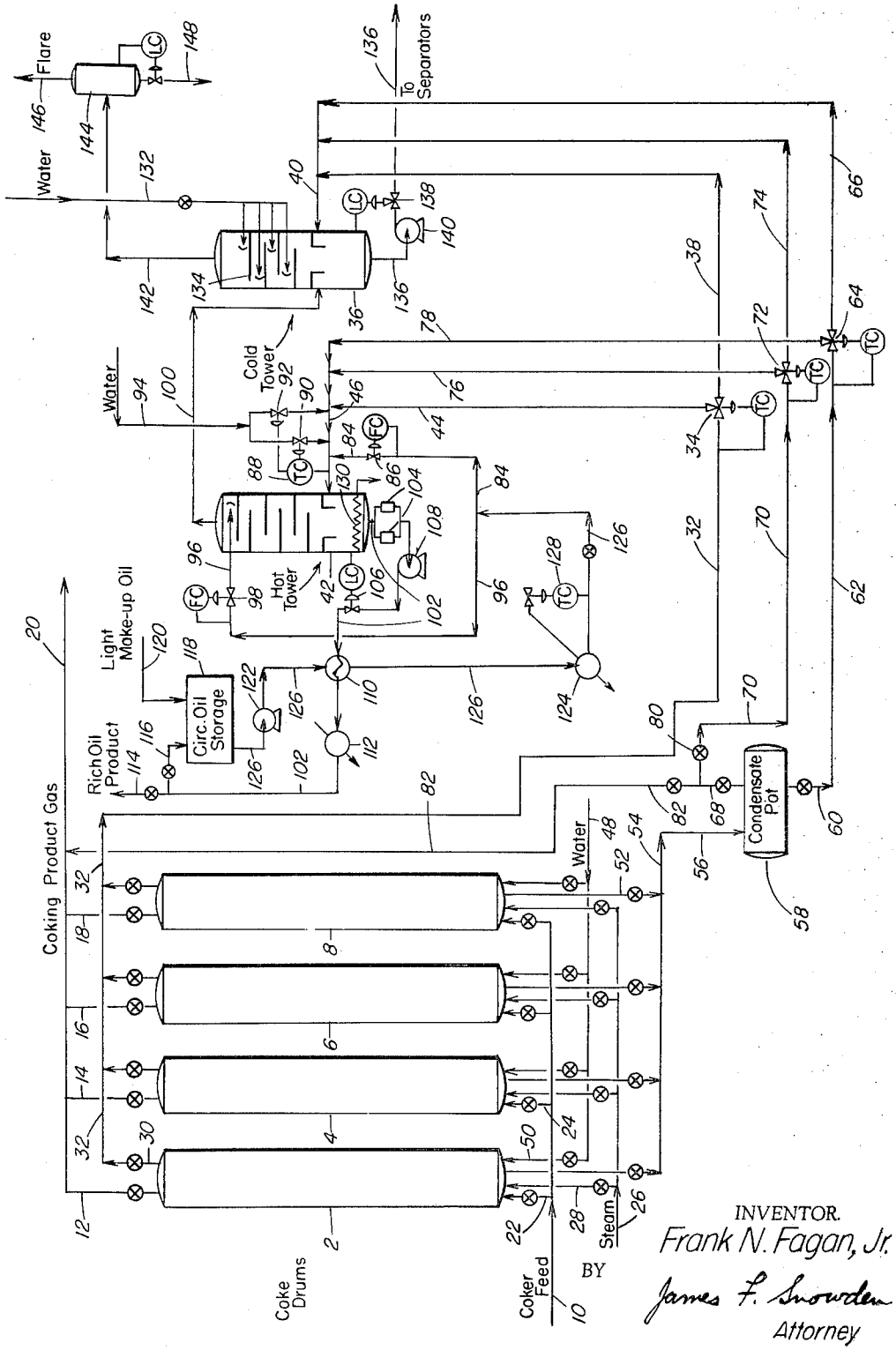
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COKER BLOW DOWN RECOVERY PROCESS

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The present invention relates to an improved process for the recovery of hydrocarbons from mixtures at varying temperatures and pressures containing hydrocarbons of varying densities and varying proportions of moisture. In particular it is concerned with the recovery of hydrocarbons from the blow down effluents from coke drums containing coke in delayed coking systems as well as from the fluids employed in preheating empty coke drums.

