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PROCESS FOR GASIFICATION OF HYDROCARBONS
TO HYDROGEN AND CARBON MONOXIDE
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3,128,164

FIGURE 1

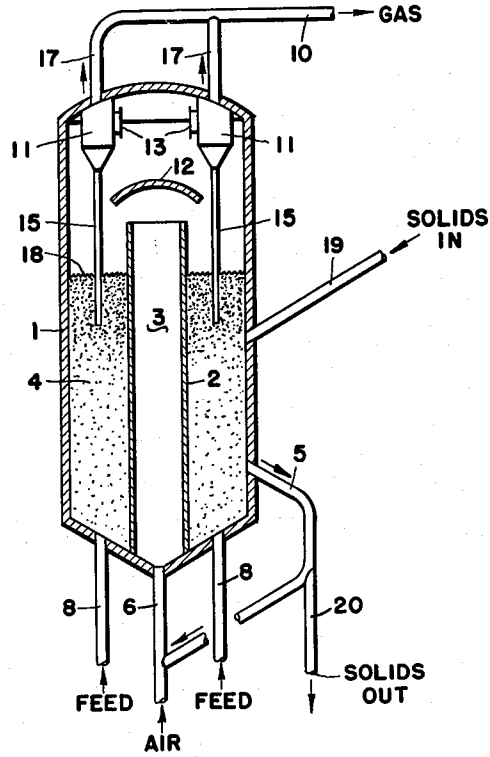
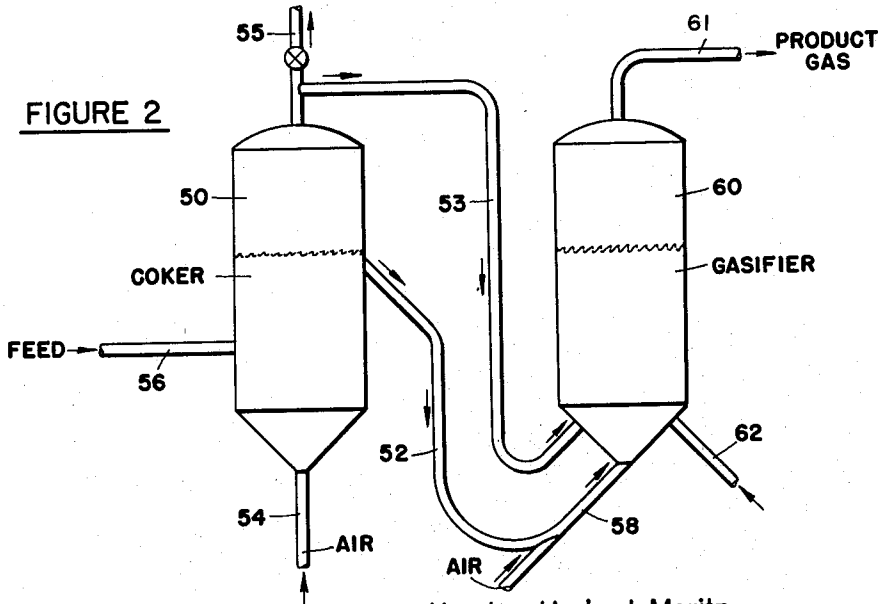


FIGURE 2



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3,128,164

PROCESS FOR GASIFICATION OF HYDROCARBONS TO HYDROGEN AND CARBON MONOXIDE

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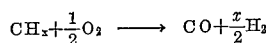
8 Claims. (Cl. 48-212)

The present invention relates to an improved process for reacting hydrocarbons in a fluidized bed of hot carbonaceous solids to make a reducing gas consisting predominantly of carbon monoxide and hydrogen. More specifically, the present invention relates to an improved process for converting hydrocarbonaceous gaseous and liquid fluids into reducing gases having high CO/CO₂ and high H₂/H₂O ratios in the presence of an oxidizing gas, such as air, and in the further presence of a fluidized bed of hot carbonaceous solids. Still more specifically, the present invention relates to an improved process for converting heavy oils and residua into a gaseous product riched in H₂ and CO than processes presently available.

