

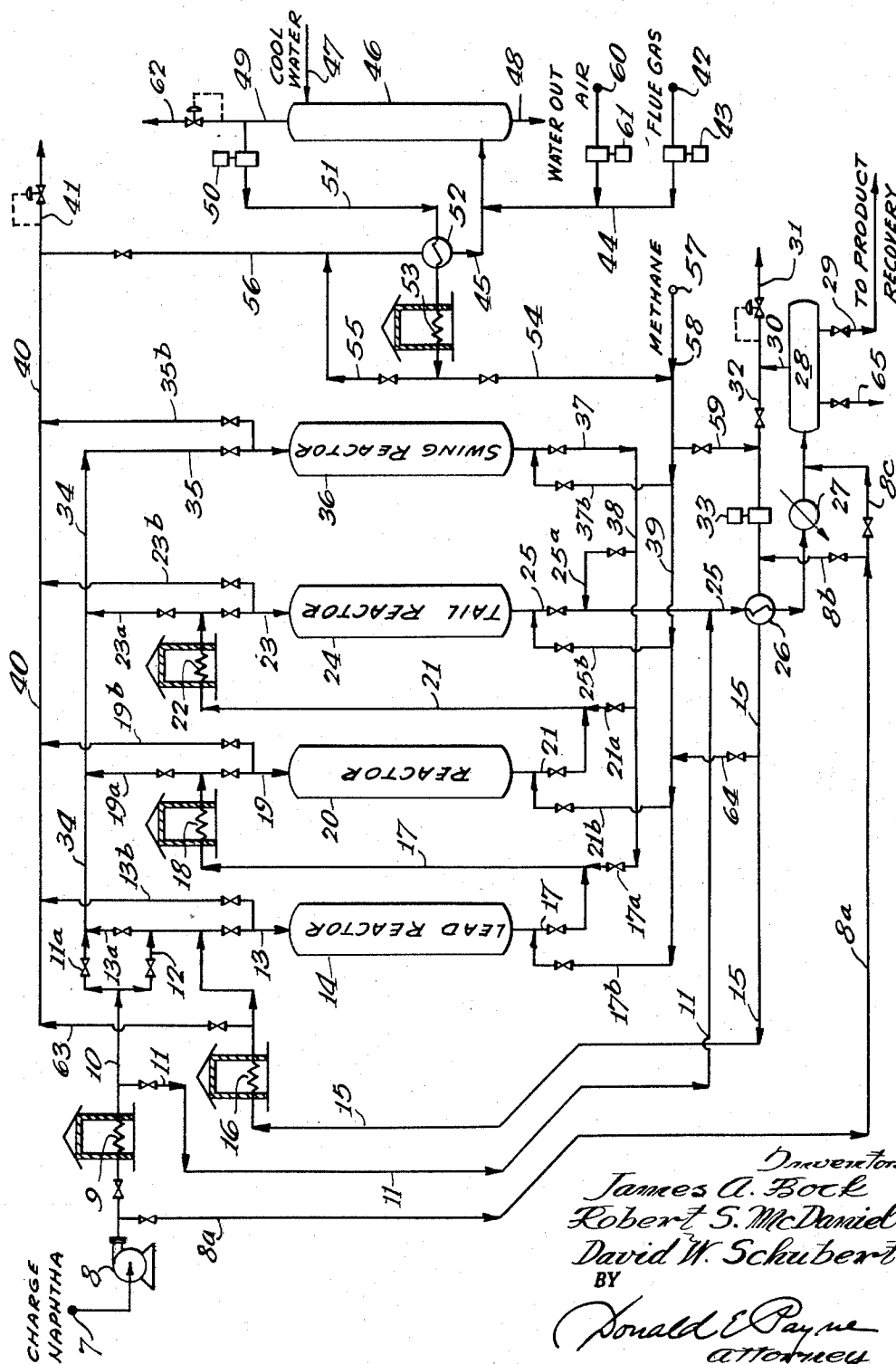
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START-UP OF PLATINUM CATALYST NAPHTHA HYDROFORMING SYSTEM

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START-UP OF PLATINUM CATALYST NAPHTHA HYDROFORMING SYSTEM

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9 Claims. (Cl. 208—146)

This invention relates to an improved start-up procedure for a platinum catalyst naphtha hydroforming system and it pertains more particularly to a method of starting up such a system in the absence of extraneous hydrogen, which method is of particular utility in starting up a regenerative system as exemplified by Ultraforming.

