comprising the bed.

In general, commercial scale gasification fluid beds will range in height from about 60 to about 125 feet and both the carbonaceous feed and carbon monoxide and hydrogen feed to the bed will, independently and generally, be introduced at a point or points along the bed and within a range of distances ranging from about 10% of the total bed height below the top of the bed to about 60% of the total bed height from the top. Such disposition will, of course, result in a maximum size reduction or a minimum bed height for any given gasification operation. It is, however, within the scope of this invention to position either the carbonaceous feed or the carbon monoxide and hydrogen introduction within the specified range with the other being introduced at or near the bottom of the bed in a manner consistent with the prior art.

It is believed that the present invention will be better understood by reference to the appended drawing. Referring to the drawing, then, the process illustrated is one for the production of a chemical synthesis gas by the gasification of carbonaceous material such as coal, lignite, coal char, coke or similar carbonaceous material with steam at an elevated temperature in the presence of a catalyst.

In the embodiment illustrated, a particularly preferred catalyst is prepared by impregnating the feed solids with a solution of an alkali metal compound or mixture of such compounds and thereafter heating the impregnated material to a temperature sufficient to produce an interaction between the alkali metal and the carbon present. Generally, the solid feed material will be finely divided to a particle size suitable for fluidization and a particle size of about 8 mesh or smaller on the U. S. Sieve Series Scale is particularly suitable. In the embodiment illustrated, the feed is passed into line 10 from a feed preparation plant or storage facility that is not shown in the drawing. The solids introduced into line 10 are fed into a hopper or similar vessel 11 from which they are passed through line 12 into feed preparation zone 14. This zone contains a screw conveyor or similar device, not shown in the drawing, that is powered by a motor 16, a series of spray nozzles or similar devices 17 for the spraying of an alkali metal-containing solution supplied through line 18 onto the solids as they are moved through the preparation zone by the conveyor, and a similar set of nozzles or the like 19 for the introduction of a hot dry gas, supplied through line 20, which serves to heat the impregnated solids and drive off the moisture. A mixture of water vapor and gas is withdrawn from zone 14 through line 21 and passed to a condenser, not shown, from which water may be recovered for use as makeup or the like. The majority of the alkali metal-containing solution is recycled through line 49 from the alkali metal recovery portion of the process, which is described hereafter. Any makeup alkali metal solution required may be introduced into line 18 via line 13.

In general, sufficient alkali metal-containing solution is introduced into preparation zone 14 to provide from about 1 to about 50 weight percent of an alkali metal compound or mixture of such compounds on the coal or other carbonaceous solids. From about 5 to about 30 percent is preferred. The dried impregnated solid particles prepared in zone 14 are withdrawn through line 24 and passed to a closed hopper or similar vessel 25 from which they are discharged through a star wheel feeder or equivalent device 26 in line 27 at an elevated pressure sufficient to permit their entrainment into a stream of high pressure steam, inert gas or other carrier gas introduced into line 29 via line 28. Moreover, it is within the scope of this invention to use all or a portion of the carbon monoxide and hydrogen as a carrier gas. The carrier gas and entrained solids are passed through line 29 into manifold 30 and introduced into the gasifier 32 through any one or more of a plurality of feed points 31, 31' and 31''. As previously indicated, the uppermost or highest feed point will be at least about 10% of the total fluid bed height below the top of the fluid bed and the lowest feed point will be no more than 60% of the total bed height below the top of the fluid bed. In lieu of or in addition to hopper 25 and star wheel feeder 26, the feed system may employ parallel lock hoppers, pressurized hoppers, aerated standpipes operated in series, or other apparatus to raise the input feed solids stream to the required pressure level.

