

HYDROCRACKING OF DIPHENYLALKANES

Field Of The Invention

5 The present invention relates to catalytic hydrocracking of a diphenylalkane to produce alkylbenzene. The present invention further relates to a process for catalytic hydrocracking of 2-methyl-2,4-diphenylpentane to produce isopropyl benzene ("cumene").

Background Of The Invention

10 Substantial quantities of diphenylalkanes are produced as a byproduct in various chemical processes involving the reaction or the production of (substituted) benzene-containing compounds such as styrene, phenol, dialkyl benzene and cumene. For example, 2-methyl-2,4-diphenylpentane and dimethyl-2,3-diphenylbutane are produced as byproducts in many processes which use cumene, such as in the cumene-phenol
15 peroxidation process, and in the process for the recovery of cumene from cumyl alcohol.

 WO01/70714, assigned to Sumitomo, describes a process relating to oxidizing isopropylbenzene (also known as "cumene") to obtain isopropylbenzene peroxide as an oxygen carrier for the epoxidation of propylene to produce propylene oxide and isopropylbenzene alcohol (cumyl alcohol). The isopropylaryl alcohol is
20 dehydrated/hydrogenated, via a hydrogenolysis step, to isopropylbenzene (cumene) and recycled for repeated use. During the hydrogenolysis step, substantial quantities of cumene dimer are produced as undesirable by-products, because cumene can stay in the hydrogenation bed after it is produced, but before exiting, resulting in the further hydrogenation or dimerization thereof.

25 U.S. Patent Number 6,455,712, assigned to Shell, describes a process for producing alkylene oxide (also known as oxirane) compounds, such as propylene oxide, by oxidizing olefin with alkylbenzene hydroperoxide obtained by oxidizing alkylbenzene with oxygen. The alkylbenzene hydroperoxide is converted to alkylaryl alcohol which is dehydrated/hydrogenated via a multi-step process of hydrogenolysis followed by
30 fractionation to separate alkylbenzene, from other side products, for reuse for making alkylbenzene. Substantial quantities of alkylaryl alcohol could be converted to dimers/oligomers of alkylbenzene as side products.

 WO02/072507, assigned to Sumitomo, describes a process for hydrocracking of 2,3-dimethyl-2,3-diphenylbutane in the presence of a copper chromium catalyst. The

conversion rate at 220 °C is only 84%. Moreover, there are usually also other cumene dimers, such as 2-methyl-2,4-diphenylpentane, produced as side-products in processes involving cumene which are not mentioned in this patent.

It is therefore desirable to develop an efficient process which would recover undesirable alkylbenzene dimers or oligomers by-products produced to higher value alkylbenzene with high yield at relatively low temperature and low costs, so that alkylbenzene can be cost effectively recycled for repeated use in the same commercial process or for other industrial use.

Summary of the Invention

The invention is directed to a hydrocracking process comprising the steps of:

reacting a diphenyl alkane having a formulation of $R_1R_2C(Ph)-(C)_n(H)_m-C(Ph)R_3R_4$ with hydrogen using a catalyst comprising a metal selected from the group consisting of Group IB and Group VIII metal compounds on an acidic support to produce alkylbenzene(s) having a structure of $R_1R_2C(Ph)R_5$ and $R_6(Ph)CR_3R_4$;

wherein the total number of carbon atoms for R_5 and R_6 is equal to n ; and,

wherein R_1 , R_2 , R_3 , R_4 each is a hydrogen or a hydrocarbon group having 1-10 carbon atoms.

Detailed Description of the Invention

The present invention relates to a process for catalytic hydrocracking of a diphenyl alkane to produce an alkylbenzene. Illustrative and non-limiting examples of suitable hydrogenation or hydrocracking catalysts include catalysts comprising Group VIII metal or Group IB metal, particularly those comprising copper, palladium, platinum and nickel.