Platinum catalyst hydroforming systems present problems very different than those encountered in molybdena hydroforming systems and while it was known that the latter could be started up without the use of extraneous hydrogen, it has heretofore been considered essential to employ extraneous hydrogen in the start-up of platinum catalyst naphtha hydroformers. The provision of extraneous hydrogen is, of course, troublesome and expensive. An object of this invention is to provide a start-up procedure for platinum catalyst hydroforming which does not require the use of extraneous hydrogen and which, at the same time, does not result in appreciable catalyst deactivation or pressure build-up across catalyst beds. A further object is to provide a start-up procedure which will minimize loss of promoter components from the catalyst and minimize corrosion problems. Another object is to provide a start-up procedure which will avoid development of hot spots during initial periods of on-stream operation. Other objects will be apparent as the detailed description of the invention proceeds.

In practicing the invention, a supported platinum catalyst is charged to each of the reactors of a naphtha hydroforming system and the system is then purged with an inert gas such as flue gas at about atmospheric pressure, e.g. 5 to 10 p.s.i.g., at a temperature below about 200° F. and then pressured to at least the desired operating pressure to check for possible leaks. After pressure testing, the system is depressured and purged with a gas which is free from carbon oxides; while nitrogen may be used for this purpose, it is preferred to employ an unreactive hydrocarbon gas such as methane (conveniently supplied as natural gas), ethane, propane or light hydrocarbon gas mixtures of the type commonly referred to as LPG or fuel gas. After carbon oxides have been purged, the hydrocarbon gas is pressured to about 30 to 200 p.s.i. and circulated through the reactor system while the heater and reheater furnaces are slowly fired in order gradually to increase the temperature of the catalyst beds and to remove any water which may be present in the system. The heating should be relatively slow at least up to about 300° F. and the total heating may require about 4 to 40 hours or more during which time the hydrocarbon gas is circulated through the system and any water picked up therein is condensed, separated from the recycled hydrocarbon gas and removed from the system. The amount of water in the hydrocarbon gas circulating through the catalyst beds should be below 5 mol percent and preferably below 1 mol percent. Aqueous condensate thus removed may be corrosive and it should be withdrawn before it comes in contact with the circulating compressor.

When the catalyst beds have been heated to about 700°

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F. by the circulating hydrocarbon gas stream and substantially all water has been removed from the system and when the pressure in the system is substantially below 200 p.s.i., naphtha charge may be slowly introduced into the circulating stream. In a regenerative system the catalyst bed in either the swing reactor or one of the regular reactors may be raised to a somewhat higher temperature of about 750° F., usually not higher than about 800° F. A preferred operation is to connect the swing reactor in tail reactor position so that while methane is being circulated in the rest of the system, naphtha preheated to about 750–800° F. is introduced directly into the swing reactor and the effluent therefrom is discharged directly through heat exchangers to the hydrogen separator. The hydrogen thus formed in the swing reactor is recycled through the whole system and its concentration in the recycled gas rapidly reaches 80 percent or more. When the hydrogen pressure reaches the desired level, e.g. about 200 p.s.i.g. or more, the incoming naphtha charge at about 800° F. transfer line temperature is cut into the first reactor for on-stream flow through regular on-stream reheaters and reactors. The catalyst in the on-stream reactors is thus protected by hydrogen initially generated in the swing reactor. As soon as the system is brought to the desired pressure of operation, e.g. about 300 p.s.i., the temperature of the initial heater and reheaters is increased from about 800° F. to about 850 to 950° F., e.g. about 900° F., and the system is thus gradually brought to on-stream operating conditions.

By initially operating at relatively low hydrogen partial pressure, i.e. about 200 p.s.i. or lower, and relatively low temperature, i.e. not more than about 800° F., any tendency toward formation of run away hot spots due to hydrocracking is avoided. Any carbon deposited on the catalyst in initial generation of hydrogen is limited to the catalyst in the swing reactor since hydrogen produced in the swing reactor will be adequate to prevent appreciable carbon formation in the remaining reactors when charge is introduced thereto. The swing reactor may then be cut out of the system, the catalyst therein may be regenerated, and the other reactors will remain on-stream without necessity for regeneration for a much longer period of time than would otherwise be possible.

