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Gasoline Sulfur Reduction in Fluid Catalytic Cracking

This invention relates to the reduction of sulfur in gasolines and other petroleum products produced by the catalytic cracking process. The invention provides a catalytic composition for reducing product sulfur and a process for reducing the product sulfur using this composition.

Catalytic cracking is a petroleum refining process which is applied commercially on a very large scale, especially in the United States where the majority of the refinery gasoline blending pool is produced by catalytic cracking, with almost all of this coming from the fluid catalytic cracking (FCC) process. In the catalytic cracking process heavy hydrocarbon fractions are converted into lighter products by reactions taking place at elevated temperature in the presence of a catalyst, with the majority of the conversion or cracking occurring in the vapor phase. The feedstock is so converted into gasoline, distillate and other liquid cracking products as well as lighter gaseous cracking products of four or less carbon atoms per molecule. The gas partly consists of olefins and partly of saturated hydrocarbons.

During the cracking reactions some heavy material, known as coke, is deposited onto the catalyst. This reduces its catalytic activity and regeneration is desired. After removal of occluded hydrocarbons from the spent cracking catalyst, regeneration is accomplished by burning off the coke and then the catalyst activity is restored. The three characteristic steps of the catalytic cracking can be therefore be distinguished: a cracking step in which the hydrocarbons are converted into lighter products, a stripping step to remove hydrocarbons adsorbed on the catalyst and a regeneration step to burn off coke from the catalyst. The regenerated catalyst is then reused in the cracking step.

Catalytic cracking feedstocks normally contain sulfur in the form of organic sulfur compounds such as mercaptans, sulfides and thiophenes. The products of the cracking process correspondingly tend to contain sulfur impurities even though about half of the sulfur is converted to hydrogen sulfide during the cracking process, mainly by catalytic decomposition of non-thiophenic sulfur compounds. The distribution of

sulfur in the cracking products is dependent on a number of factors including feed, catalyst type, additives present, conversion and other operating conditions but, in any event a certain proportion of the sulfur tends to enter the light or heavy gasoline fractions and passes over into the product pool. With increasing environmental regulation being applied to petroleum products, for example in the Reformulated Gasoline (RFG) regulations, the sulfur content of the products has generally been decreased in response to concerns about the emissions of sulfur oxides and other sulfur compounds into the air following combustion processes. Reduction of gasoline sulfur is critical not only for the SO_x emission but also for the sulfur poisoning of automobile catalytic converters. The poisoning of catalytic converters will cause other emission problems such as NO_x .

The environmental concerns have focussed extensively on the sulfur content of motor gasoline in view of its pre-eminent position as a vehicle fuel for passenger cars in the United States. These concerns have, however, also extended to the higher boiling distillate fractions including the light cycle oil (LCO) and fuel oil fractions (light fuel oil, LFO, and heavy fuel oil, HFO) obtained from the catalytic cracking process. With these products, hydrodesulfurization has long been used to reduce the levels of sulfur in the product fractions and, in general this has proved effective. The higher boiling fractions are not, however, as amenable to desulfurization as the lower boiling fractions because of the increasingly refractory nature of the sulfur compounds, especially the substituted benzothiophenes with increasing boiling point. In the LCO hydrodesulfurization process, methyl and/or alkyl substitution of benzothiophene and dibenzothiophene makes desulfurization reactivity of the organic sulfurs decline substantially and they become "hard sulfur" or "refractory sulfur". An example of LCO sulfur GC is shown in Figure 1 with sulfur GC speciation. A review by Girgis and Gates, *Ind. Eng. Chem.*, 30, 1991, 2021-2058, reported that substitution of a methyl group into the 4- position or into the 4- and 6- positions decreases the desulfurization activity by an order of magnitude. Houalla et al. reported activity debits of 1-10 orders of magnitude, M. Houalla et al., *Journal of Catalysis*, 61, 1980, 523-527. Lamure-Meille et al. suggested that steric hindrance of the alkyl group causes the low reactivity of methyl substituted dibenzothiophenes Lamure-Meille et al., *Applied Catalysis A: General* 131, 1995, 143-157. Given the utilization of the higher boiling

catalytic cracking products as feeds for hydrodesulfurization processes, the incentive to reduce the incidence of the more refractory organic sulfur compounds in these cracking fractions becomes apparent.

- 5 One approach has been to remove the sulfur from the FCC feed by hydrotreating before cracking is initiated. While highly effective, this approach tends to be expensive in terms of the capital cost of the equipment as well as operationally since hydrogen consumption is high. Another approach has been to remove the sulfur from the cracked products by hydrotreating. Again, while effective, this solution has the
10 drawback that valuable product octane may be lost when the high-octane olefins are saturated.

From the economic point of view, it would be desirable to achieve sulfur removal in the cracking process itself since this would effectively desulfurize the major component of
15 the gasoline blending pool without additional treatment. Various catalytic materials have been developed for the removal of sulfur during the FCC process cycle but, so far, most developments have centered on the removal of sulfur from the regenerator stack gases. An early approach developed by Chevron used alumina compounds as additives to the inventory of cracking catalyst to adsorb sulfur oxides in the FCC
20 regenerator; the adsorbed sulfur compounds which entered the process in the feed were released as hydrogen sulfide during the cracking portion of the cycle and passed to the product recovery section of the unit where they were removed. See Krishna et al, *Additives Improve FCC Process*, Hydrocarbon Processing, November 1991, pages 59-66. The sulfur is removed from the stack gases from the regenerator but product
25 sulfur levels are not greatly affected, if at all.

An alternative technology for the removal of sulfur oxides from regenerator removal is based on the use of magnesium-aluminum spinels as additives to the circulating catalyst inventory in the FCCU. Under the designation DESOX™ used for the
30 additives in this process, the technology has achieved a notable commercial success. Exemplary patents on this type of sulfur removal additive include U.S. 4,963,520; 4,957,892; 4,957,718; 4,790,982 and others. Again, however, product sulfur levels are not greatly reduced.

A catalyst additive for the reduction of sulfur levels in the liquid cracking products is proposed by Wormsbecher and Kim in U.S. Patents 5,376,608 and 5,525,210, using a cracking catalyst additive of an alumina-supported Lewis acid for the production of reduced-sulfur gasoline but this system has not achieved significant commercial success. The need for an effective additive for reducing the sulfur content of liquid catalytic cracking products has therefore persisted.

In Singapore Patent Application Serial No. 9904234-3, filed 31 August 1999 (Mobil IP Case 10061), we have described catalytic materials for use in the catalytic cracking process which are capable of reducing the sulfur content of the liquid products of the cracking process. These sulfur reduction catalysts comprise, in addition to a porous molecular sieve component, a metal in an oxidation state above zero within the interior of the pore structure of the sieve. The molecular sieve is in most cases a zeolite and it may be a zeolite having characteristics consistent with the large pore zeolites such as zeolite beta or zeolite USY or with the intermediate pore size zeolites such as ZSM-5. Non-zeolitic molecular sieves such as MeAPO-5, MeAPSO-5, as well as the mesoporous crystalline materials such as MCM-41 may be used as the sieve component of the catalyst. Metals such as vanadium, zinc, iron, cobalt, and gallium were found to be effective for the reduction of sulfur in the gasoline, with vanadium being the preferred metal. When used as a separate particle additive catalyst, these materials are used in combination with the active catalytic cracking catalyst (normally a faujasite such as zeolite Y, especially as zeolite USY) to process hydrocarbon feedstocks in the fluid catalytic cracking (FCC) unit to produce low-sulfur products. Since the sieve component of the sulfur reduction catalyst may itself be an active cracking catalyst, for instance, zeolite USY, it is also possible to use the sulfur reduction catalyst in the form of an integrated cracking/sulfur reduction catalyst system, for example, comprising USY as the active cracking component and the sieve component of the sulfur reduction system together with added matrix material such as silica, clay and the metal, e.g. vanadium, which provides the sulfur reduction functionality.

Another consideration in the manufacture of FCC catalysts has been catalyst stability, especially hydrothermal stability since cracking catalysts are exposed during use to repeated cycles of reduction (in the cracking step) followed by stripping with steam and then by oxidative regeneration which produces large amounts of steam from the combustion of the coke, a carbon-rich hydrocarbon, which is deposited on the catalyst particles during the cracking portion of the cycle. Early in the development of zeolitic cracking catalysts it was found that a low sodium content was required not only for optimum cracking activity but also for stability and that the rare earth elements such as cerium and lanthanum conferred greater hydrothermal stability. See, for example, *Fluid Catalytic Cracking with Zeolite Catalysts*, Venuto et al., Marcel Dekker, New York, 1979, ISBN 0-8247-6870-1.

We have now developed catalytic materials for use in the catalytic cracking process which are capable of improving the reduction in the sulfur content of the liquid products of the cracking process including, in particular, the gasoline and middle distillate cracking fractions. The present sulfur reduction catalyst are similar to the ones described in U.S. Application No. 09/144,607 in that a metal component in an oxidation state above zero is present in the pore structure of a molecular sieve component of the catalyst composition, with preference again being given to vanadium. In the present case, however, the composition also comprises one or more rare earth elements, preferably cerium. We have found that the presence of the rare earth component enhances the stability of the catalyst, as compared to the catalysts which contain only vanadium or another metal component and that in certain favorable cases, especially with cerium as the rare earth component, the sulfur reduction activity is also increased by the presence of the rare earth elements. This is surprising since the rare earth cations in themselves have no sulfur reduction activity.

The present sulfur reduction catalysts may be used in the form of an additive catalyst in combination with the active cracking catalyst in the cracking unit, that is, in combination with the conventional major component of the circulating cracking catalyst inventory which is usually a matrixed, zeolite containing catalyst based on a faujasite zeolite, usually zeolite Y. Alternatively, they may be used in the form of an integrated cracking/ product sulfur reduction catalyst system.

According to the present invention, the sulfur removal catalyst composition comprises a porous molecular sieve that contains (i) a metal in an oxidation state above zero within the interior of the pore structure of the sieve and (ii) a rare earth component.

- 5 The molecular sieve is in most cases a zeolite and it may be a zeolite having characteristics consistent with the large pore zeolites such as zeolite beta or zeolite USY or with the intermediate pore size zeolites such as ZSM-5. Non-zeolitic molecular sieves such as MeAPO-5, MeAPSO-5, as well as the mesoporous crystalline materials such as MCM-41 may be used as the sieve component of the catalyst.
- 10 Metals such as vanadium, zinc, iron, cobalt, and gallium are effective. If the selected sieve material has sufficient cracking activity, it may be used as the active catalytic cracking catalyst component (normally a faujasite such as zeolite Y) or, alternatively, it may be used in addition to the active cracking component, whether or not it has any cracking activity of itself.

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- The present compositions are useful to process hydrocarbon feedstocks in fluid catalytic cracking (FCC) units to produce low-sulfur gasoline and other liquid products, for example, light cycle oil that can be used as a low sulfur diesel blend component or as heating oil. Besides achieving a very significant reduction in the sulfur content of
- 20 the cracked gasoline fraction, the present sulfur reduction catalyst materials enable the sulfur levels of light cycle oil and fuel oil products (light fuel oil, heavy fuel oil) to be reduced. The sulfur reduction in LCO occurs with the predominantly substituted-benzothiophenes and substituted-dibenzothiophenes; the removal of these more refractory species will improve the efficiency of sulfur reduction in subsequent LCO
- 25 hydrodesulfurization process. The reduction in HFO sulfur may allow upgrading of coker products from the oil to premium coke.

- While the mechanism by which the metal-containing zeolite catalyst compositions remove the sulfur components normally present in cracked hydrocarbon products is
- 30 not precisely understood, it does involve the conversion of organic sulfur compounds in the feed to inorganic sulfur so that the process is a true catalytic process. In this process, it is believed that a zeolite or other molecular sieve provides shape selectivity

with varying pore size, and the metal sites in zeolite provide adsorption sites for the sulfur species.