As a particular embodiment of the present invention, the catalyst employed is supported by an acidic carrier (also known as "support"), particularly a carrier which is in an "acidic hydrogen form". The term of "acidic hydrogen form" means the 50% or more of ion exchangeable cations are protons (also known as " H^+ " or "hydrogen (+) ions"). Not intended to be bound by the theory, it is proposed that acidic hydrogens help promote the hydrocracking reaction and increase the acidity of the catalyst. The highly active catalyst is able to hydrocrack the diphenylalkane at relatively mild conditions, thereby minimizing the production of side products such as alkyl cyclohexanes and lower the utility costs. Non-limiting illustrative examples of acidic support include silica, silica-alumina, and zeolite, such as Mordenite, Na/H-Mordenite, H-Mordenite, beta-zeolite, H-beta-zeolite, Y-zeolite,

H-Y-zeolite, and the like. When zeolites are used, it is preferred that they are in the acidic hydrogen form.

In one particular embodiment of the present invention, the catalysts contain from 0.1 to 5%wt., particularly from 0.2 to 2 %wt, calculated as the weight of the metal on the basis of the total weight of the catalyst, of Group VIII metal or a Group VIII metal compound as principal catalytic component, alone or with promoters and modifiers such as palladium/gold, palladium/silver, cobalt/zirconium, nickel, preferably deposited on an acidic support, more preferably deposited on an acidic support in acidic hydrogen form. One illustrative example of a suitable catalytic material comprises palladium oxide or palladium, particularly 0.1 to 5.0 % by weight, and more particularly 0.2 to 2% by weight, calculated as the weight of the metal on the basis the total weight of the catalyst, supported on an appropriate support medium such as zeolite (particularly Mordenite or H-Mordenite) or silica. A preferred catalyst comprises from 0.2 to 2% by weight of palladium on H-Mordenite.

As other particular embodiments of the present invention, Group IB metals of the Periodic Table of Elements, such as copper, are used as the principle catalytic component, alone or with promoters and modifiers such as chromium, zinc, zirconium, aluminum, magnesium, a rare earth metal, Group VIII metals, etc. for hydrogenation. The Group IB metal-containing catalysts preferably contain from 10 to 80%, particularly from 30 to 75, more particularly from 50 to 70, as the weight of the oxide basis the total weight of the catalyst, of a Group IB metal, particularly on an acidic support, more particularly on a support in acidic hydrogen form. Specific non-limiting examples of such catalysts include catalysts comprising copper, Raney copper, copper/chrome, copper/zinc, copper/zinc/chrome, copper/zinc/zirconium, copper/silica, copper/alumina and other copper-based catalyst systems, particularly those supported by an acidic support, more particularly those supported by a support in an acidic hydrogen form. Some specific illustrative examples include commercially available Copper on Silica catalyst, T-366 (having approximately 54 wt.% of copper on silica as a press extrudate or formed extrudate), obtainable from Sud Chemie; Copper Chromite catalyst, G-22/2, obtainable from Sud Chemie; and Cu/Zn/Zr catalyst prepared according Example 3 of US patent 5,475,159; and the like. Combinations of these catalysts can also be used. Such catalysts were found to give good results at relatively low temperature. In one non-limiting illustrative embodiment of the present invention, these catalysts are preferably used at a