The present application is a continuation-in-part of Serial No. 40,260, filed July 1, 1960.

Coking hydrocarbons and gasifying the product with an oxygen-containing gas, such as air, oxygen or steam, is well known by this time. The processes generally involve subjecting a heavy oil in the presence of a fluidized particulate mass, such as adsorbents, coke, activated carbon, and the like, in a gasification zone to contact with an oxidizing gas. The hydrocarbonaceous material is cracked, and the cracked products are oxidized. Though the mechanism is not too clear, probably the initial reaction products are CO₂ and H₂O. These interact at least in part with carbon either present in the bed or formed during the cracking step to produce H₂ and CO. Reaction temperatures are about 1800-2000° F. and heat for this is furnished by the combustion process.

An important problem connected with this process is the control of the reaction products' ratios. In particular when the gases are to be used for reducing purposes, as in metallurgical operations, it is important to recover a gas having the highest possible CO/CO₂ ratio. Ideally, this should be accomplished by injecting air and hydrocarbons at such stoichiometric ratios that essentially all of the carbon in the feed is converted to CO, in accordance with the equation

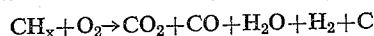


However, experience has shown that substantial portions of the oxygen in the product gas are in the form of CO₂ and H₂O rather than CO and H₂. The former gases are oxidizing constituents and are undesirable as constituents of reducing gases.

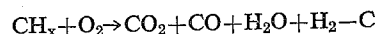
The present invention in its principal feature is based on the discovery that gas quality, i.e., the carbon recovery, is markedly improved when the gasification reactor is operated under "coke consumption conditions." "Coke consumption" means that part of the coke bed in the reactor is consumed continuously, and this is accomplished by lowering the carbon to oxygen ratio in the feed inlet stream. Thus, for example, if it is desired theoretically to convert all of the carbon in the hydrocarbon feed stream to carbon monoxide, it would be necessary, theoretically, to introduce into the reactor two moles of carbon for every mole of oxygen. Again, theoretically, the exit gas ratio of C/O₂ will then be 2. This exit gas ratio is termed the "carbon recovery" and is a convenient

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measure of reducing gas quality. If the C/O₂ ratio of the feed stream is greater than the C/O₂ ratio of the exit stream, i.e., carbon recovery, coke builds up in the reactor in accordance with the generalized reaction



The conditions promoting this reaction are referred to as "coke deposition" conditions. On the other hand, if the C/O₂ ratio of the feed is less than the C/O₂ ratio of the product, coke is consumed from the reactor in accordance with the reaction



This is referred to as "coke consumption" conditions, and coke is depleted from the reactor, i.e., it is consumed from the fluidized coke bed.

In actual practice the carbon recovery of the product is a long way from 2 and is governed by the kinetics of the gasification reaction. At low temperatures and short holding times over a relatively inert solid such as fluidized coke it may be as low as 1 or even less. At high temperatures, long holding times and very active carbonaceous solids, such as activated carbons, it may be as high as 1.9. Thus under coke deposition conditions, the following have been observed.

TABLE I

Coke.....	Carbon Recoveries		
	Fluid Coke		Activated Carbon
	15	55	
.....
Holding Time, Secs.....	15	55	15
1,800° F.....	1.05	1.36	1.45
1,900° F.....	1.22	1.55	1.65
2,000° F.....	1.39	1.74	1.85

From these data it is apparent that it is necessary to go to very high holding times and/or to use very high temperatures and/or highly activated carbons in order to obtain good carbon recoveries.

The gasification of hydrocarbons may be done for a variety of desired end products, i.e., town gas, synthesis gas, hydro gas, reducing gas, etc. It is clear that the end use of the gaseous product will determine its desired composition. In many applications, specifically in direct iron ore reduction (therefore "reducing gas") the gas must have very high ratios of CO/CO₂ and H₂/H₂O to maintain an equilibrium over the iron oxide favorable for iron formation. Ideally, the gas consists of pure hydrogen and CO. Carbon recovery is a convenient measure of the reducing efficiency of a gas. This is illustrated in the following table.