The drawings are graphs which show the performance of the present sulfur reduction compositions as described below.

FCC Process

The present sulfur removal catalysts are used as a catalytic component of the circulating inventory of catalyst in the catalytic cracking process, which these days is almost invariably the fluid catalytic cracking (FCC) process. For convenience, the invention will be described with reference to the FCC process although the present additives could be used in the older moving bed type (TCC) cracking process with appropriate adjustments in particle size to suit the requirements of the process. Apart from the addition of the present additive to the catalyst inventory and some possible changes in the product recovery section, discussed below, the manner of operating the process will remain unchanged. Thus, conventional FCC catalysts may be used, for example, zeolite based catalysts with a faujasite cracking component as described in the seminal review by Venuto and Habib, *Fluid Catalytic Cracking with Zeolite Catalysts*, Marcel Dekker, New York 1979, ISBN 0-8247-6870-1 as well as in numerous other sources such as Sadeghbeigi, *Fluid Catalytic Cracking Handbook*, Gulf Publ. Co. Houston, 1995, ISBN 0-88415-290-1.

Somewhat briefly, the fluid catalytic cracking process in which the heavy hydrocarbon feed containing the organosulfur compounds will be cracked to lighter products takes place by contact of the feed in a cyclic catalyst recirculation cracking process with a circulating fluidizable catalytic cracking catalyst inventory consisting of particles having a size ranging from 20 to 100 microns. The significant steps in the cyclic process are:

(i) the feed is catalytically cracked in a catalytic cracking zone, normally a riser cracking zone, operating at catalytic cracking conditions by contacting feed with a source of hot, regenerated cracking catalyst to produce an effluent comprising cracked products and spent catalyst containing coke and strippable hydrocarbons;

(ii) the effluent is discharged and separated, normally in one or more cyclones, into a vapor phase rich in cracked product and a solids rich phase comprising the spent catalyst;

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(iii) the vapor phase is removed as product and fractionated in the FCC main column and its associated side columns to form liquid cracking products including gasoline,

(iv) the spent catalyst is stripped, usually with steam, to remove occluded hydrocarbons from the catalyst, after which the stripped catalyst is oxidatively regenerated to produce hot, regenerated catalyst which is then recycled to the cracking zone for cracking further quantities of feed.

The feed to the FCC process will typically be a high boiling feed of mineral oil origin, normally with an initial boiling point of at least 290°C (550°F) and in most cases above 315° C (600°F). Most refinery cut points for FCC feed will be at least 345°C (650°F). The end point will vary, depending on the exact character of the feed or on the operating characteristics of the refinery. Feeds may be either distillate, typically with an end point of 550°C (1020°F) or higher, for example, 590°C (1095°F) or 620°C (1150°F) or, alternatively, resid (non-distillable) material may be included in the feed and may even comprise all or a major proportion of the feed. Distillable feeds include virgin feeds such as gas oils, e.g. heavy or light atmospheric gas oil, heavy or light vacuum gas oil as well as cracked feeds such as light coker gas oil, heavy coker gas oil. Hydrotreated feeds may be used, for example, hydrotreated gas oils, especially hydrotreated heavy gas oil but since the present catalysts are able to effect a considerable reduction in sulfur, it may be possible to dispense with initial hydrotreatment where its objective is to reduce sulfur although improvements in crackability will still be achieved.

In the present process, the sulfur content of the gasoline portion of the liquid cracking products, is effectively brought to lower and more acceptable levels by carrying out the catalytic cracking in the presence of the sulfur reduction catalyst.

FCC Cracking Catalyst

The present sulfur reduction catalyst compositions may be used in the form of a separate particle additive which is added to the main cracking catalyst in the FCCU or, alternatively, they may be used as components of the cracking catalyst to provide an integrated cracking/sulfur reduction catalyst system. The cracking component of the catalyst which is conventionally present to effect the desired cracking reactions and the production of lower boiling cracking products, is normally based on a faujasite zeolite active cracking component, which is conventionally zeolite Y in one of its forms such as calcined rare-earth exchanged type Y zeolite (CREY), the preparation of which is disclosed in U.S. Patent No. 3,402,996, ultrastable type Y zeolite (USY) as disclosed in U.S. Patent No. 3,293,192, as well as various partially exchanged type Y zeolites as disclosed in U.S. Patents Nos. 3,607,043 and 3,676,368. Cracking catalysts such as these are widely available in large quantities from various commercial suppliers. The active cracking component is routinely combined with a matrix material such as silica or alumina as well as a clay in order to provide the desired mechanical characteristics (attrition resistance etc.) as well as activity control for the very active zeolite component or components. The particle size of the cracking catalyst is typically in the range of 10 to 100 microns for effective fluidization. If used as a separate particle additive, the sulfur reduction catalyst (and any other additive) is normally selected to have a particle size and density comparable to that of the cracking catalyst so as to prevent component separation during the cracking cycle.

Sulfur Reduction System - Sieve Component

According to the present invention, the sulfur removal catalyst comprises a porous molecular sieve which contains a metal in an oxidation state above zero within the interior of the pore structure of the sieve. The molecular sieve is in most cases a zeolite and it may be a zeolite having characteristics consistent with the large pore zeolites such as zeolite Y, preferably as zeolite USY, or zeolite beta or with the intermediate pore size zeolites such as ZSM-5, with the former class being preferred.

The molecular sieve component of the present sulfur reduction catalysts may, as noted above, be a zeolite or a non-zeolitic molecular sieve. When used, zeolites may be selected from the large pore size zeolites or intermediate pore zeolites (see *Shape Selective Catalysis in Industrial Applications*, Chen et al, Marcel Dekker Inc., New York 1989, ISBN 0-8247-7856-1, for a discussion of zeolite classifications by pore size according to the basic scheme set out by Frillette et al in *J. Catalysis* 67, 218-222 (1981)). The small pore size zeolites such as zeolite A and erionite, besides having insufficient stability for use in the catalytic cracking process, will generally not be preferred because of their molecular size exclusion properties which will tend to exclude the components of the cracking feed as well as many components of the cracked products. The pore size of the sieve does not, however, appear to be critical since, as shown below, both medium and large pore size zeolites have been found to be effective, as have the mesoporous crystalline materials such as MCM-41.

Zeolites having properties consistent with the existence of a large pore (12 ring) structure which may be used to make the present sulfur reduction catalysts include zeolites Y in its various forms such as Y, REY, CREY, USY, of which the last is preferred, as well as other zeolites such as zeolite L, zeolite beta, mordenite including de-aluminated mordenite, and zeolite ZSM-18. Generally, the large pore size zeolites are characterized by a pore structure with a ring opening of at least 0.7 nm and the medium or intermediate pore size zeolites will have a pore opening smaller than 0.7 nm but larger than 0.56 nm. Suitable medium pore size zeolites which may be used include the pentasil zeolites such as ZSM-5, ZSM-22, ZSM-23, ZSM-35, ZSM-50, ZSM-57, MCM-22, MCM-49, MCM-56 all of which are known materials. Zeolites may be used with framework metal elements other than aluminum, for example, boron, gallium, iron, chromium.

The use of zeolite USY is particularly desirable since this zeolite is typically used as the active cracking component of the cracking catalyst and it is therefore possible to use the sulfur reduction catalyst in the form of an integrated cracking/sulfur reduction catalyst system. The USY zeolite used for the cracking component may also, to advantage, be used as the sieve component for a separate particle additive catalyst as it will continue to contribute to the cracking activity of the overall catalyst present in

the unit. Stability is correlated with low unit cell size with USY and, for optimum results, the UCS for the USY zeolite in the finished catalyst should be from 2.420 to 2.460 nm, preferably 2.420 to 2.455 with the range of 2.420 to 2.445 nm, preferably 2.435 to 2.440 nm, being very suitable. After exposure to the repeated steaming of the FCC cycles, further reductions in UCS will take place to a final value which is normally within the range of 2.420 to 2.430 nm.

In addition to the zeolites, other molecular sieves may be used although they may not be as favorable since it appears that some acidic activity (conventionally measured by the alpha value) is required for optimum performance. Experimental data indicate that alpha values in excess of 10 (sieve without metal content) are suitable for adequate desulfurization activity, with alpha values in the range of 0.2 to 2,000 being normally suitable¹. Alpha values from 0.2 to 300 represent the normal range of acidic activity for these materials when used as additives.

Exemplary non-zeolitic sieve materials which may provide suitable support components for the metal component of the present sulfur reduction catalysts include silicates (such as the metasilicates and titanosilicates) of varying silica-alumina ratios, metalloaluminates (such as germaniumaluminates), metallophosphates, aluminophosphates such as the silico- and metalloaluminophosphates referred to as metal integrated aluminophosphates (MeAPO and ELAPO), metal integrated silicoaluminophosphates (MeAPSO and ELAPSO), silicoaluminophosphates (SAPO), gallogermanates and combinations of these. A discussion on the structural relationships of SAPO's, AIPO's, MeAPO's, and MeAPSO's may be found in a number of resources including *Stud. Surf. Catal.* 37 13 - 27 (1987). The AIPO's contain aluminum and phosphorus, whilst in the SAPO's some of the phosphorus and/or some of both phosphorus and aluminum is replaced by silicon. In the MeAPO's various metals are present, such as Li, B, Be, Mg, Ti, Mn, Fe, Co, An, Ga, Ge, and As, in addition to aluminum and phosphorus, whilst the MeAPSO's additionally contain silicon. The negative charge of the $\text{Me}_a\text{Al}_b\text{P}_c\text{Si}_d\text{O}_e$ lattice is compensated by cations,

¹ The alpha test is a convenient method of measuring the overall acidity, inclusive of both its internal and external acidity, of a solid material such as a molecular sieve. The test is described in U.S. Pat. No. 3,354,078; in the *Journal of Catalysis*, Vol. 4, p. 527 (1965); Vol. 6, p. 278 (1966); and

where Me is magnesium, manganese, cobalt, iron and/or zinc. Me_xAPSO's are described in U.S. Pat. No. 4,793,984. SAPO-type sieve materials are described in U.S. Pat. No. 4,440,871; MeAPO type catalysts are described in U.S. Pat. Nos. 4,544,143 and 4,567,029; ELAPO catalysts are described in U.S. Pat. No. 4,500,651, and ELAPSO catalysts are described in European Patent Application 159,624. Specific molecular sieves are described, for example, in the following patents: MgAPSO or MAPSO-U.S. Pat. No. 4,758,419. MnAPSO-U.S. Pat. No. 4,686,092; CoAPSO-U.S. Pat. No. 4,744,970; FeAPSO-U.S. Pat. No. 4,683,217 and ZnAPSO U.S. Pat. No. 4,935,216. Specific silicoaluminophosphates which may be used include SAPO-11, SAPO-17, SAPO-34, SAPO-37; other specific sieve materials include MeAPO-5, MeAPSO-5.

Another class of crystalline support materials which may be used is the group of mesoporous crystalline materials exemplified by the MCM-41 and MCM-48 materials.

These mesoporous crystalline materials are described in U.S. Patents Nos. 5,098,684; 5,102,643; and 5,198,203. MCM-41, which is described in U.S. 5,098,684, is characterized by a microstructure with a uniform, hexagonal arrangement of pores with diameters of at least 1.3 nm: after calcination it exhibits an X-ray diffraction pattern with at least one d-spacing greater than 1.8 nm and a hexagonal electron diffraction pattern that can be indexed with a d100 value greater than 1.8 nm which corresponds to the d-spacing of the peak in the X-ray diffraction pattern. The preferred catalytic form of this material is the aluminosilicate although other metasilicates may also be utilized. MCM-48 has a cubic structure and may be made by a similar preparative procedure.

