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PROCESS FOR HYDROREFINING AND CRACKING GAS OILS TO PRODUCE GASOLINE

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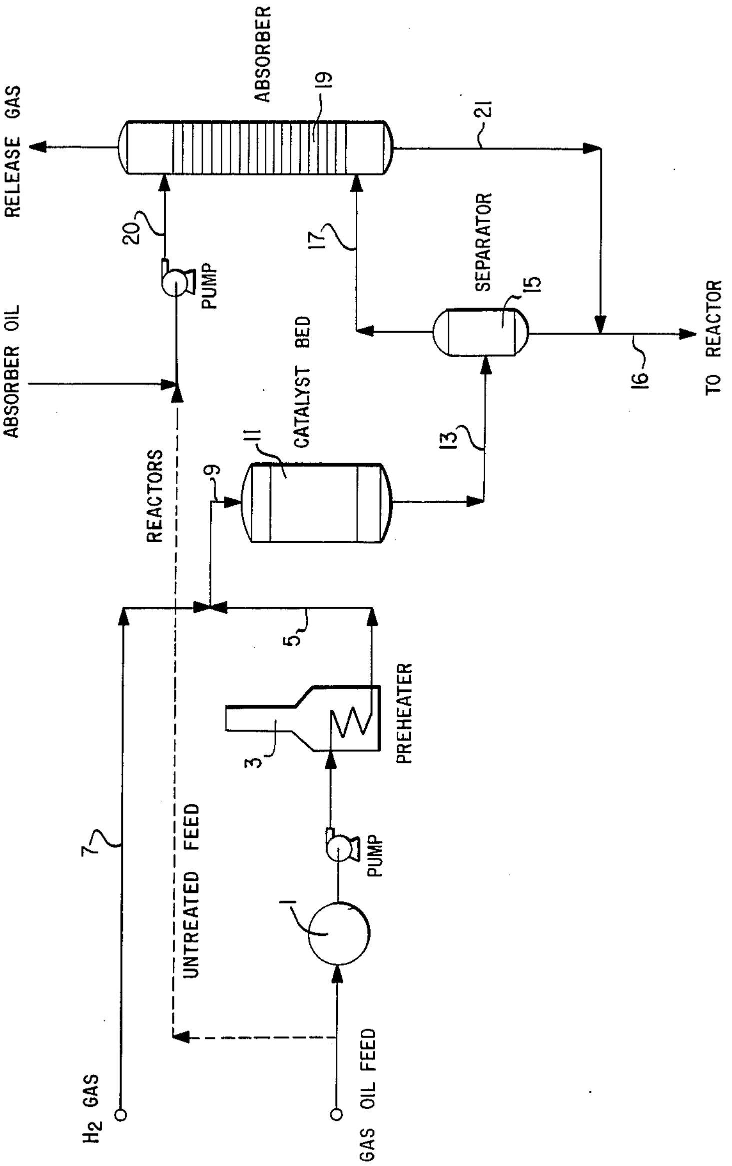

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PROCESS FOR HYDROREFINING AND CRACKING GAS OILS TO PRODUCE GASOLINE

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The present invention relates to a process for hydrotreating and catalytically cracking hydrocarbon feedstocks. More specifically this invention relates to an improved process wherein the hydrotreatment is "integrated" with the cracking facilities.

Catalytic cracking feedstocks such as gas oils ordinarily contain contaminants such as sulfur and organo-metallic compounds as well as coke precursors. The presence of significant amount of sulfur compounds in cracking feeds is undesirable in that the gasoline and other products from the catalytic cracking unit will also contain sulfur. Organo-metallic impurities, usually compounds of metals such as nickel, vanadium, iron, etc., tend to poison the cracking catalysts and thereby increase the cost of the operation by decreasing the useful catalyst life. Coke precursors are disadvantageous in that the coke production they effect reduces conversion levels. Consequently, it has often been proposed to prepare these cracking feedstocks prior to their introduction into cracking units by subjecting them to a hydrotreating operation. Hydrotreating of these feeds is known to reduce their contaminating metal, sulfur content and coke precursors.