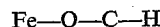
TABLE II

Iron Ore Reduction at 1800° F.

Carbon Recovery	Moles Fe Reduced/Mole C in Gas
1.29.....	0
1.6.....	0.3
2.0.....	0.53

Based on C/H ratio of gas=1.6.

These numbers are determined solely by the



equilibrium and show that for iron ore reduction at 1800° F. to proceed at all, the gas must have a carbon recovery of better than 1.3. For economical reasons it

should be better than 1.6. It is important to realize that the gasification of hydrocarbons with air is not thermodynamically limited above 1800° F. At 1800° F. and above the thermodynamic equilibrium, carbon recovery is essentially equal to 2, i.e., greatly favors formation of CO and H₂. Despite this favorable thermodynamic equilibrium Table I points out that high gas qualities acceptable for iron ore reduction are obtained only with great difficulty over fluid coke and with decreased difficulty over activated carbon. In other words, this is a rate limited process affected by properties of the solid, over which gasification is conducted. This invention teaches the unexpected result that coke consumption conditions in the reactor greatly enhance the rate of conversion of hydrocarbons to CO and H₂. This means that a smaller reactor and lower temperatures can be utilized.

It is therefore an important object of the present invention to set forth a novel and improved process for making reducing gases from hydrocarbons and in particular from heavy hydrocarbon oils, and obtain high yields of CO and H₂.

It is a further object of the present invention to set forth a two-stage process for effectively converting hydrocarbons into reducing gases.

Other and further objects and advantages of the present invention will be more clear hereinafter.

It has now been found that gas quality, in terms of carbon recovery, is markedly improved when the gasification reactor is operated under coke consumption conditions. In fact, improvement of conversion of over 45% has been realized when, under the same conditions of temperature and residence time, the gasification reactor has been operated in coke consumption rather than coke deposition conditions, i.e., with coke being consumed from the fluidized bed rather than being deposited therein. These results are unexpected for the following obvious reasons. The rate of solid-gaseous reactions is proportional to the time of exposure over the active surface. That this is so for the gasification system is shown in Table I. At the same temperature, longer holding times over fluid coke gave better gas qualities. And accordingly, when the holding time is decreased, poorer gas qualities are obtained.

Under coke consumption conditions, however, the reverse is found to be true. Due to the fact that coke is consumed from the reactor, the fluid bed level decreases. Thus, the holding time of the reactants over the coke is decreased. It would have been expected therefore that under coke consumption conditions the gas quality deteriorates rather than improves. That the reverse is true is shown by the following run record obtained under coke consumption conditions.

TABLE III

Run Hour	Feed Rate, g./hr.	Air Rate, s.c.f.h.	Coke Holdup, hr.	Carbon Recovery
3.....	244	70	29.8	1.49
5.....	244	70	28.4	1.50
7.....	244	70	27.1	1.49
9.....	244	70	25.8	1.53
11.....	244	70	24.6	1.55
13.....	244	70	23.5	1.58
15.....	244	70	22.3	1.66

The reasons why operating a gasification zone under coke consumption conditions should so markedly affect the product gas quality are not exactly known. One explanation may lie in the observation that, under coke consumption conditions, the surface area of the coke comprising the fluidized bed is markedly increased, and this more active coke may be more efficiently utilized by the reacting gases. Similarly, under coke deposition conditions it has been found that the coke surface area is destroyed.

In accordance with the present invention, therefore, a

gasification zone wherein an oxidizing gas, such as air, and a hydrocarbonaceous fluid such as heavy oils, Bunker C, tars and the like, are contacted at very high temperatures in the presence of a fluidized bed of coke is operated under coke consumption conditions.

