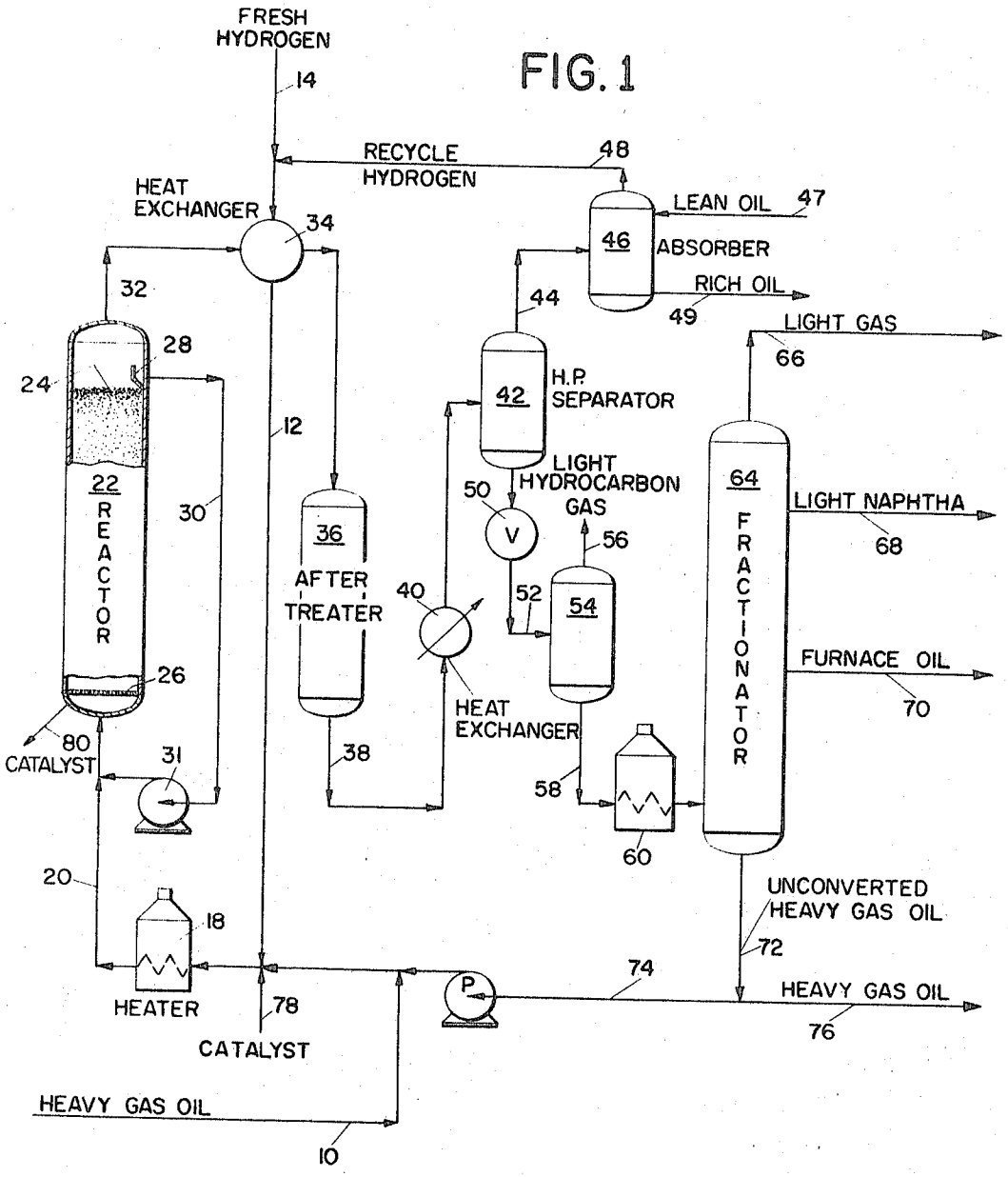
May 30, 1967 M. C. CHERVENAK ETAL 3,322,665