Metal Components

Two metal components are incorporated into the molecular sieve support material to make up the present catalytic compositions. One component is a rare earth such as lanthanum or a mixture of rare earth elements such as cerium and lanthanum. The other metal component can be regarded as the primary sulfur reduction component

although the manner in which it effects sulfur reduction is not clear, as discussed in Application Serial No. 09/144,607, to which reference is made for a description of sulfur reduction catalyst compositions containing vanadium and other metal components effective for this purpose. For convenience this component of the composition will be referred to in this application as the primary sulfur reduction component. In order to be effective, this metal (or metals) should be present inside the pore structure of the sieve component. Metal-containing zeolites and other molecular sieves can be prepared by (1) post-addition of metals to the sieve or to a catalyst containing the sieve(s), (2) synthesis of the sieve(s) containing metal atoms in the framework structure, and by (3) synthesis of the sieve(s) with trapped, bulky metal ions in the zeolite pores. Following addition of the metal component, washing to remove unbound ionic species and drying and calcination should be performed. These techniques are all known in themselves. Post-addition of the metal ions is preferred for simplicity and economy, permitting available sieve materials to be converted to use for the present additives. A wide variety of post-addition methods of metals can be used to produce a catalyst of our invention, for example, aqueous exchange of metal ions, solid-state exchange using metal halide salt(s), impregnation with a metal salt solution, and vapor deposition of metals. In each case, however, it is important to carry out the metal(s) addition so that the metal component enters the pore structure of the sieve component.

It has been found that when the metal of the primary sulfur reduction component is present as exchanged cationic species in the pores of the sieve component, the hydrogen transfer activity of the metal component is reduced to the point that hydrogen transfer reactions taking place during the cracking process will normally maintained at an acceptably low level with the preferred metal components. Thus, coke and light gas make during cracking increase slightly but they remain within tolerable limits. Since the unsaturated light ends can be used in any event as alkylation feed and in this way recycled to the gasoline pool, there is no significant loss of gasoline range hydrocarbons incurred by the use of the present additives.

Because of the concern for excessive coke and hydrogen make during the cracking process, the metals for incorporation into the additives should not exhibit

hydrogenation activity to a marked degree. For this reason, the noble metals such as platinum and palladium which possess strong hydrogenation-dehydrogenation functionality are not desirable. Base metals and combinations of base metals with strong hydrogenation functionality such as nickel, molybdenum, nickel-tungsten, cobalt-molybdenum and nickel-molybdenum are not desirable for the same reason. The preferred base metals are the metals of Period 4, Groups 5, 8, 9, 12, 13 (IUPAC classification, previously Groups VB, VIII, IIB, IIIA) of the Periodic Table. Vanadium, zinc, iron, cobalt, and gallium are effective with vanadium being the preferred metal component. It is surprising that vanadium can be used in this way in an FCC catalyst composition since vanadium is normally thought to have a very serious effect on zeolite cracking catalysts and much effort has been expended in developing vanadium suppressers. See, for example, Wormsbecher et al, *Vanadium Poisoning of Cracking Catalysts: Mechanism of Poisoning and Design of Vanadium Tolerant Catalyst System*, J. Catalysis 100, 130-137 (1986). It is believed that the location of the vanadium inside the pore structure of the sieve immobilizes the vanadium and prevents it from becoming vanadic acid species which can combine deleteriously with the sieve component; in any event, the present zeolite-based sulfur reduction catalysts containing vanadium as the metal component have undergone repeated cycling between reductive and oxidative/steaming conditions representative of the FCC cycle while retaining the characteristic zeolite structure, indicating a different environment for the metal.

Vanadium is particularly suitable for gasoline sulfur reduction when supported on zeolite USY. The yield structure of the V/USY sulfur reduction catalyst is particularly interesting. While other zeolites, after metals addition, demonstrate gasoline sulfur reduction, they tend to convert gasoline to C_3 and C_4 gas. Even though much of the converted C_3 and C_4 can be alkylated and re-blended back to the gasoline pool, the high C_4 wet gas yield may be a concern since many refineries are limited by their wet gas compressor capacity. The metal-containing USY has similar yield structure to current FCC catalysts; this advantage would allow the V/USY zeolite content in a catalyst blend to be adjusted to a target desulfurization level without limitation from FCC unit constraints. The vanadium on Y zeolite catalyst, with the zeolite represented by USY, is therefore a particularly favorable combination for gasoline sulfur reduction

in FCC. The USY which has been found to give particularly good results is a USY with a unit cell size in the range from 2.420 to 2.460 nm, preferably in the range 2.420 to 2.450, e.g. 2.435 to 2.450 nm (following treatment). Combinations of base metals such as vanadium/zinc as the primary sulfur reduction component may also be
5 favorable in terms of overall sulfur reduction.

The amount of the primary sulfur reduction metal component in the sulfur reduction catalyst is normally from 0.2 to 5 weight percent, typically 0.5 to 5 weight percent, (as metal, relative to weight of sieve component) but amounts outside this range, for
10 example, from 0.10 to 10 weight percent may still be found to give some sulfur removal effect. When the sieve is matrixed, the amount of the primary sulfur reduction metal component expressed relative to the total weight of the catalyst composition will, for practical purposes of formulation, typically extend from 0.1 to 5, more typically from 0.2 to 2 weight percent of the entire catalyst.

15 The second metal component of the sulfur reduction catalyst composition comprises a rare earth metal or metals which is present within the pore structure of the molecular sieve and is thought to be present in the form of cations exchanged onto the exchangeable sites present on the sieve component. The rare earth (RE) component
20 significantly improves the catalyst stability in the presence of vanadium. For example, higher cracking activity can be achieved with RE+V/USY catalyst compared to a V/USY catalyst, while comparable gasoline sulfur reduction is obtained. Rare earth elements of the lanthanide series from atomic number from 57 to 71 such as lanthanum, cerium, dysprosium, praseodymium, samarium, europium, gadolinium,
25 ytterbium and lutetium may be used in this way but from the point of view of commercial availability, lanthanum and mixtures of cerium and lanthanum will normally be preferred. Cerium has been found to be the most effective rare earth component from the viewpoint of sulfur reduction as well as catalyst stability and its use is therefore preferred although good results may also be achieved with other rare
30 earth elements, as shown below..

The amount of rare earth is typically from 1 to 10 wt. percent of the catalyst composition, in most cases from 2 to 5 wt. percent. Relative to the weight of the

sieve, the amount of the rare earth will normally be from 2 to 20 weight percent and in most cases from 4 to 10 weight percent of the sieve, depending on the sieve:matrix ratio. Cerium may be used in amounts from 0.1 to 10, normally 0.25 to 5, weight percent of the catalyst composition and relative to the sieve, will normally be from 0.2 to 20, in most cases from 0.5 to 10, weight percent.

The rare earth component can suitably be incorporated into the molecular sieve component by exchange onto the sieve, either in the form of the unmatrixed crystal or of the matrixed catalyst. When the composition is being formulated with the preferred USY zeolite sieve, a very effective manner of incorporation is to add the rare earth ions to the USY sieve (typically 2.445 – 2.465 nm unit cell size) followed by additional steam calcination to lower the unit cell size of the USY to a value typically in the range of 2.420 to 2.460 nm., after which the primary metal component may be added if not already present. The USY should have a low alkali metal (mainly sodium) content for stability as well as for satisfactory cracking activity; this will normally be secured by the ammonium exchange made during the ultrastabilization process to a desirable low sodium level of not more than 1 weight percent, preferably not more than 0.5 weight percent, on the sieve.

The metal components are incorporated into the catalyst composition in a way which ensures that they enter the interior pore structure of the sieve. The metals may be incorporated directly into the crystal or into the matrixed catalyst. When using the preferred USY zeolite as the sieve component, this can suitably be done as described above, by recalcining a USY cracking catalyst containing the rare earth component to ensure low unit cell size and then incorporating the metal, e.g. vanadium, by ion exchange or by impregnation under conditions which permit cation exchange to take place so that the metal ion is immobilized in the pore structure of the zeolite. Alternatively, the primary sulfur reduction component and the rare earth metal component can be incorporated into the sieve component, e.g. USY zeolite or ZSM-5 crystal, after any necessary calcination to remove organics from the synthesis after which the metal-containing component can be formulated into the finished catalyst composition by the addition of the cracking and matrix components and the formulation spray dried to form the final catalyst.

When the catalyst is being formulated as an integrated catalyst system, it is preferred to use the active cracking component of the catalyst as the sieve component of the sulfur reduction system, preferably in the form of a faujasite for example, zeolite USY, both for simplicity of manufacture but also for retention of controlled cracking properties. It is, however, possible to incorporate another active cracking sieve material such as zeolite ZSM-5 into an integrated catalyst system and such systems may be useful when the properties of the second active sieve material are desired, for instance, the properties of ZSM-5. The impregnation/exchange process should in both cases be carried out with a controlled amount of metal so that the requisite number of sites are left on the sieve to catalyze the cracking reactions which may be desired from the active cracking component or any secondary cracking components which are present, e.g. ZSM-5.

15 *Use of Sulfur Reduction Catalyst Composition*

Normally the most convenient manner to use the sulfur reduction catalyst will be as a separate particle additive to the catalyst inventory. In its preferred form, with zeolite USY as the sieve component, the addition of the catalyst additive to the total catalyst inventory of the unit will not result in significant reduction in overall cracking because of the cracking activity of the USY zeolite. The same is true when another active cracking material is used as the sieve component. When used in this way, the composition may be used in the form of the pure sieve crystal, pelleted (without matrix but with added metal components) to the correct size for FCC use. Normally, however, the metal-containing sieve will be matrixed in order to achieve adequate particle attrition resistance and to maintain satisfactory fluidization. Conventional cracking catalyst matrix materials such as alumina or silica-alumina, usually with added clay, will be suitable for this purpose. The amount of matrix relative to the sieve will normally be from 20:80 to 80:20 by weight. Conventional matrixing techniques may be used.

Use as a separate particle catalyst additive permits the ratio of sulfur reduction and cracking catalyst components to be optimized according to the amount of sulfur in the

feed and the desired degree of desulfurization; when used in this manner, it is typically used in an amount from 1 to 50 weight percent of the entire catalyst inventory in the FCCU; in most cases the amount will be from 5 to 25 weight percent, e.g. 5 to 15 weight percent. About 10 percent represents a norm for most practical purposes. The additive may be added in the conventional manner, with make-up catalyst to the regenerator or by any other convenient method. The additive remains active for sulfur removal for extended periods of time although very high sulfur feeds may result in loss of sulfur removal activity in shorter times.

The alternative to the use of the separate particle additive is to use the sulfur reduction catalyst incorporated into the cracking catalyst to form an integrated FCC cracking/gasoline sulfur reduction catalyst. If the sulfur reduction metal components are used in combination with a sieve other than the active cracking component, for example, on ZSM-5 or zeolite beta when the main active cracking component is USY, the amount of the sulfur reduction component (sieve plus metals) will typically be up to 25 weight percent of the entire catalyst or less, corresponding to the amounts in which it may be used as a separate particle additive, as described above.

Other catalytically active components may be present in the circulating inventory of catalytic material in addition to the cracking catalyst and the sulfur removal additive. Examples of such other materials include the octane enhancing catalysts based on zeolite ZSM-5, CO combustion promoters based on a supported noble metal such as platinum, stack gas desulfurization additives such as DESOX™ (magnesium aluminum spinel), vanadium traps and bottom cracking additives, such as those described in Krishna, Sadeghbeigi, *op cit* and Scherzer, *Octane Enhancing Zeolitic FCC Catalysts*, Marcel Dekker, New York, 1990, ISBN 0-8247-8399-9. These other components may be used in their conventional amounts.

The effect of the present additives is to reduce the sulfur content of the liquid cracking products, especially the light and heavy gasoline fractions although reductions are also achieved in the higher boiling distillate products including the light cycle oil and light and heavy fuel oil fractions, making this more amenable to hydrodesulfurization techniques by removal of the more refractory sulfur compounds. The distillate

fractions may then be hydrodesulfurized under less severe, more economical conditions to produce distillate products suitable for use as a diesel or home heating oil blend component.