By using a methane or other unreactive hydrocarbon gas for drying the system and establishing circulation and initial preheating conditions, catalyst activity is protected. It is important that all oxygen be purged from the system before methane or other hydrocarbon gas is introduced because the platinum catalyzes combustion even at relatively low temperatures. By initiating the introduction of charging stock at low temperature and low pressure, carbon formation on the catalyst is minimized and, in a non-regenerative system, the initial dehydrogenation may be effected at temperatures somewhat lower than those hereinabove stated, i.e. about 650° to about 700° F. Start-up temperatures in the range of 700 to 800° F. are preferable in regenerative platinum hydroforming systems and by operating with one of the reactors at a higher temperature than the remaining reactors, any carbon deposition can be largely restricted to the higher temperature reactor; as soon as desired high hydrogen content of recycle gases and operating pressures have been established, the reactor which was initially at high temperature may be cut out of the system and the catalyst therein may be regenerated while the remaining reactors continue in on-stream operation.

The invention will be more clearly understood by reference to the following example read in conjunction with the accompanying drawing which is a schematic flow diagram of an Ultraforming system in which our start-up procedure is particularly advantageous.

In normal operation of an Ultraforming system a

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naphtha charge such, for example, as the 150 to 360° F. fraction of M-C virgin naphtha is introduced from source 7 by pump 8 through preheater 9 and transfer line 10 from which the preheated charge may be by-passed by line 11 to the product recovery system during start-up procedure. In on-stream operation transfer line 10 will discharge through lines 12 and 13 to reactor 14 along with recycled hydrogen from line 15 which is preheated in heater 16. Effluent from reactor 14 passes through line 17, reheater 18 and transfer line 19 to reactor 20. Effluent from reactor 20 passes through line 21, reheater 22 and transfer line 23 to tail reactor 24. It should be understood that more than three reheater-reactor stages may be employed in the system.

Effluent from the tail reactor flows through line 25, heat exchanger 26 and cooler 27 to separator 28 from which hydroformed product is withdrawn through line 29 to a stabilizer and/or conventional product recovery system. A part of the hydrogen withdrawn from the separator through line 30 may be vented through line 31 but usually about 4,000 to 10,000 cubic feet per barrel of charge is recycled through line 32 by means of circulating compressor 33 to line 15.

Transfer lines 11a, 13a, 19a and 23a may be selectively connected to header 34 for discharging through line 35 to swing reactor 36, the effluent from which passes through line 37 to header 38 and thence through line 17a to line 17, line 21a to line 21 or line 25a to line 25. With the valves in lines 11, 12, 13a and 19a closed and the valves in lines 23, 23a and 11a opened, preheated naphtha may be introduced directly to the swing reactor without interrupting circulation of gases through the remaining reactors and a part of the circulating gases may be passed to the swing reactor. When on-stream flow is established, the valves in lines 11 and 11a remain closed and the valve in line 12 remains open, valves in lines 13, 17, 19, 21, 23 and 25 being open and valves in lines 13a, 17a, 19a, 21a, 23a and 25a being closed when the swing reactor is cut out of on-stream position. The swing reactor may be substituted for the lead reactor by opening valves in lines 13a, 35, 37 and 17a and closing valves 13 and 17. Alternatively, it may be substituted for intermediate reactor 20 by opening valves in lines 19a, 35, 37 and 21a and closing the valves in lines 19 and 21. The swing reactor may take the place of the tail reactor by opening valves in lines 23a, 35, 37 and 25a and closing valves in lines 23 and 25. It will thus be seen that each of the reactors may be taken off-stream for regeneration and replaced by the swing reactor and that, alternatively, the swing reactor may be connected to operate in parallel with any of the other on-stream reactors during periods when no regeneration is required.

In some Ultraforming systems the recycled hydrogen and the naphtha charge are heated in the same preheater. In such systems the charge introduced by pump 8 may be introduced by lines 8a and 8b to line 15 just ahead of heat exchanger 26 during normal operation and may be introduced by line 8a and line 8c to the line entering separator 28 during start-up.