HIGH CONVERSION HYDROGENATION OF HEAVY GAS OIL

Filed May 18, 1965 3 Sheets-Sheet 1

FIG. 1



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HIGH CONVERSION HYDROGENATION OF HEAVY GAS OIL

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FIG. 2

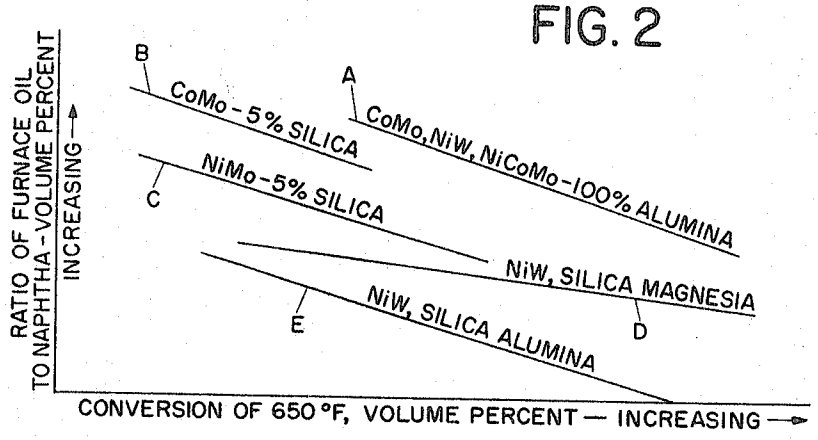
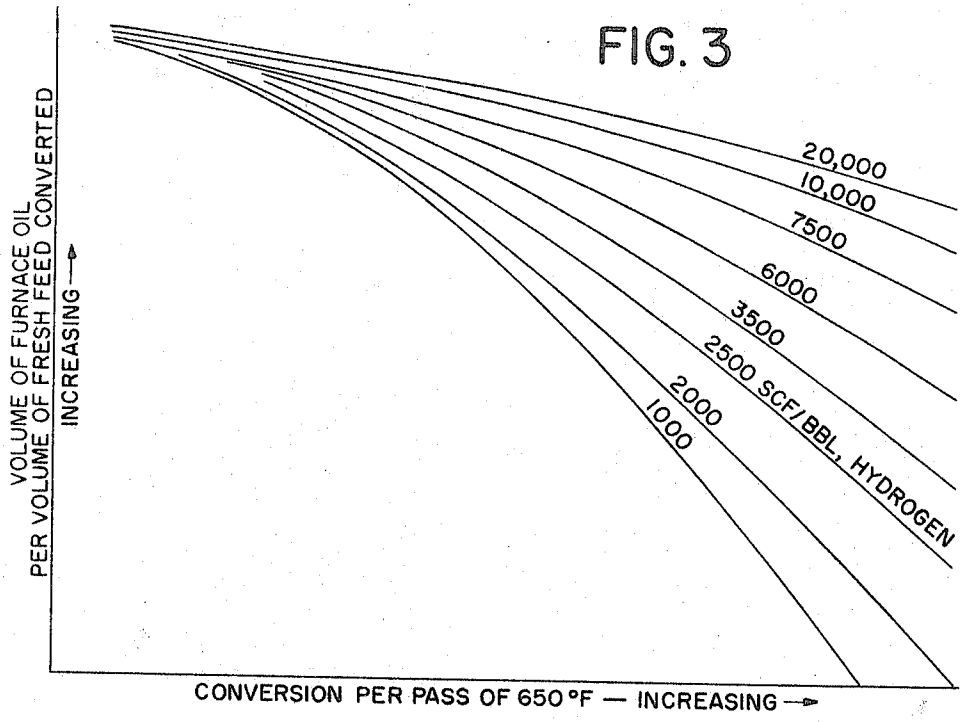


