

Description

The present invention relates to a hydrogenation catalyst comprising one or more Group VIII noble metals, preferably platinum and/or palladium, supported on a modified zeolite Y. The invention further comprises the use of the catalyst for hydrogenating hydrocarbon feedstocks, particularly for hydrogenating the aromatics in such feedstocks to saturated compounds.

The need for relatively non-aromatic liquid hydrocarbons boiling in the range of about 125 °C to about 625 °C (e.g., aviation turbine fuels, diesel fuels, flashed distillates, solvents, medicinal and other white oils) continues to increase. These products usually, or at least often, contain relatively high percentages of olefins and monocyclic and polycyclic aromatics so that further hydrogenation is desirable to produce higher valued products. The presence of aromatics in diesel fuels may contribute to air pollution. The hydrofinishing of these types of hydrocarbons to reduce unsaturates is becoming ever more important commercially.

Diesel fuels, which typically have a boiling range between about 200 °C to about 360 °C, are under ever increasing environmental pressure to have their aromatics levels decreased. The presence of aromatics in diesel fuel can also significantly reduce the quality of the fuel. The accepted measure of diesel fuel quality is the Cetane Number ("C.N.") which is directly related to the types of molecules that are found in the fuel. Running a diesel engine on a lower cetane fuel than it was designed for will upset the combustion process due to a longer ignition delay period. This will result in poorer cold starting performance and higher levels of noise, smoke and legislated emissions. The cetane number of molecules within a class (e.g., normal paraffins) increases with the number of carbon atoms in the molecule. Molecular classes may be ranked in terms of their cetane number for a specific carbon number; normal paraffins have the highest cetane number followed by normal olefins, isoparaffins and lastly by monocyclic naphthenes. Aromatic molecules have the lowest cetane number. For example, naphthalene has a C.N. of about -15; tetrahydronaphthalene (tetralin) has a C.N. of about 7; decahydronaphthalene (decalin) has a C.N. of 48; butylcyclohexane has a C.N. of about 50; and decane has a C.N. of 76. Thus, a process which would saturate the aromatics in diesel fuel and also convert the resulting saturates to acyclics without significant hydrocracking would improve the quality of the diesel fuel with regard to environmental quality as well as efficiency.

The hydrogenation of unsaturated hydrocarbons, particularly aromatic hydrocarbons, to corresponding saturated hydrocarbons using a catalyst comprising platinum and/or palladium supported on alumina is disclosed in U.S. patent no. 3,943,053. An improvement over this catalyst can be found in European Patent Application no. 303,332, published February 15, 1989, wherein as a support is utilized a Y-type of zeolite which has been modified to have a higher silica to alumina ratio and a smaller unit cell size than an unmodified zeolite Y. Similar modified Y-type zeolites are known to be used in hydrocracking operations. See, for example, U.S. patent no. 4,401,556, issued August 30, 1983. The problem in utilizing the modified Y-type zeolite as in European Patent Application No. 303,332, is that some hydrocracking activity is present which can hydrocrack the hydrocarbon feedstocks being hydrofinished, resulting in a reduced liquid yield.

A primary object of this invention is to provide a catalyst and process for hydrogenating petroleum fractions containing aromatics as well as olefins while concomitantly minimizing hydrocracking.

Another object of this invention is to provide a catalyst and process for hydrogenating the aromatics and olefins in diesel fuel stocks, converting the saturated aromatics at least in part to acyclic hydrocarbons while concomitantly minimizing hydrocracking thereby increasing the cetane number of the diesel fuel stock.

Another object of this invention is to provide a catalyst for the second stage of a low pressure two stage process wherein a sulphur- and nitrogen-containing diesel boiling range hydrocarbon feedstock is hydrotreated in the first stage, the product of the first stage is separated from the gaseous phase and passed to a second stage wherein it is hydrogenated to reduce the aromatics and olefins, producing additional acyclic hydrocarbons therefrom with minimum hydrocracking, resulting in a product with an increased cetane number.

The present invention therefore relates to a catalyst composition suitable for hydrogenating aromatics and olefins in hydrocarbon feedstocks boiling between about 125 °C and about 625 °C. The catalyst composition comprises one or more Group VIII noble metal(s) on a support wherein the support comprises a zeolite prepared by contacting a starting Y-type zeolite having a unit cell size less than 24.65 angstroms, a silica to alumina molar ratio greater than 5, and a sum of alkali and alkaline earth metal content of less than 0.3 percent by weight (basis metal) with at least one aqueous solution comprising alkali metal ions and/or alkaline earth metal ions under conditions sufficient to provide an increased sum of final alkali and alkaline earth metal content (measured in terms of gram equivalent weight per gram of zeolite), which sum of final alkali and alkaline earth metal content is at least 1.5 times greater than the sum of alkali and alkaline

earth metal content of the starting zeolite (measured in terms of gram equivalent weight per gram of zeolite). Preferably the Group VIII noble metal comprises platinum and/or palladium and most preferably comprises a combination of platinum and palladium. The instant invention also comprises the use of the instant catalysts to hydrogenate hydrocarbons boiling in the range of 125-625 °C, and particularly with contracted diesel range hydrocarbon boiling in the range of 125-425 °C, in order to saturate substantial portions of the aromatics and olefins.

The term "hydrocracking" as used herein refers to a reaction wherein a hydrocarbon molecule is broken at carbon-carbon bonds into two or more lower molecular weight hydrocarbon molecules with the concomitant addition of hydrogen to the molecules. Hydrocracking does not include the ring opening of a cyclic hydrocarbon molecule that is not accompanied by fragmentation into two or more molecules of lower molecular weight. The term "alkali(ne earth) metal" as used herein refers to a metal selected from the group consisting of alkali metal, alkaline earth metal and mixtures thereof, that is, it refers to alkali metal and/or alkaline earth metal and includes one or more alkali metals, one or more alkaline earth metals and two or more of a mixture of alkali metal(s) and alkaline earth metal(s). The term "ion" refers to an electrically charged moiety. It is understood that ions do not exist in vacuo, but are found in combination with charge-balancing counter ions. The term "alkali metal" is used herein as a descriptor of the elements of Group IA of the Periodic Table of the Elements (Li, Na, K, Rb, Cs.) . The term "alkaline earth metal" is used herein as a descriptor of the elements of Group IIA of the Periodic Table of the Elements (Be, Mg, Ca, Sr, Ba). Alkali(ne-earth) metal herein does not refer to the element in the metallic or zero valent state, but rather is a shorthand use for the element in the positive valent state. The gram equivalent weight of alkali(ne earth) metal per gram of catalyst is determined by dividing the weight of alkali(ne earth) metal in a gram of catalyst by the atomic weight of the alkali(ne earth) metal.

The term "basis metal" refers to the definition of the alkali(ne earth) metal content rather than that of the alkali(ne earth) metal oxides.

An important aspect of the instant invention is the use as supports of Y-type zeolites having relatively low alkali metal contents, say less than about 0.3, preferably less than about 0.15 percent by weight basis metal and which have been ion exchanged to increase their alkali(ne earth) metal content. "Y-type zeolites" are zeolites which have the same general crystal structure as zeolite Y but which have contracted unit cells when compared to zeolite Y. The Y-type zeolite is frequently referred to in the art as simply Y zeolite or zeolite Y. These zeolites having contracted unit cell sizes are also known as ultrastable or ultrastabilized Y zeolites or USY zeolites. Irrespective of the terminology applied in the art, the zeolites used to prepare the instant catalysts are zeolites having the general crystal structure of zeolite Y and which are specifically defined by their unit cell size, silica to alumina molar ratio and alkali(ne earth) metal content.

The zeolitic materials which can be used as starting materials to produce the supports used to prepare the catalyst composition of the present invention comprise readily available Y-type zeolites such as zeolite Y, ultra-stable zeolite Y and very ultra-stable zeolite Y which have been modified by using processes known in the art to produce the base materials having the required unit cell size dimension together with the required silica to alumina molar ratios and low alkali(ne earth) metal content. Such modifications of unit cell size and silica to alumina molar ratio also necessarily produce zeolites having low alkali(ne earth) metal contents. Suitable modification processes comprise ion-exchange techniques, say one or more ion exchange steps with ammonium compounds, followed by one or more calcination stages, optionally in the presence of steam. See for example U.S. Patent No. 3,929,672, issued December 30, 1975. Normally, Y-type zeolites already partially modified are subjected to a so-called dealumination technique to reduce the amount of alumina present in the system. Dealumination techniques are described extensively in the art and comprise inter alia the use of acid extraction, the use of silicon halides or other suitable chemical treating agents, chelates as well as the use of chlorine or chlorine-containing gases at high temperatures. Good results have been obtained using materials which have been subjected to one or more acid-leaching procedures, but other techniques can be applied as well.

The starting zeolite should be an Y-type zeolite having a unit cell size less than 24.65 angstroms, a silica to alumina molar ratio of greater than 5 and an alkali(ne earth) metal content of less than 0.3 percent by weight basis metal.

In a preferred embodiment the unit cell size of the starting zeolite should be less than 24.4 angstroms, preferably less than 24.35 angstroms and more preferable less than 24.30 angstroms. More preferably the unit cell size will range between 24.2 and 24.3 angstroms, and most preferably between 24.22 and 24.28 angstroms. The silica to alumina molar ratio of the preferred zeolite should be greater than 25, more preferably greater than 35, even more preferably greater than 50 and most preferably greater than 60. The processes used to dealuminate zeolites to obtain the high silica to alumina molar ratios of the starting zeolites result in zeolites having alkali(ne earth) metal contents that are relatively low compared to zeolite Y.

The alkali(ne earth) metal contents of the preferred starting zeolites are less than 0.15, preferably less than 0.075 and more preferably less than 0.04 percent by weight of the zeolite basis the alkali(ne earth) metal. Preferably, the starting Y-type zeolite comprises an alkali metal. While the alkali metal can be any of the alkali metals from lithium through cesium, sodium is preferred with potassium being the second most preferred. The alkali metal may be a mixture of two or more different alkali metals. The commercial manufacture of zeolite Y normally results in a sodium-exchanged zeolite Y. Using different manufacturing conditions, Y zeolites containing alkaline earth metal ions could be prepared in which case the alkaline earth metals would be considered equivalent to alkali metal ions when the valence difference is considered.

The starting zeolites are contacted with one or more solutions, preferably aqueous solutions, comprising one or more alkali(ne earth) metal ions. These solutions are prepared by dissolving one or more dissolvable alkali(ne earth) metal salts in a suitable solvent. The alkali(ne earth) metal salts that are suitable are any salts that can be dissolved in a suitable impregnating solution. Illustrative but non-limiting examples of suitable salts are alkali(ne earth) metal bicarbonates, carbonates, chlorates, perchlorates, cyanides, hydroxides, iodates, nitrates, nitrites, sulphates, hydrogen sulphates, sulphites, dithionates, thiosulphates, alkoxides, carboxylates, sulphonates, halides and the like. Preferred salts for use when water is the solvent include the nitrates, halides, sulphates and lower carboxylates. The contact with the solution of alkali(ne earth) metal ions can occur at any point in the catalyst manufacturing process. It can be applied to a catalyst article after Group VIII metals incorporation or to a catalyst precursor pellet prior to Group VIII metals incorporation or to the zeolite prior to being formed into a catalyst (precursor) pellet. Preferably it is applied to the zeolite simultaneously with the application of the catalytic metals.