temperature of from 100 to 250 °C. Such catalysts may, but is not limited to, comprise from 5 percent by weight to 80 percent by weight, calculated as the oxide, basis the total weight of the catalyst, of copper. Further, such catalysts may contain from 10 percent by weight to 80 percent by weight, calculated as the oxide, basis the total weight of the catalyst, of zinc, from 15 to 85 percent by weight of acidic carrier, particularly a support in acidic hydrogen form. A particular example of the catalyst contains from 10 percent by weight to 80 percent by weight, calculated as the oxide, basis the total weight of the catalyst, of copper, from 10 percent by weight to 80 percent by weight, calculated as the oxide, basis the total weight of the catalyst, of zinc, from 0.1 percent by weight to 20 percent by weight, calculated as the oxide, basis the total weight of the catalyst, of rare earth and from 10 to 80 weight percent of acidic carrier, particularly a support in acidic hydrogen form. A further particular catalyst contains from 10 percent by weight to 80 percent by weight, calculated as the oxide, basis the total weight of the catalyst, of copper, from 10 percent by weight to 80 percent by weight, calculated as the oxide, basis the total weight of the catalyst, of zinc, and from 0.05 percent by weight to 30 percent by weight, basis the total weight of the catalyst, of aluminum. A further particular catalyst contains from 10 percent by weight to 80 percent by weight, calculated as the oxide, basis the total weight of the catalyst, of copper, from 10 percent by weight to 80 percent by weight, calculated as the oxide, basis the total weight of the catalyst, of zinc, from 0.05 percent by weight to 30 percent by weight, basis the total weight of the catalyst, of zirconium, and 10 to 80 weight percent of acidic carrier, particularly a support in acidic hydrogen form. Another preferred catalyst contains from 10 percent by weight to 80 percent by weight, calculated as the oxide, basis the total weight of the catalyst, of copper, from 10 percent by weight to 80 percent by weight, calculated as the oxide, basis the total weight of the catalyst, of zinc, from 0.05 percent by weight to 30 percent by weight, basis the total weight of the catalyst, of zirconium, and from 0.05 percent by weight to 30 percent by weight, basis the total weight of the catalyst, of aluminum, and 10 to 80 weight percent of acidic carrier, particularly a support in acidic hydrogen form. And a further preferred catalyst contains from 10 percent by weight to 80 percent by weight, calculated as the oxide, basis the total weight of the catalyst, of copper, from 10 percent by weight to 80 percent by weight, calculated as the oxide, basis the total weight of the catalyst, of zinc, from 0.05 percent by weight to 30 percent by weight, basis the total weight of the catalyst, of magnesium, from 0.1 percent by weight to 20 percent by weight, calculated as the oxide,

basis the total weight of the catalyst, of rare earth, and 10 to 80 weight percent of acidic carrier, particularly a support in acidic hydrogen form.

As a particular embodiment of the present invention, the afore-mentioned hydrocracking catalysts are reduced by hydrogen prior to being loaded in the reactor. As a non-limiting illustrative example, a hydrocracking catalyst is crushed and sized into appropriate size, e.g. 6-20 mesh, particles. The catalyst is introduced into a reactor and slowly reduced by heating the catalyst particles to a temperature of e.g. 150-250 °C at a rate of from 1 to 10°C, particularly from 1.5 to 5 °C per minute, while flowing 0.001 to 0.1, specifically 0.02 – 0.10 wt. % hydrogen in nitrogen at a rate of 1-200, specifically 2-30 L/Hr. The catalyst is allowed to reduce at 150-250°C for 1-10 hours and then the hydrogen content in the nitrogen is doubled every 1-5 hours until the gas is 1-10, specifically 2-5 wt.% hydrogen in nitrogen. Catalysts containing copper are preferably reduced at a temperature of between 150-200 °C to minimize sintering. The catalyst is reduced for a final one to five hour period and then cooled while maintaining gas flow. After cooling, the reactor is capped without allowing any air to enter and the gas flow is stopped. The reactor is opened in a nitrogen filled environment and the catalyst.

The particles of reduced catalyst, prepared by the afore-mentioned procedure are loaded onto a reactor over a bed supports, e.g. made of porous plate/tray or screen, optionally in a nitrogen filled environment. The reduced catalysts are sized and shaped to stay above the bed supports.

Without limiting the scope of the present invention, the diphenylalkane has a formulation of $R_1R_2C(Ph)-(C)_n(H)_m-C(Ph)R_3R_4$ and the alkylbenzene produced has a structure of $R_1R_2C(Ph)R_5$ and $R_6(Ph)CR_3R_4$; wherein the total number of carbon atoms for R_5 and R_6 is equal to n ; n is from 0 to 10, preferable from 0 to 5; wherein R_1 , R_2 , R_3 , R_4 each is a H or a hydrocarbon group having 1-10 carbon atoms. The number of hydrogen is equal or less than $2n$, dependent on degree of branching of the $(C)_n(H)_m$;

As one particular embodiment of the present invention, the diphenylalkane is 2-methyl-2,4-diphenylpentane. Another particular embodiment relates to a mixture of 2-methyl-2,4-diphenylpentane and dimethyl-2,3-diphenylbutane.