- 5 The cracking process itself will be carried out in the normal manner with the addition of the sulfur reduction catalyst, either in the form of an additive or as an integrated catalytic cracking/sulfur reduction catalyst (single particle catalyst). Cracking conditions will be conventional in nature.
- 10 The sulfur removed during the cracking process by the use of the catalyst is converted to inorganic form and released as hydrogen sulfide which can be recovered in the normal way in the product recovery section of the FCCU in the same way as the hydrogen sulfide conventionally released in the cracking process. The increased load of hydrogen sulfide may impose additional sour gas/water treatment requirements but
- 15 with the significant reductions in gasoline sulfur achieved, these are not likely to be considered limiting.

Very significant reductions in cracked product sulfur can be achieved by the use of the present catalysts, in some cases up to 50% relative to the base case using a

- 20 conventional cracking catalyst, at constant conversion, using the preferred form of the catalyst described above. Gasoline sulfur reduction of 25 % is readily achievable with many of the additives according to the invention, as shown by the Examples below. For the middle distillate fractions including the LCO fraction, reductions of up to 25 percent may also be achieved, as shown below in the Examples, coupled with a
- 25 reduction in the level of refractory sulfur compounds including the alkyl substituted benzothiophenes and dibenzothiophenes. The extent of sulfur reduction may depend on the original organic sulfur content of the cracking feed, with the greatest reductions achieved with the higher sulfur feeds. The metals content of the equilibrium catalyst in the unit may also affect the degree of desulfurization achieved, with a low metals
- 30 content, especially vanadium content, on the equilibrium catalyst favoring greater desulfurization. Desulfurization will be very effective with E-catalyst vanadium contents below 1,000 ppm although the present catalysts remain effective even at much higher vanadium contents. Sulfur reduction may be effective not only to

improve product quality but also to increase product yield in cases where the refinery cracked gasoline end point has been limited by the sulfur content of the heavy gasoline fraction; by providing an effective and economical way to reduce the sulfur content of the heavy gasoline fraction, the gasoline end point may be extended without the need to resort to expensive hydrotreating, with a consequent favorable effect on refinery economics. Removal of the various thiophene derivatives which are refractory to removal by hydrotreating under less severe conditions is also desirable if subsequent hydrotreatment is contemplated.

Example 1

Preparation of Catalyst Series 1

All samples in Catalyst Series 1 were prepared from a single source of spray dried material, consisting of 50% USY, 21% silica sol and 29% clay. The USY had a starting unit cell size of 2.454 nm, $\text{SiO}_2/\text{Al}_2\text{O}_3$ mol ratio of 5.46 and a total surface area of $810 \text{ m}^2\text{g}^{-1}$.

A V/USY catalyst, Catalyst A, was prepared by slurring the above spray dried catalyst with NH_4OH at a pH of 6, followed by filtration, ammonium sulfate exchange and washing with water. The catalyst was calcined in the presence of steam at 1300°F for 2 hours and impregnated with vanadyl oxalate. The steam calcination lowered the unit cell size of the zeolite and improved its stability in the presence of vanadium.

A V/USY catalyst, Catalyst B, was prepared in the same way as Catalyst A, with the exception that the initial slurring of the catalyst was performed at a pH between 3.2 and 3.5.

Two RE+V/USY catalysts, Catalyst C and D were prepared in the same way as Catalyst B, with the exception that after ammonium sulfate exchange the catalysts were exchanged with solutions of rare earth chloride to add 2 and 4 wt% RE_2O_3 onto the catalyst, respectively. The rare earth solution that was used had some of its Ce^{3+} extracted out, thus contains only a little Ce ions.

A Ce+V/USY catalyst, Catalyst E was prepared in the same way as Catalyst B, with the exception that after ammonium sulfate exchange the catalyst was exchanged with a solution of cerium chloride to add 5% cerium (as CeO₂) onto the catalyst.

5

These catalysts were then steamed deactivated, to simulate catalyst deactivation in an FCC unit, in a fluidized bed steamer at 770°C (1420°F) for 20 hours using 50% steam. The physical properties of the calcined and steam deactivated catalysts are summarized in Table 1.

10

Table 1

Physical Properties of the V, RE+V, and Ce+V USY/ Silica Sol Catalysts (Series 1)

	V/USY Cat. A	V/ USY Cat. B	RE+V/ USY Cat. C	RE+V/ USY Cat. D	Ce+V/ USY Cat. E
Calcined Cat.					
V loading, wt%	0.36	0.37	0.39	0.38	0.39
RE ₂ O ₃ loading, wt%	N.A.	N.A.	2.0	4.1	5.1
Ce ₂ O ₃ , wt%	N.A.	N.A.	0.49	0.95	4.95
La ₂ O ₃ , wt%	N.A.	N.A.	0.96	1.83	0.03
Na ₂ O, wt%	0.30	0.24	0.42	0.21	0.19
Unit cell size, nm	2.433	2.433	2.442	2.443	2.442
Deactivated Cat. (CPS 770°C 20 hrs)					
Surface area, m ² g ⁻¹	255	252	249	248	284
Unit cell size, nm	2.425	2.424	2.426	2.428	2.428

15 Example 2

Preparation of Catalyst Series 2

A V/USY catalyst, Catalyst F, was prepared using a USY zeolite with a silica-to-alumina ratio of 5.4 and unit cell size of 2.435 nm. A fluid catalyst was prepared by spray drying an aqueous slurry containing 50 wt% of the USY crystals in a silica sol/clay matrix. The matrix contained 22-wt% silica sol and 28-wt% kaolin clay. The spray-dried catalyst was exchanged with NH_4^+ by an exchange with a solution of ammonium sulfate and then dried. Then the USY catalyst was impregnated with a solution of vanadium oxalate to target 0.5 wt% V.

A RE+V/USY catalyst, Catalyst G, was prepared using a USY zeolite with a silica-to-alumina ratio of 5.5 and a unit cell size of 2.454 nm. The USY was exchanged with NH_4^+ by an exchange with a solution of ammonium sulfate. The NH_4^+ exchanged USY was then exchanged with rare earth cations (e.g., La^{3+} , Ce^{3+} , etc.) by exchange with a solution of mixed rare earth chlorides. The rare earth solution that was used had most of its Ce^{3+} extracted out, thus contained very little Ce. The RE-exchanged USY was further washed, dried, and calcined in the presence of steam in a rotary calciner at 760°C (1400°F). The steam calcination lowered the unit cell size of the zeolite to 24.40 Å and improved its stability in the presence of vanadium. A fluid catalyst was prepared by spray drying an aqueous slurry containing 50 wt% of the RE-USY crystals in a silica sol/clay matrix. The matrix contained 22-wt% silica sol and 28-wt% kaolin clay. The spray-dried catalyst was exchanged with NH_4^+ by an exchange with a solution of ammonium sulfate and was then dried and calcined at 540°C (1000°F) for 2 hours. Following calcination, the RE/USY catalyst was impregnated with a $\text{VO}(\text{SO}_4)_2$ solution.

Catalyst H, was prepared using similar procedures as for Catalyst G except a solution of mixed RECl_3 containing mostly CeCl_3 was used to exchange the USY. Catalyst H was prepared using a commercial USY zeolite with a silica-to-alumina ratio of 5.5 and a unit cell size of 2.454 nm. The USY was exchanged with NH_4^+ by an exchange with a solution of ammonium sulfate. The NH_4^+ -exchanged USY was then exchanged with a solution of CeCl_3 containing some lanthanum. The exchanged USY was further washed, dried, and calcined in the presence of steam in a rotary calciner at 760°C (1400°F). The steam calcination lowered the unit cell size of the zeolite to 2.440 nm. A fluid catalyst was prepared by spray drying an aqueous slurry containing 50 wt% of

the rare earth containing USY crystals in a silica sol/clay matrix. The matrix contained 22-wt% silica sol and 28-wt% kaolin clay. The spray-dried catalyst was exchanged with NH_4^+ by an exchange with a solution of ammonium sulfate and was then dried and calcined at 540°C (1000°F) for 2 hours. Following calcination, the catalyst was impregnated with a VOSO_4 solution. Physical properties of the calcined catalysts are summarized in Table 2.

Table 2

Physical Properties of V/USY, RE+V/USY Silica-Sol Catalysts (Series 2)

	V/USY Catalyst F	RE+V/ USY Catalyst G	RE+V/ USY Catalyst H
Calcined Cat.			
V loading, wt%	0.5	0.43	0.44
RE_2O_3 loading, wt%	N.A.	1.93	2.66
CeO_2 loading, wt%	N.A.	0.21	2.42
Na_2O , wt%	0.13	0.16	0.20
Surface area, m^2g^{-1}	327	345	345
Unit cell size, nm	2.435	-	-

Example 3

Preparation of Catalyst Series 3

A V/USY catalyst, Catalyst I, was prepared using a commercial H-form USY (crystal) with a bulk silica-to-alumina ratio of 5.4 and 2.435 nm unit cell size. A fluid catalyst was prepared by spray drying aqueous slurry containing 40 wt% of the USY crystals, 25 wt% silica, 5 wt% alumina, and 30 wt% kaolin clay. The spray-dried catalyst was calcined at 540°C (1000°F) for 3 hours. The resulting H-form USY catalyst was impregnated with a vanadium oxalate solution to target 0.4 wt% V by incipient wetness impregnation. The impregnated V/USY catalyst was further air calcined at 540°C (1000°F) for 3 hours. The final catalyst contains 0.39% V.

A Ce+V/USY catalyst, Catalyst J, was prepared from the same, spray-dried H-form USY catalyst intermediate as Catalyst I. The H-form USY catalyst was impregnated with a solution of $\text{Ce}(\text{NO}_3)_3$ to target 1.5wt% Ce loading using an incipient wetness impregnation method. Resulting Ce/USY catalyst was air calcined at 540°C (1000°F) for 3 hours followed by steaming at 540°C (1000°F) for 3 hours. Then the catalyst was impregnated with a vanadium oxalate solution to target 0.4 wt% V by incipient wetness impregnation. The impregnated Ce+V/USY catalyst was further air calcined at 540°C (1000°F) for 3 hours. The final catalyst contains 1.4% Ce and 0.43% V.

Table 3

Physical Properties of the V and Ce+V USY/ Silica-Alumina-Clay Catalysts (Series 3)

	V/USY Catalyst I	Ce+V/ USY Catalyst J
Calcined Cat.		
V loading, wt%	0.39	0.43
Ce loading, wt%	N.A.	1.4
Surface area, m^2g^{-1}	302	250
Alpha	130	12
UCS, nm	2.436	2.437

Example 4

Preparation of Catalyst Series 4

All samples in Catalyst Series 4 were prepared from a single source of spray dried material, consisting of 50% USY, 21% silica sol and 29% clay. The starting USY had a bulk silica-to-alumina ratio of 5.4 and 2.435 nm unit cell size. The spray dried catalyst was slurried with a solution of $(\text{NH}_4)_2\text{SO}_4$ and NH_4OH at pH of 6 to remove Na^+ , followed by washing with water and air calcination at 650°C (1200°F) for 2 hours.

A V/USY catalyst, Catalyst K, was prepared using the above H-form USY catalyst.

The H-form USY catalyst was impregnated with a vanadium oxalate solution to target

0.5 wt% V by incipient wetness impregnation. The impregnated V/USY catalyst was further air calcined at 650°C (1200°F) for 2 hours. The final catalyst contains 0.53% V.

A Ce+V/USY catalyst, Catalyst L, was prepared from the above H-form USY catalyst. The H-form USY catalyst was exchanged with a solution of CeCl_3 to target 0.75 wt% Ce loading. Resulting Ce/USY catalyst was air calcined and impregnated with a vanadium oxalate solution to target 0.5 wt% V by incipient wetness impregnation. The impregnated Ce+V/USY catalyst was further air calcined. The final catalyst contains 0.72% Ce and 0.52% V.