Delayed coking systems have long been employed with recognized success for thermally cracking a wide variety of heavy petroleum fractions including residua, tars, asphalts, etc., into synthetic crudes having hydrocarbons ranging from gases to material boiling under atmospheric pressure at 1200° F. or higher as well as coke. These systems commonly employ a battery of coke drums, as for instance, two to six drums, so that the coking process with delivery of the gaseous phase to a fractionating tower may continue without interruption as each coke drum becomes filled to the level of the space allowed for foaming by merely switching the coker feed to a preheated empty coke drum. The full coke drum is depressured from about 30 pounds per square inch gage (hereinafter designated p.s.i.g.) to atmospheric pressure and cooled from about 900° F. to about 150° F. (all temperatures herein being expressed in degrees Fahrenheit unless otherwise stated) in a blow down operation which also serves to purify the coke by the removal of substantially all of the volatile matter therein and to recover as much of the hydrocarbons in said volatile matter as possible.

More troubles have probably been encountered in the blow down step than in any other phase of the operation of delayed cokers. The volatile matter in the coke bed ranges from hydrogen and methane to hydrocarbons having a boiling point of about 1200° F. or higher and lighter hydrocarbons and hydrogen are produced during blow down by cracking in the presence of steam. Most of the difficulties have involved the tars or heavy oils in the charge to the blow down system, as these have produced extremely stable dispersions in water. For simplicity these emulsions of liquid oils in water and/or suspensions of solid hydrocarbons in water are hereinafter termed emulsions. Hydrocarbons having a density approximating or greater than water form particularly stable emulsions. The plugging of towers, pipes and other equipment in the blow down system with coker tars has also occurred. It has generally not been commercially feasible to break the stable emulsions and these emulsions carry heavy materials, such as coke and tars, a portion of which often drops out later in sewer lines and obstructs the flow therethrough. There is, of course, a distinct economic loss in the hydrocarbons which are carried away in the waste water, but the pollution of rivers and other surface waters with these hydrocarbon emulsions is considered more objectionable by most persons. Preventing and minimizing water pollution is continually becoming more important as nations become more industrialized, more highly populated, and more concerned with public health; consequently, there is an increasing demand for preventing or minimizing the pollution of water.

The problem of operating a blow down recovery system efficiently with a minimum plant investment is intensified by the continually changing nature of the streams charged to the unit. As is pointed out in detail hereinafter, these

streams vary greatly in composition, volume and temperature; the variations are sudden at times and gradual at other times, irregular at one moment and cyclic at another; and the changes occurring in one stream may differ from those taking place in another charge stream in magnitude or in direction or both.

An object of the invention is an improved method for the recovery of hydrocarbons from fluid streams of changing characteristics.

Another object of the invention is to provide an improvement in the recovery of hydrocarbons from mixtures of hydrocarbon and moisture of varying composition at varying temperatures.

A further object of the invention is to provide an improved process for the recovery of the hydrocarbons from the blow down effluent of a delayed coker system.

Still another object of the invention is to minimize the pollution of surface waters.

A still further object of the invention is to provide for the recovery of hydrocarbons from mixtures thereof with water without the formation of stable emulsions.

Yet another object is to provide for the recovery of an increased percentage of volatile hydrocarbons from a coke bed in a delayed coker unit.

Other objects and advantages of the invention will be apparent to those skilled in the art upon consideration of the detailed disclosure hereinbelow.

The above objects, as well as other objects and advantages, of the invention are obtained in the present method for the recovery of hydrocarbons from mixtures of moisture therewith of the type described by the improvement which comprises separating hydrocarbons of liquid densities approximating that of water in a first separation zone maintained at a temperature above that at which free water exists in the liquid phase, discharging from said first zone a liquid phase containing high density hydrocarbons and a vapor phase containing moisture and light hydrocarbons, condensing at least a substantial part of said vapor phase and separating vapor from a liquid mixture of water and light hydrocarbons in a second separation zone, generating a function or signal representative of the temperature of said charge mixture and diverting said charge mixture to said second separation zone responsive to values of said function representative of temperatures not substantially above the boiling point of water under the prevailing conditions.

Other aspects of the invention relate to regulating the heating of a liquid hydrocarbon absorbent to maintain the desired temperature in the first separation zone, for example, between about 250 and 375° F. and preferably between about 275 and 350° F.; employing a plurality of charge streams and diverting each to the second separation zone before its temperature drops below 275° F. or preferably before dropping to about 325° F.; maintaining the charge to the first separation zone below about 400° F. by quenching as necessary with the hydrocarbon absorbent and/or water; maintaining the density of the enriched absorbent liquid substantially lighter than water, and keeping the viscosity of the enriched liquid below about 500 Saybolt seconds Universal (SSU) measured at 300° F. for ready pumpability.