The contact of the zeolite with the solution of alkali(ne earth) metal ions encompasses ion exchange, impregnation and mixtures thereof. Ion exchange typically encompasses contacting the zeolite with one or more solutions of metal ions under conditions which allow ion exchange, followed by washing of the zeolite to remove unexchanged ions. Impregnation encompasses contacting the zeolite with a solution of metals ions and not following with a washing step. Thus, unexchanged ions will remain in the zeolite. Although it is possible that unexchanged metal ions will remain in the zeolite, it is to be understood that the sum of the amount of unexchanged metal ions and any metal cation located in ion-exchange positions in the zeolite lattice should not be in excess of that required to provide a fully metal cation-exchanged zeolite.

The zeolite is contacted with the solution of alkali(ne earth) metal ions under conditions of temperature and time sufficient to cause an increase of alkali(ne earth) metal in the final or processed zeolite of greater than 1.5 times, preferably greater than 2 times, more preferably greater than 5 times the amount of alkali(ne earth) metal originally present in the starting zeolite, when measured as gram equivalent weights of alkali(ne earth) metal per gram of zeolite. For example, if the starting zeolite contained 0.05 percent by weight of sodium oxide, then contact with a sodium ion-containing solution to provide a sodium content greater than 1.5 times would require an increase to greater than 0.075 weight percent sodium oxide, greater than 2 times would require an increase to greater than 0.1 weight percent of sodium oxide, etc. Solution contact temperatures will typically range from 10°C to 100°C. Times will be generally be in excess of 0.1 hours. The processed zeolite will have an alkali(ne earth) metal content ranging from about 0.00004 to about 0.0004 gram equivalent weights of metal per gram of zeolite. In a preferred embodiment wherein the alkali(ne earth) metal is sodium, potassium or mixtures thereof, the processed zeolite will have an alkali metal content ranging from 0.1 to 1.4 percent by weight, basis metal, more preferably from 0.1 to 0.8 weight percent, basis metal, for sodium; from 0.2 to 1.4 weight percent, basis metal, for potassium; and 0.1 to 1.4 weight percent, basis metal, for the mixture.

In a preferred embodiment the zeolite is admixed with a binder material, such as alumina, silica, silica-alumina, aluminophosphates, silicoaluminophosphates, magnesia, titania, clays or zirconia and mixtures thereof, more preferably alumina.

In general terms the catalysts of the instant invention are prepared by forming the zeolite, optionally with the binder, into particles, chunks, pieces, pellets, rings, spheres, wagon wheels, trilobes, tetralobes and the like of a size suitable for employment in a reactor. Pelleting and extruding, preferably the latter, are operations used to shape the particles. After shaping, the particles are usually calcined at temperatures ranging from 300°C to 850°C in order to convert the particles into a durable catalyst precursor pellet. Calcining times are usually in excess of 0.1 hours and generally between 0.1 and 20 hours, although longer times can be used but may not be economically desirable. After calcining, the catalytic Group VIII noble metals are incorporated into the catalyst precursor.

The Group VIII noble metals to be used in the process according to the present invention comprise ruthenium, rhodium, palladium, osmium, iridium and platinum. Very good results have been obtained with platinum, palladium and with combinations of platinum and palladium. The use of catalysts containing both platinum and palladium is preferred since such catalysts allow relatively low hydrogenation temperatures

and pressures. The noble metals are suitably applied in amounts from 0.05 to 3%w, basis total catalyst. Preferably amounts are used in the range from 0.2 to 2%w. When platinum and palladium are used as the noble metals, a platinum/palladium molar ratio ranging from 0.1 to 1.0 is preferably applied.

The catalyst compositions according to the invention are suitably prepared by incorporating the desired amount of noble metal(s) into the appropriate zeolite, optionally together with a binder, by means of impregnation and/or ion-exchange techniques followed by drying, calcining and a reduction treatment, suitably with hydrogen-containing gas(es) to produce the noble metal(s) in the desired state.

The zeolite/(binder) is impregnated by spraying, dipping or otherwise applying the solution of the Group VIII metal salt(s) or compound(s) to the zeolite/(binder) to get the desired concentration, or alternatively the solution may be applied several times with optional drying and/or calcining between applications. Other methods of catalyst preparative techniques are also available, such as mulling one or more Group VIII metal salt(s) or compound(s) with water and a zeolite/binder mixture prior to extrusion. Suitable methods comprise impregnation of the appropriate zeolite/(binder) material with an aqueous solution of suitable Group VIII metal(s) salts. Suitable palladium salts include, for example, a palladium(II) halide, preferably palladium(II)-chloride or palladium nitrate or the appropriate ammine-complex such as tetraammine palladium nitrate. Suitable platinum compounds to be used include hexahaloplatinic acid, in particular hexachloroplatinic acid optionally in the presence of hydrochloric acid or the appropriate platinum ammine complexes, such as platinum ammine hydroxide and tetraammine platinum nitrate.

After the impregnation, the catalyst is typically dried to remove the solvent for the Group VIII metal salt(s) or compound(s), usually water. Drying will typically range in temperature from 25 °C to 200 °C when water is used as a solvent. Drying may be carried out in air, under forced draft, under vacuum or by use of other traditional methods. When drying to remove water is carried out in an air oven, typically the drying temperature will range from 100 °C to 200 °C. Drying is carried out for sufficient time to remove the solvent, typically between 0.1 and 100 hours. The drying step can be combined with the calcining step.

After drying, the catalyst containing the Group VIII noble metal typically is calcined. Calcining typically is carried out in a neutral or oxidizing atmosphere, such as nitrogen or air, respectively. Air is the preferred calcining atmosphere. Calcining temperatures will range from 200 °C to 850 °C, preferably from 250 °C to 600 °C. Times are not critical and typically will be in excess of 0.1 hours and usually will range between 0.1 and 20 hours.

The final step in the preparation of the catalyst comprises activating the catalyst by heating in a reducing atmosphere at a temperature ranging from 150 °C to 600 °C, preferably at a temperature from 200 °C to 500 °C. Times typically will be in excess of about 0.1 hours and usually range from 0.1 to 20 hours. The preferred reducing atmosphere substantially comprises hydrogen. The final activation may be carried out as one of the steps of the catalyst preparation per se, or may be carried out in situ in a chemical reactor. Since the catalyst compositions are utilized for the reduction of organic compounds with hydrogen, the in situ activation is quite convenient and economical. The activation is preferably carried out at elevated pressure, typically above 30 bar and more preferably between 30 and 250 bar. As used herein the term "catalyst" refers to the Group VIII metals-containing zeolitic materials, optionally with binder, both before activation and after.

Depending on the sulphur and nitrogen levels of the feedstock(s) to be treated it may be advantageous to subject the feedstocks to one or more pre-treatments to reduce the levels of sulphur and nitrogen accordingly. Although the catalysts to be used in the present process are much more sulphur and nitrogen tolerant than prior art catalysts, it will be clear that best results will be obtained using feedstocks containing rather limited amounts of sulphur and nitrogen. It has been found that feedstocks containing up to about 1200 parts per million by weight of sulphur and up to about 225 parts per million by weight of nitrogen can be processed adequately. Preferably feedstocks with up to about 500 parts per million by weight of sulphur and up to about 100 parts per million by weight of nitrogen are processed and most preferably feedstocks with up to about 100 parts per million by weight of sulphur and up to about 25 parts per million by weight of nitrogen are processed. The lower the operating pressure, the lower will be the amounts of sulphur and nitrogen in the feedstock that can be tolerated in the hydrogenation process. For example at the lower operating pressures of around 35 bar feedstocks with up to 100 parts per million by weight of sulphur and up to 25 parts per million by weight of nitrogen are suitable. At operating pressures of around 100 bar feedstocks with up to 500 parts per million by weight of sulphur and up to 100 parts per million by weight of nitrogen are suitable. It will be appreciated that both sulphur and nitrogen present in the pre-treated feedstocks will be converted substantially under the prevailing hydrogenation conditions since they are present in rather minute concentrations, the major process being the hydrogenation of the remaining aromatic compounds present in the feedstocks. A hydrogenation process in which at least some conversion of nitrogen and sulphur compounds takes place, is sometimes referred to as a hydrofining process. For the

purpose of this application it is to be understood that whenever reference is made to a hydrogenation process, this includes such a hydrofining process.

In most cases a pretreatment will be applied to the feedstocks that will be processed over the instant catalysts. The pretreatment which can be applied in the process according to the invention normally comprises a hydrotreating. Hydrotreating processes are well known in the art. Normally Group VIB and/or Group VIII metal compounds on an amorphous carrier such as alumina or silica-alumina can be suitably applied. Examples of such metals comprise nickel, cobalt, tungsten and molybdenum. One typical commercial hydrotreating catalyst comprises nickel and molybdenum supported on an alumina carrier. The catalysts are preferably in sulphided form. Typically, temperatures between 200°C and 425°C may be applied in hydrotreating processes.

The catalyst compositions are utilized to hydrogenate aromatics and olefins in hydrocarbon feedstocks, particularly the diesel boiling range hydrocarbons. Such feedstocks typically have a boiling range from 125°C to 625°C, that is substantially all, i.e., greater than 90 percent of weight of its components boil between 125°C and 625°C. A preferred boiling range of feedstocks is from 125°C to 425°C. The instant process is particularly suitable for hydrogenating (hydrofinishing) diesel fuels, kerosenes and/or cycle oils. Kerosenes which can be advantageously applied comprise straight run kerosenes having a boiling range between 130°C and 275°C as well as hydrotreated kerosenes. The process according to the present invention is in particular suitable for the hydrogenation of so-called cycle oils, normally obtained from catalytic cracking operations. Light cycle oils, medium cycle oils as well as heavy cycle oils and mixtures of various cycle oils can be processed advantageously using the process according to the present invention. The process is particularly applied to diesel fuels and flashed distillates with the goal of reducing their aromatics content below that required by environmental regulations. On-road diesel fuels with boiling ranges from 200°C to 360°C are particularly desired end products of the instant process.

The hydrogenation process according to the present invention is normally carried out at a temperature between 150°C and 400°C, in particular between 200°C and 400°C and preferably between 225°C and 375°C. It should be noted that the temperature to be applied is dependent on the nature of the feedstock to be hydrogenated. Typically a temperature will be chosen which allows a substantial hydrogenation of the hydrogenatable components in the feedstock, say at least 60% by weight of the total amount of components to be hydrogenated. It is preferred to carry out the process under conditions which allow at least 75% conversion by hydrogenation of the hydrogenatable components. By a proper choice of temperature and pressure more than 90% by weight of the hydrogenatable components can be hydrogenated without causing substantial hydrocracking at the same time. It is preferred to carry out the hydrogenation at rather low temperatures which will reduce hydrocracking side reactions in any event. The amount of feedstock that is hydrocracked during the instant process should preferably be below 30% by weight, more preferably below 20% by weight and most preferably below 10% by weight.

The hydrogen partial pressure to be applied typically ranges between 20 and 250 bar, in particular between 30 and 150 bar, preferably between 35 and 100 bar. Total pressures, which depend on the temperature of operation and the amount of hydrogen present, may run about 10-20% higher than the hydrogen partial pressure and sometimes as much as 50% or more. Typically, liquid hourly space velocities ("LHSV") between 0.1 and 10 litres of feed per litre of catalyst per hour ("l/h") can be applied. Hydrogen/feedstock ratios (NI/kg) between 200 and 2000 can be typically applied, preferably between 400 and 1500 NI/kg.