Since coke is depleted from the fluidized bed, it must be replenished for a continuous operation. This can readily be effected by adding fresh coke to the reactor concomitantly with the addition of the hydrocarbon and the oxidizing gas. In one modification, a low temperature coking process furnishes the coke necessary to maintain the coke inventory in the gasification zone.

In operation of the gasification zone, the fluidizable coke particles have a size of about 40 to 300 microns and form a turbulent dense bed when the fluidizing gas has a linear velocity in the range of 0.5 to 4 feet per second through the bed.

For the gasification reaction in which the hydrocarbon feed material, liquid or gaseous, dispersed in the fluidized solids bed, the bed temperatures are in the range of 1750° to 2400° F., preferably 1750° to 2100° F., with a small temperature gradient preferably less than 100° F. throughout the bed.

Pressures in the gasification reaction zone are generally slightly above atmospheric, e.g. 0 to 40 p.s.i.g., but, the pressure can be increased.

The hydrocarbon feed should be admitted into the fluidized solids sufficiently above the bottom of the bed, i.e., where the air enters the bed, to avoid direct oxidation of the hydrocarbons by oxygen which causes formation of hot spots that in turn may cause deactivation and results in a fine carbon black formation. Such fine carbon black is less than 1 micron in size generally. Also, the hydrocarbon entrance into the bed should not be too far up toward the top of the bed, but where sufficient time of contact is allowed for the gaseous hydrocarbons from the feed to decompose in the presence of the coke-bearing fluid solids, at a temperature above 1750° F. For example, the hydrocarbons could be made to enter the bed about 1 foot above a grid, supporting the bed and through which air is distributed into the bed while a remaining 3 to 10 or more feet of the bed is above the level where the hydrocarbon feed enters. Thus, the turbulently agitated solids receiving carbon and coke deposits undergo sufficient rapid backmixing into contact with the air entering the bottom of the bed to obtain oxidation of such deposits, thus maintaining practically a uniform temperature throughout the bed.

The process of the present invention can be more clearly understood in connection with the drawing, wherein FIGURE 1 shows an apparatus suitable for carrying out the process of the present invention, and FIGURE 2 shows an embodiment employing a two vessel system including a coker and a gasifier.

Turning now to FIGURE 1, this corresponds to FIGURE 1 of Serial No. 40,260, filed July 1, 1960, of which the present application is a continuation-in-part.

The reactor shown in FIGURE 1 comprises vessel 1 containing fluidized solids and having communicating parts. Vessel 1 may be a heat-resistant steel shell lined with heat-resistant refractory material.

A central draft tube 2 constructed of suitable refractory material; e.g., silicon carbide, ceramics, metal alloy, or metal with refractory lining, is disposed in vessel 1 to partition off an interior reaction zone 3 where oxidation of coke or carbon deposits on the circulated solids takes place.

Carbon containing solids are passed from fluidized bed 4 of the hydrocarbon decomposition zone by way of pipe 5 and inlet pipe 6 into zone 3.

A free oxygen containing gas; e.g., air or enriched air, for oxidation of the coke deposits on the solids circulated to zone 3 is supplied from line 7. This air may be pre-heated; e.g., to a temperature of 400° to 1000° F.

The fluid hydrocarbon feed, such as methane, liquid

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hydrocarbon, or both gaseous and liquid hydrocarbons, is introduced through feed line 8 into solids bed 4 where the hydrocarbon material is decomposed to form hydrogen and coke deposits at a temperature in the range of 1750° to 2000° F. A hydrocarbon feed introduced as liquid may have a preheat temperature of 200° to 650° F. Gaseous hydrocarbon feeds may have a higher preheat temperature; e.g., up to 1200° F. The hydrocarbon feed may be a liquid hydrocarbon which is atomized into the cracking zone with a gas substantially free of oxygen.