An important feature of the invention is the avoidance of condensing or absorbing the heavier and denser hydrocarbons, especially those denser than water, in the presence of water in the liquid state as such avoidance precludes formation of the highly objectionable and stable emulsions described hereinbefore. This is accomplished by absorbing or condensing the high density hydrocarbons in a stripping or separation zone maintained at temperature conditions such that free moisture cannot exist in the liquid state therein. Suitable temperature

control is effected by techniques including regulating the temperature of the charge mixture to said zone, regulating the temperature of the liquid absorbent in that zone and also of the gaseous effluent therefrom. The gaseous effluent containing lighter hydrocarbons and steam is condensed in large part at a substantially lower temperature, usually close to ambient temperature, in a second separation zone by a cooling medium along with any charge stream which is too cool to be introduced into the first zone.

The absorption liquid in the first zone also serves to dissolve the high density hydrocarbons and bring their viscosities in solution into a range suitable for pumping with conventional pumps and, more importantly, to produce solutions having densities distinctly below that of water. The solution containing the high density hydrocarbons may thereafter be freely mixed with water as it does not tend to form the undesirable stable emulsions therewith. The mixture of low density hydrocarbons and water obtained as a liquid phase effluent of the second separating zone readily separates into two liquid layers and thus also is not subject to the formation of stable emulsions.

For a better understanding of the nature and objects of this invention reference should be had to the accompanying drawing, which is a schematic representation or flow sheet of the portion of a delayed coking unit involved in the recovery of volatile matter from the coke drums.

Turning now to the drawing a battery of coke drums **2**, **4**, **6** and **8** of a delayed coker is operated to coke a conventional feed comprising a mixture of sweet and sour stocks have the following characteristics:

Gravity, ° API		11.6
Initial boiling point	° F.	695
5 vol. percent boiling point	° F.	893
10 vol. percent boiling point	° F.	940
Sulfur	Wt. percent	2.0
Conradson carbon residue	do	13.1

This feed enters through conduit **10** and a branch conduit connected to the bottom of each of the coke drums. The coking product gas is withdrawn from the operating drums through the valve lines **12**, **14**, **16**, and **18** into conduit **20** which carried the vaporized product to a fractionating tower (not shown). The products of this coking operation are essentially hydrocarbons and coke in the following boiling ranges and percentages based on the weight of feed:

	Percent
Dry gas	6.0
Butanes	2.1
Other C_4–380° F.	15.7
380–500° F.	4.4
500–650° F.	15.0
650–925° F.	24.4
925–approx. 1200° F.	9.4
Dry coke	23.0

As the coking operation proceeds, the drums which are on stream gradually fill up with coke and high boiling liquids or semisolids, such as tars, to the bottoms of their foam spaces. At this point, the drums must then be taken off stream and emptied. The battery of four coke drums is operated in a cycle in which two of the drums are on-stream and another two off-stream. A typical cycle comprises coking for 24 hours, a blow down period of 5 hours, draining and coke removal 8.5 hours, preheating 8 hours, and idle time 2.5 hours.

The blow down operation includes four periods, but the effluent from the coke drum which is being blown down is passed to the blow down recovery system only during the last three of these periods. However, during the preheat period in an empty drum, both steam and then effluent vapors from an operating or on-stream coke drum are passed into the blow down recovery system.

The sequential stages of the blow down operation are known as the "little steam," "big steam," "little water," and "big water" stages.

To describe the operation when coke drum **2** has just become full of coke, it is taken off-stream by closing the valve in feed line **22**. At the same time the clean, empty coke drum **4**, preheated to a temperature of about 800° F., is put on-stream by opening the valves in feed line **24** and product line **14**. Drum **6** is in the middle of an on-stream coking period, and the drum **8** is undergoing a decoking operation with the cutting of coke therefrom by drills and high pressure jets of water in conventional fashion.

In the first or little steam stage of the blow down operation, the easily strippable portion of volatiles in the coke bed are distilled off by admitting steam at any suitable pressure, typically 100 p.s.i.g. or more, via conduits **26** and **28** at the relatively slow rate of 5,000 lbs./hr. into the approximately 50 feet deep bed of hot coke in the drum **2** which has an internal diameter of 19 feet. The effluent is gaseous which term is employed herein to designate vaporized substances as well as materials that are gases at normal temperatures, and it is taken off via conduits **12** and **20** to the fractionating tower in the coker system, for this initial blow down effluent is quite similar in composition to the effluent of an on-stream coker. This step requires about thirty minutes and the temperature on the coke bed drops from about 900 to 850° during that time while the pressure in the drum is held at 15–20 p.s.i.g. or about half of the pressure in an on-stream coke drum.