FIG. 3



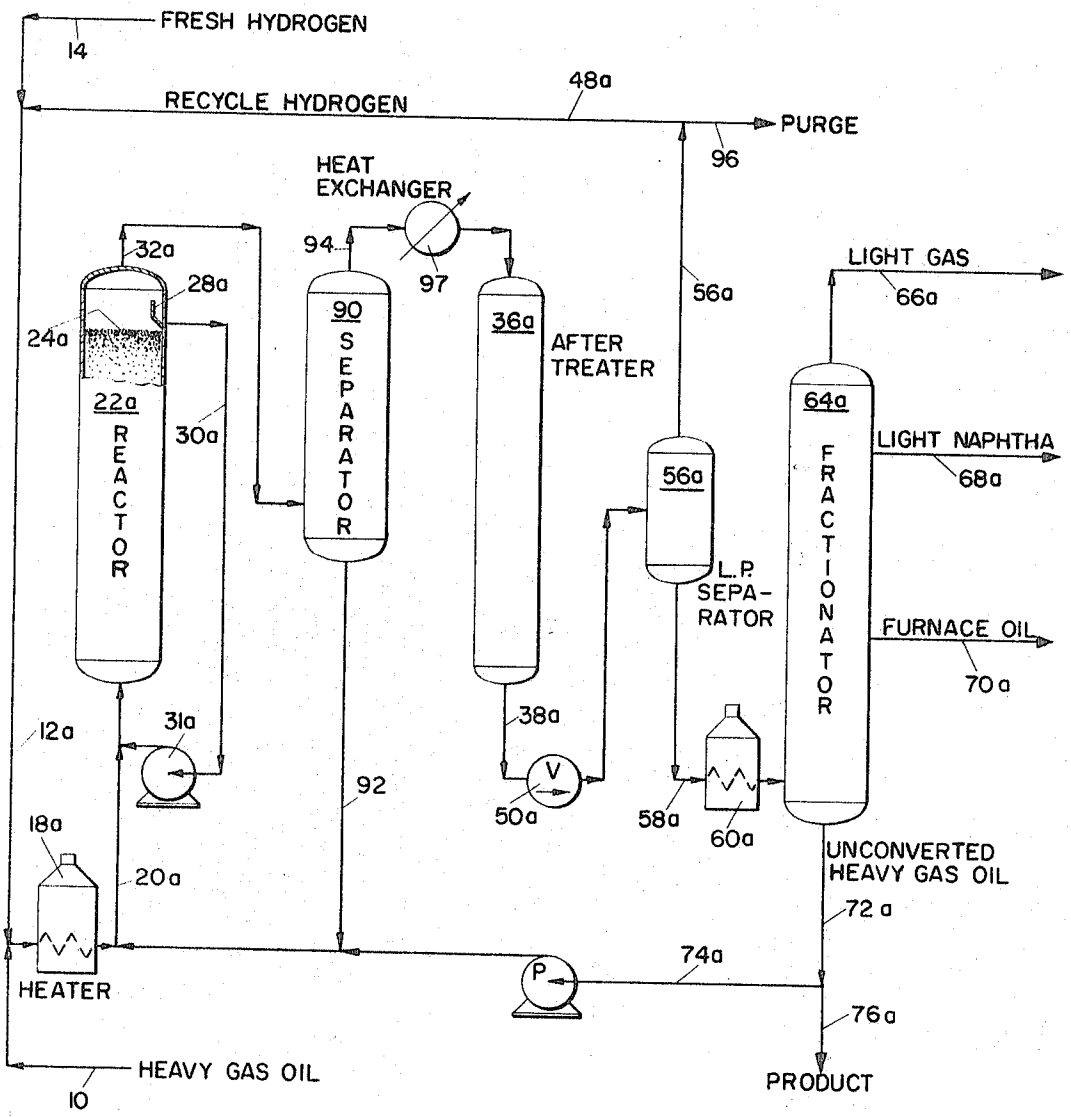
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HIGH CONVERSION HYDROGENATION OF HEAVY GAS OIL

Filed May 18, 1965 3 Sheets-Sheet 3

FIG. 4



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3,322,665

HIGH CONVERSION HYDROGENATION OF HEAVY GAS OIL

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2 Claims. (Cl. 208—97)

This invention relates to improvements in the hydrogenation of heavy gas oils. Our present invention concerns a high conversion hydrogenation operation wherein more than 40% of the charge stock boiling above 650° F. is converted to products boiling below 650° F. More specifically, the purpose of the invention is to produce the highest ratio of furnace oil to naphtha, operating at the above said high conversions. This is accomplished by careful control of catalyst and operating conditions, and by a completely unexpected and unusual effect of using high hydrogen gas rates as will be shown.

The heavy gas oil charge stocks utilized in this invention are fractions containing at least 50 volume percent boiling between about 650° F. to about 1100° F. These fractions are those which can be removed from the crude by subsequent distillation (including distillation under vacuum) after removal of the lower boiling furnace oil and naphtha cuts. In about half of the world, this fraction is either catalytically cracked or hydrocracked to produce primarily naphtha or gasoline. However, in the remaining half of the world, primarily in Europe, the Middle East, and the rest of continental Asia, a surplus of naphtha or gasoline exists and a process is urgently needed to convert such heavy gas oils to furnace oil rather than to naphtha or gasoline.

As is well known to those in the art, furnace oil is the broad middle fraction of an oil comprising kerosene, No. 2 fuel, and diesel fuel; this fraction boils roughly between 400° F. and 650° F., but in some cases, the lower limit of this boiling range may be as low as 300° F. and the higher limit about 800° F. As is similarly well known, naphtha is the lightest normally liquid fraction of an oil and is usually incorporated or reformed into gasoline; this fraction generally has an initial boiling point of about 100° F. and a final boiling point slightly less than 400° F. However, in cases where there is an excellent market for kerosene or furnace oil, a naphtha fraction may be withdrawn with an end point of only about 300° F.

For the purpose of clarity in the presentation of this invention, we shall define naphtha as a fraction boiling between 100° F. and 400° F., furnace oil as the fraction boiling between 400° F. and 650° F., and heavy gas oil as the fraction boiling between 650° F. and about 1100° F., and this invention provides a means of converting heavy gas oil to furnace oil with a minimum yield of naphtha as approximately delineated by the above boiling ranges. However, it should be noted that the invention is applicable and valuable to the aforementioned object if any reasonable boiling ranges are chosen for the naphtha, furnace oil, and heavy gas oil.