In general, the gasifier 32 will be operated at a pressure between about 100 and 1500 psia, preferably at a pressure within the range of about 200 and 800 psia for any desired product distribution. The carrier gas may be preheated to a temperature in excess of about 300°F, but below the initial softening point of the coal or other feed material employed. Feed particles may be suspended in the carrier gas in a concentration between about 0.2 and about 5.0 pounds of solid feed material per pound of carrier gas. The optimum ratio for a particular system will depend in part upon the particle size and density, the molecular weight of the gas employed, the temperature of the solid feed material and the input gas stream, the amount of alkali metal compound employed and other factors. In general, ratios between about 0.5 and about 4.0 pounds of solid feed material per pound of carrier gas are preferred.

Gasifier 32 contains a fluidized bed of carbonaceous solids extending upward within the vessel above an internal grid or similar distribution device not shown in the drawing. The bed is maintained in the fluidized state by means of steam introduced through bottom inlet 36. The bed may be partially maintained with carbon monoxide and hydrogen introduced through line 33, manifold 34 and peripherally spaced injection lines and nozzles 35, when a part or all of carbon monoxide and hydrogen are introduced at the bottom of the bed. As previously indicated, however, hydrogen retards the desirable gasification reactions and is, therefore, preferably introduced into the fluid bed at a higher elevation. When this is done, the carbon monoxide and hydrogen may be introduced through any one or more of a plurality of injection points 35'-35'' which are supplied by manifold 33' and which are independently positioned in the same portion of the fluid bed as the carbonaceous feed points 31, 31' and 31''. Alternatively, all or a part of the carbon monoxide and hydrogen could be introduced into the fluid bed with the carbonaceous feed, as previously indicated, through line 28.

The particular injection system shown in the drawing is not critical and hence other methods for injecting the steam, hydrogen and carbon monoxide may be employed. In some instances, for example, it may be preferred to introduce the gases through multiple nozzles to obtain more uniform distribution of the injected fluid and reduce the possibility of channeling and related problems. The space velocity of the rising gases within the fluidized bed will normally be between about 100 and about 3000 actual volumes of steam, hydrogen and carbon monoxide per hour per volume of fluidized solids.

With the fluidized bed in gasifier 32, the carbonaceous solids are subjected to a temperature within the range between about 1000°F and about 1500°F, preferably between about 1200°F and 1400°F. At such a temperature the carbon-alkali metal catalyst will equilibrate the gas phase reactions occurring during gasification to produce additional methane and at the same time supply substantial amounts of additional exothermic heat in situ.

Due to the gas phase equilibrium conditions existing as a result of the carbon-alkali metal catalyst and due to the presence of added hydrogen and carbon monoxide, there will be a net heat production. Moreover, competing reactions, that in the absence of the catalyst and the added hydrogen and carbon monoxide would ordinarily tend to produce additional hydrogen and carbon monoxide, are suppressed. The heat produced tends to balance the endothermic heat consumed by the reaction of the steam with carbon and as the amount of carbon monoxide and hydrogen added increases, an overall thermoneutral reaction can be approached. When the gasifier is basically in heat balance, the heat required to preheat the feed to the reaction temperature and compensate for heat losses from the gasifier is supplied for the most part by excess heat in the gases introduced into the gasifier through line 36. In the absence of the exothermic heat provided by the catalyzed gas phase reactions, these gases would have to be heated to substantially higher temperatures than those normally employed here.

The gas leaving the fluidized bed in gasifier 32 passes through the upper section of the gasifier, which serves as a disengagement zone where the particles too heavy to be entrained by the gas leaving the vessel are returned to the bed. If desired, this disengagement zone may include one or more cyclone separators or the like for removing relatively large particles from the gas. The gas withdrawn from the upper part of the gasifier through line 37 will normally contain an equilibrium mixture at reaction temperature and pressure of methane, carbon dioxide, hydrogen, carbon monoxide and unreacted steam. Hydrogen sulfide, ammonia and other contaminants formed from sulfur and nitrogen contained in the feed material may also be present in this gas and entrained fines may also be present.