Next follows the big steam step in which the steam is supplied at the same pressure but at the higher rate of about 15,000 lbs. of steam per hour for a period of 1.5 hours by opening the valve further in conduit **28**. As the steam rate is increased, the flow of vapor to the coker fractionator is discontinued by closing the valve in line **12** and opening the valve in conduit **30** to admit the vapor effluent into the main blow down conduit **32** which terminates in the automatic temperature controlled three-way valve **34**. The three-way valve directs the blow down effluent to the cold blow down tower **36** via lines **38** and **40** or alternatively to the hot blow down tower **42** via conduits **44** and **46** depending on whether the stream in line **32** is below or above the temperature setting of valve **34**. At the optimum setting of 350°, the stream in conduit **32** is initially cooled so much by its passage through the cold conduit that it is directed to cold tower **36**, but the high temperature of the blow down effluent quickly heats up line **32**. The stream is switched to the hot tower **42** when it reaches 350° at the temperature sensing point immediately upstream of valve **34**. If necessary or desirable, the interval during which the effluent flows to the cold tower may be reduced by passing steam or a hot gas, such as the effluent of an on-stream coke drum, through conduit **32** for a sufficient preheating period before the blow down effluent is admitted to that line.

The big steam interval is 90 minutes long and the temperature of the coke bed gradually decreases to about 700° by the end of this period. The temperature of the stream of blow down effluent does not decrease at a constant rate as a result of such variables as uneven porosity in the coke bed or the steam occasionally reaching hot spots in the coke bed, that is regions wherein the temperature is substantially higher than adjacent material. Also there is a considerable formation of light gases during this step, especially the last half of the period and the subsequent little water operation, as a result of the cracking of coke or hydrocarbons in the presence of steam.

As the heavy steaming continues, the concentration of the more volatile normally liquid hydrocarbons in the blow down effluent gradually decreases, and there is an increase in the proportion of vaporized hydrocarbons

boiling in the middle range 500–800° F. as well as in the extremely high boiling material, that is the heavier hydrocarbons with densities extending up to and beyond the density of water, for example, having specific gravities as high as 1.10 to 1.15. These denser hydrocarbons are the most troublesome in forming difficult emulsions with water.

In the next period of the cycle, the steaming is discontinued by closing the valve in line 28 and water is admitted at a carefully controlled low rate of about 50 gallons per minute through conduits 48 and 50 for one hour. The water admitted to the drum during this little water stage of course quickly flashes into steam upon reaching the hot coke in drum 2 and a careful control of the low water rate is necessary for two reasons. First, to avoid creating excessive steam pressures in the coke drum, and secondly, to avoid generating steam in a volume too great for the blow down system to handle. During this step the flow of vapor is still entirely through main blow down condiut 32 and the automatic valve 34 continues to direct the stream to either the hot or cold tower according to any fluctuations in temperature as described earlier. There is a decrease in concentration of hydrocarbons in the effluent as this stage progresses. Also, the rate of cooling the coke bed increases substantially and the temperature drops off to about 500° F. Nevertheless, a sizable proportion of the highest density hydrocarbons is carried over in the blow down vapors at this time at temperatures far below their atmospheric boiling points under the influence of steam distillation.

In the final or big water stage, the flow of water through the valve line 50 is increased substantially to a rate of about 500 gallons per minute and allowed to continue at this rate until the drum fills up with water and overflows briefly into the blow down conduit 32. This requires about two hours and the drum temperature drops to a 175°–200° F. range which is sufficiently low to permit the safe removal of the drum heads and extraction of the coke in a later operation which forms no part of the present invention. The hydrocarbon content of the blow down stream continues to decrease during this interval and the effluent which flows through line 32 mainly to the hot tower 42 at the begining of the step is diverted to the cold tower most of the time during the middle and latter parts of the big water interval.

Meanwhile, the empty coke drum 8 is being preheated over an extended period and substantial amounts of the heating fluids are delivered to the blow down system for recovery. During the final half hour of the big steam and first half hour of the little water operations in coke drum 2, drum 8 is steamed to raise its temperature to about 300° F. and also check the pressure tightness of the drum after the reheading operation at pressures around 50 p.s.i.g. or 15 p.s.i. above normal coking pressures. Steam condenses on the cold walls of drum 8 and the condensate is drawn off via conduits 52, 54 and 56 into the condensate pot 58. The liquid phase is withdrawn from the pot, preferably under control of a liquid level controller (not shown) via conduits 60 and 62 connect with an automatic temperature controlled three-way valve 64. The latter is similar to the valve 34 described earlier in both construction and operation. During this stage of the preheating, valve 64 directs all of the aqueous condensate containing any residual hydrocarbons from the condensate pot and drum 8 to cold tower 36 via lines 66 and 40. This liquid is far too cold to be introduced into the hot tower as that would promote the formation of some extremely stable oil in water emulsions there. At the same time, vapor from the condensate pot is being taken via conduits 68 and 70 to automatic control valve 72 which is of the same type as control valves 34 and 64 and is also adjusted for operation at the same temperature of 350°. Since the present steam temperature in line 70 is below that figure, all of this steam is conducted to cold tower 36 via lines 74 and 40.

