DUAL CATALYST SYSTEM

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FIELD OF THE INVENTION

This application is related to a process for hydrocracking of heavy naphtha feeds, more particularly to a process in which aromatics saturation occurs in a separate stage from hydrocracking, by the placement of a low activity aromatics saturation catalyst prior to a high activity conversion catalyst.

BACKGROUND OF THE INVENTION

Heavy naphtha hydrocracking, even in adiabatic situations, has been found to operate with more stability if the aromatics saturation and hydrocracking reactions are decoupled, with aromatics saturation occurring prior to hydrocracking, although both may occur in the same reactor. Attempts to operate this process with only a conversion catalyst (such as high activity Pt/zeolite β) have been unsuccessful. The high heat release from aromatic saturation coupled with the high cracking activity of the conversion catalyst frequently results in reactor runaway. It is difficult to initiate the aromatics saturation reactions without also initiating the cracking reactions. Processes for upgrading naphtha, particularly heavy naphtha, involving both single and multiple steps, are well known in the refining arts.

U.S. Pat. No 5,690,810 (Lawrence et. al.) discloses a single step process to upgrade naphthas to an improved gasoline blending stock. It is an alternative to catalytic reforming, employing milder conditions. Simultaneous saturation of aromatics, paraffin isomerization, and selective cracking of heavier

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hydrocarbons occurs. The catalyst comprises a solid acid support, and at least one zeolite, such as zeolite beta or ZSM-5. The catalyst further comprises a Group VIII metal. In the instant invention, a high activity catalyst is employed, which encourages cracking at the expense of isomerization.

U.S. Pat. No. 4,906,353 (Breckenridge et.al.) discloses a dual-mode hydrocarbon conversion process which comprises reforming a sulfur-, nitrogen- and/or olefin-containing hydrocarbon feedstock, e.g., an FCC gasoline, in a conversion unit which is operated under reforming conditions. A noble metal-containing crystalline silicate having a Constraint Index of not greater than about 2 and a framework SiO_2/Al_2O_3 ratio of at least about 50 is employed as a catalyst to provide a relatively high yield of high octane reformate and a relatively low yield of C_{3-4} hydrocarbons prior to or following hydrocracking the feedstock in the unit. The conversion unit is alternately operated under hydrocracking conditions in the presence of the aforesaid catalyst to provide a relatively low yield of high octane hydrocrackate and a relatively high yield of C_{3-4} hydrocarbons. The latter can be separated from the liquid product and processed in a gas plant to provide LPG products.

U.S. Pat. No. 5,831,139 (Schmidt et.al.) discloses a process combination to selectively upgrade heavy naphtha to more aliphatic gasolines. Such gasolines contain lower concentrations of aromatics and have lower end points with concomitant reduced harmful automotive emissions. This process combination converts the higher-boiling portion of the naphtha, yields isobutane and other isoparaffins which are particularly suitable for upgrading or blending, and reduces cyclics in intermediate processing steps. The heavy naphtha fraction is treated with a solid acid isomerization catalyst. Isobutanes are removed with a separation step, and the remainder of the heavy naphtha feed is treated with a ring cleavage catalyst such as an L-zeolite. In contrast, the instant

invention does not require a separation step, and hydrocracking with a high activity zeolite is employed for upgrading purposes, rather than isomerization with a solid acid catalyst.

U.S. Pat. No. 5,364,514 (Sanborn et.al., hereinafter Sanborn). This invention discloses an integrated process for converting a hydrocarbon feedstock into liquid fuel products. Feedstock is passed to one or more hydrocracking zones to effect decomposition of organic sulfur and/or nitrogen compounds. A portion of the hydrocracked product is passed to an aromatics saturation zone, and is subsequently passed from the hydrocracking zone and the aromatics saturation zone to one or more fractionating zones wherein the products are separated into a top fraction and a bottoms fraction, with the tops fraction being separated into light gasoline, naphtha, jet fuel and diesel fuel products, and a portion or all of the bottoms fraction being recycled to the hydrocracking zone and/or the aromatics saturation zone following the optional removal of heavies and polynuclear aromatics. There are numerous separation steps in this procedure, which are not employed in the instant invention. Furthermore, saturation in Sanborn occurs following hydrocracking, rather than prior to it, as in the instant invention.

U.S. Pat. No. 3,923,641 (Morrison et.al.) discloses a process for hydrocracking naphthas with zeolite beta to produce more isobutanes. There is no preliminary aromatic saturation stage, as in the instant invention, however.