A Ce+V/USY catalyst, Catalyst M, was prepared from the above H-form USY catalyst by an exchange with a solution of CeCl_3 to target 3 wt% Ce loading. Resulting Ce/USY catalyst was air calcined and impregnated with a vanadium oxalate solution to target 0.5 wt% V by incipient wetness impregnation. The impregnated Ce+V/USY catalyst was further air calcined. The final catalyst contains 1.5% Ce and 0.53% V.

A Ce+V/USY catalyst, Catalyst N, was prepared from the above H-form USY catalyst by an incipient wetness impregnation with a solution of CeCl_3 to target 1.5 wt% Ce loading. Resulting Ce/USY catalyst was air calcined and impregnated with a vanadium oxalate solution to target 0.5 wt% V by incipient wetness impregnation. The impregnated Ce+V/USY catalyst was further air calcined. The final catalyst contains 1.5% Ce and 0.53% V.

These catalysts were then steam deactivated to simulate catalyst deactivation in an FCC unit, in a fluidized bed steamer at 770°C (1420°F) for 20 hours using 50% steam and 50% gas. The gas stream was changed from air, N_2 , propylene and N_2 mix, and to N_2 for every ten minutes, then circled back air to simulate the coking/ regeneration cycle of a FCC unit (cyclic steaming). Two sets of deactivated catalyst samples were generated: the steam deactivation cycle was ended with air-burn (ending-oxidation) for one batch of catalysts, and the other ended with propylene (ending-reduction). The coke content of the "ending-reduction" catalyst is less than 0.05% C. The physical properties of the calcined and steam (ending oxidation) catalysts are summarized in Table 4.

Table 4
Physical Properties of V and Ce+V USY/ Silica Sol Catalysts (Series 4)

	V/USY Catalyst K	Ce+V/ USY Catalyst L	Ce+V/ USY Catalyst M	Ce+V/ USY Catalyst N
Calcined Cat.				
V loading, wt%	0.53	0.52	0.53	0.53
Ce loading, wt%	N.A.	0.72	1.5	1.5
Na, ppm	890	1190	1190	1260
Deactivated Cat. (CPS 770°C, 20 hrs)				
Surface area, m ² g ⁻¹	237	216	208	204
Unit cell size, nm	2.425	2.423	2.425	2.425

5

Example 5**Preparation of Catalyst Series 5**

10 All samples in Catalyst Series 5 were prepared from a single source of spray dried material, consisting of 40% USY, 30% colloidal silica sol, and 30% clay. The starting H-form USY had a bulk silica-to-alumina ratio of 5.4 and 2.435nm unit cell size. The spray-dried catalyst was air calcined at 540°C (1000°F) for 3 hours.

15 A Ce/USY catalyst, Catalyst O, was prepared using the above H-form USY catalyst. The H-form USY catalyst was impregnated with a solution of Ce(NO₃)₃ to target 1.5wt% Ce loading using an incipient wetness impregnation method. Resulting Ce/USY catalyst was air calcined at 540°C (1000°F) followed by steaming at 540°C (1000°F) for 3 hours.

20 A Ce+V/USY catalyst, Catalyst P, was prepared from Catalyst O. The Ce/USY catalyst was impregnated with a vanadium oxalate solution to target 0.5 wt% V by incipient wetness impregnation. The impregnated Ce+V/USY catalyst was dried and

air calcined at 540°C (1000°F) for 3 hours. The final catalyst contains 1.4% Ce and 0.49% V.

A Ce+V/USY catalyst, Catalyst Q, was prepared from Catalyst O by an exchange with a solution of VOSO₄ at pH ~3 to target 0.5 wt% V loading. Resulting Ce+V/USY catalyst was dried and air calcined at 540°C (1000°F) for 3 hours. The final catalyst contains 0.9% Ce and 0.47% V. Physical properties of calcined catalysts are summarized in Table 5.

Table 5

Physical Properties of the Ce and Ce+V USY/ Silica-Clay Catalysts (Series 5)

	Ce/USY Catalyst O	Ce+V/ USY Catalyst P	Ce+V/ USY Catalyst Q
Calcined Cat.			
V loading, wt%	N.A.	0.49	0.47
Ce loading, wt%	1.6	1.4	0.9
Na, ppm	-	-	940
Surface area, m ² g ⁻¹	284	281	272
Alpha	-	10	14
Unit cell size, nm	2.435	2.436	2.436

Example 6

Preparation of RE+V/USY/Silica-Sol Catalyst

A RE+V/USY catalyst, Catalyst R, was prepared using a NaY zeolite with a silica-to-alumina ratio of 5.5 and a unit cell size of 2.465 nm. The zeolite Y was exchanged with NH₄⁺ by an exchange with a solution of ammonium sulfate. The NH₄⁺-exchanged Y was then exchanged with rare earth cations (e.g., La³⁺, Ce³⁺, etc.) by exchange with a solution of mixed rare earth chlorides from which most of the Ce³⁺ had been

removed by extraction, so that the solution contained very little cerium. The RE-exchanged Y was further washed, dried, and calcined in the presence of steam at 705°C (1300°F) for 2 hours. The steam calcination lowered the unit cell size of the zeolite and improved its stability in the presence of vanadium. A fluid catalyst was prepared by spray drying an aqueous slurry containing 50 wt% of the RE-USY crystals in a silica sol/clay matrix. The matrix contained 22-wt% silica sol and 28-wt% kaolin clay. The spray-dried catalyst was exchanged with NH_4^+ by an exchange with a solution of ammonium sulfate and was then dried and calcined at 540°C (1000°F) for 1 hour. Following calcination, the RE/USY catalyst was impregnated with a vanadium oxalate solution. Physical properties of the calcined catalyst are summarized in Table 6.

Table 6

Physical Properties of RE+V USY/ Silica-Sol Catalyst

	<u>RE+V/ USY</u> Catalyst R
<u>Calcined Cat.</u>	
V loading, wt%	0.43
RE_2O_3 loading, wt%	1.93
CeO ₂ loading, wt%	0.21
Na ₂ O, wt%	0.16
Surface area, m ² /g	345
Unit cell size, Å	24.58

Example 7**Preparation of Ce+V/USY/Silica-Sol Catalyst**

A Ce+V/USY catalyst, Catalyst S, was prepared using a NaUSY zeolite with a silica-to-alumina ratio of 5.5 and a unit cell size of 2.454 nm. The USY was exchanged with NH_4^+ by an exchange with a solution of ammonium sulfate. The NH_4^+ exchanged

USY was then exchanged with Ce^{3+} cations by exchange with a solution of cerium chloride containing a small amount of other rare earth ions (e.g., La^{3+} , Pr, Nd, Gd, etc.). The Ce-exchanged USY was further washed, dried, and calcined in the presence of steam at 705°C (1300°F) for 2 hours. The steam calcination lowered the unit cell size of the USY and improved its stability in the presence of vanadium. A fluid catalyst was prepared by spray drying an aqueous slurry containing 50 wt% Ce/USY crystals in a silica sol/clay matrix. The matrix contained 22-wt% silica sol and 28-wt% kaolin clay. The spray-dried catalyst was exchanged with NH_4^+ by an exchange with a solution of ammonium sulfate and was then dried and calcined at 540°C (1000°F) for 1 hour. Following calcination, the Ce/USY catalyst was impregnated with a vanadium sulfate solution. Physical properties of the calcined catalyst are summarized in Table 7.

Table 7

Physical Properties of Ce+V/USY/ Silica-Sol Catalyst

	<u>Ce+V/USY</u> <u>Catalyst S</u>
<u>Calcined Cat.</u>	
V loading, wt%	0.44
RE_2O_3 loading, wt%	2.66
CeO_2 loading, wt%	2.42
Na_2O , wt%	0.20
Surface area, m^2g^{-1}	345
Unit cell size, nm	2.446

Experimental Procedure: Cracking Performance Evaluation

The catalysts from Examples 1 through 7 were evaluated for FCC and the results are reported in Examples 8 through 15. The blended catalysts (a candidate plus equilibrium catalyst) were tested for gas oil cracking activity and selectivity with the

ASTM microactivity (MAT) test, modified from ASTM procedure D-3907, using a vacuum gas oil (VGO) feed. Some of the additive catalysts were evaluated in a circulatory riser pilot unit using a vacuum gas oil or a hydrotreated feed. The compositions of the three VGO feeds and one severely cat feed hydrotreated feed (CFHT) used in these Examples are shown in Table 8.

Table 8
Properties of Cracking Feeds

Charge Properties	VGO No. 1	VGO No. 2	VGO No. 3	CFHT
API Gravity	26.6	22.5	24.2	23.6
Aniline Point, C	83	73	187	164
CCR, wt%	0.23	0.25	0.6	0.09
Sulfur, wt%	1.05	2.59	1.37	0.071
Nitrogen, ppm	600	860	900	1200
Basic nitrogen, ppm	310	340	290	380
Ni, ppm	0.32	-	0.2	0.2
V, ppm	0.68	-	0.1	0.2
Fe, ppm	9.15	-	0	0.3
Cu, ppm	0.05	-	0	0
Na, ppm	2.93	-	0.6	1.2
Sim. Dist., °C				
IBP,	181	217	192	172
50wt%,	380	402	430	373
99.5%,	610	553	556	547

A range of cracking conversions was obtained by varying the catalyst-to-oil ratios with the reactions run at 527°C (980°F). The ranges of cut points for the cracked product were:

Gasoline	C ₅ ⁺ to 220°C (125 - 430°F)
5 Light LCO	220° to 310°C (430°F - 590°F),
Heavy LCO	310° to 370°C (590° - 700°F)
Light Fuel Oil (LFO)	220° to 370°C (430° - 700°F)
Heavy Fuel Oil (HFO)	370°C ⁺ (700°F ⁺) .

- 10 The gasoline range product from each material balance was analyzed with a sulfur GC (AED) to determine the gasoline sulfur concentration. To reduce experimental errors in the sulfur concentration associated with fluctuations in distillation cut point of gasoline, the sulfur species ranging from thiophene to C₄-thiophenes in syncrude (excluding benzothiophene and higher boiling S species) were quantitated and the
- 15 sum was defined as "cut-gasoline S."

To determine the sulfur content of the distillate fractions boiling above the gasoline range (Examples 14 and 15), the syncrude was subjected to atmospheric distillation to separate the gasoline. The bottoms fraction was then further vacuum distilled to

20 generate two LFO/LCO fractions (light LCO and heavy LCO) and HFO. The sulfur species in the LCO sample were quantitated using a GC equipped with J&W 100 m DB-Petro column and Sievers 355B sulfur detector. The concentration of each LCO sulfur species was calculated based on the percentage of the sulfur species from the GC and the total sulfur content measured by XPS method.

25

Example 8

Fluid Catalytic Cracking Evaluation of Series 1 Catalysts

The catalysts from Example 1 were steam deactivated in a fluidized bed steamer at

30 770°C (1420°F) for 20 hours using 50% steam and 50% gas as described in Example 4 above, ending with an air-burn (ending-oxidation). Twenty-five weight percent of the steamed additive catalysts were blended with an equilibrium catalyst of very low metals level (120 ppm V and 60 ppm Ni) from an FCC unit.

The catalytic cracking performances of the catalysts were evaluated using VGO No. 1 as cracking feed in the microactivity test. The performances of the catalysts are summarized in Table 9, where the product selectivity was interpolated to a constant conversion, 65wt% conversion of feed to 220°C or below (430°F) material.