After about one hour the steam to drum 8 is shut off and a portion of the effluent of the on-stream coke drums 4 and 6 is admitted to the top of drum 8 from the coking product line 20 by gradually and cautiously opening the valve in line 18. Care is exercised in admitting this stream to drum 8 in order to avoid any undesirable transfer of liquids from drums 4 and 6 to drum 8. This stage of the heat-up operation requires several hours, and the flow through drum 8 is increased until eventually about 10 to 15 percent by volume of the product stream in conduit 20 from the two operating coke drums is flowing into drum 8. At first, much of this gaseous phase material is condensed in the warm drum 8 which is at a temperature of about 250° F. and the condensate is withdrawn into the pot 58 from which both the vapor and liquid phases are transferred to the cold tower 36 as described earlier in reference to the steam condensate. This continues until the streams passing through conduits 70 and 62 respectively attain a temperature of 350° as the condensate pot becomes hotter and hotter; then, each of the streams is directed via conduits 76 and 78 respectively to the inlet line 46 of the hot tower. During this prolonged heating with the coker effluent, the drum 8 and pot 58 may reach temperatures slightly above 400°; then it is generally preferable to deliver the vapor stream from condensate pot 58 to coking product line 20 and thence to the coker fractionator as a synthetic crude rather than to the hot blow down tower 42 which is intended for the recovery of slop oil. This is performed by closing valve 80 and opening the valve in conduit 82.

For optimum results, the temperatures of any streams being charged into inlet manifold 46 of the hot tower from conduits 44, 76 and/or 78 are above 350°. As one of the factors for insuring that the denser hydrocarbons in the streams are separated in the liquid phase in the hot tower, the inlet manifold is quenched with one or more streams as may be necessary to keep the overall temperature of the charge from rising above 375°. A circulating oil at 300°, which is flowing at all times that charge material is being introduced into the hot tower, is admitted to inlet conduit 46 from line 84 under the regulation of an automatic control valve 86 to maintain the desired conditions in hot tower 42. It is sprayed into the charge mixture, preferably but not necessarily countercurrent to the flow through line 46, in order to mix the two streams as thoroughly as possible. This of course reduces the temperature in inlet manifold 46 somewhat and further regulation is obtained as necessary by setting the temperature controller 88 to the desired maximum tower inlet temperature of 375°. This temperature controller operates two automatic valves 90 and 92 to spray quenching streams of cold water from the conduit 94 into the inlet manifold 46 at two different points therein.

A larger stream of the circulating oil is introduced at 300° through the conduit 96 under the regulation of a flow control valve 98 into the top of the tower 42 to cascade downward over the trays countercurrent to the rising gaseous phase of the charge. Tower 42 is a hot absorption tower where the denser hydrocarbons are separated in the liquid phase by absorption in the circulating or flux oil while substantially all of the free moisture and the lighter hydrocarbons pass overhead via line 100 to the cold tower 36. This is accomplished by maintaining the hot tower under conditions such that free moisture cannot exist there in the liquid state. Water forms stable emulsions with solid or liquid particles of the extremely dense hydrocarbons but not with the hydrocarbons of distinctly lighter specific gravity than water such as naphthas and gas oils. The term "free moisture" is not used herein to include the very small amount of moisture dissolved in the liquid hydrocarbon stream withdrawn at the bottom of the hot tower. At the temperatures selected, such dissolved moisture only amounts to a few parts per million which is insufficient to promote any emulsion forming tendencies.

The enriched circulating oil with the heavier or denser

hydrocarbons dissolved therein is taken off at the bottom of the tower in line **102** through the twin strainers **104** in response to regulation by the automatic valve **106** set to maintain a constant level of liquid in the bottom of tower **42**. Particles of coke and any undissolved tar are filtered off in the strainers. The pump **108** circulates the enriched oil through the heat exchanger **110** and cooler **112** which is cooled to about 150° for safety purposes in storage and transfer. It is then divided between lines **114** and **116** and about ⅕ of the oil (corresponding to the net absorption) is withdrawn as a rich oil product in line **114** and recycled either to an oil storage tank in the refinery where this slop oil of about 15° API gravity may be used for coker feed or refinery fuel oil or charged directly to a process unit such as a coker. The balance of the flux oil is carried to the storage tank **118** where it is mixed with a stream of light make up oil entering in conduit **120**. This make up oil is typically a gas oil of about 30° API gravity. It is desirably obtained from the liquid phase of the bottoms of the cold blow down tower but may also be taken from other sources, such as a coker net product stream or other process unit streams. From tank **118** the circulating oil is returned to the hot tower and inlet line via pump **122** and heat exchangers **110** and **124** in conduit **126** and thence split between the lines **96** and **84**. It is important to maintain the absorption oil at the proper inlet temperature for the tower and this is accomplished by means of a temperature controller **128** which governs the supply of a heating medium **3** such as steam, to the final heat exchanger **124**. For maximum economy the recirculation of the absorption oil to the hot tower is reduced or increased proportionately to the volume of material charged to the tower in order to minimize pumping and heating costs. A heating coil **130** for steam or other suitable heating medium is located in the bottom of hot tower **42**. With the battery of four coke drums described here, the blow down system is operated intermittently with idle periods in the cycle wherein the temperature of the flux oil in the tower drops considerably. It is important to keep the hot tower at a temperature such that free moisture cannot exist there in the liquid state at all times, not merely usually, when a charge mixture is entering the tower, for a small amount of free water can form difficult emulsions which may affect the entire contents of tank **118** or the refinery storage tank. The coil **130** is used to bring the liquid in the bottom of the blow down tower back to the desired operating temperature of 325° after an idle period or shutdown.