SUMMARY

An unexpected increase in liquid product selectivity has been observed following partial or complete aromatic saturation prior to hydrocracking. The increased selectivity was observed using hydrocracking catalysts loaded with base metals as well as using those loaded with noble metals. Utilization of a dual catalyst system to decouple the aromatics saturation and hydrocracking reactions has been applied successfully over a wide total pressure range (350-1200 psig). The utilization of aromatics saturation catalyst followed by base metal hydrocracking catalyst provides some of the benefits of noble metal hydrocracking catalyst (decreased aromatics content, increased isobutane selectivity, and increased liquid product yield) at reduced catalyst cost. Increased process stability (decreased runaway tendency) is also obtained by decoupling the aromatics saturation and hydrocracking reactions. The high heat release associated with aromatic saturation reactions provides an additional heat source which magnifies any process instability. Decoupling the aromatics saturation and cracking reactions greatly improves the operability of the hydrocracking process.

Utilization of the dual catalyst system (aromatics saturation/cracking) provides an improvement in yields when either noble metal or base metal hydrocracking catalysts are employed. In the case of noble metal catalyst systems an optimized dual catalyst system (noble metal aromatics saturation/base metal hydrocracking combination) may be utilized to obtain the same yield and product property benefits obtained with a complete fill of noble metal catalyst while reducing catalyst cost. Improved process stability was also obtained via decoupling of the aromatic saturation and cracking reactions. Utilization of a dual catalyst system for either improved yields or operability is applicable to fuels and lubes hydrocracking. Fuels hydrocracking generally

involves conversion levels greater than 30%, and lubes hydrocracking generally involves conversion levels below 30%.

BRIEF DESCRIPTION OF THE FIGURES

Figure 1 illustrates product yield v. conversion of feed for single and dual catalyst systems.

Figure 2 illustrates hydrogen consumption v. conversion of feed for single and dual catalyst systems.

Figure 3 illustrates hydrogen consumption v. catalyst system employed.

Figure 4 illustrates the products yields provided v. catalyst system employed.

Figure 5 illustrates volumetric product yield v. catalyst system employed.

Figure 6 illustrates distribution of total liquid product component v. catalyst system employed.

Figure 7 illustrates the aromatics content of the liquid recycle v. catalyst system employed.

Figure 8 illustrates ratio of isoparaffins to normal paraffins v. catalyst system employed.

Figure 9 illustrates the amount of hydrogen consumed v. the percent noble metal catalyst in the reactor.

Figure 10 illustrates the relationship of light naphtha product yield to the percent noble metal catalyst in the reactor.

Figure 11 illustrates the relationship of heavy naphtha product yield to the percent noble metal catalyst in the reactor.

Figure 12 illustrates yields by volume of C_3-C_5 products v. percent noble metal catalyst in reactor.

Figure 13 illustrates yields by volume of C_6+ products v. percent noble metal catalyst in reactor.

Figure 14 illustrates the composition of total liquid product v. percent noble metal catalyst in the reactor.

Figure 15 illustrates the aromatics content of liquid recycle v. percent noble metal catalyst in the reactor.

Figure 16 illustrates the ratio of iso to normal hydrocarbons v. percent noble metal catalyst in the reactor.

Figure 17 illustrates start of cycle activity for different catalyst systems.

DETAILED DESCRIPTION OF THE INVENTION

Feedstocks

The feedstock for the process can be straight-run, thermal or catalytically cracked naphtha. Naphthas derived from shales, tar sands and coal may also be treated. Typically naphthas boil at 25° to 260°C. While the process can accept any naphtha in this boiling range, it generally shows its greatest advantage on feedstocks which boil between 50° and 260°C. Since one of the features of the invented process is saturation of aromatic hydrocarbons, the feedstock naphtha will contain aromatic hydrocarbons, generally from 1 to 40 volume per cent. In order to obtain full benefit from the isomerization and cracking functions of the process the feed will also contain from 5 to 40 per cent n-paraffins. Preferred feedstocks will boil between 70° and 250°C and will contain 5 to 25 percent aromatic hydrocarbons and 10 to 30 percent n-paraffins.

This process converts naphthas to gasoline blending stock by saturating and thus removing benzene and other aromatics. Some paraffin isomerization occurs, also. Higher boiling hydrocarbon components are selectively cracked. The heavier hydrocarbons are converted to gasoline range and lighter components.