Table 9

Catalytic Cracking Performance of Series 1 Catalysts, VGO No. 1

	ECat	+ 25%	+ 25%	+ 25%	+ 25%	+ 25%
		V/USY	V/USY	RE+V/ USY	RE+ V/USY	RE+ V/USY
	Base Case	Cat A	Cat B	Cat C	Cat D	Cat E
MAT Product						
Yields						
Conversion, wt%	65	65	65	65	65	65
Cat/Oil	3.0	3.3	3.3	2.9	3.0	2.9
		<i>Incremental Yield</i>				
H ₂ yield, wt%	0.03	+0.05	+0.05	+0.04	+0.02	+0.04
C ₁ + C ₂ Gas, wt%	1.1	+0.1	+0.1	+0	+0.1	+0
Total C ₃ Gas, wt%	4.3	+0.1	+0.1	-0.1	+0	-0.2
C ₃ ⁼ yield, wt%	3.7	+0.1	+0.1	+0	+0	-0.1
Total C ₄ Gas, wt%	9.3	+0.1	+0.2	-0.1	+0	-0.3
C ₄ ⁼ yield, wt%	4.7	+0.3	+0.4	+0.4	+0.1	+0
C ₅ ⁺ Gasoline, wt%	47.6	-0.6	-0.4	+0.4	+0	+0.5
LFO, wt%	29.6	+0	+0.2	+0	+0.1	+0
HFO, wt%	5.4	+0	-0.2	+0	-0.1	+0
Coke, wt%	2.4	+0.3	+0.0	-0.2	-0.1	-0.1
Cut Gasoline S, ppm	618	377	366	369	382	352
% Reduction in Cut Gasoline S	Base	39.0	40.8	40.4	38.3	43.1

- 5 The cat-to-oil ratios in Table 9 show that the blends of deactivated V/USY and ECat require higher cat-to-oil ratio than the 100% equilibrium catalyst base case to achieve

65% conversion (3.3 vs. 3.0 Cat/Oil, i.e., 10% reduction in activity). It is due to lower cracking activity of V/USY catalysts relative to the equilibrium catalyst. In comparison, addition of the RE+V/USY catalysts did not increase the cat-to-oil ratio to achieve 65% conversion. These cat-to-oil results indicate that the RE+V/USY catalysts are more stable and maintain their cracking activity better than the V/USY catalysts.

Compared to the equilibrium catalyst base case, addition of V/USY and RE+V/USY catalyst made small changes in the overall product yield structure. There were slight increases in hydrogen and coke yields. Also a small changes in C₄ gas, gasoline, light cycle oil and heavy fuel oil yields were observed. Addition of the V/USY and RE+V/USY catalysts changed the gasoline S concentration substantially. When 25 wt% of each of Catalyst A or B (V/USY reference catalysts) was blended with the equilibrium FCC catalyst, 39.0 and 40.8 % reduction in gasoline sulfur concentration was achieved. When 25wt% of RE/USY catalysts (Catalyst C and D) was added to the equilibrium catalyst, the gasoline sulfur reduction activities are comparable to the reference catalysts (38-40%). The RE+V/ USY catalyst containing mostly cerium as the rare earth metal (Catalyst E) gave a 43.1% reduction in gasoline S, to reduce the gasoline S content by 4% additionally, i.e., 10% improvement over the V/USY and mixed RE/USY catalysts. All catalysts have a comparable vanadium loading (0.36 – 0.39%).

These results show that addition of rare earth elements improves the cracking activity of a V/USY catalyst. Changes in the cracked product yields are minor. Among rare earth ions, cerium exhibits a unique property in that the Ce+V/USY catalyst not only exhibits higher cracking activity but also exhibits increased gasoline sulfur reduction activity at fluid catalytic cracking conditions. The rare earth RE/USY catalyst without major amounts of cerium has no added benefit over V/USY for gasoline S reduction whereas the presence of cerium further lowered the gasoline sulfur level of the V/USY or RE/USY (without major cerium levels) catalysts.

Example 9**Comparison of Cracking Activity of Series 2 Catalysts.**

The V and RE/V USY catalysts from Example 2 (Series 2 Catalysts, Catalyst F, G, H) were steam deactivated at 770°C (1420°F) for various lengths of time to compare catalyst stability. The catalysts were steamed in a fluidized bed steamer for 2.3, 5.3, 10, 20, and 30 hours using 50% steam and 50% gas (cyclic steaming ending-reduction, as described in Example 4 above). The surface area retentions of the deactivated catalysts are plotted in Figure 2.

The steam deactivated catalysts were tested for gas oil cracking activity using the ASTM microactivity test (ASTM procedure D-3907) with Vacuum Gas Oil No. 2 (above - 2.6 wt% S). At a 30-second contact time and at 545°C (980°F) reaction temperature, a weight percent conversion to 220°C- (430°F-) was measured at a constant catalyst-to-oil ratio of 4:1. Conversions as a function of steam deactivation time are plotted in Figure 3.

The surface area retentions shown in Figure 2 indicate that V/USY and RE+V/USY catalysts show comparable surface area retention upon various hydrothermal deactivation conditions suggesting that all three catalysts have comparable framework structure stability. However, the conversion plot shown in Figure 3 clearly indicates that RE+V/USY have much improved cracking activity retention as severity of hydrothermal deactivation increases. Upon the hydrothermal deactivation, the improvement in cracking activity from the V/USY to RE+V version was 15% conversion. No apparent differences were observed between the RE versions with varying amounts of cerium. These results are consistent with that of Example 8 where RE+V/USY achieved the target conversion at a lower cat-to-oil ratio than V/USY. These conversion results indicate that the RE+V/USY catalysts are more stable and maintain their cracking activity better than the V/USY catalysts. The addition of rare earth ions to the USY followed by steam calcination to lower the unit cell size of the zeolite improved catalyst stability in the presence of vanadium.

Example 10**Fluid Catalytic Cracking Evaluation of Series 3 Catalysts**

5 The V and Ce+V USY catalysts from Example 3 (Catalysts I, J) were steam deactivated in a fluidized bed steamer at 770°C (1420°F) for 20 hours using 50% steam and 50% gas, as described in Example 4 above, ending with an air-burn (ending-oxidation). Twenty-five weight percent of steamed additive catalysts were blended with the low metals FCC equilibrium catalyst (120 ppm V and 60 ppm Ni). The blended catalysts were then evaluated by the MAT test as described above, using
10 VGO No. 1 feed.

Performances of the Series 3 catalysts using the VGO No. 1 feed are summarized in Table 10, where the product selectivity was interpolated to a constant conversion, 70wt% conversion of feed to 220°C- (430°F-) material.

Table 10**Catalytic Cracking Performance of Series 3 Catalysts**

	ECat Base Case	+ 25% V/USY cat (Catalyst I)	+25% Ce+V/USY cat (Catalyst J)
MAT Product Yields			
Conversion, wt%	70	70	70
Cat/Oil	3.3	3.8	3.7
		<i>Incremental Yield</i>	
H ₂ yield, wt%	0.03	+0.04	+0.13
C ₁ + C ₂ Gas, wt%	1.4	+0.1	+0.1
Total C ₃ Gas, wt%	5.4	+0.1	-0.1
C ₃ ⁼ yield, wt%	4.5	+0.1	-0.1
Total C ₄ Gas, wt%	10.9	+0.2	-0.2
C ₄ ⁼ yield, wt%	5.2	+0.4	+0.2
iC ₄ yield, wt%	4.8	-0.2	-0.4
C ₅ ⁺ Gasoline, wt%	48.9	-0.3	-0.3
LFO, wt%	24.6	+0.5	+0.3
HFO, wt%	4.7	-0.2	-0.1
Coke, wt%	2.7	+0	+0.5
Cut Gasoline S, PPM	529	378	235
% Reduction in Cut Gasoline S	Base	29	56

Table 10 compares the FCC performances of V/USY and Ce+V/USY/ Silica-Alumina-Clay catalysts each blended with an equilibrium FCC catalyst (ECat) after cyclic steam deactivation (ending oxidation). Compared to the ECat base case, the addition of

V/USY and Ce+V/USY catalyst changes the overall product yield structure only slightly. Yield changes in C₄- gas, gasoline, light cycle oil, and heavy fuel oil are all small. Moderate increases in hydrogen and coke yields were observed. While the product yield changes were small, the V/USY and Ce+V/USY catalysts changed the gasoline S concentration substantially. When 25 wt% of Catalyst I (V/USY reference catalyst) was blended with an equilibrium FCC catalyst, 29% reduction in gasoline sulfur concentration was achieved. In comparison, Ce+V/ USY catalyst (Catalyst J) gave 56% reduction in gasoline S. Addition of Ce to the V/USY catalyst reduced the gasoline S content by additional 27%, i.e., 93% improvement over the V/USY reference catalyst. Both catalysts have comparable vanadium loadings (0.39% vs. 0.43% V). In light of the fact that Ce by itself does not have any gasoline sulfur reduction activity (see below in Example 11), these results are quite unexpected and clearly demonstrate the benefits of cerium addition.

Example 11

Fluid Catalytic Cracking Evaluation of Series 4 Catalysts after Cyclic Steaming

The performances of V and Ce+V catalysts from Example 4 are summarized in this example. The Series 4 catalysts were steam deactivated as described above in Example 4 by cyclic steaming (ending-reduction), then blended with an the low metals (120 ppm V and 60 ppm Ni) FCC equilibrium catalyst in a 25:75 weight ratio and tested using VGO No. 1 feed. The results are summarized in Table 11.

Table 11**Catalytic Cracking Performance of V vs. Ce+V/USY/ Silica-Sol Catalysts**

	ECat Base Case	+ 25% V/USY cat (Cat K)	+ 25% Ce+V/USY (Cat M)	+ 25% Ce+V/USY (Cat N)
MAT Product Yields				
Conversion, wt%	65	65	65	65
Cat/Oil	3.0	3.4	3.2	3.3
		<i>Incremental Yield</i>		
H ₂ yield, wt%	0.03	+0.02	+0.02	+0.02
C ₁ + C ₂ Gas, wt%	1.1	+0	+0	+0.1
Total C ₃ Gas, wt%	4.4	-0.1	-0.1	+0
C ₃ ⁼ yield, wt%	3.7	+0	-0.1	+0
Total C ₄ Gas, wt%	9.5	-0.1	-0.2	-0.1
C ₄ ⁼ yield, wt%	4.8	+0.1	+0.1	+0.1
iC ₄ yield, wt%	4.1	-0.2	-0.3	-0.1
C ₅ + Gasoline, wt%	47.4	+0.1	+0.5	+0.1
LFO, wt%	29.7	-0.2	+0	-0.1
HFO, wt%	5.3	+0.2	+0	+0.1
Coke, wt%	2.3	+0.1	-0.1	+0
Cut Gasoline S, PPM	516	489	426	426
% Reduction in Cut Gasoline S	Base	5.2	17.4	17.4

- 5 Table 11 compares FCC performances of V/USY and Ce+V/USY/ Silica-Sol additive catalysts after cyclic steam deactivation (ending-reduction). Compared to the ECat base case, addition of the V/USY and Ce+V/USY catalysts made very little changes in the overall product yield structure. The yields of hydrogen, C₄- gas, gasoline, light

cycle oil, heavy fuel oil and coke were changed by less than 0.2 wt% each. Additions of the V/USY and Ce+V/USY catalysts changed the gasoline S concentration to different extents. When 25 wt% of Catalyst K (V/USY - reference catalyst) was blended with the equilibrium FCC catalyst, 5.2% reduction in gasoline sulfur concentration was achieved. For comparison, Ce+V/USY catalysts (Catalysts M and N) gave 17.4% reduction in gasoline S, respectively. Addition of Ce to the V/USY catalyst reduced the gasoline S content by additional 12.3%, i.e., 237% improvement over the V/USY reference catalyst.

10 **Example 12**

Fluid Catalytic Cracking Evaluation of Series 4 Catalysts after Cyclic Steaming

The performances of the V and Ce+V catalysts from Example 4 after cyclic steam deactivation are summarized in this example. The catalysts of Example 4 were deactivated by cyclic steaming as described above in Example 4 (ending-oxidation) and were then blended with the low metals (120 ppm V and 60 ppm Ni) FCC equilibrium catalyst in 25:75 weight ratio. The evaluation results obtained with the VGO No. 1 feed are summarized in Table 12.