As is well known, basically the same gaseous effluent will be produced in the gasifier when steam is used to effect the gasification. As is also well known, the ultimate gaseous product depends upon the further processing to which this effluent is subjected. As a result, the improvement of this invention is equally applicable to any catalytic process wherein a carbonaceous material is gasified in the presence of steam.

In the embodiment illustrated, the effluent gas is introduced into cyclone separator or similar device 38 for removal of the larger fines. The overhead gas then passes through line 39 into a second separator 41 where smaller particles are removed. The gas from which the solids have been separated is taken overhead from separator 41 through line 42 and the fines are discharged downward through dip legs 40 and 43. These fines may be returned to the gasifier or passed to the alkali metal recovery portion of the process.

In the system shown in the drawing, a stream of high ash content char particles is withdrawn through line 44 from gasifier 32 in order to control the ash content of the system and permit the recovery and recycle of alkali metal constituents of the catalyst. The solids in line 44, which may be combined with fines recovered from the gasifier overhead gas through dip legs 40 and 43 and line 45, are passed to alkali metal recovery unit 46. The recovery unit will normally comprise a multistage countercurrent leaching system in which the high ash content particles are countercurrently contacted with water introduced through line 47. An aqueous solution of alkali metal compounds is withdrawn from the unit through line 48 and recycled through lines 49 and 18 to feed preparation zone 14. Ash residues from which soluble alkali metal compounds have been leached are withdrawn from the recovery unit through line 50 and may be disposed of as land fill or further treated to recover added alkali metal constituents.

The gas leaving separator 41 is passed through line 42 to gas-gas heat exchanger 51 where it is cooled by indirect heat exchange with a gaseous mixture of methane and steam introduced through line 77. The cooled gas is then passed through line 53 into waste heat boiler 54 where it is further cooled by indirect heat exchange with water introduced through line 55. Sufficient heat is transferred from the gas to the water to convert it into steam, which is withdrawn through line 56. During this cooling step, unreacted steam in the gas from exchanger 51 is condensed out and withdrawn as condensate through line 57. The cool gas leaving waste heat boiler 54 through line 58 is passed to water scrubber 59. Here the gas stream passes upward through the scrubber where it comes in contact with water injected into the top of the scrubber through line 60. The water absorbs ammonia and a portion of the hydrogen sulfide in the gas stream and is withdrawn from the bottom of the scrubber through line 61 and passed to downstream units for further processing. The water scrubbed gas stream is withdrawn from the scrubber through line 62 and is now ready for treatment to remove bulk amounts of hydrogen sulfide and other acid gases.

The gas stream is passed from water scrubber 59 through line 62 into the bottom of solvent scrubber 63. Here the gas passes upward through the contacting zone in the scrubber where it comes in contact with a downflowing stream of solvent such as monoethanolamine, diethanolamine, a solution of sodium salts of amino acids, methanol, hot potassium carbonate or the like introduced into the upper part of the solvent scrubber through line 64. If desired, the solvent scrubber may be provided with spray nozzles, perforated plates, bubble cap plates, packing or other means for promoting intimate contact between the gas and the solvent. As the gas rises through the contacting zone, hydrogen sulfide, carbon dioxide and other acid gases are absorbed by the solvent, which leaves the scrubber through line 65. The spent solvent containing carbon dioxide, hydrogen sulfide and other contaminants is passed through line 65 to a stripper, not shown in the drawing, where it is contacted with steam or other stripping gas to remove the absorbed contaminants and thereby regenerate the solvent. The regenerated solvent may then be reused by injecting it back into the top of the scrubber via line 64.

A clean gas containing essentially methane, hydrogen, and carbon monoxide in amounts substantially equivalent to the equilibrium quantities of those gases in the raw product gas withdrawn from gasifier 32 through line 37 is withdrawn overhead from the solvent scrubber via line 66. The methane content of the gas will normally range between about 20 and about 60 mole percent and the gas will be of an intermediate BTU heating value, normally containing between about 400 and about 750 BTUs per standard cubic foot.