Each of the reactors is preferably aluminized and provided with a refractory lining of low iron content. They may each contain about the same amount of catalyst although, if desired, the subsequent reactors may contain somewhat more catalyst than the initial reactors. The catalyst may be of any known type of supported platinum catalyst and the platinum is preferably supported on alumina; it may be prepared by compositing a platinum chloride with an alumina support as described, for example, in U.S. 2,659,701 and it preferably contains about .3 to .6 weight percent of platinum.

In regenerative systems the on-stream pressure is usually below about 400 p.s.i.g., i.e. in the range of about 200 to 350 p.s.i.g. or more, while in non-regenerative systems the pressure is usually about 500 p.s.i.g. or higher. The inlet temperatures to each reactor are usually in the

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range of about 850 to 970° F., e.g. about 920° F., and may be approximately the same for each reactor although it is sometimes desirable to employ somewhat lower inlet temperature to the initial reactor than to the remaining reactors. The over-all weight space velocity may be in the range of about 1 to 4 pounds of naphtha per pound of catalyst per hour. There is, of course, a pressure drop in the system so that the lead reactor may operate at about 50 to 100 p.s.i. higher pressure than the tail reactor.

For effecting purging and regeneration of the catalyst in any bed, purge gases and regeneration gases may be introduced through manifold line 39 and a selected one of lines 17b, 21b, 25b and 37b. Such purge and regeneration gases may be selectively withdrawn through lines 13b, 19b, 23b and 35b to manifold line 40 from which gases may be vented or flared through line 41. Flue gas from source 42 may be introduced to the system by compressor 43 and passed by lines 44 and 45 through a cooler 46 which is preferably a scrubbing tower into which cool water is introduced through line 47 and from which water is withdrawn through line 48. The scrubbed flue gas withdrawn from the top of the tower through line 49 is passed by compressor 50 through line 51, heat exchanger 52, heater 53 and line 54 to manifold line 39 when it is desired to introduce flue gas into the system for purging and/or regeneration. By closing the valve in line 54 and opening the valve in line 55, the flue gas may be recirculated through line 56, heat exchanger 52 and line 45 back to the scrubber.

Methane or other hydrocarbon gas may be introduced as a purging fluid from source 57 and line 58 to manifold line 39. Either flue gas or methane from manifold line 39 may be introduced to the inlet of circulating compressor 33 by line 59. Air may be introduced from source 60 by compressor 61 for effecting regeneration and/or rejuvenation of the catalyst. Excess flue gas may be vented from the system during regeneration by line 62. Hot hydrogen for stripping hydrocarbons from catalyst in a blocked-out reactor may be introduced by line 63 to manifold line 40 and thence through one of lines 13b, 19b, 23b or 35b to the selected reactor. Also, hydrogen may be introduced from line 15 to manifold line 39 by line 64.

In starting up this system in accordance with our invention the entire system is, of course, first checked for mechanical defects, cleaned out, dried, and the reactors are charged with catalyst. A layer of alumina balls is preferably placed on top of each of the catalyst beds to prevent swirling of the catalyst pellets which might otherwise lead to abrasion and production of catalyst fines. Flue gas from source 42 is next introduced into the system and passed via 44, 45, 46, 49, 50, 51, 52, 53, 54, 39 and 59 to the inlet of circulating compressor 33 and it is thereafter passed through all of the heaters, transfer lines and reactors to remove all air therefrom, the air and flue gas being at this time purged from the system through line 31. The temperature of the flue gas during this purge should be below 200° F. and the pressure is preferably about atmospheric, i.e. about 5 to 10 p.s.i.g.

After the initial flue gas purge, the system is pressured with flue gas to at least the desired operating pressure, e.g. about 350 p.s.i.g., to check for possible gas leaks. After pressure testing, the system is again depressured by opening the valve in line 31. Then the valve in line 55 is opened, the valve in line 54 is closed, methane or other inert hydrocarbon gas is introduced from source 57 by line 58, line 39 and line 59 to the inlet of circulating compressor 33. As soon as all of the flue gas has been purged from the system by methane, the valve in line 31 is set to hold a back pressure of about 30 to 200 p.s.i., e.g. about 100 p.s.i., and the methane is continuously circulated through all of the reactors while gradually firing the furnaces to supply heat through heaters 16, 18 and 22, the swing reactor preferably being connected in parallel with the tail reactor at this time. The temperature

of the circulating methane is slowly increased to effect drying of the catalyst and any further drying of the reactor linings that may be required. Water thus removed may be condensed in cooler 27, separated from the circulating methane in separator 28 and withdrawn therefrom through line 65, the valve in line 29 being closed at this time. Any condensate produced during this drying operation may be corrosive and, if desired, auxiliary drying means may be employed to prevent any moisture or corrosive liquid from entering circulating compressor 33. The heating up of the reactors should be sufficiently slow up to about 300° F. so that the gas will not pick up more than about 5 mol percent of water. After this initial drying step, the temperature may be increased more rapidly to about 700° F. and circulation should be continued at this high temperature until no further appreciable amounts of water are removed from the system.