The heat carrying solids are maintained in a fluidized state in both the cracking and oxidation zone. By fluidized is meant the maintaining of solid particles having a size in the range of 25 to 1000 microns in the form of a dense bed having a fairly distinct upper level using a fluidizing gas with a superficial gas velocity in the range of 0.2 to 10 ft./sec., preferably 0.5 to 3 ft./sec., and the maintaining of such sized particles in a more dilute entrained state at such gas velocities, with the solids in both instances occupying at least 20 vol. percent of the gas solids mixture. The fluidized solids provide turbulent mixing and excellent heat transfer between the zones. By the turbulent mixing of the solids, the temperature variation (ΔT) throughout the bed 4 is kept sufficiently small to prevent hot spot formation. Heat is added to the solids in bed 4 by conduction and radiation from tube 2, from walls of vessel 1, and from solids overflowing from tube 2 at its upper end. A deflecting baffle 12 may be located above the upper outlet of tube 2.

Gaseous product is withdrawn from vessel 1 through gas exit line 10. It consists principally of carbon monoxide formed inside tube 2 and free hydrogen rising from bed 4, which become admixed in the solids disengaging space above the zones. This gaseous product passes through one or more cyclone separators 11 having inlet ports 13, dip legs 15 for return of solids separated from the gases back to bed 4, and gas outlet lines 17 to main gas line 10 which conducts the gases to any unit or plant where they are used; e.g., for reducing an iron oxide. The product gases may be given other treatment if desired; e.g., by hot carbon to increase the concentration of carbon monoxide or a catalytic water gas shift treatment that increases the yield of hydrogen by the reaction of the carbon monoxide component with water.

Adequate turbulence of the solids in bed 4 to keep the ΔT low; i.e., below 100° F., throughout the bed is obtained by regulated flow rate of the hydrocarbon feed into the bed. Even when injecting a liquid hydrocarbon feed sufficient quantities of gaseous decomposition products, vapors and hydrogen are formed instantaneously and maintain fluidization of the solid particles. Thus, the dense bed of solid particles 4 can be kept free of undesired gases, especially oxygen or oxidizing gases.

Solid coke particles of suitable size for fluidization, e.g., petroleum coke or activated coke, are added through inlet 19 when the solids require replenishment. Used solids can be withdrawn through line 20.

The air used for reaction with carbon deposits on the solids is preferably introduced in a zone sufficiently separate from the hydrocarbon cracking zone to prevent partial oxidation of the hydrocarbon feed. In the operation of the reactor shown in FIGURE 1, air is introduced into the lower end of draft tube 2 and converts the carbon on the solids circulated from bed 4 by line 5 to carbon oxides. The oxidizing reaction in zone 3 is highly exothermic and permits a temperature above 2000° F. to be maintained in zone 3. Over 75% of the heat requirements of the gasification process of this invention are supplied by this oxidation of the carbon deposits.

The high temperatures in zone 3 favor formation of a high proportion of carbon monoxide relative to carbon dioxide. The reaction in the oxidizing zone 3 between carbon and oxygen from the air is a stepwise process in which first the carbon is believed to be oxidized to car-

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bon dioxide. This reaction is very fast and highly exothermic. As carbon dioxide is formed, a second reaction takes place between carbon dioxide and carbon to form carbon monoxide. The second reaction is endothermic and relatively slower than the first reaction and is favored by high temperatures, preferably temperatures above 2000° F. to go to completion. Thus, the conversion to carbon monoxide and the temperature in the oxidizing zone inside the draft tube are interrelated. The operation of the oxidation zone 3 and preheating requirements depend on the carbon monoxide concentration desired in the gas formed. For approximately complete conversion of the oxygen from the air to carbon monoxide, temperatures of 2200° F. and above are usually required for a convenient short holding time in the oxidation zone. In order to reach such elevated temperature, the air should be preheated to above 500° F.

While gas containing mainly nitrogen and carbon monoxide flows out at the upward open end of tube 2 when air is used as the source of oxygen, nearly pure hydrogen is produced in the bed 4 by the cracking of the hydrocarbon feed in this bed. A relatively small amount of water is formed inside the oxidation zone 3 by oxidation of hydrogen in and absorbed by the coke on the solid particles. In the upper part of the reactor a small amount of water may be formed by the reversible reaction of carbon dioxide with hydrogen.