As will be readily apparent, this intermediate BTU gas could be withdrawn as a product. When this is done a portion of the product could be separated and then subjected to steam reforming to produce the carbon monoxide and hydrogen required for improved thermal efficiency. Alternatively, the carbon monoxide and hydrogen could be provided from any of the sources therefor known in the prior art.

The intermediate BTU gas withdrawn overhead from solvent scrubber 63 through line 66 is introduced into heat transfer unit 67 where it passes in indirect heat exchange with liquid methane introduced through line 68. The methane vaporizes within the heat transfer unit and is discharged as the intermediate BTU gas, which is primarily composed of methane, hydrogen and carbon monoxide, is cooled to a low temperature approaching that required for liquefaction of the methane contained in the gas, after which the chilled gas is passed through line 70 into cryogenic unit 71. Here the gas is further cooled by conventional means until the temperature reaches a value sufficiently low to liquefy the methane under the pressure conditions existing in the unit. Compressors and other auxiliaries associated with the cryogenic unit are not shown. The amount of pressure required for the liquefaction step will depend in part upon the pressure at which the gasifier is operated and the pressure losses which are incurred in the various portions of the system. A substantially pure stream of liquefied methane is taken off through line 72 and may be withdrawn as product. In the In the embodiment illustrated, however, the methane is passed through line 68 into heat transfer unit 67 as described earlier. Hydrogen and carbon monoxide are withdrawn overhead from cryogenic unit 71 through line 80 and recovered as a chemical synthesis product gas. Normally, the cryogenic unit is operated and designed in such a manner that less than about 10 mole percent of methane, preferably less than about 5 mole percent, remains in the product gas removed through line 80. Thus, the chemical synthesis gas produced in the process is one of extremely high purity and therefore has many industrial applications.

As previously indicated, the methane could be withdrawn as product and the carbon monoxide and hydrogen separated in the cryogenic separator returned to the gasifier to facilitate thermal efficiency. In the embodiment illustrated, however, the recycle methane gas removed from heat transfer unit 67 and through line 69 is passed to compressor 73 where its pressure is increased to a value from about 25 psi to about 150 psi above the operating pressure in gasifier 32. The pressurized gas is withdrawn from compressor 73 through line 74 and passed through tubes 75 located in the convection section of steam reforming furnace 76. Here, the high pressure gas picks up heat via indirect heat exchange with the hot flue gases generated in the furnace. The methane gas is removed from the tubes 75 through line 77 and mixed with steam, which is generated in waste heat boiler 54 and injected into line 77 via line 56. The mixture of methane gas and steam is then passed through line 77 into gas-gas heat exchanger 51 where it is heated by indirect heat exchange with the raw product gas removed from separator 41. The heated mixture is removed from exchanger 51 and passed through line 78 to steam reforming furnace 76.

The preheated mixture of steam and methane gas in line 78 is introduced into the internal tubes 79 of the steam reforming furnace where the methane and steam react with one another in the presence of a conventional steam reforming catalyst. The catalyst will normally consist of metallic constituents supported on an inert carrier. The metallic constituent will normally be selected from Group VI-A and the iron group of the Periodic Table and may be chromium, molybdenum, tungsten, nickel, iron, and cobalt, and may include small amounts of potassium carbonate or a similar compound as a promoter. Suitable inert carriers include silica, alumina, silica-alumina, zeolites, and the like.

The reforming furnace is operated under conditions such that the methane in the feed gas will react with steam in the tubes 79 to produce hydrogen and carbon monoxide according to the following equation:

$$H_20 + CH_4 \longrightarrow 3H_2 + C0$$

The temperature in the reforming furnace will normally be maintained between about 1200°F and about 1800°F, preferably between about 100°F and about 300°F above the temperature in gasifier 32. The pressure will range between about 10 and about 30 psi above the pressure in the gasifier. The mole ratio of steam to methane introduced into the reactor will range between about 2:1 and about 15:1, preferably between about 3:1 and about 7:1. The reforming furnace may be fired by a portion of the methane gas removed from heat transfer unit 67 via line 69, a portion of the intermediate BTU gas removed from solvent scrubber 63 through line 66, or a similar fuel gas.