As another particular embodiments of the present invention, additional alkyl groups can be present on the aromatic rings of the diphenylalkane, and the alkylbenzene produced comprises an alkylbenzene selected from the group consisting of *para*-di(iso-propyl)benzene, *meta*-di(iso-propyl)benzene, *ortho*-di(iso-propyl)benzene, 1-ethyl-4-(iso-

propyl)benzene, 1-ethyl-3-(iso-propyl)benzene, 1-ethyl-4-(iso-propyl)benzene and mixtures thereof.

The pressure for the hydrocracking reaction should range between from 0 to 400 psig, particularly from 5-300 psig, and more particularly from 0 to 140 psig (or 1-10 bar).

5 The temperature of the hydrocracking reactor is from 140 to 300°C, particularly from 160 to 280°C, more particularly from 180 to 235°C.

The hydrogenation flow rate must be adjusted such that it is sufficient to support the hydrocracking reaction and replace hydrogen lost from the catalyst. At least a stoichiometric amount of hydrogen relative to the alkenylbenzene dimers must be present
10 in the system to be available for the reaction. As a non-limiting example, a small excess of hydrogen flow is provided to occlude the hydrogen into the liquid and to accommodate the nature of this reaction between a gas and a liquid.

The feed weight hourly space velocity (WHSV), may vary over a very wide range within the other condition perimeters, and can be from 0.1 to 100, from 0.2 to 20
15 particularly from 0.5 to 15 liters per hour. WHSV, as used herein, means the unit weight of feed per hour entering the reaction distillation reactor per unit weight of catalyst in the reactor catalyst bed.

The % conversion of the dimers of alkylbenzene to alkylbenzene, as used herein, is defined as the following:

20

$$\% \text{ conversion of dimers} = \frac{(\text{wt.\% of dimers in feed} - \text{wt.\% of dimers in product}) \times 100}{\text{Wt.\% of dimers in feed}}$$

As a particular embodiment of the present invention, from 86.0 to 100.0% by
25 weight, particularly from 90.0% to 100.0% by weight, more particularly from 94.0 to 100%, still more particular from 97.5.0 to 100.0%, and still more particular from 98.0 to 100.0 by weight of the dimers of alkylbenzene in the hydrocracking reaction media can be converted to compounds with (only)one benzene ring at a temperature from 185°C to 235 °C. The same ranges of cracking also can be accomplished at from 140°C to 300°C,
30 particularly from 160°C to 280° C. As an illustrative particular embodiment, from 97.5 to 100.0 percent by weight of dimers of alkylbenzene is converted to compounds with (only) one benzene ring at a temperature from 185°C to 235 °C. As another illustrative particular embodiment, from 97.5 to 100.0, more particularly from 98.0 to 100.0 percent by weight of

dimers of alkylbenzene is converted to compounds with (only) one benzene ring at a temperature from 185°C to 225 °C.