The process according to the present invention may be carried out in any suitable equipment. A fixed bed reactor system wherein the feedstock is passed over one or more stationary beds of catalyst is preferred. The reactor effluent is passed to a separation zone where hydrogen-rich gas is separated off and suitably recycled to the reaction zone together with make-up hydrogen as needed.

The hydrogen source used can be made of pure hydrogen or of hydrogen-containing mixtures for instance the gases produced in catalytic reforming processes.

The catalysts of the instant invention are particularly suited for processing diesel range fuel stocks which contain olefins and aromatics in order to saturate the olefins and aromatics and convert the aromatics at least in part to acyclics thereby improving the environmental quality and cetane number of the feedstock and producing a diesel fuel having a boiling range between 125°C and 425°C, preferably between 200°C and 360°C.

A particular embodiment of the instant invention comprises a two stage process for the combined hydrotreating and hydrogenating of diesel boiling range feedstocks containing nitrogen- and sulphur-containing impurities as well as relative high amounts of aromatics along with olefins to produce a diesel boiling range fuel with low nitrogen- and sulphur-containing impurities and a high cetane number. The first stage comprises contacting the feedstock with hydrogen and a conventional hydrotreating catalyst at

hydrotreating conditions. The first stage oil product is stripped of gases (hydrogen sulphide, ammonia, hydrogen and light hydrocarbons) and then mixed with fresh hydrogen and fed to the second stage which contains the catalyst composition of the instant invention. Product oils with very low aromatics levels may be obtained by this method. Because of the activity of the catalysts according to the present invention, the two stage process can be operated at relatively low pressures, preferably from 35 to 100 bar. More preferably the hydrogen partial pressure in at least the second stage ranges from 35 to 70 bar. Such low pressures can result in considerable savings in construction costs of the reactors and associated equipment.

A specific embodiment of the instant invention thus comprises a low pressure process for the combined hydrotreating and hydrogenation of hydrocarbon feedstocks containing nitrogen- and sulphur-containing impurities, aromatics and olefins and which feedstocks have a boiling range between about 125°C and about 625°C which process comprises:

(a) passing the feedstock to a first stage wherein said feedstock is contacted with hydrogen and a hydrotreating catalyst comprising a Group VIII metal selected from nickel, cobalt and mixtures thereof and a Group VIB metal selected from molybdenum, tungsten and mixtures thereof supported on a support, preferably alumina, at a temperature ranging from 200°C to 425°C, preferably 250°C to 400°C,

(b) stripping gases from the product passing from the first stage to produce a liquid product,

(c) passing said stripped liquid product to a second stage wherein said liquid product is contacted with hydrogen and a hydrogenating catalyst according to the present invention as outlined hereinbefore,

(d) stripping gases from the product passing from the second stage to produce a final liquid product.

The present invention will now be illustrated by means of the following Examples.

The following illustrative embodiments describe the typical preparation of zeolites and the resulting catalyst compositions of the present invention.

Catalyst A-1

A starting dealuminated Y-type zeolite having a molar silica to alumina ratio of 63, a sodium content (basis metal) of 0.04 weight percent and a unit cell size of 24.25 angstroms was used to prepare catalyst A-1.

The above zeolite was ion exchanged in an 8.1 weight percent solution of sodium nitrate at 82°C for 1 hour and then washed with de-ionized water. The resultant material was dried at 116°C for 16 hours to produce the zeolite product used to prepare the catalyst. The product had a sodium content (basis metal) of 0.31 weight percent and a surface area of 810 m²/g. This dried material was mulled with 20 percent by weight (dry basis) of pseudoboehmite powder (Vista Catapal B Alumina) in the presence of about 1.5 percent by weight acetic acid and sufficient water to make an extrudable mass. The extrudate was dried at 125°C and calcined at 550°C for 2 hours. The calcined extrudate was contacted with an aqueous solution of tetraammine platinum nitrate and tetraammine palladium nitrate (at a pH of 5.3) in concentrations sufficient to provide a final catalyst having 0.3 percent by weight of platinum, basis total catalyst weight, and 0.5 percent by weight of palladium, basis total catalyst weight. The resultant material was dried at 125°C and calcined at 300°C for 2 hours.

Comparative Catalyst CA-1

The above example was repeated except that the ion exchange step with the sodium nitrate was omitted.

Catalysts B-1 through B-6

A starting Y-type zeolite having a molar silica to alumina ratio of 83, a sodium content (basis metal) of 0.02 weight percent, a unit cell size of 24.26 and a surface area of 760 m²/g was used to prepare these catalysts.

The above zeolite was mulled with 20 percent by weight (dry basis) of pseudoboehmite powder (Vista Catapal B Alumina) in the presence of about 1.5 percent by weight acetic acid and sufficient water to make an extrudable mass. The extrudate was dried at 125°C and calcined at 550°C for 2 hours. The calcined extrudate was impregnated with an aqueous solution of tetraammine platinum nitrate, tetraammine palladium nitrate and sodium nitrate (at a pH of 5.3) in concentrations sufficient to provide a final catalyst having 0.3 percent by weight of platinum, basis total catalyst weight, 0.5 percent by weight of palladium, basis total

catalyst weight, and the varying weights of sodium indicated in the table below. The resultant material was dried at 125 °C and calcined at 300 °C for 2 hours.

Catalyst	Sodium Content, %wt metal
B-1	0.17
B-2	0.34
B-3	0.50
B-4	0.83
B-5	1.67
B-6	3.33

Comparative Catalyst CB-1

The above example was repeated except that sodium nitrate was omitted from the impregnation step.

Catalysts C-1 through C-2

A starting Y-type zeolite having a molar silica to alumina ratio of 82, a sodium content (basis metal) 0.02 weight percent, a unit cell size of 24.26 angstroms and a surface area of 768 m²/g was used to prepare these catalysts.

The above zeolite was ion exchanged with an 1.0 N solution of calcium nitrate, (C-1), or an 1.0 N solution of magnesium nitrate, (C-2), at 93 °C for three hours. The zeolites were filtered and washed with de-ionized water and then dried at 116 °C for 16 hours to produce zeolite products suitable for preparing catalysts. The resultant products both had surface areas of 741 m²/g and CaO or MgO contents of 0.14 weight percent (anhydrous). The dried zeolite powders were then milled with 20 percent by weight (dry basis) of pseudoboehmite powder (Vista Catapal B Alumina) in the presence of about 1.5 percent by weight acetic acid and sufficient water to make an extrudable mass. The extrudate was dried at 125 °C and calcined at 550 °C for 3 hours. The calcined extrudate was contacted with an aqueous solution of tetraammine platinum nitrate and tetraammine palladium nitrate (at a pH of 5.3) in concentrations sufficient to provide a final catalyst having 0.3 percent by weight of platinum, basis total catalyst weight, and 0.5 percent by weight of palladium, basis total catalyst weight. The resultant material was dried at 125 °C and calcined at 300 °C for 2 hours.

Catalyst	Alkaline-earth Metal Oxide Content, %wt
C-1	0.14 (CaO)
C-2	0.14 (MgO)

Catalysts D-1 through D-6

A starting Y-type zeolite having a molar silica to alumina ratio of 83, a sodium content (basis metal) of 0.02 weight percent, a unit cell size of 24.26 and a surface area of 760 m²/g was used to prepare these catalysts.

The above zeolite was milled with 20 percent by weight (dry basis) of pseudoboehmite powder (Vista Catapal B Alumina) in the presence of about 1.5 percent by weight acetic acid and sufficient water to make an extrudable mass. The extrudate was dried at 125 °C and calcined at 550 °C for 2 hours. The calcined extrudate was contacted with an aqueous solution of tetraammine platinum nitrate, tetraammine palladium nitrate and an alkali(ne earth) metal nitrate in concentrations (at a pH of 5.3) sufficient to provide a final catalyst having 0.3 percent by weight of platinum, basis total catalyst weight, 0.5 percent by weight of palladium, basis total catalyst weight, and the varying weights of alkali(ne earth) metal indicated in the table below. The resultant material was dried at 125 °C and calcined at 300 °C for 2 hours.

Catalyst	Alkali(ne-earth) Content, %wt metal
D-1	0.31 (Li)
D-2	0.31 (K)
D-3	0.46 (Cs)
D-4	0.42 (Ca)
D-5	0.25 (Ca)
D-6	0.20 (Ba)

Catalysts E-1 through E-3

A USY zeolite having a unit cell size of 24.57 angstroms, a silica to alumina molar ratio of 7.3, a sodium content of 0.15, basis metal, and a surface area of 800 m²/g was used as the starting zeolite. This starting zeolite was milled with 20 percent by weight of alumina, extruded into pellets and calcined.

Sodium nitrate in the amount shown below was dissolved in 500 millilitres of deionized water. To this solution were added 100 grams of the zeolite/alumina pellets and the mixture was stirred at the temperature listed below for 24 hours.

Catalyst	NaNO ₃ Conc. (M)	Temp., °C
E-1	0.25	25
E-2	1.00	25
E-3	2.00	100

Catalysts E-1 and E-2 were then filtered and washed with 1 litre of deionized water. Catalyst E-3 was re-exchanged with fresh NaNO₃ and heated for another 16 hours before washing with four times 100 millilitres of deionized water. All catalysts were dried at 110°C and catalyst E-3 was calcined at 500°C for 2 hours. The water pore volume of the catalysts were determined and then the catalysts were pore volume impregnated with the appropriate amounts of Pd (NH₃)₄(NO₃)₂ and Pt (NH₃)₄(NO₃)₂ dissolved in deionized water, adjusted to a pH of 5.3. All catalysts were then dried overnight at 110°C and calcined at 300°C. The final properties are summarized below.

Metal Loading, % wt.			
Catalyst	Pt	Pd	Na
E-1	0.35	0.53	0.4
E-2	0.33	0.50	1.6
E-3	0.33	0.55	2.2

Hydrogenation Process

Illustrative Embodiment P-1

The following illustrative embodiment describes the process using catalysts of the instant invention and comparative catalysts, prepared as described above, to hydrogenate (hydrofinish) two diesel feedstocks.

A stainless steel vertical tube reactor was used with gas flow entering from the top. The reactor was packed from top to bottom as follows: a layer of glass wool, a layer of 20-30 mesh silicon carbide, 18.75 grams of the catalyst admixed with 70 grams of 60-80 mesh silicon carbide, a layer of 20-30 mesh silicon carbide and a layer of glass wool. Two feedstocks were used in the testing. The physical properties of the first feedstock tested, denoted 88-50C, are listed in Table 1. The physical properties of the second feedstock, denoted 90-45, are described in Table 4. This feedstock had higher nitrogen, sulphur and aromatics content than the first feedstock. The test protocols for these two feedstocks are outlined in Tables 2 and 5, respectively. Liquid and gas products were analyzed by GC and the aromatic contents of the liquid products were determined by NMR and GC-MS.

Performance results for feedstocks 88-50C and 90-45 are shown in Tables 3 and 6, respectively.