After the system has thus been dried by hot circulating methane, the valve in line 65 is closed and the valve in line 29 is opened. The naphtha preheater 9 is fired and charged naphtha preheated to about 800° F. is introduced through by-pass line 11 (the valves in lines 11a and 12 being closed) directly to product effluent line 25 from which it passes through exchanger 26, cooler 27, separator 28 and line 29 to the product recovery system for establishing operating conditions therein. If a single preheater is employed, naphtha charge is introduced at this time to the separator through lines 8a and 8c, the valve in line 8b being closed.

The pressure of the hot circulating methane stream, if it is not already below about 200 p.s.i.g., should next be reduced to as low a pressure as is feasible, preferably to a pressure in the range of about 30 to 150 p.s.i.g. In a non-regenerative platinum hydroforming system a small amount of naphtha charge is then introduced through line 8b or through the naphtha preheater 9 and introduced with methane at the defined low pressure and at a temperature of at least about 650° but not substantially higher than 700° F. into reactors wherein the catalyst is not heated substantially above 700° F. Dehydrogenation of the charge takes place rather slowly under these conditions but in a non-regenerative system it is important to avoid deposition of carbonaceous deposits. When the hydrogen concentration in the recycled gas stream reaches about 80 percent or more and the pressure built up by generated hydrogen reaches about 500 p.s.i., the temperature of the incoming naphtha may be increased to the desired operating temperature.

In the preferred regenerative hydroforming system the catalyst in one of the reactors is preferably heated to about 750 to 800° F. and the naphtha which is introduced into this reactor is preheated to about the same or somewhat higher temperature so that the generation of hydrogen will largely be accomplished in a single vessel. When the swing reactor is employed for hydrogen generation and all catalyst beds have been heated to about 700° F. by circulating methane, the valve in line 23 may be closed, the valve in line 23a opened and the temperature of heater 22 increased to provide a transfer line temperature of about 800° F. so that the swing reactor is thus heated to about 750 to 800° F. Next, with the pressure in the circulating methane system below about 200 p.s.i. and preferably in the range of 30 to 150 p.s.i., the valves in lines 23a and 23 are opened, the valve in line 11a is opened while valves in lines 11, 12 and 13a are closed, naphtha vapors preheated to approximately 800° F. are introduced directly through lines 34 and 35 to the swing reactor while recycled gases are continuously circulated through heater 16, reactor 14, heater 18, reactor 20, heater 22 and reactors 24 and 36 (in parallel), the temperature of these heaters at this time being increased to bring the temperature of the lead and intermediate reactors up to about 800 to 850° F. The hydrogen generated by dehydrogenation in the swing reactor will quickly in-

crease the hydrogen concentration of the circulating gases to about 80 percent or more and the pressure in the system may be increased by the generated hydrogen to the desired operating pressure. As soon as the circulating gas stream is at least about 80 percent hydrogen, the valve in line 12 is opened and the valve in line 11a is closed so that the charge preheated to about 800° F. is now cut into the hot circulating hydrogen stream. If and when regeneration of the catalyst in the swing reactor is desired, sufficient hot hydrogen is available for stripping hydrocarbons therefrom via lines 63, 40 and 35b after which the valves in lines 63, 37 and 25a are closed. The transfer line temperatures of heaters 9, 16, 18 and 22 may now be increased to the desired level, e.g. about 900 to 920° F. and if the operating pressure has not already been attained, it may quickly be attained in this on-stream operation. Thus, carbon deposition on the lead, intermediate and tail reactors is substantially avoided by starting up with hydrogen produced in the swing reactor. By gradually increasing temperature and pressure during this start-up procedure, the danger of encountering preliminary hot spots (apparently caused by undesired hydrocracking) may be avoided. Since gases are continuously being circulated through the reactors, reheaters and transfer lines while hydrogen is displacing methane, there will be no abrupt temperature changes in any part of the system.