As more particularly taught in the Johanson U.S. patent, 2,987,465, hydrogenation of heavy oils is most effectively accomplished in the liquid phase, in an upflowing liquid-gas stream of sufficient velocity to place the catalyst in random motion. The flow of liquid and gas upwardly through the bed of catalyst is such that the catalyst will be expanded at least 10% over the bed volume without fluid flow. In such condition, they are described as "ebullated" in the said Johanson patent. As stated therein, it is relatively a simple matter to operate any particular process so as to cause the mass of catalyst employed to become ebullated, and to calculate the percent expansion of the ebullated mass for any given set of reactor conditions.

In most processes carried out in accordance with this invention, the expanded volume of the ebullated mass will exceed by 10% but by no more than about 100%, the volume of the settled mass.

In order to convert heavy gas oils to produce a maximum yield of furnace oil, it is critical to select the catalyst with care. We have found that many of the catalysts normally applicable to the class of reactions designated as hydrogenation or hydrodesulfurization catalysts are useful for this invention. However, we have also found that the class of catalysts designated as useful for a type of reaction generically designated as hydrocracking is of little value for this invention. Such latter catalysts commonly designated as multi-functional and commonly comprising a metallic sulfide disposed on an acidic support, produce considerably higher yields of naphtha relative to furnace oil than provided by similar contents and compositions of metallic sulfides disposed on weakly acidic or non-acidic supports. Such preferable weakly acidic or non-acidic supports include alumina, bauxite, magnesium, zirconia, and the like. Highly acidic supports which we have found undesirable include conventional cracking catalysts and other composites of silica-alumina, silica-magnesia, silica-alumina-zirconia, acid treated clays, and the like. Unsatisfactory results are likewise obtained with the synthetic metal auminum silicates commonly referred to as "molecular sieves." A preferred support is activated alumina or alumina gel, washed carefully and calcined at at least 1000° F. to minimized acidity. The catalysts will contain one or more metallic components which are deposited on the support, in the active catalytic state, these metallic components will exist as sulfides. Metallic components commonly employed are nickel, cobalt, molybdenum, and tungsten. These are deposited on the support by methods familiar to those skilled in the art. When more than one metal is provided, such as in the case of cobalt with molybdenum, or the case of nickel with tungsten, it is suspected that the metals exert a synergistic effect on each other. Thus, a catalyst containing 1–5% cobalt oxide together with 8–15% molybdenum oxide supported on activated alumina or alumina gel is a preferred catalyst in the process of this invention.

The catalyst particles utilizable in this invention may comprise cylinders, spheres, microspheres, beads and the like. They may be of a particle size broadly between ¼" and 325 mesh (Tyler). However, to obtain a suitable ebullated bed, it is preferable that the catalyst be sized within arrow tolerances. For example, more satisfactory performance will be obtained in one case using a catalyst with a particle size distribution predominantly between 3 and 20 mesh, or in another case, using a catalyst with a particle size distribution between about 100 mesh and 325 mesh, rather than using the total of broad range between 3 mesh (¼") and 325 mesh in any given case.

The reaction zone is maintained at a temperature between about 750° F. and 900° F. These temperatures are considerably higher than those commonly utilized in the process generically known as hydrocracking; the use of such high temperatures is permissible only because of the weakly acidic character of the catalyst provided in this invention. If it were attempted to operate at such high temperatures with a highly acidic hydrocracking catalyst, the charge stock would be overconverted and the catalyst bed completely coked. Thus, the operating temperature range specified in our invention, which we consider an important feature in obtaining high furnace oil yields, is a consequence of the choice of a catalyst support of low acidic activity as described above.

We have found that hydrogen partial pressures in the range of 800 p.s.i. to 2000 p.s.i. are effective to produce the desired results in this invention. A preferred range would be from 1000 p.s.i. to 1800 p.s.i. The correspond-

ing total pressures obviously depend on the purity of the hydrogen utilized but are in the broad range of from 1000 to 2500 p.s.i.g.

The space velocity of the heavy gas oil feed may be varied over the broad range of 0.25 to 5.0 volumes of fresh feed per hour per volume of total reactor space. A preferred range of feed space velocity is from 0.5 to 3.0 volumes of heavy gas oil fresh feed per hour per volume of reactor space.

An unexpected result in the development of this invention is the unusual sensitivity of the yield of furnace oil relative to naphtha with respect to the quantity of hydrogen gas supplied to the reaction zone. As is well known to those familiar with the art, such hydrogen consists of fresh feed hydrogen corresponding approximately in amount to the actual hydrogen consumed in the hydrogenation reaction, together with recycled hydrogen to supply a hydrogen excess to the reaction zone. These hydrogen streams usually contain substantial quantities of light hydrocarbon gases (predominantly methane and ethane) which are obtained from impurities in the fresh feed hydrogen, and because of incomplete separation of hydrogen from the light hydrocarbons produced by the reaction, before recycling such hydrogen back to the reaction zone. It is usually desirable to maintain hydrogen flows to the reaction zone at a minimum sufficient to supply the hydrogen consumed in the reaction zone and to provide an adequate hydrogen partial pressure in the effluent section of the reaction zone. The basic reason for such minimization is the obvious fact that the hydrogen separation and recycling systems become relatively large and expensive if excessive quantities of hydrogen are recycled to the reaction zone.