Table 1. Heavy Naphtha Feed

Run Number	1	2	3
Nitrogen, ppm	4	0.6	0.5
Sulfur, ppm	3	<20	20
Hydrogen, wt.%	13.65	13.55	13.55
API	42.2	43.6	43.4
Paraffins, wt.%	11.19	13.74	13.57
Naphthenes, wt.%	67.86	69-13	69.13
Aromatics, wt.%	20.94	17.13	17.30
Boiling Range Distribution , °F			
IBP	263	239	238
10	294	281	282
50	340	333	335
90	388	376	380
FBP	414	410	417

Catalysts

In the instant invention, both a catalyst having a aromatics saturation component and a catalyst which comprises an acidic component for cracking are employed. The cracking component is frequently PtPd USY/Al_2O_3, in highly dealuminated form or a form in which the acid sites are exchanged with a counterion. The USY is of very low acidity. The aromatics saturation function is provided by a transition metal or combination of metals. Noble metals of Group VIIIA of the Periodic Table, especially platinum or palladium are preferred. Base metals of Groups IVA, VIA and VIIIA may be used at very high pressures, however. The preferred base metals for use as aromatics saturation components are chromium, molybdenum, tungsten, cobalt and nickel, as well as combinations of metals such as nickel-molybdenum, cobalt-molybdenum, cobalt-nickel, nickel-tungsten, cobalt-nickel-molybdenum and nickel-tungsten-titanium. The acidic component comprises zeolite beta, which is described in U.S. Pat. Nos. 3,303,069 and Re.28,341 and reference is made to these patents for details of this zeolite and its preparation.

U.S. Patent Application Ser. No. 379,421 now abandoned, and its counterpart EU 94,827, disclose the use of zeolite beta in hydrocracking. Zeolite beta, in contrast to conventional hydrocracking catalysts, has the ability to attack paraffins in the feed in preference to the aromatics. The effect of this is to reduce the paraffin content of the unconverted fraction in the effluent from the hydrocracker so that it has a relatively low pour point. By contrast, conventional hydrocracking catalysts such as large pore size amorphous materials and intermediate pore crystalline aluminosilicates, are aromatic selective and tend to remove the aromatics from the hydrocracking feed in preference to the paraffins. This results in a net concentration of high molecular weight, waxy paraffins in the unconverted fraction so that the higher boiling fractions from the

hydrocracker retain a relatively high pour point (because of the high concentration of waxy paraffins) although the viscosity may be reduced (because of the hydrocracking of the aromatics present in the feed). The high pour point in the unconverted fraction has generally meant that the middle distillates from conventional hydrocracking processes are pour point limited rather than end point limited. The specification for products such as light fuel oil (LFO), jet fuel and diesel fuel generally specify a minimum initial boiling point (IBP) for safety reasons but end point limitations usually arise from the necessity of ensuring adequate product fluidity rather than from any actual need for an end point limitation in itself. In addition, the pour point requirements which are imposed effectively impose an end point limitation of about 345°C (about 650°F) with conventional processing techniques because inclusion of higher boiling fractions including significant quantities of paraffins will raise the pour point above the limit set by the specification.

When Zeolite Beta is used as the hydrocracking catalyst, however, the lower pour point of the unconverted fraction enables the end point for the middle distillates to be extended so that the volume of the distillate pool can be increased. Thus, the use of Zeolite Beta as the acidic component of the hydrocracking catalyst effectively increases the yield of the more valuable components by reason of its paraffin selective catalytic properties. A high activity zeolite beta, which possesses a relatively low silica/alumina ratio, in the range from 10 to 200, is desirable for use in the instant invention.

Process Conditions

The processing is carried out under conditions similar to those used for conventional hydrocracking. Process temperatures of 230°C to 500°C (450°F to 930°F) may conveniently be used although temperatures above 425°C (800°F) will normally not be employed as the thermodynamics of the hydrocracking reactions become unfavorable at temperatures above this point. Generally, temperatures of 300°C to 425°C (570°F to 800°F) will be employed. Total pressure is usually in the range of 500 to 20,000 kPa (58 to 2886 psig). The preferred range is from 2514 to 8375 kPa(350 to 1200 psig). The process is operated in the presence of hydrogen and hydrogen partial pressures will normally be from 600 to 6000 kPa (72 to 2305 psig). The ratio of hydrogen to the hydrocarbon feedstock (hydrogen circulation rate) will normally be from 10 to 3500 n.1.1 (56 to 19,660 SCF/bbl). The space velocity of the feedstock will normally be from 0.1 to 20 LHSV, preferably 0.1 to 10 LHSV. At low conversions, the n-paraffins in the feedstock will be converted in preference to the iso-paraffins but at higher conversions under more severe conditions the iso-paraffins will also be converted. The product is low in fractions boiling below 150°C (about 300°F) and in most cases the product will have a boiling range of 150° to 340°C (about 300° to 650°F).