As a particular embodiment of the present invention, from 86.0 to 100.0% by weight, particularly from 90.0% to 100.0%, more particularly from 94.0% to 100.0%,
5 still more particular from 97.5 to 100.0% and still more particular from 98.0 to 100.0 % by weight of the 2-methyl-2,4-di(3-isopropylphenyl)pentane can be converted to compounds with only one benzene ring such as isopropylbenzene at a temperature from 180°C to 235°C. As another particular embodiment of the present invention, from 86.0 to 100.0% by weight, particularly from 90.0% to 100.0%, more particularly from 94.0% to 100%, still
10 more particular from 97.5 to 100%, and still more particular from 98.0 to 100.0% by weight of the the 2-methyl-2,4-di(3-isopropylphenyl)pentane can be converted to isopropylbenzene at a temperature from 185°C to 235°C. The same ranges of cracking also can be accomplished at from 140°C to 300°C, particularly from 160° C to 280° C. As an illustrative particular embodiment, from 97.5 to 100.0 percent by weight of 2-methyl-2,4-
15 di(3-isopropylphenyl)pentane is converted to cumene at a temperature from 185°C to 235 °C. As another illustrative particular embodiment, from 97.5 % by weight, and more particularly from 98.0 to 100.0 percent by weight of 2-methyl-2,4-di(3-isopropylphenyl)pentane is converted to cumene at a temperature from 185°C to 225 °C.

As a particular embodiment of the present invention, from 86.0 to 100% by weight, particularly from 90.0% to 100.0%, more particularly from 94.0% to 100.0%, still
20 more particularly from 97.5% to 100%, and still more particularly from 98.0 to 100.0% by weight of the mixture of 2-methyl-2,4-di(3-isopropylphenyl)pentane and 2,3-dimethyl-2,3-diphenylbutane can be converted to compounds with only one benzene ring at a temperature from 180° C to 235 °C. The same ranges of cracking also can be accomplished
25 at from 140°C to 300°C, particularly from 160° C to 280° C. As an illustrative particular embodiment, from 97.5 to 100.0 percent by weight of mixtures of 2-methyl-2,4-di(3-isopropylphenyl)pentane and 2,3-dimethyl-2,3-diphenylbutane is converted to cumene at a temperature from 185°C to 235 °C. As another illustrative particular embodiment, from 97.5 to 100.0%, and more particularly from 98.0 to 100.0 percent by weight of a mixture of
30 2-methyl-2,4-di(3-isopropylphenyl)pentane and 2,3-dimethyl-2,3-diphenylbutane is converted to cumene at a temperature from 185°C to 225 °C.

The feed steam may come from a fixed bed reactor (recovered after fractional distillation), from the bottom stream of a catalytic distillation reactor of a

alkylphenylalcohol hydrogenation reaction, optionally fractionated to remove lighter weight alkylphenylalcohol etc., or from a cumene-phenol peroxidation process. These streams are optionally diluted with alkylbenzene prior to be fed to the hydrocracking reactor to moderate the heat generated in the hydrocracking reaction thereby improving the control of the reaction in the hydrocracker .

As an illustrative example, the feed stream contains from 0.1 to 100, particularly from 0.1 to 10, and more particularly from 0.2 to 2 % by weight of diphenylalkane; from 0 to 99, particularly from 25 to 95, and more particularly from 60 to 90% by weight of alkylbenzene; from 0 to 20, particularly from 0 to 5, and more particularly from 0 to 1 % by weight of alkenylbenzene; and from 0 to 25, particularly from 0 to 10, and more particularly from 0 to 5 % by weight of alkylphenyl alcohol. As specific illustrative example of a particular embodiment of the present invention, the feed stream contains from 0.1 to 100, particularly from 0.1 to 1-, and more particularly from 0.2 to 2% by weight of 2,3-dimethyl-2,3-diphenyl butane and/or 2-methyl-2,4-diphenyl pentane; from 0 to 99, particularly from 25 to 95, and more particularly from 60 to 90 % by weight of cumene; from 0 to 20, particularly from 0 to 5, and more particularly from 0 to 1 % by weight of alpha-methyl styrene; from 0 to 5 weight percent of ethyl benzene; from 0 to 5 weight percent of di-, tri-isopropyl benzene, ethyl benzene, propyl-benzene, ethyl-isopropyl benzene, etc., or mixtures thereof; and from 0 to 25, particularly from 0 to 10, and more particularly from 0 to 5 % by weight of cumyl alcohol.

The invention will be illustrated by the following illustrative embodiments which are provided for illustration purpose only and are not intended to limit the scope of the instant invention.