The establishment of a satisfactory quantity of total hydrogen flow in the reaction zone is complex, and depends to a large extent on the charge stock and on the type of hydrogenation reactions being practiced. In the simple desulfurization of furnace oils, hydrogen flow rates are of the order of 1000 standard cubic feet per barrel of fresh feed. In the deep hydrogenation of light and heavy oils consuming considerably more hydrogen, total hydrogen flow rates may be as high as 5000 standard cubic feet per barrel of fresh feed oil. However, it was completely unexpected to find that using a non-acidic catalyst at conditions such as those described above, that hydrogen rates in excess of 5000 standard cubic feet per barrel improved the yield of furnace oil obtainable from a heavy gas oil to an extent not before thought possible, and that the beneficial effects from such higher total hydrogen rates were sustained at least up to hydrogen rates of up to 20,000 standard cubic feet per barrel. At any total hydrogen rate between 3,000 and 20,000 standard cubic feet per barrel, the benefits obtained were markedly greater than extra attendant costs which would result from the use of such high rates.

Whereas we do not wish to be bound by any theoretical explanation for such an unexpected result of utilizing high hydrogen rates, it is of interest to speculate on such an explanation. Since at a given conversion of a heavy gas oil, the use of high hydrogen rates results in a higher yield of furnace oil relative to naphtha, it is obvious that such high gas rates have not enhanced any catalytic effect in the reaction zone; actually, from the result obtained, it must be concluded that the high gas rates inhibit the further conversion of furnace oil to the undesirable naphtha. Since it is well known that high hydrogen gas rates promote catalytic effects by reducing coke yields on catalysts, the explanation for the overall inhibiting effect of the high hydrogen rates must stem from some other basis than catalytic reaction phenomena. It is reasonable that when high hydrogen rates are used, material in the furnace oil fraction which would otherwise exist as liquid in the reaction zone, becomes vaporized and, in the upflow reaction system employed, passes through the reactor much more rapidly than it would have otherwise in the liquid phase. Thus, this furnace oil in the vapor phase does not obtain the necessary contact time for it to be further cracked to naphtha, and superior furnace oil yields are obtained at a given conversion of heavy gas oil. If this speculation were true, it is apparent that gas other than hydrogen would exert a similar vaporization effect. However, experiments in this direction have shown that although inert gas (such as methane, for example) does promote such a vaporization effect, the loss in hydrogen partial pressure so obtained by using inert gas is so deleterious that the vaporization effect is completely negated. Thus, we prefer to use from 5000 to 20,000 standard cubic feet of hydrogen per barrel of fresh feed heavy gas oil in the practice of this invention; although inerts such as methane may be present (and in all cases probably will be present) with such hydrogen, our specified range is based on the flow of the hydrogen chemical species, rather than the total flow of gas.

In essence then, the distinguishing features of our invention comprise the production of high yields of furnace oil and correspondingly low yields of naphtha from heavy gas oil using a hydrogenation catalyst deposited on a non-acidic or weakly acidic support at temperatures and hydrogen flow rates higher than those conventionally employed, and with the catalyst in a state of random motion as described above.

Since in some uses of our invention the products desired in the furnace oil boiling range must be of premium quality with respect to color, stability, smoke point, diesel index, and other criteria, it is convenient and effective to obtain such premium quality furnace oils by passing the effluent from the reaction zone as described, after cooling said effluent to about 750° F. or lower to a conventional fixed bed hydrogen treating apparatus in series with the reaction system as described. This adjunct to the process of our invention is not costly, since the effluent is fully compressed and no intermediate separations are made before it is passed to the aftertreating step. In addition, the aftertreater reactor is usually very small since a relatively high space velocity can be employed in this operation. Preferred conditions for such an aftertreating step include pressures from 1000 to 2500 p.s.i.g., reaction temperatures from 550 to 750° F., space velocities from 1 to 10 v./hr./v., hydrogen partial pressures from 800 to 2000 p.s.i., and hydrogen flow rates from 5000 to 20,000 cubic feet per barrel. It is obvious that in most cases total pressure, hydrogen partial pressure and hydrogen flow rates in the aftertreating step will be the same as those in the primary conversion. A conventional catalyst composed of nickel or cobalt together with molybdenum supported on alumina is completely satisfactory for the aftertreating step.

