

**PROCESSES FOR REGENERATING CATALYSTS AND FOR UPGRADING
ALKANES AND/OR ALKYL AROMATIC HYDROCARBONS**

CROSS-REFERENCE TO RELATED APPLICATIONS

- 5 **[0001]** This application claims priority to and the benefit of U.S. Provisional Application No. 63/195,966 having a filing date of June 02, 2021, and U.S. Provisional Application No. 63/328,923 having a filing date of April 08, 2022, the disclosures of all of which are incorporated herein by reference in their entireties.

FIELD

- 10 **[0002]** This disclosure relates to processes for regenerating a catalyst and for upgrading alkanes and/or alkyl aromatic hydrocarbons.

BACKGROUND

- 15 **[0003]** Catalytic dehydrogenation, dehydroaromatization, and dehydrocyclization of alkane and/or alkyl aromatic hydrocarbons are industrially important chemical conversion processes that are endothermic and equilibrium-limited. The dehydrogenation of alkanes, *e.g.*, C₂-C₁₂ alkanes, and/or alkyl aromatics, *e.g.*, ethylbenzene, can be done through a variety of different supported catalyst systems such as the Pt-based, Cr-based, Ga-based, V-based, Zr-based, In-based, W-based, Mo-based, Zn-based, and Fe-based systems. Among the existing propane dehydrogenation processes, certain process uses an alumina supported chromia catalyst that
- 20 provides one of the highest propylene yields of approximately 50% (55% propane conversion at 90% propylene selectivity), which is obtained at a temperature of approximately 560°C to 650°C and at a low pressure of 20 kPa-absolute to 50 kPa-absolute. It is desirable to increase the propylene yield without having to operate at such low pressure to increase the efficiency of the dehydrogenation process.
- 25 **[0004]** Increasing the temperature of the dehydrogenation process is one way to increase the conversion of the process according to the thermodynamics of the process. For example, at 670°C, 100 kPa-absolute, in the absence of any inert/diluent, the equilibrium yield propylene yield has been estimated via simulation to be approximately 74%. At such high temperature, however, the catalyst deactivates very rapidly and/or the propylene selectivity becomes
- 30 uneconomically low. The rapid catalyst deactivation is believed to be caused by coke depositing onto the catalyst and/or agglomeration of the active phase. Coke can be removed by combustion using an oxygen-containing gas, however, agglomeration of the active phase is believed to be exacerbated during the combustion process, which rapidly reduces the activity and stability of the catalyst.

[0005] There is a need, therefore, for improved processes for regenerating at least partially deactivated catalysts and processes for dehydrogenating, dehydroaromatizing, and/or dehydrocyclizing alkane and/or alkyl aromatic hydrocarbons. This disclosure satisfies this and other needs.

5 SUMMARY

[0006] Processes for regenerating an at least partially deactivated catalyst and processes for upgrading a hydrocarbon are provided. In some embodiments, the process can be used to regenerate an at least partially deactivated catalyst that can include a Group 10 element, an inorganic support, and a contaminant. The Group 10 element can have a concentration in the
10 range of from 0.001 wt% to 6 wt%, based on the weight of the inorganic support. The process can include (I) heating the at least partially deactivated catalyst using a heating gas mixture that can include H₂O at a concentration of greater than 5 mol%, based on the total moles in the heating gas mixture to produce a precursor catalyst. The process can also include (II) providing an oxidative gas that can include no greater than 5 mol% of H₂O, based on the total moles in
15 the oxidative gas. The process can also include (III) contacting the precursor catalyst at an oxidizing temperature in a range of from 620°C to 1,000°C with the oxidative gas for a duration of at least 30 seconds, preferably at least 1 minute, preferably at least 5 minutes, to produce an oxidized precursor catalyst. The process can also include (IV) obtaining a regenerated catalyst from the oxidized precursor catalyst.