Illustrative Embodiments

I. Illustrative Embodiment I:

I(A) Preparation of Hydrocracking Catalyst Pd-Mordenite Catalyst

5 A mixture of 1500 grams of sodium mordenite, (having the following properties: a surface area of 430 square meters per gram; an average crystallite size of around 1 micron; a cyclohexane adsorption uptake of 7.6 cc/g; and a molar silica to alumina ratio of 11.1), 9000 grams of ammonium nitrate and 15 liters of 1.5 M nitric acid was heated to 50 °C. and stirred for five hours. The solid material was filtered off and washed with 25 liters of
10 deionized water. This treatment of the Mordenite with ammonium nitrate in nitric acid was repeated twice with fresh ammonium nitrate and nitric acid each time. After each treatment the solid material was filtered off and washed with water and dried overnight at 120 °C. Palladium was added to the zeolite to a level of 0.35 percent by weight by treatment with an aqueous solution containing Tetraamine palladium nitrate and an excess
15 of ammonium nitrate prepared by dissolving 6.55 grams of tetramine palladium nitrate in 308 grams of deionized water and adding to this solution 4.92 grams of ammonium nitrate. The palladium solution was then co-mulled with 1083 grams of dealuminated mordenite having an LOI (loss of ignition at 750 °C for 2 hours) of 10.6%. The palladium-containing mordenite was uniformly mixed and then 338 grams of pseudoboehmite
20 alumina (Catapal B which is commercially available from Vista Chemical Company) having an LOI of 28.4% was added and allowed to mix. The mixture was extruded and the 1.6 mm extrudates were dried in air for 16 hours at 125 °C, and then calcined in flowing air at 500°C. for two hours. The catalyst was crushed and sized to 6-20 mesh particles and then reduced using the procedure as described in IIA below.

IB. Hydrocracking of Cumene Dimer Using Pd on H-Mordenite Catalyst

The bottom stream from a catalytic distillation column, for the catalytic distillation of cumyl alcohol to produce cumene, was distilled to yield a cumene dimer rich mixture that was diluted with cumene and fed into a fixed bed hydrogenation loaded with the acidic
30 palladium on H-Mordenite catalyst as described in I(A) for hydrocracking under the condition as provided in TABLE 1 below. The results are shown in TABLE 2 below.

Table 1

Feedrate	33.5 g/hr
Reaction Temperature	220°C
Pressure	10 bar
Hydrogen Flowrate	4 L/Hr
Catalyst Weight	33.5 g (before reduction)

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Table 2

Fixed Bed Cumene Dimers To Cumene Results
With Palladium On H-Mordenite Catalyst At 220°C

Component	Cumene dimers in cumene (FEED)	Fixed Bed Product
2,3-Dimethyl-2,3-diphenylbutane, (wt%)	1.94	0.06
2-methyl-2,4-diphenylpentane, (wt%)	1.03	0.02
Cumene, (wt%)	96.75	99.82
Isopropylcyclohexane, (wt%)	0.08	0.08
<i>Alpha</i> -Methyl styrene, (wt%)	0.20	0.02

10 II. Illustrative Embodiment II

IIA Preparation of Hydrocracking Catalyst T-366 Catalyst

A commercially available copper on silica catalyst, T-366, available from Sud Chemie, having 54 wt.% of Cu on silica extruded into 3.2 mm extrudate, is further processed using the following procedure for the catalytic cracking experiments.

15 Five grams of Sud Chemie T-366 copper on silica catalyst (3mm tablets) was crushed and sized into 6-20 mesh particles. The catalyst was mixed with 45 grams of 80 mesh silicon carbide and centered inside a 69 cm long stainless steel reactor tube between beds of 20 mesh SiC and glass wool. The reactor tube had an internal diameter of 1.5cm.