Furtrer objects and advantages of our invention will appear from the following description of preferred forms of embodiment thereof when taken with the drawings attached, and in which:

FIGURE 1 is a schematic flow diagram of a preferred form of embodiment of the invention.

FIGURES 2 and 3 are graphs of approximate conversion conditions under specific catalyst and hydrogen conditions, and

FIGURE 4 is a schematic flow diagram of a modified form of embodiment of the invention for a higher conversion of heavy gas oil.

The feed heavy gas oil in FIG. 1 under the necessary pressure in line **10**, is mixed with hydrogen from line **12** obtained from fresh hydrogen **14** and recycle hydrogen as will be described, and passes through heater **18** and through line **20** to the reactor **22**. Reactor **22** is a simple high pressure cylindrical vessel constructed in accordance with standard engineering principles. The reactor is filled with catalyst indicated at **24** and the latter maintained in an ebullating condition by the upflow velocities of liquid and gas passing upwardly through the reactor. A distributor deck **26** is suitably employed to initiate contact of

the liquid and gaseous feeds with the catalyst in the reactor. Above the expanded bed level of the ebullating catalyst, the reactor is provided with a suitable disengaging device 28 whereby part of the liquid effluent from the reaction zone may be recycled in line 30 to the inlet of the reactor without substantial cooling as shown in FIG. 1 by pump 31 and line 20.

The total liquid and vapor issue from the reactor through line 32, are cooled in exchanger 34 and then enter the aftertreating reactor 36. The aftertreating reactor 36 is also a simple cylindrical vessel containing catalyst disposed in a fixed bed. The direction of flow through the reactor in this case may be upflow or downflow. The products leave the aftertreating reactor through line 38 and pass through exchanger 40 into the high pressure separator 42.

The gaseous effluent from the high pressure separator 42 issues through line 44, is cooled, and passes into the absorber-scrubber 46. In the absorber-scrubber, a suitable oil 47, preferably obtained from among the products of the process, is utilized to obtain a rich hydrogen stream which issues through line 48, is combined with a stream of fresh hydrogen from 14 which becomes the hydrogen feed line 12 for reactor 22. In accordance with this invention, the total hydrogen content of the gas in line 12 corresponds to between 5000 and 20,000 standard cubic feet per barrel of fresh feed oil supplied at 10. The rich oil 49 from absorber-scrubber 36 passes to a stripper not shown. The non-volatile effluent from high pressure separator 42 also passes through valve 50 and line 52 to low pressure separator 54. Light hydrocarbon gases are obtained as an overhead product at 56 from the low pressure separator. The liquid products issue from the bottom of the low pressure separator through line 58 and pass through heater 60 into the fractionator 64. The fractionator separates out four streams (a) a light gas product shown issuing at 66, (b) a naphtha product shown issuing at 68, (c) a furnace oil product shown issuing at 70, and (d) unconverted heavy gas oil shown issuing at 72. The unconverted heavy gas oil may be recycled to extinction through line 74, partially recycled through line 74 with the remainder being taken as a product shown at 76, or all taken as a product at 76.

When a premium furnace oil is not desired, the aftertreating step may be omitted with the system then essentially the same as that shown in FIG. 1 with the exception that cooler 34 and aftertreater reactor 36 are omitted.

It is obvious that makeup catalyst may be continuously or periodically added to reactor 22 by using a suitable catchpot not shown. In such a case, a catalyst withdrawal catchpot, also not shown, would be provided. When relatively fine catalyst is employed, small amounts of makeup catalyst may be added with the oil feed at 78 before or after heater 18. If aftertreater 36 is employed, spent catalyst corresponding in quantity to the makeup would be withdrawn as by line 80 through a catchpot not shown from reactor 22. However, if such an aftertreating step were not used, it might be more desirable to allow such small amounts of makeup catalyst (generally of the order of 0.001 to 0.1 pound per barrel) to pass from the top of the reactor with the other products, and ultimately to be withdrawn together with the unconverted heavy gas oil product.

If relatively fine catalyst is employed, inclusion of the hot recycle loop (comprising line 30 and pump 32), might not be necessary, since the feed might provide sufficient velocity to suitably ebullate the catalyst 24 in the reactor 22. However, using catalyst in the coarse limits of the specified range, such a hot liquid recycle stream would probably be required. In the latter case, the recycle loop might be outside of the reactor as shown in FIG. 1, or within the reactor.

It is also obvious that instead of fractionator 64, the product liquids can pass to a simpler vessel in which the heavy gas oil is separated from the remaining products, the heavy gas oil then being recycled or drawn off as shown in FIG. 1, with the remaining products then passing to a separate fractionator for fine fractionation to finished products. Other modifications for recovery of the product and recycle of the required hydrogen are apparent to those skilled in the art and need not be discussed here.

The examples provided below further illustrate the various features and advantages of our invention and may serve to further differentiate the invention from existing art.