The hydrogenation unit operation commonly employed in this preparatory hydrotreating step effects partial heating of the feedstock by indirect heat exchange contact with the hydrogenation reactor effluent, heating the feedstock to reaction temperature in a hydrogenation reactor preheater and then introducing the feedstock into a hydrogenation reactor provided with a sulfur-resistant hydrogenation catalyst. The effluent from the hydrogenation reactor is cooled and flashed, the liquid portion being sent to a stripper which removes the hydrogen sulfide. The vapor portion from the flashing operation is again cooled and collected in a second flash drum, after which the resulting condensate is sent to a stripper overhead receiver where a decrease in pressure releases the dissolved hydrogen sulfide. The overhead from the stripper is cooled and also sent to the overhead receiver. The hydrogenated and stripped liquid effluent products from the stripper are then sent to the cracking unit preheater.

Hydrogenation units of the type described, however, are very expensive and thereby reduce the economic attractiveness of the process. A big disadvantage is the heating requirements involved in these systems, that is, the heating of the feed for hydrogenation, cooling and reheating for cracking. It is evident, therefore, that a process that reduces the capital requirements of these hydrogenation facilities and avoids the expense of cooling and reheating would be in large demand.

It has now been found that integrating the hydrogenation unit operation with a catalytic cracking unit in accordance with the process of the present invention materially reduces the requirements for heat exchanging, cooling, flashing, stripping and heating the various streams in the system. Moreover, cooling and reheating the liquid hydrogenated product is avoided, and as a result, elevated temperatures are maintained throughout the integrated hydrogenation process of the present invention.

The process of the present invention can be best described by the following example with reference to the accompanying drawing wherein is shown diagrammatically an integrated hydrogenation process embodying the present invention.

Referring to the drawing, 37,000 barrels per day of a virgin gas oil from a mixture of 60% Kuwait and 40% North Texas crude oil is heated by indirect heat exchange in heat exchanger **1** with the products from a fluidized catalytic cracking reactor (not shown). The partially heated feedstock is then pumped into the preheater **3**. The virgin gas oil feedstock has the following physical properties:

Gravity, °API	22.8
Percent H	12.24
Percent S	2.22
Percent N	0.11
Carbon residue (Rams)	0.468
ASTM 200 ml. Vac. Dist. IBP	615
50%	839
95%	1028
Metals:	
NiO, p.p.m.	0.29
V_2O_5, p.p.m.	1.36
Cr, p.p.m.	0.17
Fe, p.p.m.	1.3

The feedstock leaves preheater **3** at about 760° F. and is joined with cool hydrogen (as recycle H_2) via line **7** at a rate of 250 s.c.f. of H_2 per barrel of feed. The combined hydrogen-feedstock stream is then introduced by means of line **9** into hydrogenation reactor **11** provided with a fixed bed of cobalt molybdate on alumina catalyst. The catalyst analyzes 2.7% CoO and 11.9% MoO_3 on alumina. The hydrogenation conditions employed in the reactor **11** are as follows:

Temperature, °F.	725
Pressure, p.s.i.g.	500
WHSV	1.0
H_2 rate, 1,000 s.c.f. H_2/bbl. of feed	250

The hydrogenation reactor effluent is withdrawn from reactor **11** and directed by line **13** into a separator **15**, wherein the effluent is separated at approximately reaction temperature and pressure to yield a vapor phase and a liquid phase. The separation temperature is within about 10° F. and the separation pressure is within about 10 p.s.i. of the conditions at the outlet of reactor **11**. The resulting liquid product is analyzed and the yield determined after 116 and 212 hours, respectively, of on stream operation. The results are shown in Table I below.