In the cold tower **36**, through which is cascading a large flow of cold water introduced from the conduit **132** on to several of the trays **134**, light hydrocarbons and most of the vaporized moisture from the inlet conduits **40** and **100** are condensed. This includes most of the C_4 and heavier hydrocarbons which are withdrawn from the bottom of the tower in line **136** in response to an automatic valve **138** regulated by a level controller to maintain a constant level at the bottom of tower **36**. The pump **140** passes this liquid phase to one or more separating tanks (not shown) where the relatively light liquid hydrocarbons are decanted from the top of the aqueous phase and a small amount of heavy solids collected by a raking device is occasionally withdrawn from the bottom of the water layer. The quantity of heavier hydrocarbons accumulating in the cold side of the blow down system is so small as to create no problems as to forming stable emulsions. The decanted light oil is a gas oil of about 30° API gravity. It is a suitable make up oil for the flux oil circulated through the hot tower.

The flow of cold water through line **132** is regulated by the valve therein in response to the quantity and temperature of the streams entering cold tower **36** in conduits **40** and **100** to maintain a substantially constant exit temperature of preferably 100° F. or less in conduit **136**. This exit temperature is of course dependent on the quantity and temperature of the cold water supply. The gaseous effluent from the tower in line **142** is at substantially the same temperature as the bottoms. It is passed to the knockout pot or separator **144** from which hydrocarbon and other uncondensible gases pass overhead to recovery facilities such as a gas plant or to a flare via line **146** while a liquid stream **148** is withdrawn at the bottom under the regulation of a level controller **150** and passed to a sour water stripper for the recovery of any hydrocarbons therein.

From the foregoing description, it is apparent that the coker blow down system operates under almost continually varying conditions. These variations occur in the temperatures of the various mixtures of hydrocarbons and moisture flowing from the coke drums or condensate pot. Also the ratio of hydrocarbons to moisture in the streams charged to the blow down system is constantly changing sometimes increasing and sometimes decreasing. In addition, the proportions of hydrocarbons of different densities do not remain constant. The pressure on the coke drum undergoing blow down decreases from about 30 p.s.i.g. to almost atmospheric pressure. Moreover, there are obviously great changes in the volume in the materials charged to the blow down system. While some of these changes are gradual over most of the period, there are occasional irregularities as a result of various occurrences, such as water or steam reaching a hot spot in the coke bed. Also there are variations in direction as well as magnitude with one of the streams to the blow down system gradually decreasing from a relatively high temperature during the big steam period, while the other two lines may be carrying relatively cool condensate and vapor therefrom from the condensate pot at a relativley low but rising temperature as an empty coke drum is being subjected to preheating.

The process of the present invention has the capability and particularly the necessary flexibility of handling these many variations with improved efficiency and without the difficulties experienced in the past.

Temperature control in the hot tower is essential in order to avoid any free moisture being deposited there in the liquid state at all times when material is being charged to this tower. Inasmuch as the blow down system operates under a pressure of only a few pounds above atmospheric pressure, in theory it is only necessary to maintain temperatures slightly above 212° F. in the charge and circulating scrubbing oil. However, in sizable commercial equipment, it is common for one part of a tower to be 10 or 20° hotter or colder than another location in that tower. Also, in the event of any free moisture being deposited here, a sizable temperature differential or heat potential is necessary for quickly vaporizing that water before it is drawn off in the liquid phase due to the high rate of oil circulation. Accordingly, from a practical standpoint temperatures throughout the hot tower must be substantially above the boiling point of water and it has been found that 250° F. is about the minimum feasible for large scale operations. Temperatures above 300° F. are preferred. The temperature of the material in inlet line **46** is desirably maintained between about 275° F. and 400° F. and it is preferred to hold this material within the range of about 325–400°. Blow down fluids at lower heat levels are, of course, customarily diverted to the cold tower. While the circulating oil may be introduced into the tower at about 250–375° F., a range of about 275–350° is preferred. The gaseous effluent taken overhead in conduit **100** should be kept between 275 and 375° to avoid free water as a liquid on one hand and carry-over of relatively dense hydrocarbons on the other hand, keeping in mind the steam distillation effect, and the preferred temperature is between 300 and 350°.