Example 1

This example shows the effect of the acidity of the catalyst support on the ratio of furnace oil to naphtha produced in the hydrogenation of a Kuwait heavy virgin gas oil. The data are plotted in FIG. 2.

The charge stock has an API gravity of 22.1, a sulfur content of about 2.63, and boils over the range from 686° F. to 1010° F.

FIG. 2 illustrates the utilization of the operating conditions of this invention on various catalysts. All of the results are obtained at a pressure in the range of 1500 to 1900 p.s.i.g., and with a hydrogen rate of 10,000 standard cubic feet per barrel of fresh feed. The abscissa indicates the extent of conversion of material boiling above 650° F. It is calculated simply by subtracting the volume percent of product boiling above 650° F. from 100. The ordinate indicates the volume ratio of furnace oil relative to naphtha in the products. Furnace oil is taken to be the product boiling between 400 to 650° F., and naphtha the product boiling to 400° F. including butanes.

FIG. 2 illustrates that as conversion of heavy gas oil is increased, the ratio of furnace oil relative to naphtha which is produced decreases for all catalysts. However, more importantly, FIG. 2 shows that as the acidity of the support is increased, the ratio of furnace oil to naphtha markedly decreases. Furthermore, in most cases, this effect transcends effects due to the use of different metallic components on the catalyst.

Thus, as shown in FIG. 2, essentially pure alumina supports (curve A) yield the best selectivity. As the acidity of the catalyst support is increased slightly by the addition of 5% silica (curves B and C), selectivity declines. As the acidity is drastically increased by a support composition consisting of 75% silica and 25% alumina (generally similar to the composition of a cracking catalyst) (curve E) selectivity declines markedly. The silica-magnesia support (curve D) known to be less acid than normal silica-alumina cracking catalysts, but much more acid than 100% alumina by itself, gives results intermediate between those of silica-alumina and 100% alumina.

Actual yields from examples correlated in FIG. 2 further show the marked effect of the acidity of the catalyst support. For example, at an overall heavy gas oil conversion of 60%, the following results are obtained:

	Yields, volume percent	
	Support, 100% Alumina	Support, Silica-Alumina
Naphtha	19.3	31.1
Furnace Oil	44.7	35.9
Heavy Gas Oil	40.0	40.0

Thus, in a 10,000 barrel per day plant the use of a silica-alumina support would produce 3,110 barrels per day of undesirable naphtha with 3590 barrels per day of highly desirable furnace oil, whereas the use of a 100% alumina support would provide only 1930 barrels per day of the undesirable naphtha, and 4470 barrels per day of furnace oil.

Example 2

This example shows the effect of hydrogen gas rate on selectivity to furnace oil relative to naphtha. FIG. 3 shows

the results and is a correlation from a large mass of experimental data.

The charge stock used to obtain the results shown in FIG. 3 is the same Kuwait heavy gas oil used in FIG. 2 and Example 1. The catalysts employed for these runs include cobalt-molybdenum on 100% alumina, and nickel-tungsten on 100% alumina. The catalysts were maintained in an ebullated bed condition for all runs which were 1500–2250 p.s.i.g. and over a reaction temperature range of from 800 to 860° F. Some of the runs were made single pass, with the remainder, including an external recycle of material boiling above 650° F.

The abscissa in FIG. 3 indicates conversion per pass of material boiling above 650° F. Conversion per pass is obtained by dividing the total conversion of material boiling above 650° F. by 1 plus the recycle ratio. The ordinate of FIG. 3 indicates the volume of furnace oil obtained per volume of fresh feed converted expressed as a percentage. (Thus, for example, at 5% conversion per pass, all gas velocities give the result that about 73% of the converted feed will be converted to furnace oil; the remainder is obviously to naphtha and light gases.) The various lines shown in FIG. 3 indicate the magnitude of various hydrogen gas flows used in the experimental work. The results are indicated as standard cubic feet of hydrogen per barrel of fresh feed.

As previously, the furnace oil boiling range has been taken to be between 400° F. and 650° F.

As shown in FIG. 3, as conversion per pass increases, selectivity to furnace oil decreases. This illustrates the beneficial effect of external recycle of material boiling above 650° F. in the conversion practiced herein.

Unfortunately, it is not possible to operate at a low conversion per pass, since the distillation column providing the recycle is a major piece of equipment which is very costly, and because of greatly increased utility costs. Economic studies show that a conversion per pass below 30% is practically prohibitive, and that it is highly desirable to operate above 50% conversion per pass.

FIG. 3 illustrates the large differences in furnace oil selectively obtained at relatively high conversions per pass by varying hydrogen rates. For example, at 60% conversion per pass, selectivity to furnace oil is 61.4% with a hydrogen gas flow of 2500 s.c.f./bbl. of fresh feed, 65.4% with a gas flow of 5000 s.c.f./bbl. and 69.6% with a gas flow of 20,000 s.c.f./bbl.