Instead of employing the swing reactor for initially generating hydrogen, the lead reactor, the tail reactor or in fact any of the on-stream reactors may be preheated to a temperature of about 750 to 800° F. by raising the temperature of the appropriate heater and the system may be brought on-stream by introducing charging stock at about 700° F. to all of the reactors except the preheated reactor for which the start-up inlet temperature is preferably about 750 to 800° F. With this method of start-up, line 11a is not required and the preheated naphtha may be introduced by transfer lines 10 and 12 when a separate preheater 9 is employed or may be introduced into line 15 by line 8b when hydrogen and naphtha are preheated in the same heating coil. In this embodiment one reactor, e.g. the swing reactor, is preferably blocked out during the initial start-up and the inlet temperatures to the reactors which are at about 700° F. should not be increased until the hydrogen concentration in the circulating gas is at least about 80 percent. When the circulating methane has thus been essentially replaced by hydrogen and the pressure of the circulating gases has been increased to at least about 200 p.s.i., all transfer line temperatures may be increased to the desired operating level and any further increase in pressure may be built up so that on-stream operating conditions are fully established. In this case, the reactor which was operated at the initially high temperature and in which most of the carbon deposition has occurred may be cut out of the system for regeneration and replaced by the reactor which was initially blocked out.

The method of effecting catalyst regeneration will be described as applied to the swing reactor but it will be understood that the same procedure may be employed for any one of the other reactors when it is blocked out. When the charge inlet valve in line 35 is closed and while the valve in line 37 remains open, hot hydrogen is introduced by line 63 to manifold line 40 and thence through line 35b to strip out any hydrocarbons that may remain in the reactor, this stripped material being discharged through lines 38 and 25a to line 25. Next, the valve in lines 63 and 37 are closed and reactor 36 is depressured by opening the valve in line 41. Next, the reactor is purged to eliminate hydrogen therefrom either by introducing flue gas from line 54 or methane from source 57 via lines 39 and 37b, the purge gases being vented through lines 35b, 40 and 41. If methane is employed as a purge, it is followed by a flue gas purge and after the flue gas purge, the valve in line 41 is closed

and introduction of flue gas from source 42 is continued to pressure the reactor with flue gas to approximately the same pressure as that employed in on-stream processes, i.e. about 300 p.s.i. This flue gas, under such pressure, is circulated through the reactor by compressor 50 and sufficient heat is supplied to the circulating gas by heater 53 to bring the temperature of the catalyst bed to about 650 to 750° F. preparatory to initiating regeneration. Next, controlled amounts of air are introduced from source 60 by compressor 61 into the circulating flue gas stream at a rate to effect combustion of carbonaceous deposits without exceeding a combustion zone temperature of about 1050° F. The hot flue gas leaving the reactor at about this temperature passes by lines 35b, 40 and 56 through heat exchanger 52 and thence through line 45 to scrubber 46 wherein the gas is scrubbed with cool water for condensing and eliminating most of the water formed by combustion of hydrocarbonaceous deposits. The net amount of flue gas production is vented from the system through line 62, the valve in which is set to maintain the desired back pressure of about 300 p.s.i. The cooled flue gas which is recirculated by compressor 50 may be further dried by passing through a desiccant bed (not shown) before it is returned through heat exchanger 52 to heater 53 which, during regeneration, maintains a transfer line temperature of approximately 700° F.

The transfer line temperature of heater 53 is increased to about 950° F. in order to reheat the catalyst bed to that temperature. If rejuvenation is required (and it usually is not required until the catalyst has been regenerated many times) the introduction of flue gas is stopped and the introduction of air is continued so that the catalyst is treated with a circulating air stream at a pressure of about 300 p.s.i. and a temperature of about 950° F. or more for a period of about one-half hour to twelve hours or more depending upon the extent of rejuvenation required.

After the regeneration (or after rejuvenation if rejuvenation has been effected) the introduction of air is stopped, the system is depressured by slowly opening the valve in line 41 and flue gas is introduced from source 42 to purge all oxygen from the swing reactor and from the regeneration system. After the flue gas purge, the system may be purged with methane introduced from source 57 through line 37b. The flue gas (or methane, if employed) may then be purged from the system with hydrogen introduced through lines 64, 39 and 37b. When a methane purge is employed, however, it is not always necessary to employ a hydrogen purge.