The catalyst was slowly reduced by heating the catalyst particles at a rate of 3°C per minute from 20°C to 180°C while flowing 0.05 wt. % hydrogen in nitrogen at a rate of 10 L/Hr. The catalyst was allowed to reduce at 180°C for 2 hours and then the hydrogen content in the nitrogen was doubled every 2 hours until the gas was 3.2 wt.% hydrogen in nitrogen. The catalyst was reduced for a final two-hour period and then cooled while maintaining gas flow. After cooling, the reactor was capped without allowing any air to enter and the gas flow was stopped. The reactor was opened in a nitrogen filled glove box and the catalyst and silicon carbide were separated by screen sieve.

IIB. Hydrocracking of Cumene Dimers Using T-366 Copper on Silica Catalyst

The bottom stream from a catalytic distillation column, for the catalytic distillation of cumyl alcohol to produce cumene, was distilled to yield a cumene dimer rich mixture that was diluted with cumene and fed into a fixed bed hydrogenation loaded with the T-366 catalyst as described in II(A) for hydrocracking under the condition as provided in TABLE 3 below. The results are shown in TABLE 4 below.

Table 3

Feedrate	33.5 g/hr
Reaction Temperature	260°C
Pressure	10 bar
Hydrogen Flowrate	4 L/Hr
Catalyst Weight	33.5 g (before reduction)

Table 4

Fixed Bed Cumene Dimer To Cumene Results
With T-366 Catalyst At 260°C

Component	Cumene dimers in cumene (FEED)	Fixed Bed Product
2,3-Dimethyl-2,3-diphenylbutane, (wt%)	1.94	0.05
2-methyl-2,4-diphenylpentane, (wt%)	1.03	0.09
Cumene, (wt%)	96.75	99.54
Isopropylcyclohexane, (wt%)	0.08	0.18
alpha-Methyl styrene, (wt%)	0.20	0.09

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III. Comparative Example

IIIA. Preparation of Hydrogenation Catalyst Palladium on Carbon

Pressed granules of 0.5 wt.% of palladium on carbon, available from the Calsicat division of Mallinckrodt Incorporated was mixed with silicon carbide and reduced according the same procedure as described in IIA above.

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IIIB. Hydrocracking of Cumene Dimers with Palladium on Carbon Catalyst

The bottom stream from a catalytic distillation column, for the catalytic distillation of cumyl alcohol to produce cumene, was distilled to yield a cumene dimer rich mixture that was diluted with cumene and fed into a fixed bed hydrogenation loaded with the acidic palladium on carbon catalyst as described in III(A) for hydrocracking under the condition as provided in TABLE 5 below. The results are shown in TABLE 6 below.

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Table 5

Feedrate	33.5 g/hr
Reaction Temperature	220 °C
Pressure	10 bar
Hydrogen Flowrate	4 L/Hr
Catalyst Weight	33.5 g (before reduction)

Table 6

Fixed Bed Cumene Dimers To Cumene Results
With Palladium on Carbon Catalyst at 220°C

Component	Cumene dimers in cumene (FEED)	Fixed Bed Product
2,3-Dimethyl-2,3-diphenylbutane, (wt%)	1.94	0.10
2-methyl-2,4-diphenylpentane, (wt%)	1.03	0.10
Cumene, (wt%)	96.75	99.60
Isopropylcyclohexane, (wt%)	0.08	0.16
<i>Alpha</i> -Methyl styrene, (wt%)	0.20	0.04

IV. Conclusion

As shown in Illustrative Embodiment IB, when a catalyst of Pd loaded on a highly acidic support is used for hydrocracking cumene dimers 2,3-dimethyl-2,3-diphenylbutane and 2-methyl-2,4-diphenylpentane to cumene, the conversion rate at 220°C C is 98.0% and the selectivity to cumene is close to 100%. As shown in Illustrative Embodiment IIB, when a catalyst of Cu loaded on silica (an acidic support) is used for hydrocracking of cumene

dimers (2,3-dimethyl-2,3-diphenylbutane and 2-methyl-2,4-diphenylpentane) to cumene, the conversion rate is 95.3%, and the selectivity is 100%.