Maintaining the necessary conditions to avoid stable emulsions in hot blow down tower **42** involves maintaining the necessary temperature conditions under the substantially atmospheric pressure of only 1 or 2 p.s.i.g. or slight-

ly higher therein including the proper control of the temperature and flow rate circulating oil used for absorption. This oil has a plurality of important functions. It serves to keep the specific gravity of the liquid hydrocarbon phase below that of water by dissolving the denser hydrocarbons in the relatively light circulating oil. Also, it reduces the viscosity of the hydrocarbons that are deposited in the tower to the extent that the enriched liquid may be easily circulated by the conventional pumps. Preferably the viscosity of the enriched liquid is kept below 500 SSU measured at 300° F. The oil introduced into the top of the tower controls the temperature throughout the tower in combination with the temperature control of the charged material. The circulating oil also serves as an absorbing or scrubbing medium for removing the denser hydrocarbons from the charge by intimate contact therewith. The circulating oil introduced into the inlet manifold ahead of the tower also serves to at least partially quench or control the inlet temperature of the charge. Spraying the circulating oil into the charge stream in that manifold also performs the function of substantially increasing the effective volumetric capacity of the tower by absorbing substantial quantities of hydrocarbon vapors into the more concentrated liquid stage before reaching the tower thereby greatly reducing the gaseous volume of the feed to the tower.

The density and viscosity of the high density hydrocarbons in the blow down effluent need to be reduced for several reasons. Upon introduction into the light medium or heavy cycle stocks commonly employed in refineries, these heavy materials do not blend readily but settle instead of dissolving and thus tend to plug the equipment with tarry deposits. However, they blend well with the flux oil in the present process as a result of intimate contact therewith at elevated temperature in the absence of free water and especially in the finely divided spray of oil in the inlet manifold to the hot tower. Once dissolved, the high density materials no longer tend to promote emulsions with water and no troubles are encountered in storing the enriched scrubbing oil in a storage tank which also may contain water.

The present invention has been described in connection with certain specific apparatus, operating conditions and schedules for purposes of illustration but it obviously may be employed with many other different types of equipment under varying conditions and operating schedules. Accordingly, it should not be presumed to be restricted in any such aspects except in regard to the recitation of specific limitations in the appended claims.

I claim:

1. In a method for recovery of hydrocarbons from a charge mixture of hydrocarbons of varying densities and moisture supplied during a separation period at varying temperatures, the improvement which comprises separating hydrocarbons of high liquid densities approximating that of water in a first separation zone maintained at a temperature above that at which free water exists in the liquid phase in said first zone, discharging from said first zone a liquid phase containing said high density hydrocarbons and a vapor phase containing moisture and light hydrocarbons, condensing at least a substantial part of said vapor phase and separating vapor from a liquid mixture of water and light hydrocarbons in a second separation zone, directing said charge mixture to said first zone when the temperature of said charge mixture is sufficiently high to prevent condensation of water in said first zone, and directing said charge mixture to said second zone when the temperature of said charge mixture is not sufficiently high to prevent condensation of water in said first zone.

2. A process according to claim 1 in which said charge mixture is supplied in a plurality of individual streams of different compositions, each said stream is directed to said first zone when the temperature of the stream is sufficiently high to prevent condensation of water in said first zone, and each said stream is directed to said second zone when the temperature of the stream is not sufficiently high to prevent condensation of water in said first zone.

3. A process according to claim 1 in which heavier hydrocarbons including hydrocarbons of greater liquid density than water are stripped from said charge mixture in said first zone by absorption in a liquid hydrocarbon absorbent of substantially lower density and the resulting liquid phase is discharged as a substantially homogeneous enriched liquid lighter than water.

4. A process according to claim 1 in which a liquid hydrocarbon absorbent is passed into intimate contact with said charge mixture in said first separation zone to produce said liquid phase containing said high density hydrocarbons and heating said absorbent prior to introduction into said first zone to a temperature sufficient to assist in maintaining said first zone at a temperature sufficiently high to prevent condensation of water therein.