The conversion may be conducted by contacting the feedstock with a fixed stationary bed of catalyst, a fixed fluidized bed or with a transport bed. A simple configuration is a trickle-bed operation in which the feed is allowed to trickle through a stationary fixed bed. With such a configuration, it is desirable to initiate the reaction with fresh catalyst at a moderate temperature which is raised as the catalyst ages, in order to maintain catalytic activity. The catalyst may be regenerated by contact at elevated temperature with hydrogen gas, for example. or by burning in air or other oxygen-containing gas.

A preliminary hydrotreating step to remove nitrogen and sulfur and to saturate aromatics to naphthenes without substantial boiling range conversion will usually improve catalyst performance and permit lower temperatures, higher space velocities, lower pressures or combinations of these conditions to be employed. The initial hydrotreating step is operated at a lower temperature range (200-500°F) than the hydrocracking steps which follow it. The temperature of the aromatics saturation catalyst would be dependent on the aromatic content of the feed (heat release achieved) as well as the temperature required to initiate the saturation reactions. The bed of aromatics saturation catalyst would be at a lower temperature than the beds of hydrocracking catalyst.

The hydrocracking may be operated either in a naphtha directing mode under conditions of moderate to high severity or under conditions of low to moderate severity to produce a relatively higher proportion of product boiling in the middle distillate range.

DATA

An unexpected selectivity advantage was observed when the aromatics saturation and hydrocracking reactions in the proprietary low pressure heavy naphtha hydrocracking process and in conventional second stage hydrocracking operation were decoupled.

The evaluation of this catalyst strategy in low pressure heavy naphtha hydrocracking required comparing data collected using an adiabatic pilot plant with data collected using an isothermal pilot plant. The heavy naphtha feedstock used in this evaluation is described in Table 1, above. The catalysts employed in the adiabatic pilot plant include a low cracking activity

aromatic saturation catalyst which comprises Pt/Pd USY/Al_2O_3 and a catalyst for heavy naphtha conversion which comprises Pt/zeolite beta/Al_2O_3. The process conditions for the adiabatic pilot plant evaluation of the dual catalyst system were 335 psig, 1.5 WHSV (total catalyst), and 4100 SCF/bbl.

The isothermal pilot plant employed a single catalyst system. The conversion catalyst was nominally 0.6 wt.% Pt, 65 wt.% zeolite beta and 35 wt.% alumina. The alpha prior to metals addition was 350 and the SiO_2/Al_2O_3 ratio was approximately 50. The process conditions for the isothermal pilot plant evaluation of the single catalyst system were 350 psig, 2 WHSV, and 3100 SCF/bbl.

The yields with the dual catalyst system relative to the single catalyst system are shown in Figure 1 and the hydrogen consumption is shown in Figure 2. Figure 1 shows that the light gas (C_1-C_4) production with the dual catalyst system is decreased relative to the single catalyst system, thereby resulting in an increase in C_5-300°F product. Similarly, Figure 2 shows that the hydrogen consumption of the dual catalyst system is decreased relative to the single catalyst system at the same 300°F^+ conversion.

The adiabatic pilot plant evaluation of heavy naphtha hydrocracking required decoupling the aromatics saturation and hydrocracking reactions for stable operation. Attempts to operate this process with only the conversion catalyst (Pt/zeolite beta) were unsuccessful. The high heat release from aromatic saturation coupled with the high cracking activity of the conversion catalyst (Pt/zeolite beta) resulted in reactor runaway. It was not possible to initiate the aromatics saturation reactions without also initiating the cracking reactions. After the aromatics saturation and hydrocracking reactions

were decoupled, via placement of a aromatics saturation catalyst with low acid activity before the high activity conversion catalyst, stable operation with 300°F$^+$ conversion up to 98% (single pass) was observed.

Evaluation of the dual catalyst strategy for conventional second stage hydrocracking required that both catalysts be placed in the same reactor, therefore, both catalyst systems had to be operated at the same temperature. Typically the aromatic saturation catalyst would be operated in a lower temperature range; however, due to the equipment limitations of the recycle-to-extinction pilot plant this could not done. The process conditions for the application of this concept in second stage hydrocracking application were 1200 psig, 1 LHSV (total feed), 5000 SCF/bbl, and 400°F recyle cut point. The properties of the feedstock for these evaluations are given in Table 2.