The gaseous effluent stream from the steam reforming furnace, which will normally be a mixture consisting primarily of hydrogen, carbon monoxide, and unreacted steam, is passed, preferably without substantial cooling, through lines 81 to manifolds 33 and/or 34' and ultimately into gasifier 32. This stream will be the primary source of the hydrogen, carbon monoxide and steam required in the gasifier. In a preferred embodiment, therefore, it is desirable that the reforming furnace effluent contain sufficient carbon monoxide and hydrogen to provide the desired thermal balance.

As pointed out previously, substantial quantities of exothermic heat are released in the gasifier as a result of the reaction of hydrogen with carbon oxides and the reaction of carbon monoxide with steam. Thus, the carbon monoxide and hydrogen in the reformer effluent stream comprises a substantial portion of the heat input into the gasifier. To supply the desired amounts of hydrogen and carbon monoxide in the effluent, sufficient methane should normally be present in the feed to the reforming furnace so that enough carbon monoxide and hydrogen is produced by steam reforming the methane to compensate for the amount of hydrogen and carbon monoxide removed in the chemical synthesis product gas withdrawn from the process overhead of cryogenic unit 71 through line 80.

In a preferred embodiment of the present invention, coal will be gasified with steam in the presence of an alkali metal catalyst and at a temperature within the range from about 1200 to about 1400°F and at a pressure within the range from about 200 to about 600 psia. The gasification will be accomplished in a fluid bed having a bed height within the range from about 60 to about 130 feet. The fluid bed will be maintained with steam introduced at the bottom of the gasification vessel and distributed through a suitable grid. The coal feed will be introduced into the fluid bed at one or more points located within the range from about 20 percent of the total height below the top of the bed to about 50 percent of the total height below the bed. The catalytic process will be operated so as to produce a substitute natural gas and substantially all of the carbon monoxide and hydrogen contained in the gaseous effluent from the gasifier will be recovered and recycled to the gasification vessel. The recycled carbon monoxide and hydrogen will be introduced into the fluid bed at one or more points positioned along the fluid bed within from about 20 percent of the total height from the top of the bed to about 50 percent of the total height from the top of the bed.

In the preferred embodiment, the amount of carbon monoxide and hydrogen recycled will be equal to the amount of carbon monoxide and hydrogen which would be produced as a result of the steam gasification of the coal if no carbon monoxide and hydrogen were introduced and when sufficient nominal holding time is provided to permit equilibration of the gaseous effluent from the gasifier. Also in the preferred embodiment, the exact point or points of the coal feed introduction will be optimized as a function of the activity of the coal to steam gasification.

Having thus broadly described the invention and set forth a preferred embodiment thereof, it is believed that the invention will be even better understood by reference to the following Example.

EXAMPLE

In this example, a series of steam gasifications was completed over a range of gasification temperatures of 1275°F and pressures of 5000 psia and at a steam to coal ratio of 1.58. From these tests, a kinetic model was developed and from this model, it has been predicted that the optimum fluid bed volume can be reduced by 11 percent by raising the coal feed point to a height within the range from about 10% below the top of the bed to about 60 percent of the total height below the top of the bed. It has also been predicted from this model that the reactor volume can be reduced by 27 percent if an equilibrium mixture of carbon monoxide and hydrogen is introduced at a point or points located at a point below the top of the bed by an amount equal to about 10 percent of the total height. It has further been predicted that the total volume can be reduced by 42 percent if both the feed point and the carbon monoxide and hydrogen point or points of introduction are both relocated. Based on predictions from the results obtained by relocating each feed point separately, it was anticipated that only a 35 percent reduction would have been realized by relocating both feed points simultaneously.