Table 1: Properties of Feedstock 88-50C

Spec. Gravity:	0.8794 g/cc at 16°C
% Wt. Hydrogen:	12.61
% Wt. Carbon:	87.32
% Wt. Oxygen:	0.024
Sulphur:	440 ppm
Nitrogen:	2 ppm
Aromatics:	34.2% vol. (GC-MS)

13.3% Aromatic Carbon (UV)

15.9% Aromatic Carbon (NMR)

TBP-GC (°C)	IBP	171
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5%	224
----	-----

10%	237
-----	-----

30%	263
-----	-----

50%	285
-----	-----

70%	310
-----	-----

90%	384
-----	-----

FBP	419
-----	-----

Table 2: Test Protocol for Feedstock 88-50C

- A. Feedstock: Table 1
 Unit Pressure: 50 bar
 Temperature (°C): 320, 300, 290
 LHSV, (hr⁻¹): 1.25, (except as otherwise noted)
 Hydrogen to Oil: 1000 Nl/kg
 Catalyst Wt. 18.75 g
- B. Day 1: The catalyst is loaded into the reactor, the hydrogen flow is started and the pressure is raised to 50 bar while the reactor is heated to 400°C at a ramp rate of 55°C/hr. The catalyst is reduced for 8 hours at 400°C and then the reactor is cooled to 320°C at which point the feedstock is cut in.
- Days 2-8: Unit line-out.
- Day 9: Balance period started at 320°C.
- Day 10: Balance period started at 300°C.
- Day 11: Balance period started at 290°C.
- Day 12: Shut off feed, strip catalyst and cool reactor.

Table 3

Performance Results on Feedstock 88-50C				
Catalyst	% Arom. Conv		204 °C + Yield, %wt.	
	290 °C	320 °C	290 °C	320 °C
A-1	55*	70*	96.3	91.5
CA-1	56*	75*	88.8	55.4
B-1	76	90	92.9	70.4
B-2	73	86	93.5	80.1
B-3	69	83	94.8	85.2
B-4	44	60	95.3	91.6
B-5	20	28	97.0	96.4
B-6	6	8	97.5	97.5
CB-1	71	89	91.1	59.6
C-1	73	90	92.1	71.8
D-3	68	84	92.6	64.3

* Aromatics conversion measured at 1.4 LHSV

It can be seen from the data in Table 3 that at substantially comparable conversions a significant increase in the yield is obtained using the alkali(ne earth) metal enhanced zeolites that is, the catalyst composition according to the present invention, when compared to the non-enhanced zeolites of catalysts CA-1 and CB-1.

Table 4: Properties of Feedstock 90-45

Spec. Gravity:	0.8916 g/cc at 16°C
% Wt. Hydrogen:	12.27
% Wt. Carbon:	87.63
% Wt. Oxygen:	0.03
Sulphur:	640 ppm
Nitrogen:	25 ppm
Aromatics:	38.1% vol. (GC-MS)
	16.0% Aromatic Carbon (UV)
	25.2% Aromatic Carbon (NMR)
TBP-GC (°C)	IBP 149
	5% 208
	10% 223
	30% 260
	50% 288
	70% 316
	90% 358
	FBP 416

Table 5: Test Protocol for Feedstock 90-45

A. Feedstock:	Table 4
Unit Pressure:	48.3 bar
Temperature (°C):	329, 316, 302
LHSV, WHSV (hr ⁻¹):	1.00, 1.62 (except as otherwise noted)
Hydrogen to Oil:	1000 Nl/kg
Catalyst Wt.	18.75 g

Table 5: Test Protocol for Feedstock 90-45 (cont'd)

B. Day 1: The catalyst is loaded into the reactor, the hydrogen flow is started and the pressure is raised to 50 bar while the reactor is heated to 400°C at a ramp rate of 55°C/hr. The catalyst is reduced for 8 hours at 400°C and then the reactor is cooled to 320°C at which point the feedstock is cut in.

Days 2-8: Unit line-out.

Day 9: Balance period started at 316°C.

Day 10: Balance period started at 302°C.

Day 11: Balance period started at 329°C.

Day 12: Shut off feed, strip catalyst and cool reactor.

Table 6

Performance Results on Feedstock 90-45				
Catalyst	% Arom. Conv		204°C+ Yield, %wt.	
	302°C	316°C	302°C	316°C
B-2	54	66	92.3	90.3
CB-1	51	66	90.7	84.7
D-1	29	40	93.2	92.1
D-2	55	67	90.9	87.5
D-4	53	65	92.0	88.4
D-6	55	67	89.5	84.5

Illustrative Embodiment P-2

Catalysts E-1, E-2, E-3 and a catalyst prepared as described above for B-2 were tested in a vertical tubular reactor (down flow) using Feedstock 90-45 (See Table 4 for properties). The test protocol is outlined in Table 7. The liquid and gas products were analyzed by GC. The aromatics contents of the liquid products were determined by NMR and GC-MS. The results are tabulated in Table 8.

Table 7: TEST CONDITIONS AND PROCEDURECONDITIONS

Feedstock	:	90-45
Unit Pressure (psig)	:	48.3 bar
Temperature (°C)	:	316, 302, 329
LHSV (hr-1)	:	1.00
Hydrogen to Oil	:	1000 Nl/kg
Catalyst Weight (gm)	:	24.76
Catalyst Volume (cc)	:	40

PROCEDURE

Day 1:

Catalyst is mixed with 70 gm silicon carbide and loaded in the reactor. Hydrogen flow is started and the pressure is raised to 50 bar while the reactor is heated to 400°C at a rate of 55°C/hr. The catalyst is reduced for 8 hours at 400°C and then the reactor is cooled to 316°C and the feed is cut in.

Day 2 - 8:

Unit line-out

Day 9:

Balance period at 316°C

Day 10:

Line-out at 302°C

Day 11:

Balance period at 302°C

Day 12:

Line-out at 329°C

Day 13:

Balance period at 329°C.

Table 8: CATALYSTS PERFORMANCE

Catalyst	Zeolite Na(%wt)	Temperature (°C)						
		% Arom. Conv.			204+°C Yield(%wt)			
		302	316	329	302	316	329	
B-2	0.34	59	69	70	92.3	90.3	87.8	
E-1	0.4	43	57	58	86.5	75.7	61.0	
E-2	1.6	26	47	44	93.1	89.7	84.3	
E-3	2.2	24	40	39	92.5	89.7	87.4	

Illustrative Embodiment P-3

The following illustrative embodiment describe the process using catalyst B-2 described above to hydrogenate (hydrofinish) a 100% coker light gas oil (Feed 90-21) whose properties are shown in Table 9. It also illustrates the use of the instant hydrogenation process with a conventional hydrotreating process to provide a low pressure two stage process.

A commercial alumina supported nickel molybdenum catalyst was used as a hydrotreating catalyst. The feed stock was hydrotreated at 360 °C, 44.8 bar total unit pressure, LHSV of 0.4 hr⁻¹ and a hydrogen flow rate of 356 NI/l. After hydrotreatment the hydrotreated feed had the properties shown in Table 9.

The hydrotreated feed and hydrogen were passed over catalyst B-2 of the present invention at 300 °C, 44.8 bar total unit pressure, LHSV of 1.0 hr⁻¹ and a hydrogen flow rate of 1000 NI/kg. The final product ("saturated product") had the properties listed in Table 9.

Table 9 illustrates that the saturated product contained very low level of aromatics. The two stage process also resulted in a dramatic improvement in the cetane number. For this feed stock, hydrotreating led to a 22% increase in the cetane number (from 36 to 44) and saturation led to a further increase of 19% in the cetane number to 51.

**Table 9: Properties of Feed 90-21, Hydrotreated
Product and Saturated Product**

	<u>Feed</u>	<u>Hydrotreated Product</u>	<u>Saturated Product</u>
%H	12.20	13.45	14.39
%C	85.67	86.58	85.60
ppm S	14200	40	1
ppm N	1880	1	1
FIA (D-1319) (ASTM)			
Saturates	45.5	74.6	98.9
Olefins	12.2	0.0	0.0
Aromatics	42.2	25.4	1.1
Simulated Distillation D-2887 (ASTM)			
IBP	104°C	97°C	53°C
5%	172	155	114
10%	200	179	139
30%	231	219	197
50%	254	240	219
70%	275	264	243
90%	304	298	278
95%	316	311	294
FP	359	355	327
Cetane No.	36	44	51
D-613 (ASTM)			

Feed 90-21, the hydrotreated product and the saturated product were analyzed by GC-MS in order to determine the ring content. These results are shown in Table 10 below. The results clearly show that the process according to the present invention is effective at ring opening to produce higher cetane number product.

**Table 10: Distribution of Ring Numbers for
Feed 90-21. Hydrotreated Product
and Saturated Products**

	<u>Feed</u>	<u>Hydrotreated Product</u>	<u>Saturated Product</u>
1 ring	28.28	32.95	34.97
2 rings	35.35	32.21	30.06
3 rings	9.74	8.20	4.78
4 rings	0.97	0.23	0.05
Alicyclic- Saturates	<u>25.67</u>	<u>26.41</u>	<u>30.14</u>
	100.01	100.00	100.00

Illustrative Embodiment P-4

The following illustrative embodiment describes the process using catalyst B-2 described above to hydrogenate (hydrofinish) a feedstock (90-83) having a very high final boiling point. It also illustrates the use of the instant hydrogenation process with a conventional hydrotreating process to provide a two stage process. A commercial hydrotreating catalyst, alumina supported nickel/molybdenum, was sulphided and used. The feedstock was a blend of 66%wt diesel and 33%wt heavy flashed distillate and had the properties listed in the table below. The feedstock was hydrotreated with the sulphided Ni/Mo catalyst at 385°C and 1500 psig total pressure with a hydrogen rate of 400 NL/kg and LHSV of 0.5 hr⁻¹. The hydrotreated feed had the properties listed in the table below.

Table 11: Properties of Feedstock 90-83 Before
and After Hydrotreating

	<u>Feedstock Properties</u>	<u>Hydrotreated Feed Properties</u>
	<u>S & N Content</u>	
ppmw S	640	15
ppmw N	1340	108
Specific Gravity, (16°C)	0.9053 g/cc	0.8795 g/cc
%wt Hydrogen	12.178	12.92
%wt Carbon	87.592	87.16
	<u>Aromatics</u>	
NMR	19.83%	10.5
UV	15.91%	8.2
	<u>Distillation</u>	
IBP	224°C	216°C
10%	265	254
20%	280	274
30%	298	292
40%	316	311
50%	342	331
60%	370	359
70%	435	408
80%	512	493
90%	549	541
99%	596	583

The hydrotreated feed and hydrogen were passed over catalyst B-2 of the instant invention under the conditions listed in the table below for four separate runs yielding products 1, 2, 3 and 4 with the properties listed in the table below. Product 5 was produced by substituting B-2 with a commercially available sulphided nickel/molybdenum on alumina catalyst.

Table 12: Properties of Saturated Product

	Product	Product	Product	Product	Product
	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>
<u>Operating Conditions</u>					
H ₂ Rate, Nl/kg	660	1100	1000	1160	1080
LHSV, hr ⁻¹	1.5	0.9	1.0	0.8	0.9
Total Press., bar	48.3	48.3	82.7	103.4	48.3
Temp., °C	343	343	343	343	343
<u>S & N Content</u>					
ppmw S	16	13	4	3	5
ppmw N	44	35	29	24	65
<u>Aromatics</u>					
NMR	6.4	5.5	2.2	1.6	10.4
UV	4.9	4.6	1.7	1.1	8.5

Products 2 and 5 were obtained under essentially the same conditions, but it can be seen that the use of the instant catalyst results in a significant decrease in the aromatics content of the final product.