Table 2. Stripper Downcomer Feed

Run	1	2	3
Nitrogen, ppm	0.3	<0.05	<0.5
Sulfur, ppm	24	20	21
Hydrogen, wt.%	12.8	12.78	12.75
API	31.52	31.54	31.5
Gravity, g/mi	0.8611	0.8633	0.8635
Boiling Range Distribution, F			
IBP	204	207	206
10	403	401	400
50	536	536	535
90	653	656	655
FBP	766	770	770

Comparisons are demonstrated in Figures 3-8 of the performance of a base metal hydrocracking catalyst system, NiMo/USY/Al_2O_3(System 1), a dual catalyst system, System 2 [20% aromatics saturation (Pt/Pd low acidity USY/Al_2O_3/80% base metal hydrocracking catalyst) and a noble metal hydrocracking catalyst system, Pd/USY/Al_2O_3, System 3. Shown in Figures 3-8 are comparisons of the hydrogen consumption, product yield (wt.% and vol.%), Total Liquid Product (TLP) component distribution, liquid recycle aromatic concentration, and C_4-C_5 iso/normal ratios, respectively. It was assumed that the increased activity of the aromatics saturation component (noble metal) was the major contributor to the observed differences in performance between the various catalyst systems. To analyze these effects the observed differences were evaluated as a function of the percent of noble metal

aromatics saturation component in the catalyst system (System 1 = 0, System 2 = 20%, System 3 = 100%) and the results are shown in Figures 9-16, respectively.

The dual catalyst system, System 2, exhibits increased hydrogen consumption relative to the base metal catalyst system (Figure 3) as expected due to the incorporation of 20% noble metal aromatics saturation catalyst. The increased hydrogen consumption of the dual catalyst system is greater than if it is assumed that the hydrogen consumption is linear between 0 and 100% noble metal aromatics saturation component (Figure 9). This suggests that with optimization of the ratio of aromatics saturation to conversion catalyst, complete aromatic saturation could be obtained with a dual catalyst system having an aromatics saturation catalyst which contains a smaller percentage of noble metal than is found on a commercial noble metal hydrocracking catalyst. It would therefore be less expensive to employ a dual catalyst system in a hydrocracking process than a single hydrocracking catalyst which is loaded with noble metals.

As seen in Figure 4, use of System 2 as opposed to System 1 resulted in decreased C_5- and light naphtha production while the amount of heavy naphtha produced relative to the base metal hydrocracking catalyst employed was increased. These yield shifts are in the same direction as the yield shift observed with the noble metal relative to base metal hydrocracking catalyst. If it is assumed that the yield shift is dependent primarily on the aromatics saturation function we observed larger shifts than would be predicted for increasing the activity of 20% of the aromatics saturation component (Figure 10-11).

Slightly higher C_3-C_5 volumetric yields were observed using System 2 as opposed to using System 1. This is inconsistent with what has been observed for noble metal hydrocracking catalyst relative to base metal hydrocracking catalyst. When comparing noble metal hydrocracking catalyst to base metal hydrocracking catalyst, less light gas (C_3-C_5 volumetric yield) production is generally observed with the noble metal HDC catalyst. Surprisingly for System 2, which employs a noble metal catalyst in combination with a base metal catalyst, more C_3-C_5 volumetric yield was observed than was observed for System 1, which employed the base metal hydrocracking catalyst alone.

As seen in Figure 5, however, it is only the C_4 volumetric yield which is increased for the dual catalyst system relative to the base metal hydrocracking catalyst used alone. Butanes (particularly isobutane) have much higher value to the refinery than other light gases; therefore, the observed increase might be beneficial (see Figure 8). The C_3 and C_5 volumetric yields are decreased for the dual catalyst system relative to the base metal catalyst system, which is consistent with the trends observed for noble metal hydrocracking catalyst relative to base metal. The C_6+ volumetric yield observed with the dual catalyst system is greater than that observed with the base metal hydrocracking catalyst alone. This is directionally consistent with the noble metal hydrocracking catalyst yield shift. If the argument used previously is applied to this data, a larger increase in C_6+ yield is observed than would have been predicted based on a 20% increase in the aromatics saturation activity of the catalyst (Figure 13).