Next, the valve in line 37b is closed and the reactor is pressured with hot hydrogen introduced by lines 63, 40 and 35b. When the reactor is thus brought to desired operating pressure, the valves in lines 63 and 35b are closed and the reactor may be placed on-stream by opening valves in lines 35 and 37.

While our invention has been described as applied to a particular Ultraforming system, it should be understood that it is applicable to other types of regenerative platinum catalyst hydroforming systems and it is also applicable to non-regenerative systems such as Platforming. Various alternative arrangements and operating conditions will be apparent from the above description to those skilled in the art.

We claim:

1. The method of starting up a platinum catalyst naphtha hydroforming system including lead, intermediate and tail reactors, heating zones, a gas separation zone, a compressor and lines for recycling separated gas, which method comprises purging catalyst beds and reactors with an inert gas to remove oxygen therefrom, purging inert gas from the reactors with an unreactive hydrocarbon gas, circulating said hydrocarbon gas through the reactors while heating it to increase the reactor temperatures to

about 700° F., introducing preheated naphtha charging stock at a temperature in the range of about 700° to 800° F. into one of the reactors while the pressure of the circulating hydrocarbon gas is in the range of about 30 to 200 p.s.i.g., continuing the introduction of naphtha charging stock at a temperature in said range until the hydrogen concentration in recycled gas reaches about 80 percent and then increasing the temperature of the introduced naphtha charge to at least about 850° F.

2. The method of claim 1 which includes the step of simultaneously increasing the pressure of the recycled gas while the hydrogen concentration thereof is being increased.

3. The method of claim 1 wherein the unreactive hydrocarbon gas is chiefly methane.

4. The method of claim 1 which includes the steps of introducing the unreactive hydrocarbon gas at the inlet of said compressor, passing it through said heating zones and reactors, cooling the effluent gas leaving the tail reactor to condense water and separating water from the circulating gas in the separating zone.

5. The method of claim 1 which includes the step of initially introducing preheated naphtha charging stock into a reactor which is at a higher temperature than the remaining reactors.

6. The method of claim 5 which includes the step of initially passing preheated naphtha only through the swing reactor of a regenerative hydroforming system containing such a swing reactor.

7. The method of starting up a platinum catalyst naphtha hydroforming system including lead, intermediate and tail reactors, heating zones, a gas separation zone, a compressor and lines for recycling gas from said separation zone, which method comprises purging free oxygen from said system with a non-combustible gas, purging said non-combustible gas from the system by introducing thereto an unreactive hydrocarbon gas, circulating said hydrocarbon gas through the reactors at gradually increasing reactor temperatures up to about 650 to 700° F. while removing from the circulating gas any water removed from the system by said gas, heating the catalyst in one of said reactors to a temperature in the range of about 700 to 800° F., introducing preheated naphtha charge into said reactor at a temperature in said range and at an initial pressure normal operating pressure whereby dehydrogenation of naphtha is effected and the hydrogen concentration in the recycled gas is increased, and increasing the temperature of the introduced naphtha to at least about 850° F. in all of the on-stream reactors after sufficient hydrogen has been built up in the recycled gas stream to substantially prevent carbon formation on the catalyst.

8. The method of claim 7 in which the hydroforming system is non-regenerative and in which the introduced preheated naphtha charging stock is at a temperature which is initially not substantially higher than 700° F.

9. The method of claim 7 in which the hydroforming system is regenerative and contains a swing reactor, which method includes the steps of initially introducing naphtha at a temperature substantially higher than 700° F. into a selected reactor for providing a hydrogen build-up in the system and subsequently blocking out said selected reactor and regenerating catalyst therein.

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UNITED STATES PATENT OFFICE
CERTIFICATE OF CORRECTION

Patent No. 2,910,430

October 27, 1959

James A. Bock et al.

It is hereby certified that error appears in the printed specification of the above numbered patent requiring correction and that the said Letters Patent should read as corrected below.

Column 8, line 45, after "pressure" insert "and" below "and".

Signed and sealed this 10th day of May 1960.

(SEAL)

Attest:

KARL H. AXLINE

Attesting Officer

ROBERT C. WATSON
Commissioner of Patents