Claims

Claims for the following Contracting States : BE, CH, DE, DK, FR, GB, IT, LI, NL, SE

- A catalyst composition comprising one or more Group VIII noble metal(s) on a support wherein the support comprises a zeolite obtainable by contacting a starting Y-type zeolite having a unit cell size of less than 24.65 angstroms, a silica to alumina molar ratio greater than 5, and a sum of alkali and alkaline earth metal content of less than 0.3 weight percent (basis metal) with at least one aqueous solution comprising alkali metal ions and/or alkaline earth metal ions under conditions sufficient to provide an increased sum of final alkali and alkaline earth metal content (measured in terms of gram equivalent weight per gram of zeolite), which sum of final alkali and alkaline earth metal content is at least 1.5 times greater than the sum of the alkali and alkaline earth metal content of the starting zeolite (measured in terms of gram equivalent weight per gram of zeolite).
- The composition of claim 1 wherein the starting Y-type zeolite has a unit cell size of less than 24.4 angstroms, preferably less than 24.35 angstroms.
- The composition as claimed in claim 1 or 2 wherein the starting zeolite has a unit cell size of between 24.2 and 24.3 angstroms, preferably a unit cell size of between 24.22 and 24.28 angstroms.
- Composition according to any one of the preceding claims wherein the starting Y-type zeolite has a silica to alumina molar ratio greater than 25, preferably greater than 35.



Table 12: Properties of Saturated Product

	Product	Product	Product	Product	Product
	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>
<u>Operating Conditions</u>					
H ₂ Rate, Nl/kg	660	1100	1000	1160	1080
LHSV, hr ⁻¹	1.5	0.9	1.0	0.8	0.9
Total Press., bar	48.3	48.3	82.7	103.4	48.3
Temp., °C	343	343	343	343	343
<u>S & N Content</u>					
ppmw S	16	13	4	3	5
ppmw N	44	35	29	24	65
<u>Aromatics</u>					
NMR	6.4	5.5	2.2	1.6	10.4
UV	4.9	4.6	1.7	1.1	8.5

Products 2 and 5 were obtained under essentially the same conditions, but it can be seen that the use of the instant catalyst results in a significant decrease in the aromatics content of the final product.

Claims

Claims for the following Contracting States : BE, CH, DE, DK, FR, GB, IT, LI, NL, SE

1. A catalyst composition comprising one or more Group VIII noble metal(s) on a support wherein the support comprises a zeolite obtainable by contacting a starting Y-type zeolite having a unit cell size of less than 24.65 angstroms, a silica to alumina molar ratio greater than 5, and a sum of alkali and alkaline earth metal content of less than 0.3 weight percent (basis metal) with at least one aqueous solution comprising alkali metal ions and/or alkaline earth metal ions under conditions sufficient to provide an increased sum of final alkali and alkaline earth metal content (measured in terms of gram equivalent weight per gram of zeolite), which sum of final alkali and alkaline earth metal content is at least 1.5 times greater than the sum of the alkali and alkaline earth metal content of the starting zeolite (measured in terms of gram equivalent weight per gram of zeolite).
2. The composition of claim 1 wherein the starting Y-type zeolite has a unit cell size of less than 24.4 angstroms, preferably less than 24.35 angstroms.
3. The composition as claimed in claim 1 or 2 wherein the starting zeolite has a unit cell size of between 24.2 and 24.3 angstroms, preferably a unit cell size of between 24.22 and 24.28 angstroms.
4. Composition according to any one of the preceding claims wherein the starting Y-type zeolite has a silica to alumina molar ratio greater than 25, preferably greater than 35.



5. Composition according to claim 4 wherein the starting Y-type zeolite has a silica to alumina molar ratio greater than 50, preferably greater than 60.
6. Composition according to any one of the preceding claims wherein the starting Y-type zeolite has an alkali and alkaline earth metal content of less than 0.15 percent by weight, preferably less than 0.075 percent by weight.
7. Composition according to any one of the preceding claims wherein the starting Y-type zeolite comprises an alkali metal.
8. Composition according to any one of the preceding claims wherein the final alkali and alkaline earth metal content is at least 2 times greater than the alkali and alkaline earth metal content of the starting zeolite, preferably at least 5 times greater than the alkali and alkaline earth metal content of the starting zeolite.
9. Composition according to any one of the preceding claims wherein the final alkali and alkaline earth metal is selected from sodium, potassium and mixtures thereof and its content ranges from 0.1 to 1.4 percent by weight, basis metal.
10. Composition of claim 9 wherein the alkali and alkaline earth metal is sodium and ranges from 0.1 to 0.8 percent by weight, basis metal.
11. Composition according to any one of the preceding claims wherein the catalyst additionally comprises a binder selected from the group consisting of alumina, silica, silica-alumina, aluminophosphates, silicoaluminophosphates, magnesia, titania, zirconia, clays and mixtures thereof.
12. Composition according to any one of the preceding claims wherein the Group VIII noble metal(s) is (are) selected from the group consisting of platinum, palladium and mixtures thereof, preferably in an amount from 0.05 to 3 percent by weight of the total catalyst.
13. Composition according to claim 12 wherein the Group VIII noble metal(s) comprise platinum and palladium in a platinum to palladium molar ratio ranging from 0.1 to 1.0.
14. Composition according to any one of the preceding claims which has been calcined at a temperature ranging from 200 °C to 850 °C, preferably 250 °C to 600 °C.
15. The composition of claim 14 which has further been activated at a temperature ranging from 150 °C to 600 °C, preferably from 200 °C to 500 °C, in a hydrogen-containing atmosphere at an elevated pressure, preferably greater than 30 bar.
16. A process for the hydrogenation of hydrocarbon feedstocks containing hydrogenatable components and which have a boiling range between 125 °C and 625 °C which process comprises contacting said feedstock at a temperature ranging between 150 °C and 400 °C with hydrogen and a catalyst as claimed in any one of claims 1-15.
17. Process as claimed in claim 16 wherein the contact temperature ranges between 200 °C and 400 °C, preferably between 225 and 375 °C.
18. Process as claimed in claim 16 or 17 wherein the catalyst has been calcined at a temperature ranging from 200 °C to 850 °C prior to use in the process.
19. Process as claimed in claim 18 wherein the catalyst has further been activated at a temperature ranging from 150 °C to 600 °C, preferably from 200 °C to 500 °C in a hydrogen-containing atmosphere at an elevated pressure, preferably greater than 30 bar prior to use in the process.
20. Process according to any one of claims 16-19, wherein a diesel range fuel feedstock containing aromatics and olefins is hydrogenated to produce a product having a boiling range between 125 °C and 425 °C whereby said olefins and aromatics are substantially converted to saturates and the saturated



aromatics are converted at least in part to acyclics with hydrocracking of the feedstock being substantially absent.

- 5 21. A process for the combined hydrotreating and hydrogenation of hydrocarbon feedstocks containing nitrogen- and sulphur-containing impurities, aromatics and olefins and which feedstocks have a boiling range between 125 °C and 625 °C which process comprises:
- 10 (a) passing the feedstock to a first stage wherein said feedstock is contacted with hydrogen and a hydrotreating catalyst comprising a Group VIII metal sulphide selected from nickel, cobalt and mixtures thereof and a Group VIB metal sulphide selected from molybdenum, tungsten and mixtures thereof supported on a support at a temperature ranging from 200 °C to 425 °C,
- (b) stripping gases from the product passing from the first stage to produce a liquid product,
- (c) passing said liquid product to a second stage wherein said liquid product is contacted with hydrogen and a hydrogenating catalyst as claimed in any one of claims 1-15,
- 15 (d) stripping gases from the product passing from the second stage to produce a final liquid product.
22. The process of claim 21 wherein the hydrogen partial pressure in both stages ranges from 35 to 100 bar.
23. The process of claim 22 wherein the hydrogen partial pressure in the second stage ranges from 35 to 20 70 bar.

Claims for the following Contracting State : ES

- 25 1. Process for the preparation of a catalyst composition comprising one or more Group VIII noble metal(s) on a support which process comprises bringing the Group VIII metal(s) onto a support comprising a zeolite obtainable by contacting a starting Y-type zeolite having a unit cell size of less than 24.65 angstroms, a silica to alumina molar ratio greater than 5, and a sum of alkali and alkaline earth metal content of less than 0.3 weight percent (basis metal) with at least one aqueous solution comprising alkali metal ions and/or alkaline earth metal ions under conditions sufficient to provide an increased sum of final alkali and alkaline earth metal content (measured in terms of gram equivalent weight per gram of zeolite), which sum of final alkali and alkaline earth metal content is at least 1.5 times greater than the sum of the alkali and alkaline earth metal content of the starting zeolite (measured in terms of gram equivalent weight per gram of zeolite).
- 30 2. The process of claim 1 wherein the starting Y-type zeolite has a unit cell size of less than 24.4 angstroms, preferably less than 24.35 angstroms.
3. The process as claimed in claim 1 or 2 wherein the starting zeolite has a unit cell size of between 24.2 and 24.3 angstroms, preferably a unit cell size of between 24.22 and 24.28 angstroms.
- 40 4. Process according to any one of the preceding claims wherein the starting Y-type zeolite has a silica to alumina molar ratio greater than 25, preferably greater than 35.
5. Process according to claim 4 wherein the starting Y-type zeolite has a silica to alumina molar ratio greater than 50, preferably greater than 60.
- 45 6. Process according to any one of the preceding claims wherein the starting Y-type zeolite has an alkali and alkaline earth metal content of less than 0.15 percent by weight, preferably less than 0.075 percent by weight.
- 50 7. Process according to any one of the preceding claims wherein the starting Y-type zeolite comprises an alkali metal.
8. Process according to any one of the preceding claims wherein the final alkali and alkaline earth metal content is at least 2 times greater than the alkali and alkaline earth metal content of the starting zeolite, preferably at least 5 times greater than the alkali(ne earth) metal content of the starting zeolite.
- 55