From the foregoing, it is believed readily apparent that elevation of either the feed point or the carbon monoxide and hydrogen introduction point will result in a significant reduction in reactor size or required bed height. It is also believed readily apparent that if both of these feed points are elevated a significant and synergistic reduction in total bed height is realized.

While the present invention has been described and illustrated by reference to a particular embodiment thereof, it will be appreciated by those of ordinary skill in the art that the same lends itself to variations not necessarily illustrated herein.

CLAIMS

1. A gasification process wherein a solid carbonaceous material is gasified in a fluidized bed in the presence of steam, added hydrogen, CO and a catalyst and wherein either the solid carbonaceous material or the added hydrogen and carbon monoxide or all three are introduced into the fluid bed at a point or points along the bed and within a range of distances ranging from about 10 percent of the total bed height below the top of the bed to about 60 percent of the total bed height from the top.

2. A process according to claim 1 wherein an alkali metal catalyst is present during gasification.

3. A process according to either of claims 1 and 2 wherein the amount of hydrogen and CO added to the fluid bed is sufficient to provide the desired thermal balance.

4. A process according to any one of the preceding claims wherein the amount of carbon monoxide and hydrogen added to the fluid bed is equal to the amount of carbon monoxide and hydrogen which would be produced during the steam gasification of the solid carbonaceous material if no carbon monoxide and hydrogen were introduced and when sufficient nominal holding time is provided to permit equilibration of the gaseous effluent from the gasifier.

5. A process according to any one of the preceding claims wherein the added hydrogen and carbon monoxide is recovered from the gaseous effluent from the gasification reactor.

6. A process according to any one of the preceding claims wherein substantially all of the hydrogen and carbon monoxide contained in the gasification reactor effluent is separated and recycled to the fluid bed.

7. A process according to any one of the preceding claims wherein the added hydrogen and carbon monoxide is obtained by reforming at least a portion of the hydrocarbon contained in the gasification reactor effluent.

8. A process according to any one of the preceding claims wherein all of the hydrocarbon gas contained in the gasification reactor effluent is separated, reformed and recycled to the fluid bed as a mixture of hydrogen and carbon monoxide.

9. A process according to any one of the preceding claims wherein the carbonaceous material is introduced at one or more points positioned along the fluid bed within a distance from about 20 percent of the total height below the top of the fluid bed to about 50 percent of the total height below the top of the fluid bed.

10. A process according to any one of the preceding claims wherein said carbonaceous material and the hydrogen and carbon monoxide are separately and independently introduced into said fluid bed at said point or points along the bed.