Further indication of the benefits of System 2 is observed in Figure 6 which evaluates the component distribution of the total liquid product (TLP). The dual catalyst approach of System 2 exhibits increased i-paraffin and naphthene content and decreased aromatic content in TLP, relative to the TLP content when base metal hydrocracking catalyst (System 1) is used alone. These changes are directionally consistent with the changes in component distribution observed as the amount of noble metal aromatics saturation catalyst increases relative to the amount of base metal hydrocracking catalyst. As observed previously these changes are greater than would have been predicted for increasing the aromatics saturation activity of 20% of the catalyst (Figure 14).

Similarly, the aromatic concentration of the liquid recycle (Figure 7) was evaluated for the various catalyst systems. The aromatics content of the liquid recycle from System 2 is intermediate between that produced with the base metal hydrocracking catalyst alone (System 1) and noble metal hydrocracking catalyst (System 3). The reduction observed with System 2 is greater than would have been predicted for increasing the aromatics saturation activity of 20% of the catalyst system (Figure 15). The decreased aromatic concentration of the liquid recycle stream indicates that there should be an improvement in catalyst deactivation rate for System 2 relative to the deactivation rate when the base metal catalyst is used alone (System 1).

The iso/normal (i/n) ratio of the C_4 and C_5 hydrocarbons produced with the various catalyst systems (Figure 8) was evaluated. Surprisingly, the greatest C_4 and C_5 i/n ratio was observed for System 2. The increased iso/n ratio of the C_4 and C_5 hydrocarbons increases the value of these products to the refinery. Based on the severity of the process with the various catalysts, the

largest iso/n ratios had been expected for the noble metal hydrocracking catalyst employed in System 3. As shown in Figure 16, the iso/n ratio of the C_4 and C_5 hydrocarbons is significantly greater than would have been predicted by use of a aromatics saturation catalyst comprising 20% noble metal.

Figure 17 shows the normalized average reactor temperature required for 60% per pass conversion in recycle-to-extinction operation with Systems 1, 2, and 3. Utilization of 20% aromatic saturation catalyst resulted in a 40% increase in conversion activity (System 2) relative to the base metal hydrocracking catalyst used alone (System 1). This increased activity may be of value to the refiner due to increased cycle length and decreased energy costs.

CLAIMS:

1. A process for hydrocracking a hydrocarbon feedstock, whereby process stability is promoted and liquid product selectivity is increased, by contacting said feed stock under superatmospheric hydrogen partial pressures, a hydrogen circulation rate from 56 to 19,660 SCF/bbl, and a space velocity from 0.1 to 20 LHSV, at a temperature in the range from 230 to 500°C, with hydrotreating catalyst, followed by a dual catalyst system, said catalyst system comprising an aromatics saturation catalyst, followed by a hydrocracking catalyst.

2. The process of claim 1, wherein both aromatics saturation and hydrocracking catalysts are placed in the same reactor, but in separate fixed beds.

3. The process of claim 1, wherein the aromatics saturation catalyst of the dual catalyst system comprises at least one noble metal and the hydrocracking catalyst comprises at least one base metal.

4. The process of claim 3, wherein the aromatics saturation catalys comprises from 20% to 100 wt.% noble metal.

5. The process of claim 1, whereby the percentage of conversion of the feed by the dual catalyst system is greater than 30 wt.%, thereby enhancing fuels product selectivity.

6. The process of claim 1, whereby the catalysts of the catalyst system are placed in separate fixed beds, whereby the bed or beds containing aromatics saturation catalyst is operated at a lower temperature than the bed or beds containing hydrocracking catalyst.

7. The process of claim 1, wherein the feedstock is a naphtha feedstock which boils between 70° and 250°C and contains from 5 to 25 percent aromatic hydrocarbons and from 5 to 50 percent n-paraffins.

8. The process of claim 1, wherein the hydrocracking catalyst comprises zeolite beta.

9. The process of claim 1, wherein the hydrocracking catalyst possesses a silica/alumina ratio in the range from 10 to 200.

10. The process of claim 3, wherein the noble metal of the aromatics saturation catalyst is selected from the group consisting of platinum and palladium.

11. The process of claim 1, wherein the hydroben partial pressure is in the range from 72 to 2305 psig.

12. The process of claim 8, whereby the alpha value of the zeolite catalyst is in the range from 20-400.

13. The process of claim 6 wherein the bed or beds containing aromatices saturaiton catalyst would be operated in the range from 250-550°F and the bed or beds containing hydrocracking catalyst would be operated in the range from 550-800°F.