9. Process according to any one of the preceding claims wherein the final alkali and alkaline earth metal is selected from sodium, potassium and mixtures thereof and its content ranges from 0.1 to 1.4 percent by weight, basis metal.
10. Process of claim 9 wherein the alkali and alkaline earth metal is sodium and ranges from 0.1 to 0.8 percent by weight, basis metal.
11. Process according to any one of the preceding claims wherein the catalyst additionally comprises a binder selected from the group consisting of alumina, silica, silica-alumina, aluminophosphates, silicoaluminophosphates, magnesia, titania, zirconia, clays and mixtures thereof.
12. Process according to any one of the preceding claims wherein the Group VIII noble metal(s) is (are) selected from the group consisting of platinum, palladium and mixtures thereof, preferably in an amount from 0.05 to 3 percent by weight of the total catalyst.
13. Process according to claim 12 wherein the Group VIII noble metal(s) comprise platinum and palladium in a platinum to palladium molar ratio ranging from 0.1 to 1.0.
14. Process according to any one of the preceding claims wherein the catalyst composition is calcined at a temperature ranging from 200 °C to 850 °C, preferably 250 °C to 600 °C.
15. The process of claim 14 wherein the catalyst composition is further activated at a temperature ranging from 150 °C to 600 °C, preferably from 200 °C to 500 °C, in a hydrogen-containing atmosphere at an elevated pressure, preferably greater than 30 bar.
16. A process for the hydrogenation of hydrocarbon feedstocks containing hydrogenatable components and which have a boiling range between 125 °C and 625 °C which process comprises contacting said feedstock at a temperature ranging between 150 °C and 400 °C with hydrogen and a catalyst as claimed in any one of claims 1-15.
17. Process as claimed in claim 16 wherein the contact temperature ranges between 200 °C and 400 °C, preferably between 225 and 375 °C.
18. Process as claimed in claim 16 or 17 wherein the catalyst has been calcined at a temperature ranging from 200 °C to 850 °C prior to use in the process.
19. Process as claimed in claim 18 wherein the catalyst has further been activated at a temperature ranging from 150 °C to 600 °C, preferably from 200 °C to 500 °C in a hydrogen-containing atmosphere at an elevated pressure, preferably greater than 30 bar prior to use in the process.
20. Process according to any one of claims 16-19, wherein a diesel range fuel feedstock containing aromatics and olefins is hydrogenated to produce a product having a boiling range between 125 °C and 425 °C whereby said olefins and aromatics are substantially converted to saturates and the saturated aromatics are converted at least in part to acyclics with hydrocracking of the feedstock being substantially absent.
21. A process for the combined hydrotreating and hydrogenation of hydrocarbon feedstocks containing nitrogen- and sulphur-containing impurities, aromatics and olefins and which feedstocks have a boiling range between 125 °C and 625 °C which process comprises:
- passing the feedstock to a first stage wherein said feedstock is contacted with hydrogen and a hydrotreating catalyst comprising a Group VIII metal sulphide selected from nickel, cobalt and mixtures thereof and a Group VIB metal sulphide selected from molybdenum, tungsten and mixtures thereof supported on a support at a temperature ranging from 200 °C to 425 °C,
 - stripping gases from the product passing from the first stage to produce a liquid product,
 - passing said liquid product to a second stage wherein said liquid product is contacted with hydrogen and a hydrogenating catalyst as claimed in any one of claims 1-15,
 - stripping gases from the product passing from the second stage to produce a final liquid product.



22. The process of claim 21 wherein the hydrogen partial pressure in both stages ranges from 35 to 100 bar.
23. The process of claim 22 wherein the hydrogen partial pressure in the second stage ranges from 35 to 70 bar.

Patentansprüche

Patentansprüche für folgende Vertragsstaaten : BE, CH, DE, DK, FR, GB, IT, LI, NL, SE

1. Katalysatorzusammensetzung bestehend aus einem bzw. mehreren auf einen Träger aufgetragenen Edelmetall bzw. Edelmetallen aus Gruppe VIII, wobei der Träger aus einem Zeolith besteht, der durch Kontaktieren eines Ausgangszeoliths vom Y-Typ mit einer Elementarzellengröße von Weniger als 24,65 Å, einem Siliciumdioxid-Aluminiumoxid-Molverhältnis von mehr als 5 und einer Summe aus Alkali- und Erdalkalimetallgehalt von Weniger als 0,3 Gew.-% (bezogen auf das Metall) mit mindestens einer wäßrigen Lösung aus Alkali- und/oder Erdalkalitionen unter zur Herstellung einer erhöhten Summe aus Alkali- und Erdalkaliendgehalt (gemessen als Gramm-Äquivalentgewicht pro Gramm Zeolith), die mindestens 1,5 mal so hoch ist wie die Summe aus Alkali- und Erdalkalimetallgehalt des Ausgangszeoliths (gemessen als Gramm-Äquivalentgewicht pro Gramm Zeolith), ausreichenden Bedingungen erhältlich ist.
2. Zusammensetzung nach Anspruch 1, wobei der Ausgangszeolith vom Y-Typ eine Elementarzellengröße von weniger als 24,4 Å, vorzugsweise von weniger als 24,35 Å besitzt.
3. Zusammensetzung nach Anspruch 1 oder 2, wobei der Ausgangszeolith eine Elementarzellengröße zwischen 24,2 und 24,3 Å, vorzugsweise eine Elementarzellengröße zwischen 24,22 und 24,28 Å besitzt.
4. Zusammensetzung nach einem der vorhergehenden Ansprüche, wobei der Ausgangszeolith vom Y-Typ ein Siliciumdioxid-Aluminiumoxid-Molverhältnis von mehr als 25, vorzugsweise von mehr als 35 besitzt.
5. Zusammensetzung nach Anspruch 4, wobei der Ausgangszeolith vom Y-Typ ein Siliciumdioxid-Aluminiumoxid-Molverhältnis von mehr als 50, vorzugsweise von mehr als 60 besitzt.
6. Zusammensetzung nach einem der vorhergehenden Ansprüche, wobei der Ausgangszeolith vom Y-Typ einen Alkali- und Erdalkalimetallgehalt von weniger als 0,15 Gew.-%, vorzugsweise von weniger als 0,075 Gew.-% besitzt.
7. Zusammensetzung nach einem der vorhergehenden Ansprüche, wobei der Ausgangszeolith vom Y-Typ ein Alkalimetall enthält.
8. Zusammensetzung nach einem der vorhergehenden Ansprüche, wobei der Alkali- und Erdalkaliendgehalt mindestens das 2-fache des Alkali- und Erdalkalimetallgehalts des Ausgangszeoliths, vorzugsweise mindestens das 5-fache des Alkali- und Erdalkalimetallgehalts des Ausgangszeoliths beträgt.
9. Zusammensetzung nach einem der vorhergehenden Ansprüche, wobei das Erdalkali- und -erdalkalimetall aus Natrium, Kalium und deren Mischungen ausgewählt ist und sein Anteil im Bereich von 0,1 bis 1,4 Gew.-% bezogen auf das Metall liegt.
10. Zusammensetzung nach Anspruch 9, wobei das Alkali- und Erdalkalimetall Natrium ist und in einem Anteil von 0,1 bis 0,8 Gew.-% bezogen auf das Metall vorliegt.
11. Zusammensetzung nach einem der vorhergehenden Ansprüche, wobei der Katalysator zusätzlich ein aus der Gruppe Aluminiumoxid, Siliciumdioxid, Siliciumdioxid-Aluminiumoxid, Alumophosphate, Silicoalumophosphate, Magnesiumoxid, Titanoxid, Zirkonoxid, Tone und deren Mischungen ausgewähltes Bindemittel enthält.
12. Zusammensetzung nach einem der vorhergehenden Ansprüche, wobei das Edelmetall bzw. die Edelmetalle aus Gruppe VIII aus der Gruppe Platin, Palladium und deren Mischungen, vorzugsweise in



einer Menge von 0,05 bis 3 Gew.-% des Gesamtkatalysators ausgewählt ist bzw. sind.

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13. Zusammensetzung nach Anspruch 12, wobei das Edelmetall bzw. die Edelmetalle aus Gruppe VIII aus Platin und Palladium bestehen und in einem Platin-Palladium-Molverhältnis im Bereich von 0,1 bis 1,0 vorliegen.
14. Zusammensetzung nach einem der vorhergehenden Ansprüche, die bei einer Temperatur im Bereich von 200 °C bis 850 °C, vorzugsweise 250 °C bis 600 °C kalziniert wurde.
- 10 15. Zusammensetzung nach Anspruch 14, wobei deren weitere Aktivierung bei einer Temperatur im Bereich von 150 °C bis 600 °C, vorzugsweise von 200 °C bis 500 °C, in einer wasserstoffhaltigen Atmosphäre bei erhöhtem Druck, vorzugsweise bei mehr als 30 bar erfolgte.
- 15 16. Verfahren zur Hydrierung von Kohlenwasserstoff-Einsatzstoffen, die hydrierbare Bestandteile enthalten und einen Siedebereich zwischen 125 °C und 625 °C aufweisen, wobei man den Einsatzstoff bei einer Temperatur im Bereich zwischen 150 °C und 400 °C mit Wasserstoff und einem Katalysator nach einem der Ansprüche 1-15 kontaktiert.
- 20 17. Verfahren nach Anspruch 16, wobei die Kontaktiertemperatur im Bereich zwischen 200 °C und 400 °C, vorzugsweise zwischen 225 °C und 375 °C liegt.
18. Verfahren nach Anspruch 16 oder 17, wobei der Katalysator vor seinem Einsatz in dem Verfahren bei einer Temperatur im Bereich von 200 °C bis 850 °C kalziniert wurde.
- 25 19. Verfahren nach Anspruch 18, wobei die weitere Aktivierung des Katalysators vor seinem Einsatz in dem Verfahren bei einer Temperatur im Bereich von 150 °C bis 600 °C, vorzugsweise von 200 °C bis 500 °C, in einer wasserstoffhaltigen Atmosphäre bei erhöhtem Druck, vorzugsweise bei mehr als 30 bar erfolgte.
- 30 20. Verfahren nach einem der Ansprüche 16-19, wobei ein Einsatzstoff aus aromaten- und olefinhaltigem Dieselkraftstoff zu einem Produkt mit einem Siedebereich zwischen 125 °C und 425 °C hydriert wird, wodurch die Olefine und Aromaten im wesentlichen zu gesättigten Verbindungen und die gesättigten Aromaten zumindest teilweise zu acyclischen Verbindungen umgesetzt werden, ohne daß es zu nennenswertem Hydrocracking kommt.
- 35 21. Verfahren zum kombinierten Hydrotreating und Hydrieren von Kohlenwasserstoff-Einsatzstoffen mit stickstoff- und schwefelhaltigen Verunreinigungen sowie Aromaten und Olefinen und einem Siedebereich zwischen 125 °C und 625 °C, wobei man:
- 40 (a) den Einsatzstoff einer ersten Stufe zuführt, in der der Einsatzstoff mit Wasserstoff und einem auf einen Träger aufgebrachten Hydrotreating-Katalysator aus einem Sulfid eines aus Nickel, Cobalt und deren Mischungen ausgewählten Metalls aus Gruppe VIII und einem Sulfid eines aus Molybdän, Wolfram und deren Mischungen ausgewählten Metalls aus Gruppe VIB bei einer Temperatur im Bereich von 200 °C bis 425 °C kontaktiert wird,
- 45 (b) aus dem aus der ersten Stufe austretenden Produkt Gase abzieht, wobei ein flüssiges Produkt entsteht,
- (c) das flüssige Produkt einer zweiten Stufe zuführt, in der das flüssige Produkt mit Wasserstoff und einem Hydrierkatalysator nach einem der Ansprüche 1-15 kontaktiert wird und
- (d) aus dem aus der zweiten Stufe austretenden Produkt Gase abzieht, wodurch ein flüssiges Endprodukt entsteht.
- 60 22. Verfahren nach Anspruch 21, wobei der Wasserstoff-Partialdruck in beiden Stufen im Bereich von 35 bis 100 bar liegt.
- 55 23. Verfahren nach Anspruch 22, wobei der Wasserstoff-Partialdruck in der zweiten Stufe im Bereich von 35 bis 70 bar liegt.