The location of the oxidation zone is inside the vertical draft tube 2 of FIGURE 1 so that it is separated from, but in good heat transfer relationship to, the hydrocarbon cracking zone. Much of the heat formed in the oxidation of the coke deposits can be conducted through the tube wall and then be radiated into the surrounding bed 4 on all sides. Thus in the hydrocarbon cracking zone 3, a temperature in the range of 1750° to 2000° F. can be maintained with low variation in temperature throughout the bed.

The height and diameter of the draft tube 2 is designed for maximum efficiency in oxidizing carbon and coke deposits on the solids to carbon monoxide with a linear gas flow rate up through the tube in the range of 3 to 10 feet per second. With this rate of flow, the gas products, mainly carbon monoxide with a small amount of carbon dioxide, carry solids in suspension up to the top of the tube where the solids spill over into the surrounding cracking zone bed where the hot solids add heat, serve to promote the cracking reaction and accumulate fresh coke deposits from the cracking of the hydrocarbon feed.

Vessel 1 is operated, in accordance with the present invention, to maintain carbon consumption conditions. Table I lists pertinent operating conditions and provides a specific example thereof, drawn with reference to FIGURE 1.

TABLE IV
Operating Conditions for Coke Consumption

	Broad	Preferred
Temperature, ° F.....	1,700-2,200	1,800-2,000
Pressure, p.s.i.g.....	0-100	20-50
Feed, C/O ₂ mole ratio.....	0.5-1.5	1.0-1.5
Superficial velocity, Ft./Sec.....	0.5-4.0	1.0-2.5
Lbs. solids holdup/mole O ₂ in air/hr.....	100-2,000	300-1,000

Solids may be any form of coke or carbonaceous solids.

In FIGURE 2 there is shown diagrammatically a simplified scheme for providing coke to replenish that consumed in the operation of the gasification step. Hydrocarbon feed, such as Bunker C oil, is precoked in vessel 50 at normal coking temperatures, in the range of 900° to 1300° F., and in a conventional manner well known per se, enough air is admitted to coker 50 solely to maintain a heat balance. Thereafter, both coke and product gases, mostly hydrocarbons with some oxides of carbon, water, nitrogen, etc., are passed via lines 52 and 53 respectively to gasification vessel 60. Here the temperature

is raised to the desired gasification temperature by further air addition through line 58. In this manner the gasification stage is always operated under coke consumption conditions. A further advantage of this two stage system is that the H₂/CO ratio of the product gases may be varied at will by withdrawing portions of the gaseous product from vessel 50 through line 55 or by admitting further hydrocarbon feed to vessel 60 through line 62. No heat losses are incurred because the over-all balance remains the same.

The process of the present invention may be further illustrated by the following specific examples.

EXAMPLE 1

In a gasification process wherein a high Conradson carbon petroleum fraction was gasified by injection into a bed of fluidized coke, a 45% increase in conversion was obtained when, under the same conditions of temperature and holding time, the gasification was carried out under coke consumption conditions rather than coke deposition conditions.

Carbon Recovery Over Fluid Coke at 1915° F.

	Coke Consumption	Coke Deposition
Holding Time.....	10 secs.	10 secs.
Carbon Recovery.....	1.57	1.13

At 2000° F., the following results were obtained:

	Coke Consumption	Coke Deposition
Feed Rate, w./h./w.....	0.017	0.017
Carbon Recovery.....	1.85	1.74

These data were obtained with 35% coke consumption, based on feed. Depending on the C/O₂ ratio of the feed, the coke in the reactor may be consumed rapidly or slowly.