5. In a delayed coker blow down recovery process, the improvement which comprises supplying at least one charge stream containing moisture and hydrocarbons from a coke drum to an absorption-condensation system, passing each said charge stream in which the temperature is not substantially above the boiling point of water under the prevailing conditions as the charge to a condensation operation, separating each said charge stream in which the temperature is substantially above the boiling point of water under the prevailing conditions as the charge for an oil absorption operation, maintaining said oil absorption charge at a temperature below about 400° F., stripping the denser hydrocarbons from said oil absorption charge by absorption in a liquid hydrocarbon absorbent maintained at a temperature substantially above the boiling point of water but not exceeding about 375° F. while the lighter hydrocarbons and free moisture are withdrawn in a gaseous effluent, condensing substantial proportions of normally liquid hydrocarbons and of water vapor from said gaseous effluent and from said condensation charge to produce a liquid phase readily separable into water and said condensed light hydrocarbons, and regulating the flow of the liquid hydrocarbon absorbent in said oil absorption operation to maintain the gaseous effluent from said operation at a temperature substantially above the boiling point of water but not exceeding about 375° F. and to absorb the denser hydrocarbons from the oil absorption charge will maintaining the density of the enriched absorbent substantially lighter than that of water.

6. A process according to claim 5 in which said absorption operation is carried out with a stream of said liquid hydrocarbon absorbent passing downwardly countercurrent to the gaseous phase of said oil absorption charge and said condensation operation is carried out with a stream of water passing downwardly countercurrent to the flow of gaseous material from which water vapor and said lighter hydrocarbons are condensed.

7. A process according to claim 5 in which a portion of said liquid hydrocarbon absorbent is intimately mixed with said oil absorption charge prior to the entry of said charge into an absorption zone in which the gaseous phase of said charge is passed countercurrent to the flow of another portion of said absorbent.

8. A process according to claim 5 in which said oil absorption charge is maintained at a temperature below about 400° F. by at least intermittently quenching said charge with a vaporizable liquid introduced at a temperature substantially below 400° F.

9. A process according to claim 5 in which the flow of liquid hydrocarbon absorbent is regulated to maintain the viscosity of the enriched liquid below about 500 Saybolt seconds Universal measured at 300° F.

10. A process according to claim 5 in which a portion of the enriched liquid hydrocarbon absorbent is withdrawn as a product of the process and the remainder is recirculated to said oil absorption operation after replacing a substantial portion of the liquid withdrawn as product

with a hydrocarbon liquid of about 20 to 40° API gravity and essentially boiling within the range of about 400° F. to 700° F.

11. In a delayed coker blow down recovery process, the improvement which comprises supplying at least one charge stream containing moisture and hydrocarbons from a coke drum to an absorption-condensation system, passing each said charge stream in which the temperature is below about 275° F. to a condensation operation, separating each said charge stream in which the temperature is at least about 275° F. as the charge for an oil absorption operation, maintaining said oil absorption charge at a temperature below about 400° F., stripping the denser hydrocarbons from said oil absorption charge by intimate mixing and absorption in a liquid hydrocarbon absorbent maintained between about 250 and 375° F. while the lighter hydrocarbons and free moisture are withdrawn in a gaseous effluent, condensing water vapor and substantial proportions of the normally liquid hydrocarbons from said gaseous effluent and from each said charge stream in which the temperature is below 275° F. by intimate contact with a liquid maintained at a temperature below the boiling point of water, withdrawing a liquid mixture readily separable into water and said condensed lighter hydrocarbons, and regulating the flow of liquid hydrocarbon absorbent in said oil absorption operation to maintain the gaseous effluent from said operation at a temperature between about 275 and 375° F. and to absorb the denser hydrocarbons from the oil absorption charge while maintaining the density of the enriched absorbent substantially lighter than that of water.

12. In a delayed coker blow down recovery process, the improvement which comprises supplying at least one charge stream containing moisture and hydrocarbons from a coke drum to an absorption-condensation system, passing each said charge stream in which the temperature is below 325° F. to a water condensation operation, separating each said charge stream in which the temperature is at least 325° F. as the charge for an oil absorption operation, maintaining said oil absorption charge at a temperature below about 400° F., stripping the denser hydrocarbons from said oil absorption charge by intimate mixing and absorption in a liquid hydrocarbon absorbent maintained between 275 and 350° F. while the lighter hydrocarbons and substantially all free moisture are withdrawn in a gaseous effluent, condensing a predominant proportion of the normally liquid hydrocarbons from said gaseous effluent and from each said charge stream in which the temperature is below 325° F. by intimate contact with water, withdrawing a liquid mixture readily separable into water and said condensed lighter hydrocarbons, and regulating the flow of liquid hydrocarbon absorbent in said oil absorption operation to maintain the gaseous effluent from said operation at a temperature between 300 and 350° F. and to absorb the denser hydrocarbons from the oil absorption charge while maintaining the density of the enriched liquid substantially lighter than water.

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