Patentansprüche für folgenden Vertragsstaat : ES

1. Verfahren zur Herstellung einer Trägerkatalysatorzusammensetzung bestehend aus einem bzw. mehreren auf einen Träger aufgetragenen Edelmetall bzw. Edelmetallen aus Gruppe VIII, wobei man das Metall bzw. die Metalle aus Gruppe VIII auf einen Träger aus einem Zeolith aufbringt, der durch Kontaktieren eines Ausgangszeoliths vom Y-Typ mit einer Elementarzellengröße von weniger als 24,65 Å, einem Siliciumdioxid-Aluminiumoxid-Molverhältnis von mehr als 5 und einer Summe aus Alkali- und Erdalkalimetallgehalt von weniger als 0,3 Gew.-% (bezogen auf das Metall) mit mindestens einer wäßrigen Lösung aus Alkali- und/oder Erdalkalitionen unter zur Herstellung einer erhöhten Summe aus Alkali- und Erdalkaliendgehalt (gemessen als Gramm-Äquivalentgewicht pro Gramm Zeolith), die mindestens 1,5 mal so hoch ist wie die Summe aus Alkali- und Erdalkalimetallgehalt des Ausgangszeoliths (gemessen in Gramm-Äquivalentgewicht pro Gramm Zeolith), ausreichenden Bedingungen erhältlich ist.
2. Verfahren nach Anspruch 1, wobei der Ausgangszeolith vom Y-Typ eine Elementarzellengröße von weniger als 24,4 Å, vorzugsweise weniger als 24,35 Å besitzt.
3. Verfahren nach Anspruch 1 oder 2, wobei der Ausgangszeolith eine Elementarzellengröße zwischen 24,2 und 24,3 Å, vorzugsweise eine Elementarzellengröße zwischen 24,22 und 24,28 Å besitzt.
4. Verfahren nach einem der vorhergehenden Ansprüche, wobei der Ausgangszeolith vom Y-Typ ein Siliciumdioxid-Aluminiumoxid-Molverhältnis von mehr als 25, vorzugsweise von mehr als 35 besitzt.
5. Verfahren nach Anspruch 4, wobei der Ausgangszeolith vom Y-Typ ein Siliciumdioxid-Aluminiumoxid-Molverhältnis von mehr als 50, vorzugsweise von mehr als 60 besitzt.
6. Verfahren nach einem der vorhergehenden Ansprüche, wobei der Ausgangszeolith vom Y-Typ einen Alkali- und Erdalkalimetallgehalt von weniger als 0,15 Gew.-%, vorzugsweise von weniger als 0,075 Gew.-% besitzt.
7. Verfahren nach einem der vorhergehenden Ansprüche, wobei der Ausgangszeolith vom Y-Typ ein Alkalimetall enthält.
8. Verfahren nach einem der vorhergehenden Ansprüche, wobei der Alkali- und Erdalkaliendgehalt mindestens das 2-fache des Alkali- und Erdalkalimetallgehalts des Ausgangszeoliths, vorzugsweise mindestens das 5-fache des (Erd)alkalimetallgehalts des Ausgangszeoliths beträgt.
9. Verfahren nach einem der vorhergehenden Ansprüche, wobei das Erdalkali- und -erdalkalimetall aus Natrium, Kalium und deren Mischungen ausgewählt ist und sein Anteil im Bereich von 0,1 bis 1,4 Gew.-% bezogen auf das Metall liegt.
10. Verfahren nach Anspruch 9, wobei das Alkali- und Erdalkalimetall Natrium ist und in einem Anteil von 0,1 bis 0,8 Gew.-% bezogen auf das Metall vorliegt.
11. Verfahren nach einem der vorhergehenden Ansprüche, wobei der Katalysator zusätzlich ein aus der Gruppe Aluminiumoxid, Siliciumdioxid, Siliciumdioxid-Aluminiumoxid, Alumophosphate, Silicoalumophosphate, Magnesiumoxid, Titanoxid, Zirkonoxid, Tone und deren Mischungen ausgewähltes Bindemittel enthält.
12. Verfahren nach einem der vorhergehenden Ansprüche, wobei das Edelmetall bzw. die Edelmetalle aus Gruppe VIII aus der Gruppe Platin, Palladium und deren Mischungen, vorzugsweise in einer Menge von 0,05 bis 3 Gew.-% des Gesamtkatalysators ausgewählt ist bzw. sind.
13. Verfahren nach Anspruch 12, wobei das Edelmetall bzw. die Edelmetalle aus Gruppe VIII aus Platin und Palladium bestehen und in einem Platin-Palladium-Molverhältnis im Bereich von 0,1 bis 1,0 vorliegen.
14. Verfahren nach einem der vorhergehenden Ansprüche, wobei die Katalysatorzusammensetzung bei einer Temperatur im Bereich von 200 °C bis 850 °C, vorzugsweise 250 °C bis 600 °C kalziniert wird.



15. Verfahren nach Anspruch 14, wobei die weitere Aktivierung der Katalysatorzusammensetzung bei einer Temperatur im Bereich von 150 °C bis 600 °C, vorzugsweise von 200 °C bis 500 °C, in einer wasserstoffhaltigen Atmosphäre bei erhöhtem Druck, vorzugsweise bei mehr als 30 bar erfolgt.
- 5 16. Verfahren zur Hydrierung von Kohlenwasserstoff-Einsatzstoffen, die hydrierbare Bestandteile enthalten und einen Siedebereich zwischen 125 °C und 625 °C aufweisen, wobei man den Einsatzstoff bei einer Temperatur im Bereich zwischen 150 °C und 400 °C mit Wasserstoff und einem Katalysator nach einem der Ansprüche 1-15 kontaktiert.
- 10 17. Verfahren nach Anspruch 16, wobei die Kontaktiertemperatur im Bereich zwischen 200 °C und 400 °C, vorzugsweise zwischen 225 und 375 °C liegt.
18. Verfahren nach Anspruch 16 oder 17, wobei der Katalysator vor seinem Einsatz in dem Verfahren bei einer Temperatur im Bereich von 200 °C bis 850 °C kalziniert wurde.
- 15 19. Verfahren nach Anspruch 18, wobei die weitere Aktivierung des Katalysators vor seinem Einsatz in dem Verfahren bei einer Temperatur im Bereich von 150 °C bis 600 °C, vorzugsweise von 200 °C bis 500 °C, in einer wasserstoffhaltigen Atmosphäre bei erhöhtem Druck, vorzugsweise bei mehr als 30 bar erfolgte.
- 20 20. Verfahren nach einem der Ansprüche 16-19, wobei ein Einsatzstoff aus aromaten- und olefinhaltigem Dieseldieselkraftstoff zu einem Produkt mit einem Siedebereich zwischen 125 °C und 425 °C hydriert wird, wodurch die Olefine und Aromaten im wesentlichen zu gesättigten Verbindungen und die gesättigten Aromaten zumindest teilweise zu acyclischen Verbindungen umgesetzt werden, ohne daß es zu nennenswertem Hydrocracking kommt.
- 25 21. Verfahren zum kombinierten Hydrotreating und Hydrieren von Kohlenwasserstoff-Einsatzstoffen mit Stickstoff- und schwefelhaltigen Verunreinigungen sowie Aromaten und Olefinen und einem Siedebereich zwischen 125 °C und 625 °C, wobei man:
 - 30 (a) den Einsatzstoff einer ersten Stufe zuführt, in der der Einsatzstoff mit Wasserstoff und einem auf einem Träger aufgebrachteten Hydrotreating-Katalysator aus einem Sulfid eines aus Nickel, Cobalt und deren Mischungen ausgewählten Metalls aus Gruppe VIII und einem Sulfid eines aus Molybdän, Wolfram und deren Mischungen ausgewählten Metalls aus Gruppe VIB bei einer Temperatur im Bereich von 200 °C bis 425 °C kontaktiert wird,
 - 35 (b) aus dem aus der ersten Stufe austretenden Produkt Gase abzieht, wobei ein flüssiges Produkt entsteht,
 - (c) das flüssige Produkt einer zweiten Stufe zuführt, in der das flüssige Produkt mit Wasserstoff und einem Hydrierkatalysator nach einem der Ansprüche 1-15 kontaktiert wird und
 - 40 (d) aus dem aus der zweiten Stufe austretenden Produkt Gase abzieht, wodurch ein flüssiges Endprodukt entsteht.
22. Verfahren nach Anspruch 21, wobei der Wasserstoff-Partialdruck in beiden Stufen im Bereich von 35 bis 100 bar liegt.
- 45 23. Verfahren nach Anspruch 22, wobei der Wasserstoff-Partialdruck in der zweiten Stufe im Bereich von 35 bis 70 bar liegt.

Revendications

Revendications pour les Etats contractants suivants : BE, CH, DE, DK, FR, GB, IT, LI, NL, SE

- 50 1. Une composition catalytique comprenant un ou plusieurs métaux nobles du groupe VIII sur le support, où le support comprend une zéolite pouvant être obtenue par mise en contact d'une zéolite de départ du type Y ayant une dimension des cellules unitaires de moins de 24,65 angströms, un rapport molaire de la silice à l'alumine supérieur à 5 et une somme des teneurs en métaux alcalins et en métaux alcalino-terreux de moins de 0,3 pour cent en poids (calculée en métaux) avec au moins une solution aqueuse comprenant des ions de métaux alcalins et/ou des ions de métaux alcalino-terreux dans des conditions suffisantes pour donner une somme accrue des teneurs finales en métaux alcalins et alcalino-terreux (mesurées en équivalents-grammes par gramme de zéolite), laquelle somme des
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teneurs finales en métaux alcalins et alcalino-terreux est au moins 1,5 fois supérieure à la somme des teneurs en métaux alcalins et alcalino-terreux de la zéolite de départ (mesurées en équivalents-grammes par gramme de zéolite).

- 5 2. La composition selon la revendication 1 dans laquelle la zéolite de type Y de départ a une dimension des cellules unitaires de moins de 24,4 angströms, de préférence moins de 24,35 angströms.
3. La composition selon la revendication 1 ou 2 dans laquelle la zéolite de départ a une dimension des cellules unitaires comprise entre 24,2 et 24,3 angströms, de préférence entre 24,22 et 24,28 angströms.
- 10 4. Composition selon l'une quelconque des revendications précédentes, dans laquelle la zéolite de type Y de départ a un rapport molaire de la silice à l'alumine supérieur à 25, de préférence supérieur à 35.
- 15 5. Composition selon la revendication 4, dans laquelle la zéolite de type Y de départ a un rapport molaire de la silice à l'alumine supérieur à 50, de préférence supérieur à 60.
6. Composition selon l'une quelconque des revendications précédentes, dans laquelle la zéolite de type Y de départ a une teneur en métaux alcalins et alcalino-terreux de moins de 0,15 pour cent en poids, de préférence moins de 0,075 pour cent en poids.
- 20 7. Composition selon l'une quelconque des revendications précédentes, dans laquelle la zéolite de type Y de départ comprend un métal alcalin.
- 25 8. Composition selon l'une quelconque des revendications précédentes, dans laquelle la teneur finale en métaux alcalins est au moins 2 fois plus forte que la teneur en métaux alcalins de la zéolite de départ, de préférence au moins 5 fois plus forte que la teneur en métaux alcalins de la zéolite de départ.
9. Composition selon l'une quelconque des revendications précédentes, dans laquelle le métal alcalin final est choisi parmi le sodium, le potassium et leurs mélanges et sa teneur est comprise entre 0,1 et 1,4 pour cent en poids, calculée en métal.
- 30 10. Composition selon la revendication 9, dans laquelle le métal alcalin est du sodium et est présent à raison de 0,1 à 0,8 pour cent en poids, en calculant en métal.
- 35 11. Composition selon l'une quelconque des revendications précédentes, dans laquelle le catalyseur comprend en outre un liant choisi dans le groupe constitué par l'alumine, la silice, la silice-alumine, les aluminophosphates, les silicoaluminophosphates, la magnésie, l'oxyde de titane, la zircone, les argiles et leurs mélanges.
- 40 12. Composition selon l'une quelconque des revendications précédentes, dans laquelle le ou les métaux nobles du groupe VIII sont choisis dans le groupe constitué par le platine, le palladium et leurs mélanges, de préférence dans une quantité de 0,05 à 3 pour cent en poids par rapport au catalyseur total.
- 45 13. Composition selon la revendication 12, dans laquelle le ou les métaux nobles du groupe VIII comprennent du platine et du palladium dans un rapport molaire du platine au palladium compris entre 0,1 et 1,0.
- 50 14. Composition selon l'une quelconque des revendications précédentes qui a été calcinée à une température comprise entre 200 °C et 850 °C, de préférence entre 250 °C et 600 °C.
15. La composition selon la revendication 14 qui de plus a été activée à une température comprise entre 150 °C et 600 °C, de préférence entre 200 °C et 500 °C, dans une atmosphère contenant de l'hydrogène à une pression élevée, de préférence supérieure à 30 bars.
- 55 16. Un procédé pour l'hydrogénation de charges de départ hydrocarbonées contenant des composants hydrogénables et qui ont un intervalle d'ébullition compris entre 125 °C et 625 °C, lequel procédé



comprend la mise en contact de la charge de départ à une température comprise entre 150 °C et 400 °C avec de l'hydrogène et un catalyseur selon l'une quelconque des revendications 1-15.