EXAMPLE 2

The coke holdup in the reactor is 30.9 lbs. The feed is a Bunker C fuel oil with a gravity of 12.6° A.P.I., a C/H of 8.36 and a Conradson carbon of 16.0 wt. percent. It is atomized into the reactor, one foot above the grid, with an inert N₂ stream at a rate of 445 g./hr. Air is admitted to the reactor below the grid at a rate of 70 s.c.f.h. This corresponds to a C/O₂ molar feed ratio of 1.5 and an inverse space velocity (lbs. C/mole O₂/hr.) of 800. Reactor temperature is 1900° F. Superficial velocity through the reactor, based on outlet gas rate, is 0.67 ft./sec. The average product dry gas analysis is:

H ₂	8.8
N ₂	72.6
CH ₄	1.0
CO	12.5
CO ₂	5.1
H ₂ O*	3.2

*Based on wet gas.

This corresponds to a carbon recovery of 1.35.

The carbon/O₂ ratio of the feed, therefore, is somewhat higher than the C/O₂ ratio of the product. Due to losses of carbon black and entrainment, however, the coke level in the reactor remained constant at about 30 to 31 lbs.

The feed rate is now lowered to 250 grams/hour, with the air rate remaining constant. Over the first four to five hours no change in product gas quality is noted, in agreement with previous experience which indicated that product gas quality is dependent on the inverse oxygen space velocity (lbs. C/mole O₂/hr.) only, and independent of hydrocarbon feed rate.

After several hours, however, the coke level in the

reactor begins to drop significantly because of the excess O₂, thereby decreasing the inverse space velocity. The expected decrease in gas quality does not take place, but on the contrary steadily improves. After 15 hours of running time the coke holdup in the reactor has dropped to 22.3 lbs., with an inverse space velocity of 574 lbs. C/mole O₂/hr. The dry product gas now analyzes as follows:

H ₂	7.6
N ₂	70.6
CH ₄	0.3
CO	18.8
CO ₂	2.7
H ₂ O*	1.7

*On wet basis.

This is a radical improvement over the initial gas product, corresponding to a carbon recovery of 1.66. Notice that the CO concentration has risen from 12.5 to 18.8 vol. percent, while CO₂ and water fell from 5.1 and 3.2 to 2.7 and 1.7 respectively. This corresponds to an increase of CO/CO₂ ratio from 2.5 to 7.0. This has taken place despite the fact that the total residence time of the vapors over the coke has decreased by 30%.

What is claimed is:

1. In a process wherein a fluidized bed of coke formed by coking a hydrocarbonaceous liquid feed is gasified in a gasification zone in the presence of air at a temperature of 1700° to 2200° F. and with the further presence of components formed in said coking stage produces a reducing gas containing H₂ and CO, wherein a portion of the coke undergoes combustion, the improvement which comprises maintaining the gasification zone under conditions such that the carbon/oxygen ratio of said feed components to said gasification zone is less than the carbon/oxygen ratio of the gasification product so that coke carbon is consumed from said bed at a higher rate than it is deposited by cracking of the feed hydrocarbon, wherein direct contact between said air and said feed is avoided and wherein replenishment coke is added to said bed to replace that consumed.

2. An improved process for producing a gas stream of high reducing capacity rich in H₂ and CO which comprises maintaining a dense turbulent bed of fluidized carbonaceous solids in a reaction zone, introducing a hydrocarbonaceous feed into said bed, cracking said feed to form hydrogen and coke deposits on said fluidized solids, introducing air into said bed to supply oxygen for converting said coke to oxides of carbon, avoiding direct contact between said air and said feed, maintaining said bed at a temperature of from about 1700° to 2200° F. and 0 to 100 p.s.i.g. pressure and operating said gasification under conditions such that the carbon/oxygen ratio of said feed and said air is less than the carbon/oxygen ratio of the product gases the effect of said ratios being that coke carbon is consumed from the bed at a higher rate than it is deposited by cracking of said feed and wherein replenishment coke is added to said bed to replace that consumed.