In contrast, as shown in Comparative Example IIIB, when a non acidic catalyst of 0.5 wt.% Pd loaded on a carbon support is used for hydrocracking of cumene dimers (2,3-
5 dimethyl-2,3-diphenylbutane and 2-methyl-2,4-diphenylpentane) to cumene, the conversion rate is 93.3, and the selectivity is 100 %.

In view of the foregoing, Pd on a highly acidic support (H-Mordenite) presents superior property of higher hydrocracking conversion rate, compared to the same loaded on a non-acidic support.

10 The ranges and limitations provided in the instant specification and claims are those which are believed to particularly point out and distinctly claim the instant invention. It is, however, understood that other ranges and limitations that perform substantially the same function in substantially the same manner to obtain the same or substantially the same result are intended to be within the scope of the instant invention as defined by the instant
15 specification and claims.

C L A I M S

1. A hydrocracking process comprising the steps of:
reacting a diphenyl alkane having a formulation of $R_1R_2C(Ph)-(C)_n(H)_m-$
 $C(Ph)R_3R_4$ with hydrogen using a catalyst comprising a metal selected from the
group consisting of Group IB and Group VIII metal compounds on an acidic
support to produce alkylbenzene(s) having a structure of $R_1R_2C(Ph)R_5$ and
 $R_6(Ph)CR_3R_4$;
wherein the total number of carbon atoms for R_5 and R_6 is equal to n ; and,
wherein R_1, R_2, R_3, R_4 each is a hydrogen or a hydrocarbon group having 1-10
carbon atoms.
2. The process according to claim 1, wherein the catalyst comprises from 0.1 % to 5
% by weight, preferably from 0.2 % to 2 % by weight of a metal selected from the
Group VIII metal compounds, preferably palladium; the acidic support is selected
from the group consisting of silica, silica-alumina, zeolites, and mixtures thereof,
preferably H-mordenite; the reaction is performed at a temperature of from 140°C
to 300°C, preferably of from 180°C to 235°C; wherein from 86.0% to 100.0%,
preferably 90.0% to 100.0% by weight of the diphenyl alkane is converted to
compounds with only one benzene ring, preferably isopropylbenzene.
3. The process according to claim 1 or 2, wherein the catalyst comprises from 10 % to
80 % by weight, preferably from 30 % to 75%, more preferably from 50% to 70%,
based on the total weight of catalyst, of a Group IB metal, preferably copper.
4. The process according to any of claims 1 to 3, wherein the diphenyl alkane
comprises a mixture of 2-methyl-2,4-di(3-isopropylphenyl)pentane and 2,3-
dimethyl-2,3-di(3-isopropylphenyl) butane; and, wherein from 86.0% to 100.0% by
weight of the combined weight of 2-methyl-2,4-di(3-isopropylphenyl)pentane, 3-
dimethyl-2,3-di(3-isopropylphenyl) butane is converted to isopropylbenzene.
5. The process according to any of claims 1 to 3 wherein the diphenyl alkane
comprises 2-methyl-2,4-di(3-isopropylphenyl)pentane.

6. The process according to any of the previous claims, wherein said alkylbenzene produced comprises an alkylbenzene selected from the group consisting of *para*-di(iso-propyl) benzene, *meta*-di(iso-propyl)benzene, *ortho*-di(iso-propyl)benzene, 1-ethyl-4-(iso-propyl)benzene, 1-ethyl-3-(iso-propyl)benzene, and 1-ethyl-4-(iso-propyl)benzene.
7. The process according to any of the previous claims, wherein the catalyst is reduced by hydrogen prior to reacting the diphenyl alkane with hydrogen.
8. The process according to claim 7 wherein the catalyst is reduced by heating to a temperature of 150-250 °C at a rate of from 1 °C to 10 °C, preferably from 1.5 °C to 5 °C per minute, while flowing 0.001 % to 0.1 % wt, preferably 0.02 % to 0.10 % wt, hydrogen in nitrogen at a rate of 1 to 200 L/hr, preferably 2 to 30 L/hr.