1/1

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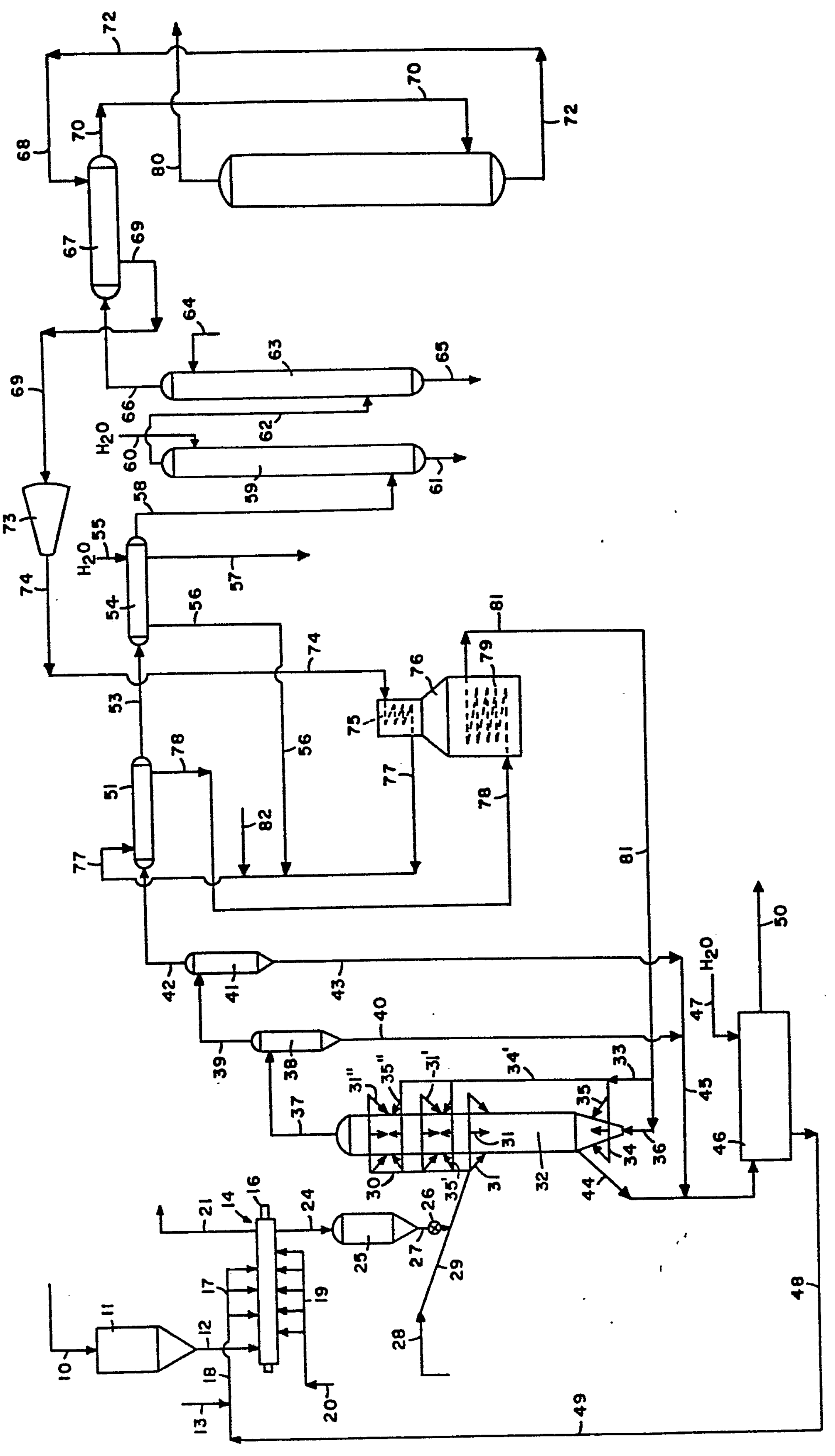

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European Patent Office

EUROPEAN SEARCH REPORT

Application number
EP 81 30 0493

DOCUMENTS CONSIDERED TO BE RELEVANT

Category	Citation of document with indication, where appropriate, of relevant passages	Relevant to claim
	US - A - 4 200 495 (LISS) * Column 7, lines 31-47; column 8, lines 1-69; column 14, lines 2-60 * --	1,9,10
	US - A - 4 211 669 (EAKMAN) * Columns 11-14; claims * --	1-8
A	DE - A - 2 741 805 (RHEINISCHE BRAUNKOHLENWERKE) * Page 1, claim 4; page 2 claim 6; page 8, lines 13-19 * --	1
A	US - A - 3 847 567 (KALINA)	
A D	FR - A - 2 381 820 (EXXON) & US - A - 4 118 204 ----	

CLASSIFICATION OF THE APPLICATION (Int Cl.³)

C 10 J 3/54

TECHNICAL FIELDS SEARCHED (Int. Cl.³)

C 10 J 3/54
3/56

CATEGORY OF CITED DOCUMENTS

X: particularly relevant
A: technological background
O: non-written disclosure
P: intermediate document
T: theory or principle underlying the invention
E: conflicting application
D: document cited in the application
L: citation for other reasons

&: member of the same patent family corresponding document

The present search report has been drawn up for all claims

Place of search	Date of completion of the search	Examiner
The Hague	15-10-1981	WENDLING

EPO Form 1503 1 06.78