3. A process for producing a hot gas stream containing principally N₂, CO, and H₂, which comprises supplying carbonaceous solids to form a turbulent dense fluidized solids bed, introducing a hydrocarbon feed into a cracking zone of said bed where the hydrocarbon feed is cracked to form H₂ and coke deposits on said particles maintained at a temperature in the range of 1700° to 2200° F. by heat radiated and conducted from an adjacent combustion zone, fluidizing said particles carrying coke deposits in said combustion zone by air of which oxygen reacts with said coke deposits to form carbon oxide gases and to heat the particles to a temperature in the range of 1700° to 2200° F., wherein direct contact between said air and said feed is avoided, turbulently admixing heated particles fluidized from the combustion zone into said cracking zone, contacting gaseous hydro-

carbon and hydrogen from the cracking zone and gases containing CO, CO₂, with N₂ from the combustion zone at a temperature in the range of 1700° to 2200° F. with coke deposits of upwardly fluidized activated particles from the cracking zone, making said gases richer in CO and H₂, and separating a resulting hot gas stream containing mainly N₂, CO and H₂ with relatively small amounts of CO₂, H₂O and gaseous hydrocarbon, said gasification being carried out such that the rate of coke deposition on said bed from cracking of said feed is less than the coke consumed from said bed wherein replenishment coke is added to said bed to replace the coke consumed, and wherein the consumption of coke is regulated by controlling the ratio of carbon in said hydrocarbon feed to oxygen.

4. A process for producing a hot gas stream containing H₂, CO and N₂ as principal components, which components, which comprises maintaining a dense turbulent bed of fluidized carbonaceous solids in a reaction zone, introducing a hydrocarbon feed into an intermediate part of said bed between an upper major part and a bottom part of said bed, cracking the hydrocarbon feed to form hydrogen and coke deposits on said solids in said intermediate part, introducing air into said bottom part of said reaction zone to supply oxygen which converts said coke deposits to gaseous oxides of carbon with evolution of heat for maintaining the bed of solids at a temperature in the range of 1700° to 2200° F., avoiding direct contact between said air and said feed, admixing gaseous carbon oxides and N₂ from the bottom part with hydrogen and hydrocarbon gases in the intermediate part, passing the resulting gas mixture up through the upper major part of the bed for further reaction that enriches the gas mixture in H₂ and CO, maintaining gasification conditions within said bed such that the carbon/oxygen ratio of the hydrocarbon feed is less than the carbon/oxygen ratio of the product gas stream so that coke is consumed from said bed at a higher rate than it is deposited from the cracking of said feed and supplying replenishment coke to said bed to replace that consumed from said bed.

5. The process defined in claim 4 wherein the hydrocarbon feed is a liquid hydrocarbon that is atomized into the cracking zone with a gas substantially free of oxygen.

6. The process of claim 4 wherein said coke consumption conditions include a temperature in the range of 1800° to 2000° F., a pressure of 20 to 50 p.s.i.g., a carbon/oxygen mole ratio in the feed of 1.0 to 1.5, a superficial velocity of 1.0 to 25 feet/second, and a solids hold-up/mol O₂ in air/hour ratio of 30 to 1000 lbs.

7. The process of claim 4 wherein said solids are fluidized coke.

8. An improved two-stage process for producing a gas stream of high reducing capacity rich in H₂ and CO which comprises passing a hydrocarbon stream into a fluidized coking zone, maintaining a bed of fluidized carbonaceous solids in said zone, maintaining a coking temperature of 900° to 1300° F. in said bed by addition of only enough air sufficient to maintain heat balance, thereafter passing product gases and fluidized coke to a gasification zone, passing air into said zone so as to oxidize at least a portion of said carbonaceous solids, while avoiding contact between said air and said stream, maintaining temperatures of 1700° to 2200° F. in said zone, maintaining said gasification zone under coke consumption conditions so that the coke formed from said feed is consumed at a higher rate than it is formed and supplying coke to said bed to replenish that consumed and wherein coke consumption conditions are regulated by adjusting the hydrocarbon stream carbon to air oxygen ratio.

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