Table 12

Catalytic Cracking Performance of V vs. Ce+V/USY/ Silica-Sol Catalyst

	ECat Base Case	+ V/USY cat (Cat K)	25% + 25% Ce+V/USY (Cat L)	+ 25% Ce+V/USY (Cat M)
MAT Product Yields				
Conversion, wt%	70	70	70	70
Cat/Oil	2.8	3.7	3.6	3.4
		<i>Incremental</i>		
		<i>Yield</i>		
H ₂ yield, wt%	0.03	+0.12	+0.13	+0.12
C ₁ + C ₂ Gas, wt%	1.5	+0.2	+0.2	+0.1
Total C ₃ Gas, wt%	5.5	+0.1	+0	-0.1
C ₃ ⁼ yield, wt%	4.7	+0	+0	-0.1
Total C ₄ Gas, wt%	11.1	+0	+0	-0.2
C ₄ ⁼ yield, wt%	5.8	+0.1	+0.1	+0
iC ₄ yield, wt%	4.6	-0.1	-0.2	-0.2
C ₅ + Gasoline, wt%	49.4	-1.0	-0.9	-0.5
LFO, wt%	25.6	-0.1	+0	+0.2
HFO, wt%	4.4	+0.1	+0	-0.2
Coke, wt%	2.3	+0.6	+0.5	+0.5
Cut Gasoline S, PPM	579	283	243	224
% Reduction in Cut Gasoline S	Base	51.1	58.1	61.3

Table 12 compares the FCC performances of V/USY and Ce+V/USY/ Silica-Sol additive catalysts after cyclic steam deactivation (ending-oxidation). Compared to the ECat base case, addition of V/USY and Ce+V/USY catalyst made slight changes in the overall product yield structure. There were moderate increases in hydrogen and coke yields. Also a small changes in C₄- gas yield gasoline, light cycle oil and heavy fuel oil were observed. Addition of the V/USY and Ce+V/USY catalysts changed the gasoline S concentration substantially. When 25 wt% of Catalyst K (V/USY - reference catalyst) was blended with an equilibrium FCC catalyst, 51.1 % reduction in gasoline sulfur concentration was achieved. In comparison, the Ce+V/USY catalysts (Catalysts L and M) gave 58.1% and 61.3% reduction in gasoline S, respectively. Addition of Ce to the V/USY catalyst reduced the gasoline S content by additional 7.0 – 10.2%, i.e., up to 20% improvement over the V/USY reference catalyst.

The product yields data of the V/USY and Ce+V/USY catalysts indicate that the yield changes from the ECat is due to addition of vanadium to the USY catalyst. The product yields of V/USY catalyst is comparable to those of Ce+V/USY catalysts except the gasoline S level. These results suggest that Ce increases the gasoline sulfur reduction activity of the V/USY additive catalyst with little effect on product yields.

Example 13

Fluid Catalytic Cracking Evaluation of Series 5 Catalysts, Study of Promotional Effects

Performances of the Ce and Ce+V catalysts from Example 5 after cyclic steaming deactivation (ending-reduction) as described above, are summarized in this example.

The deactivated catalysts were blended with the low metals (120 ppm V and 60 ppm Ni) FCC equilibrium catalyst in 25:75 weight ratio. The results obtained with the VGO No. 1 feed are summarized in Table 13.

Table 13

Catalytic Cracking Performance of Ce vs. Ce+V/USY/ Silica-Clay Catalysts

	ECat Base Case	+ 25% Ce/USY (Cat O)	+ 25% Ce+V/USY (Cat P)	+ 25% Ce+V/USY (Cat Q)
MAT Product Yields				
Conversion, wt%	70	70	70	70
Cat/Oil	3.2	3.4	3.7	3.9
		<i>Incremental Yield</i>		
H ₂ yield, wt%	0.04	+0	+0.07	+0.07
C ₁ + C ₂ Gas, wt%	1.5	+0.1	+0.1	+0.1
Total C ₃ Gas, wt%	5.7	+0.1	-0.1	+0
C ₃ ⁼ yield, wt%	4.8	+0	+0	+0
Total C ₄ Gas, wt%	11.5	+0	-0.3	-0.1
C ₄ ⁼ yield, wt%	5.7	+0	+0.1	+0.1
iC ₄ yield, wt%	4.9	+0	-0.3	-0.2
C ₅ + Gasoline, wt%	48.8	-0.2	-0.2	-0.7
LFO, wt%	25.5	+0	+0.1	+0.1
HFO, wt%	4.5	+0	-0.1	-0.1
Coke, wt%	2.4	+0	+0.3	+0.6
Cut Gasoline S, PPM	486	487	341	351
% Reduction in Cut Gasoline S	Base	0	29.8	27.7

Table 13 compares FCC performances of Ce/USY and Ce+V/USY/ Silica-clay additive catalysts after cyclic steam deactivation (ending-reduction). Compared to the ECat base case, addition of the Ce/USY catalyst made almost no changes in overall

product yields. Addition of the Ce+V/USY catalyst made slight changes in the overall product yield structure. There were moderate increases in hydrogen and coke yields, as well as slight changes in C4- gas, gasoline, light cycle oil and heavy fuel oil yields. Addition of the Ce/USY catalysts made a no change in the gasoline S concentration.

- 5 In contrast, Ce+V/USY catalysts (Catalysts P and Q, invention) gave 29.8% and 27.7% reduction in gasoline S, respectively. These results indicate that cerium by itself does not have any gasoline sulfur reduction activity. Cerium appears to have a promotional effect for vanadium to increase the activity of a V/USY gasoline sulfur reduction additive catalyst.

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Example 14

Catalyst R was evaluated in a circulating riser pilot unit (Davison Circulating Riser) in combination with a typical FCC equilibrium catalyst using VGO No. 3 feed (Table 8).

- 15 Prior to evaluation, the RE+V USY catalyst was steam deactivated at 770°C (1420°F) for 20 hours using 50% steam and 50% gas. Twenty-five weight percent of steamed additive catalyst was blended with an equilibrium catalyst (530 ppm vanadium and 330 ppm Ni) from an FCC unit.

- 20 FCC performances of the catalysts are summarized in Table 14, where the product selectivity was interpolated to a constant conversion, 75wt% conversion of feed to 220°C- (430°F-) material.

Table 14**Catalytic Cracking Performance of RE+V/USY/ Silica-Sol Catalyst**

	<u>ECat</u> <u>Base Case</u>	<u>+ 25% RE+V/USY</u> <u>(Cat R)</u>
<u>Riser Product Yields</u>		
Conversion, wt%	75	75
Cat/Oil	7.0	6.7
		<u>Incremental Yield</u>
H ₂ yield, wt%	0.03	+0.01
Total C ₃ , wt%	6.5	-0.1
Total C ₄ , wt%	12.1	-0.1
C ₅ ⁺ Gasoline, wt%	49.4	-0.1
LFO, wt%	18.3	-0.1
HFO, wt%	6.7	+0.1
Coke, wt%	4.1	+0.3
Cut Gasoline S, ppm	735	589
LCO Sulfur, wt%	2.36	2.16
% Reduction in Gasoline Sulfur	Base	20
% Reduction in LCO Sulfur	Base	9

- 5 Compared to the base case, addition 25 wt% of the RE+V/USY catalyst made small changes in the overall product yield structure. There were negligible increases in H₂ and coke yields. Also little changes in C₄- gas, gasoline, light cycle oil and heavy fuel oil yields were observed. Addition of the RE+V/USY catalyst changed the gasoline sulfur concentration substantially, 20 % reduction in gasoline sulfur concentration
- 10 being achieved. In addition to the gasoline sulfur reduction, a substantial sulfur reduction in LCO was observed, equivalent to an overall reduction of 9% LCO sulfur.

The sulfur species in LCO are shown in Table 15 and Figure 4. The LCO contains wide ranges of benzothiophene and dibenzothiophene sulfur species. The sulfur

reductions were more prominent for substituted-benzothiophenes and substituted-dibenzothiophenes such as C₃⁺-benzothiophenes, and C₁- through C₄⁺-dibenzothiophenes. The results are quite unexpected since substituted dibenzothiophenes are bulkier and expected to be harder to crack and desulfurize in FCC.

Table 15

Sulfur Speciation of LCO from VGO Feed No. 3

(wt% sulfur in LCO)	<u>ECat</u>	<u>+ 25% RE+V/USY</u>
	<u>Base Case</u>	<u>(Cat I)</u>
Benzothiophene	0.04	0.04
C ₁ -Benzothiophenes	0.22	0.24
C ₂ -Benzothiophenes	0.39	0.38
C ₃ ⁺ -Benzothiophenes	0.47	0.38
Dibenzothiophene	0.10	0.09
C ₁ -Dibenzothiophenes	0.36	0.32
C ₂ -Dibenzothiophenes	0.39	0.35
C ₃ -Dibenzothiophenes	0.24	0.22
C ₄ ⁺ -Dibenzothiophenes	0.16	0.14
Sum	2.36	2.16

Example 15

Fluid Catalytic Cracking Evaluation of Ce+V/ USY/ Silica-Sol Catalysts

The Ce+V/USY catalyst from Example 7 (Catalyst S) was evaluated in a circulating fluidized cracking unit for 40 days in combination with a typical FCC catalyst using the severely hydrotreated FCC feed of Table 8 (CFHT Feed). The base FCC equilibrium catalyst has very low metals level (200 ppm V and 130 ppm Ni). For the first 15 days, a 50/50 blend of fresh FCC catalyst and the Ce+V/USY catalyst from Example 7 was added to the FCC regenerator at 1.4% of the catalyst inventory pre day. From 15th to 40th day, a 85/15 blend of the fresh FCC catalyst and the Ce+V/USY catalyst was added to the FCC regenerator at 1.4% of the catalyst inventory per day. The

regenerator temperature was maintained at approximately 705°C (1300°F) throughout the evaluation. Two equilibrium catalyst (ECat) samples were collected: the first was sampled before the Ce+V/USY addition (base case) and the second at 40th day. Based on Ce and V analysis, the loading of the Ce+V/USY catalyst is estimated as
5 12%.

FCC performances of the catalysts are summarized in Table 16, where the product selectivity was interpolated to a constant conversion, 70wt% conversion of feed to 220°C- (430°F-) material.

Table 16**Catalytic Cracking Performance of Ce+V/USY/ Silica-Sol Catalyst**

	<u>ECat</u> <u>Base Case</u>	<u>+ 12% Ce+V/USY</u> <u>(Cat S)</u>
<u>Riser Product Yields</u>		
Conversion, wt%	70	70
Cat/Oil	6.5	6.4
		<u>Incremental Yield</u>
H ₂ yield, wt%	0.02	+0.0
Total C ₃ , wt%	4.8	+0.0
Total C ₄ , wt%	9.3	+0.1
C ₅ ⁺ Gasoline, wt%	51.9	-0.2
LFO, wt%	24.0	-0.2
HFO, wt%	6.1	+0.1
Coke, wt%	2.6	+0.1
Cut Gasoline S, ppm	100	79
Light LCO Sulfur, ppm	815	599
Heavy LCO Sulfur, ppm	1957	1687
HFO Sulfur, ppm	2700	1700
% Reduction in Gasoline Sulfur	Base	21
% Reduction in Light LCO S	Base	27
% Reduction in Heavy LCO S	Base	14
% Reduction in HFO Sulfur	Base	37
Wt% S of spent catalyst	<0.06	<0.06

- 5 Compared to the ECat base case, addition of the Ce+V/USY catalyst made small changes in the overall product yield structure. There were negligible increases in H₂ and coke yields. Also little changes in C₄ gas, gasoline, light cycle oil and heavy fuel

oil yields were observed. Addition of the Ce+V/USY catalyst changed the gasoline S concentration substantially. When 12 wt% of Ce+V/USY catalyst was blended in the equilibrium FCC catalyst, 21 % reduction in gasoline sulfur concentration was achieved. In addition to the gasoline sulfur reduction, substantial sulfur reductions in LCO and HFO were observed.