Table I

Hours on oil		116	212
Yields (based on 100% recovery):			
Total liquid product		97.7	97.4
H_2S		2.0	2.0
NH_3		0.1	0.1
Recovery, Wt. percent		99.1	98.6
H_2 consumption, s.c.f./bbl		175	170
Desulfurization, Wt. percent		83.3	82.4
Denitrogenation, Wt. percent		29.0	29.1
Carbon residue reduction, Wt. percent		56.0	48.0
NiO removal, Wt. percent		>73.1	56.4
V_2O_5 removal, Wt. percent		>92.8	>92.8
Gravity, °API		4.0	3.7
Refractive index		0.0108	0.0107
	Feed		
Total liquid product inspections:			
Gravity, °API	22.8	26.8	26.5
Percent H	12.24	12.59	12.57
Percent S	2.22	0.38	0.40
Percent N	0.11	0.08	0.08
Carbon residue (Rams)	0.468	0.211	0.250
Refractive index at 80° C	1.4884	1.4776	1.4777
Metals by emis. spec.—			
NiO, p.p.m	0.29	<0.08	0.13
V_2O_5, p.p.m	1.36	<0.1	<0.1
Fe	1.30	-------	7.8
Cr	0.17	-------	0.69

The improved liquid product from separator **15** is then sent via line **16** to the cracking reactor (not shown). The vapor phase from the separator **15** is directed via line **17** into absorber **19** where it contacts as it flows upwardly a downward flow of lean gas oil provided the absorber **19** by means of line **20**, which oil absorbs valuable normally liquid hydrocarbon components from the vapors from separator **15**. The temperature of the absorber oil introduced into absorber **19** is about 150° F. or below and the amount of absorber oil employed is that sufficient to absorb the liquid hydrocarbons and cool the gases. Depending on the amount of vapor entering the absorber the temperature and amount of the absorber oil are regulated to provide release gas from the absorber having a low temperature that is, below about 200° F., preferably below about 150° F., and a fat oil, that is absorber oil, containing the absorbed hydrocarbon components, exhibiting a temperature increase over the temperature of the absorber oil introduced of at least about 200° F., preferably at least about 250° F. In the instant case a lean gas oil temperature of 100° F. and an amount of about 5000 barrels per day of lean gas oil are used to absorb 1450 barrels per day of normally liquid hydrocarbons out of the vapor stream. The absorber oil of the present invention is a gas oil which can be provided by the cracking unit fractionator (not shown) as recycle or it can be untreated cracking stock such as untreated or straight run gas oil. Release gas at a temperature of 120° F. and containing H_2S is taken overhead and sent to other processing systems or, since it is rich in hydrogen, employed as recycle to the hydrogenation reactor **11**.

The fat oil at a temperature of 395° F. is withdrawn from absorber **19** through line **21** and combined with the hydrogenated liquid phase from separator **15**, which liquid phase has a temperature substantially that of the hydrocarbon reaction temperature to provide a mixture having a temperature of 695° F. The mixture is then sent via line **16** to the catalytic cracking reactor system. The mixture is combined with hot catalyst from the cracking unit catalyst regenerator to raise the hydrocarbons to reaction temperature of about 900° F.

Thus, the present invention provides a feed suitable for introduction into a catalytic cracking reaction system while maintaining an elevated temperature throughout the hydrogenation operation and without the necessity of reheating the liquid hydrogenated products from the hydrogenation operation. In the case where untreated feedstock is used as the absorber lean oil, the absorber fat oil can be joined with the feedstock being introduced into preheater **3**. In this way, all of the feedstock would be treated but the unit would not be dependent on the production of cycle oil or lighter fractions for absorber lean oil. If desired, a heat exchanger can be provided to indirectly exchange heat from the vapor from separator **15** with the absorber fat oil. Provision of this exchanger has been found to advantageously reduce the overall absorption temperature and provide a better recovery of the light components created in the hydrogenation reactor.

The preheater **3** can be any preheater commonly employed to heat the cracking feedstock to reaction temperatures prior to introduction into the catalytic cracking reactors. The hydrogen can be combined with the feedstock after the feedstock has passed through preheater **3** in which case the amount of hydrogen is regulated so as to form a combined hydrogen-feedback stream having the desired hydrogenation temperature, which temperature is lower than that required to effect substantial cracking, or the hydrogen can be combined with the feedstock prior to the preheater and thus the combined stream heated to the desired temperature.