- 5 17. Procédé selon la revendication 16, dans lequel la température de contact est comprise entre 200 °C et 400 °C, de préférence entre 225 °C et 375 °C.
18. Procédé selon la revendication 16 ou 17, dans lequel le catalyseur a été calciné à une température comprise entre 200 °C et 850 °C avant son utilisation dans le procédé.
- 10 19. Procédé selon la revendication 18, dans lequel le catalyseur a en outre été activé à une température comprise entre 150 °C et 600 °C, de préférence entre 200 °C et 500 °C dans une atmosphère contenant de l'hydrogène à une pression élevée, de préférence supérieure à 30 bars avant son utilisation dans le procédé.
- 15 20. Procédé selon l'une quelconque des revendications 16-19, dans lequel une charge de départ bouillant dans l'intervalle des combustibles pour moteurs Diesel contenant des composés aromatiques et des oléfines est hydrogénée pour donner un produit ayant un intervalle d'ébullition compris entre 125 °C et 425 °C, où les oléfines et les composés aromatiques sont substantiellement transformés en composés saturés et les composés aromatiques saturés sont transformés au moins en partie en composés acycliques, l'hydrocraquage de la charge de départ étant substantiellement absente.
- 20 21. Un procédé pour l'hydrotraitement et l'hydrogénation combinés de charges de départ hydrocarbonées contenant des impuretés contenant de l'azote et du soufre, des composés aromatiques et des oléfines, ces charges de départ ayant un intervalle d'ébullition compris entre 125 °C et 625 °C, lequel procédé comprend :
- 25 (a) le passage de la charge de départ à un premier étage dans lequel cette charge de départ est mise en contact avec de l'hydrogène et un catalyseur d'hydrotraitement comprenant un sulfure d'un métal du groupe VIII choisi parmi le nickel, le cobalt et leurs mélanges et un sulfure d'un métal du groupe VIB choisi parmi le molybdène, le tungstène et leurs mélanges déposés sur un support à
- 30 une température comprise entre 20 °C et 425 °C,
- (b) le strippage des gaz du produit sortant du premier étage afin de donner un produit liquide,
- (c) le passage de ce produit liquide à un second étage dans lequel le produit liquide est mis en contact avec de l'hydrogène et un catalyseur d'hydrogénation selon l'une quelconque des revendications 1-15,
- 35 (d) le strippage des gaz du produit sortant du second étage pour donner un produit liquide final.
22. Le procédé selon la revendication 21 dans lequel la pression partielle d'hydrogène dans les deux étages est comprise entre 35 et 100 bars.
- 40 23. Le procédé selon la revendication 22 dans lequel la pression partielle d'hydrogène dans le second étage est comprise entre 35 et 70 bars.

Revendications pour l'Etat contractant suivant : ES

- 45 1. Procédé pour la préparation d'une composition catalytique comprenant un ou plusieurs métaux nobles du groupe VIII sur un support, procédé selon lequel on dépose le ou les métaux du groupe VIII sur un support comprenant une zéolite pouvant être obtenue en mettant en contact une zéolite de départ du type Y ayant une dimension des cellules unitaires de moins de 24,65 angströms, un rapport molaire de la silice à l'alumine supérieur à 5 et une somme des teneurs en métaux alcalins et alcalino-terreux de
- 50 moins de 0,3 pour cent en poids (en calculant en métaux) avec au moins une solution aqueuse comprenant des ions de métaux alcalins et/ou des ions de métaux alcalino-terreux dans des conditions suffisantes pour fournir une somme accrue des teneurs finales en métaux alcalins et alcalino-terreux (mesurées en équivalents-grammes par gramme de zéolite), laquelle somme des teneurs finales en
- 55 métaux alcalins et alcalino-terreux est au moins 1,5 fois plus grande que la somme des teneurs en métaux alcalins et alcalino-terreux de la zéolite de départ (mesurées en équivalents-grammes par gramme de zéolite).



2. Le procédé selon la revendication 1 dans lequel la zéolite de type Y de départ a une dimension des cellules unitaires de moins de 24,4 angströms, de préférence moins de 24,35 angströms.
3. Le procédé selon la revendication 1 ou 2 dans lequel la zéolite de départ a une dimension des cellules unitaires comprise entre 24,2 et 24,3 angströms, de préférence entre 24,22 et 24,28 angströms.
4. Procédé selon l'une quelconque des revendications précédentes, dans lequel la zéolite de type Y de départ a un rapport molaire de la silice à l'alumine supérieur à 25, de préférence supérieur à 35.
5. Procédé selon la revendication 4, dans lequel la zéolite de type Y de départ a un rapport molaire de la silice à l'alumine supérieur à 50, de préférence supérieur à 60.
6. Procédé selon l'une quelconque des revendications précédentes, dans lequel la zéolite de type Y de départ a une teneur en métaux alcalins et alcalino-terreux de moins de 0,15 pour cent en poids, de préférence moins de 0,075 pour cent en poids.
7. Procédé selon l'une quelconque des revendications précédentes, dans lequel la zéolite de type Y de départ comprend un métal alcalin.
8. Procédé selon l'une quelconque des revendications précédentes, dans lequel la teneur finale en métaux alcalins et alcalino-terreux est au moins 2 fois plus forte que la teneur en métaux alcalins et alcalino-terreux de la zéolite de départ, de préférence au moins 5 fois plus forte.
9. Procédé selon l'une quelconque des revendications précédentes, dans lequel le métal alcalin et/ou alcalino-terreux final est choisi parmi le sodium, le potassium et leurs mélanges et est présent à raison de 0,1 à 1,4 pour cent en poids, en calculant en métal.
10. Procédé selon la revendication 9, dans lequel le métal alcalin et/ou alcalino terreux est le sodium et est présent à raison de 0,1 à 0,8 pour cent en poids, en calculant en métal.
11. Procédé selon l'une quelconque des revendications précédentes, dans lequel le catalyseur comprend en outre un liant choisi dans le groupe constitué par l'alumine, la silice, la silice-alumine, les aluminophosphates, les silicoaluminophosphates, la magnésie, l'oxyde de titane, la zircone, les argiles et leurs mélanges.
12. Procédé selon l'une quelconque des revendications précédentes, dans lequel le ou les métaux nobles du groupe VIII sont choisis dans le groupe constitué par le platine, le palladium et leurs mélanges, de préférence dans une quantité comprise entre 0,05 et 3 pour cent en poids par rapport au catalyseur total.
13. Procédé selon la revendication 12, dans lequel le ou les métaux nobles du groupe VIII comprennent du platine et du palladium dans un rapport molaire du platine au palladium compris entre 0,1 et 1,0.
14. Procédé selon l'une quelconque des revendications précédentes, dans lequel la composition catalytique est calcinée à une température comprise entre 200 °C et 850 °C, de préférence entre 250 °C et 600 °C.
15. Le procédé selon la revendication 14 dans lequel la composition catalytique est en outre activée à une température comprise entre 150 °C et 600 °C, de préférence entre 200 °C et 500 °C, dans une atmosphère contenant de l'hydrogène à une pression élevée, de préférence supérieure à 30 bars.
16. Un procédé pour l'hydrogénation de charges de départ hydrocarbonées contenant des composants hydrogénables et qui ont un intervalle d'ébullition compris entre 125 °C et 625 °C, lequel procédé comprend la mise en contact de la charge de départ à une température comprise entre 150 °C et 400 °C avec de l'hydrogène et un catalyseur selon l'une quelconque des revendications 1-15.
17. Procédé selon la revendication 16, dans lequel la température de contact est comprise entre 200 °C et 400 °C, de préférence entre 225 °C et 375 °C.



18. Procédé selon la revendication 16 ou 17, dans lequel le catalyseur a été calciné à une température comprise entre 200 °C et 850 °C avant son utilisation dans le procédé.
19. Procédé selon la revendication 18, dans lequel en outre le catalyseur a été activé à une température comprise entre 150 °C et 600 °C, de préférence entre 200 °C et 500 °C dans une atmosphère contenant de l'hydrogène à une pression élevée, de préférence supérieure à 30 bars, avant son utilisation dans le procédé.
20. Procédé selon l'une quelconque des revendications 16-19, dans lequel une charge de départ bouillant dans l'intervalle des combustibles pour moteurs Diesel qui contient des composés aromatiques et des oléfines est hydrogénée pour donner un produit ayant un intervalle d'ébullition compris entre 125 °C et 425 °C dans lequel les oléfines et les composés aromatiques sont substantiellement transformés en composés saturés et les composés aromatiques saturés sont transformés au moins en partie en composés acycliques, l'hydrocraquage de la charge de départ étant substantiellement absent.
21. Un procédé pour l'hydrotraitement et l'hydrogénation combinés de charges de départ hydrocarbonées contenant des impuretés contenant de l'azote et du soufre, des composés aromatiques et des oléfines, ces charges de départ ayant un intervalle d'ébullition compris entre 125 °C et 625 °C, lequel procédé comprend :
- (a) le passage de la charge de départ à un premier étage dans lequel cette charge de départ est mise en contact avec de l'hydrogène et un catalyseur d'hydrotraitement comprenant un sulfure d'un métal du groupe VIII choisi parmi le nickel, le cobalt et leurs mélanges et un sulfure d'un métal du groupe VIB choisi parmi le molybdène, le tungstène et leurs mélanges déposés sur un support à une température comprise entre 20 °C et 425 °C,
 - (b) le stripping des gaz du produit sortant du premier étage afin de donner un produit liquide,
 - (c) le passage de ce produit liquide à un second étage dans lequel le produit liquide est mis en contact avec de l'hydrogène et un catalyseur d'hydrogénation selon l'une quelconque des revendications 1-15,
 - (d) le stripping des gaz du produit sortant du second étage pour donner un produit liquide final.
22. Le procédé selon la revendication 21 dans lequel la pression partielle d'hydrogène dans les deux étages est comprise entre 35 et 100 bars.
23. Le procédé selon la revendication 22 dans lequel la pression partielle d'hydrogène dans le second étage est comprise entre 35 et 70 bars.