Thus, results at 60% conversion per pass with 100% total conversion of the heavy gas oil would be:

	Hydrogen Flow Rate, s.c.f./bbl.		
	2,500	5,000	20,000
Naphtha, v. percent	45.5	41.1	36.8
Furnace Oil, v. percent	61.4	65.4	69.6

These results show the marked effect of hydrogen gas rate on selectivity to furnace oil relative to naphtha in the hydrogenation of a heavy gas oil over a non-acidic or weakly acidic supported catalyst.

Example 3

This example illustrates the use of a fixed bed aftertreater in series with the process of this invention.

The base case run was made with Kuwait heavy gas oil at a space velocity of 1.0, a temperature of 850° F., a pressure of 1500 p.s.i.g., and a hydrogen rate of 10,000 standard cubic feet per barrel in an ebullated bed of a catalyst containing cobalt-molybdenum on alumina. The run was made single pass at a conversion of about 80%. The base case run, made without an aftertreater yielded a furnace oil with an API gravity of 33.4, an aniline point of 147° F., and a diesel index of 49.0.

At the conclusion of this run, an aftertreater was added to the process system, in this case a simple cylindrical vessel containing nickel-molybdenum on alumina. Total pressure, hydrogen pressure, and hydrogen flow rate through the aftertreater were the same as in the base case. The space velocity in the aftertreater was 2.5 v./hr./v. and the reaction temperature 750° F. The primary reactor was operated as in the base case run. However, with the aftertreater, a furnace oil was obtained with a gravity of 36.7° API, an aniline point of 154° F., and a diesel index of 56.5. Although the furnace oils obtained in both the base case and the case with the aftertreater are of superior quality, that obtained in the case with the aftertreater is of premium grade and of the very highest quality.

A modified flow diagram as shown in FIG. 4 accomplishes substantially the same results as will be accomplished by the flow diagram of FIG. 1. In this embodiment, all pieces of apparatus that are substantially the same as in FIG. 1 carry the same characters modified by the subscript *a*, i.e., the heavy gas oil at **10** together with hydrogen in line **12***a* enters heater **18***a* prior to entry into reactor **22***a*. The external recycle line **30***a* and pump **31***a* may be internal, as described in the earlier figure.

The effluent at **32***a* passes to separator **90** in this case, with a recycle of heavy ends in line **92**. The overhead **94**, as in the prior case, passes to exchanger **96** to aftertreater **36***a*. The effluent at **38***a* passes through valve **50***a* and passes to low pressure separator **54***a*. The gas overhead from the low pressure separator in line **56***a* is now conducted in part through hydrogen recycle line **48***a* back to the hydrogen feed line **14***a*, and is conducted in part to a purge line **96**.

As in the prior case, the liquid from the low pressure separator **54***a* in line **58***a* passes through heater **60***a* and is fractionated in fractionator **64***a* into a light gas overhead at **66***a*, a light naphtha at **68***a*, furnace oil at **70***a*, and an unconverted heavy gas oil bottoms at **72***a*. This latter may be recycled to extinction in recycle circuit **74***a* or withdrawn as product at **76***a*.

While we have shown preferred forms of embodiment of our invention, we are aware that other modifications may be made within the scope and spirit of our decription herein and of the claims appended hereinafter.

We claim:

1. The method of conversion of a heavy virgin gas oil to produce at least 60 percent (weight) of naphtha and furnace oil wherein the ratio of furnace oil to naphtha is in excess of 2:1 which comprises the steps of:

(a) passing a hydrogen containing gas at a rate of in excess of 5000 s.c.f./bbl., together with the gas oil at temperatures in the range of 750° F. and 900° F. and pressures in the range of 1000 to 2500 p.s.i.g. upwardly through a reaction zone containing a bed of low acidity catalyst at an upward velocity to place the catalyst in random motion in the liquid and for a time to obtain at least a 40% conversion of all gas and oil boiling above 650° F. to a fraction boiling below 650° F.;

(b) hydrotreating the effluent of said step (a) in a separate zone under substantially the temperature and pressure of the said step (a) with the hydrogen from said step (a) and in the presence of a catalyst containing nickel-molybdenum on alumina at a suitable space velocity;

(c) separating a gaseous fraction from said separate zone;

(d) fractionating the net effluent from said separate zone into a premium grade furnace oil, lighter products, and an unconverted heavy gas oil.

2. The method of conversion of heavy virgin gas oil as claimed in claim **1** wherein the effluent from the first mentioned reaction step is separated into a gaseous fraction and a liquid fraction, the liquid fraction is recycled

to the first mentioned reaction step, the gaseous fraction is cooled and then passed to the hydrotreating step (b).

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