The sulfur species in the light LCO fraction were analyzed using a sulfur GC. The concentration of each organo-sulfur class was as shown in Table 17 and Figure 5 for light LCO.

Table 17
Sulfur Speciation of Light LCO from CFHT Feed

(PPM sulfur in light LCO)	<u>ECat</u>	<u>+12% Ce+V/USY</u>
	<u>Base Case</u>	<u>(Cat J)</u>
C ₃ ⁺ Thiopenes	4	1
Benzothiophene	20	14
C ₁ -Benzothiophenes	141	106
C ₂ -Benzothiophenes	237	174
C ₃ ⁺ -Benzothiophenes	295	215
Dibenzothiophene	17	14
C ₁ -Dibenzothiophenes	47	34
C ₂ -Dibenzothiophenes	32	25
C ₃ -Dibenzothiophenes	20	13
C ₄ ⁺ -Dibenzothiophenes	3	1
Sum	815	599

As shown in Figure 5, the light LCO contains mainly benzothiophene sulfur species. The sulfur reductions, equivalent to an overall 27 percent reduction in the light LCO sulfur, were more prominent for substituted benzothiophenes such as C₁- through C₃⁺-benzothiophenes. The results are quite unexpected since substituted

benzothiophenes are bulkier, and they are expected to be harder to crack and desulfurize in FCC.

The sulfur species in the heavy LCO fraction were as shown in Table 18 and Figure 6.

- 5 The heavy LCO contains mainly dibenzothiophene sulfur species. The sulfur reductions (14 percent overall for heavy LCO) were more prominent for substituted dibenzothiophenes such as C₁- through C₄⁺-dibenzothiophenes. The results are quite unexpected since substituted dibenzothiophenes are bulkier and expected to be harder to crack and desulfurize in FCC.

Table 18

Sulfur Speciation of Heavy LCO from CFHT Feed

(PPM sulfur in heavy LCO)	ECat	+ 12% Ce+V/USY
	Base Case	(Cat S)
Benzothiophene	0	0
C ₁ -Benzothiophenes	2	1
C ₂ -Benzothiophenes	12	7
C ₃ ⁺ -Benzothiophenes	128	86
Dibenzothiophene	65	54
C ₁ -Dibenzothiophenes	361	312
C ₂ -Dibenzothiophenes	537	480
C ₃ -Dibenzothiophenes	475	425
C ₄ ⁺ -Dibenzothiophenes	378	322
Sum	1957	1687

5

The sulfur reduction catalyst is active in reducing benzothiophene and dibenzothiophene sulfur species as well as thiophene sulfur species. The reduction of sulfur takes place predominantly for substituted benzothiophenes and substituted dibenzothiophenes. These results suggest that C-S bonds in alkyl substituted-

10

The facile sulfur reduction of substituted-thiophenes, substituted-benzothiophenes, and substituted-dibenzothiophenes was quite unexpected and will enhance the effectiveness of a subsequent LCO hydrodesulfurization process. It is well known for

15 LCO hydrodesulfurization process that methyl and/or alkyl substitution of benzothiophene and dibenzothiophene makes desulfurization reactivity of the organic sulfurs decline substantially and become "hard sulfur" or "refractory sulfur". Our LCO containing a lower amount of substituted benzo and dibenzothiophenes will generate a diesel fuel with a lower sulfur content after a hydrodesulfurization process than the

20 comparable LCO generated by a conventional FCC catalyst.

**CLAIMS:**

1. A method of reducing the sulfur content of a catalytically cracked petroleum fraction, which comprises catalytically cracking a petroleum feed fraction containing
5 organosulfur compounds at elevated temperature in the presence of a cracking catalyst and a product sulfur reduction catalyst which comprises a porous molecular sieve having (i) a first metal component which is within the interior pore structure of the molecular sieve and which comprises a metal in an oxidation state greater than zero and (ii) a second metal component which is within the interior pore structure of
10 the molecular sieve and which comprises at least one rare earth, to produce liquid cracking products of reduced sulfur content.
2. A method according to claim 1 in which the product sulfur reduction catalyst comprises a large pore size or intermediate pore size zeolite as the molecular sieve
15 component and, as the first metal component, at least one of vanadium, zinc, iron, cobalt, or gallium.
3. A method according to claim 2 in which the large pore size zeolite comprises a
20 faujasite zeolite.
4. A method according to claim 2 in which the first metal component comprises vanadium.
5. A method according to claim 2 in which the second metal component
25 comprises lanthanum alone or in combination with cerium.
6. A method according to claim 2 in which the second metal component comprises cerium.
- 30 7. A method according to claim 1 in which the second metal component is present in an amount from 1 to 10 weight percent of the catalytic composition.



8. A method according to claim 1 in which the product sulfur reduction catalyst comprises a USY zeolite having a UCS of from 2.420 to 2.460 nm, a bulk silica:alumina ratio of at least 5.0 as the molecular sieve component and, as the first metal component, at least one of zinc or vanadium in an oxidation state greater than zero and, as the second metal component, cerium alone or in combination with lanthanum
9. A method according to claim 1 in which the sulfur reduction catalyst is a separate particle additive catalyst.
10. A method according to claim 1 in which the liquid cracking products of reduced sulfur content comprise a gasoline fraction, a cycle oil fraction, and a fuel oil fraction where the cycle oil and fuel oil boil above the gasoline fraction.
11. A fluid catalytic cracking process in which a heavy hydrocarbon feed containing organosulfur compounds is catalytically cracked to lighter products by contact in a cyclic catalyst recirculation cracking process with a circulating fluidizable catalytic cracking catalyst inventory consisting of particles having a size ranging from 20 to 100 microns, by:
- (i) catalytically cracking the feed in a catalytic cracking zone operating at catalytic cracking conditions by contacting feed with a source of regenerated cracking catalyst to produce a cracking zone effluent comprising cracked products and spent catalyst containing coke and strippable hydrocarbons;
 - (ii) discharging and separating the effluent mixture into a cracked product rich vapor phase and a solids rich phase comprising spent catalyst;
 - (iii) removing the vapor phase as a product and fractionating the vapor to form liquid cracking products including gasoline,
 - (iv) stripping the solids rich spent catalyst phase to remove occluded hydrocarbons from the catalyst,
 - (v) transporting stripped catalyst from the stripper to a catalyst regenerator;
 - (vi) regenerating stripped catalyst by contact with oxygen containing gas to produce regenerated catalyst; and



(vii) recycling the regenerated catalyst to the cracking zone to contact further quantities of heavy hydrocarbon feed,

in which the sulfur content of the liquid cracking products is reduced by carrying out the catalytic cracking in the presence of a product sulfur reduction catalyst which comprises a porous molecular sieve having (i) a first metal component which is within the interior pore structure of the molecular sieve and which comprises a metal in an oxidation state greater than zero and (ii) a second metal component which is within the interior pore structure of the molecular sieve and which comprises at least one rare earth.

12. A method according to claim 11 in which the cracking catalyst comprises a matrixed faujasite zeolite.

13. A method according to claim 12 in which the product sulfur reduction catalyst comprises a large pore size or intermediate pore size zeolite as the molecular sieve component, vanadium as the first metal component and cerium, alone or in combination with at least one other rare earth metal as the second metal component.

14. A method according to claim 13 in which the large pore size zeolite of the product sulfur reduction catalyst comprises a faujasite zeolite.

15. A method according to claim 11 in which the first metal component comprises vanadium.

16. A method according to claim 11 in which the second metal component comprises lanthanum alone or in combination with cerium.

17. A method according to claim 11 in which the second metal component comprises cerium.

18. A method according to claim 11 in which the second metal component is present in an amount from 1 to 10 weight percent of the catalytic composition.



19. A method according to claim 11 in which the liquid cracking products comprise a gasoline fraction of reduced sulfur content and a cycle oil fraction of reduced sulfur content boiling above the gasoline fraction.

5 20. A fluidizable catalytic cracking product sulfur reduction catalyst composition for reducing the sulfur content of a catalytically cracked gasoline fraction during the catalytic cracking process, which comprises fluidizable particles having a size ranging from 20 to 100 microns of (i) a porous molecular sieve component, (ii) a first metal component comprising a metal in an oxidation state greater than zero located within
10 the interior pore structure of the porous molecular sieve component and (iii) a second metal component comprising a rare earth metal located within the interior pore structure of the porous molecular sieve component.

21. A fluidizable catalytic cracking product sulfur reduction catalyst composition
15 according to claim 20 in which the porous molecular sieve component comprises a porous hydrocarbon cracking sieve component.

22. A fluidizable catalytic cracking product sulfur reduction catalyst composition according to claim 21 in which the porous molecular sieve component comprises
20 zeolite USY having a UCS of from 2.420 to 2.460 nm and a bulk silica:alumina ratio of at least 5.0.

23. A fluidizable catalytic cracking product sulfur reduction catalyst composition according to claim 22 in which the porous molecular sieve component comprises
25 zeolite USY having a UCS of from 2.420 to 2.435 nm and a bulk silica:alumina ratio of at least 5.0.

24. A fluidizable catalytic cracking product sulfur reduction catalyst composition according to claim 20 which contains from 0.1 to 5 weight percent vanadium as the
30 first metal component, based on the weight of the zeolite, of the first metal component.



25. A fluidizable catalytic cracking product sulfur reduction catalyst composition according to claim 24 which comprises as the second metal component, a combination of cerium and at least one other rare earth.
- 5 26. A fluidizable catalytic cracking product sulfur reduction catalyst composition according to claim 20 which comprises cerium as the second metal component.
27. A fluidizable catalytic cracking product sulfur reduction catalyst composition according to claim 20 in which the metal components have has been introduced into
10 the zeolite as exchanged cationic species within the zeolite pores.
28. A fluidizable catalytic cracking product sulfur reduction catalyst according to claim 20 which is formulated with a matrix component as a fluid cracking catalyst additive.
- 15 29. A fluidizable catalytic cracking product sulfur reduction catalyst according to claim 20 which is formulated as an integrated fluidizable catalytic cracking/product sulfur reduction catalyst for cracking a heavy hydrocarbon feed to produce liquid cracking products including gasoline and reducing the sulfur content of the
20 catalytically cracked gasoline fraction during the catalytic cracking process, which comprises fluidizable particles having a size ranging from 20 to 100 microns of a hydrocarbon cracking component which comprises a zeolitic molecular sieve which contains located within the pore structure of the zeolite (i) a first metal component which comprises at least one of at least one of vanadium, zinc, iron, cobalt, and
25 gallium and (ii) a second metal component which comprises at least one rare earth.
30. An integrated fluidizable catalytic cracking/product sulfur reduction catalyst according to claim 29 which contains from 0.1 to 5 weight percent, based on the weight of the zeolite, of vanadium as the first metal component.
- 30 31. An integrated fluidizable catalytic cracking product sulfur reduction catalyst according to claim 29 in which the second metal component comprises a combination



of cerium and at least one other rare earth in an amount from 1 to 5 weight percent of the catalyst.

32. An integrated fluidizable catalytic cracking product sulfur reduction catalyst according to claim 29 in which the second metal component comprises a combination of cerium in an amount from 1 to 5 weight percent of the catalyst.

33. An integrated fluidizable catalytic cracking product sulfur reduction catalyst according to claim 29 in which the zeolitic molecular sieve comprises zeolite USY having a UCS of from 2.420 to 2.460 nm and a bulk silica:alumina ratio of at least 5.0.

34. A fluidizable catalytic cracking product sulfur reduction catalyst composition according to claim 32 in which the porous molecular sieve component comprises zeolite USY having a UCS of from 2.420 to 2.435 nm and a bulk silica:alumina ratio of at least 5.0.

35. A fluidizable catalytic cracking product sulfur reduction catalyst according to claim 29 which is formulated as an integrated fluid cracking/sulfur reduction catalyst with a matrix component and a faujasite zeolite as the cracking component.