The hydrogenation reaction conditions employed in the hydrogenation reactor will generally fall in the following ranges: temperature, about 550 to 800° F., preferably about 700 to 750° F.; pressure, about 200 to 1500 p.s.i.g., preferably about 300 to 500 p.s.i.g.; weight hourly space velocity (WHSV), about 0.1 to 10, preferably about 0.5 to 4 and a molecular hydrogen to hydrocarbon feed rate of about 250 to 5000 standard cubic feet (s.c.f.) of hydrogen per barrel of feed. The catalyst employed can be any of the conventional sulfur-resistant hydrogenation catalysts as, for example, cobalt molybdate on alumina. The hydrogenation can be carried out in single or multiple reactors, often a series of 2 to 5 reactors.

The cracking reactor to which the absorber fat oil and the hydrogenated liquid phase from the separator are sent can be of the fixed, moving or fluidized bed type and can employ the cracking conditions and catalysts ordinarily employed in these operations. These conditions include, for instance, a temperature of about 800 to 1100° F., a pressure of about atmospheric to 50 p.s.i.g. and no substantial amount of free or molecular hydrogen is added to the cracking reactor. A particularly preferred solid refractory oxide catalyst is silica-alumina.

It is claimed:

1. A process for hydrorefining and cracking gas oil which comprises heating gas oil to hydrorefining temperature, said temperature being lower than the temperature required to effect substantial hydrocracking, contacting the gas oil and hydrogen with a hydrorefining catalyst under hydrorefining conditions of temperature and pressure to yield a hydrorefined gas oil product, separating the gas oil product at approximately the hydrorefining temperature and pressure to yield a vapor phase and liquid phase, contacting said vapor phase with gas oil to absorb normally liquid hydrocarbons from the vapor phase product, thereby yielding a fat oil, and cracking said liquid phase of the hydrorefined product for conversion to gasoline boiling range materials.

2. The process of claim **1** wherein the gas oil with which the hydrorefined vapor phase is contacted is a straight run gas oil.

3. The process of claim **2** wherein the absorber fat oil obtained is recycled and joined with the gas oil feedstock.

4. The process of claim **1** wherein fat oil resulting from said contacting is combined with the hydrorefined liquid phase sent to the cracking reaction system.

5. The process of claim **1** wherein the gas oil with which the hydrorefined vapor phase is contacted is a recycle gas oil from the cracking reaction and the fat oil resulting from the contacting is combined with the hydrorefined liquid phase sent to the cracking reaction system.

6. The process of claim **1** in which the desired hydrorefining temperature is from about 700 to about 750° F., the pressure is from about 300 to about 1500 p.s.i.g., the hydrogen is present in the range from about 250 to about 5000 standard cubic feet of hydrogen per barrel of gas oil, the hydrorefining catalyst is cobalt-molybdate on alumina and further in which the weight hourly space velocity is from 0.5 to 4 pounds of gas oil per pound of catalyst per hour.

7. A process for hydrorefining and cracking gas oil which comprises heating gas oil to hydrorefining temperature, said temperature being lower than the temperature required to effect substantial hydrocracking, contacting the gas oil and hydrogen with a hydrorefining catalyst under hydrorefining conditions of temperature and pressure to yield a hydrorefined gas oil product, separating the gas oil product at approximately the hydrorefining temperature and pressure to yield a vapor phase and liquid phase, contacting said vapor phase with gas oil at a temperature of up to 150° F. to absorb normally liquid hydrocarbons from the vapor phase product, thereby yielding a fat oil, the temperature and amount of said gas oil being regulated to provide release gas from said absorption having a temperature below about 200° F. and a fat oil exhibiting a temperature increase over the absorber gas oil of at least about 200° F., combining the fat oil with said hydrorefining liquid phase and cracking the combined feed mixture for conversion to gasoline boiling range materials.

8. The process of claim **7**, wherein the gas oil with which the hydrorefined vapor phase is contacted is recycle gas oil from the cracking reaction.

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