[0007] In other embodiments, a process for upgrading a hydrocarbon can include (I)
20 contacting a hydrocarbon-containing feed with a catalyst that can include a Group 10 element and an inorganic support to effect one or more of dehydrogenation, dehydroaromatization, and dehydrocyclization of at least a portion of the hydrocarbon-containing feed to produce an at least partially deactivated catalyst that can include the Group 10 element, the inorganic support,
25 and a contaminant and an effluent that can include one or more upgraded hydrocarbons and molecular hydrogen. The hydrocarbon-containing feed can include one or more of C₂-C₁₆ linear or branched alkanes, or one or more of C₄-C₁₆ cyclic alkanes, one or more C₈-C₁₆ alkyl aromatics, or a mixture thereof. The Group 10 element can have a concentration in the range of from 0.001 wt% to 6 wt%, based on the weight of the inorganic support. The hydrocarbon-
30 containing feed and the catalyst can be contacted at a temperature in a range from 300°C to 900°C. The one or more upgraded hydrocarbons can include at least one of a dehydrogenated hydrocarbon, a dehydroaromatized hydrocarbon, and a dehydrocyclized hydrocarbon. The process can also include (II) heating the at least partially deactivated catalyst using a heating gas mixture that can include H₂O at a concentration of greater than 5 mol%, based on the total

moles in the heating gas mixture to produce a precursor catalyst. The process can also include (III) providing an oxidative gas that can include no greater than 5 mol% of H₂O, based on the total moles in the oxidative gas. The process can also include (IV) contacting the precursor catalyst at an oxidizing temperature in a range of from 620°C to 1,000°C with the oxidative gas for a duration of at least 30 seconds to produce an oxidized precursor catalyst. The process can also include (V) obtaining a regenerated catalyst from the oxidized precursor catalyst. The process can also include (VI) contacting an additional quantity of the hydrocarbon-containing feed with at least a portion of the regenerated catalyst to produce additional at least partially deactivated catalyst and additional effluent.

BRIEF DESCRIPTION OF THE DRAWINGS

[0008] FIG. 1 shows the catalyst used in Example 6 can be effectively regenerated using a certain two-step regeneration scheme for 80+ cycles.

[0009] FIG. 2 shows that isobutane dehydrogenation with a catalyst was stable over 30+ cycles despite the catalyst being subjected to a regeneration temperature of 800°C.

15 [0010] FIG. 3 shows that the performance of a catalyst used in Example 19 was stable for over 30+ cycles despite the catalyst being subjected to a regeneration temperature of 800°C.

[0011] FIG. 4 shows that the performance of a first catalyst used in Example 20 was stable over 20+ cycles despite the catalyst being subjected to a regeneration temperature of 800°C.

20 [0012] FIG. 5 shows that the performance of a second catalyst used in Example 20 was stable over 30+ cycles despite the catalyst being subjected to a regeneration temperature of 800°C.

[0013] FIG. 6 shows that the performance of a catalyst used in Example 26 was stable over 20+ cycles despite the catalyst being subjected to a regeneration temperature of 800°C.

25 [0014] FIG. 7 shows that the performance of the comparative catalyst 1 continued to deactivate even though the regeneration temperature (620°C) was much lower than the other examples.

[0015] FIG. 8 shows a catalyst composition (catalyst 33) maintained its performance for 204 cycles.

DETAILED DESCRIPTION

30 [0016] Various specific embodiments, versions and examples of the invention will now be described, including preferred embodiments and definitions that are adopted herein for purposes of understanding the claimed invention. While the following detailed description gives specific preferred embodiments, those skilled in the art will appreciate that these embodiments are exemplary only, and that the invention may be practiced in other ways. For purposes of determining infringement, the scope of the invention will refer to any one or more

of the appended claims, including their equivalents, and elements or limitations that are equivalent to those that are recited. Any reference to the “invention” may refer to one or more, but not necessarily all, of the inventions defined by the claims.

[0017] In this disclosure, a process is described as comprising at least one “step.” It should be understood that each step is an action or operation that may be carried out once or multiple times in the process, in a continuous or discontinuous fashion. Unless specified to the contrary or the context clearly indicates otherwise, multiple steps in a process may be conducted sequentially in the order as they are listed, with or without overlapping with one or more other steps, or in any other order, as the case may be. In addition, one or more or even all steps may be conducted simultaneously with regard to the same or different batch of material. For example, in a continuous process, while a first step in a process is being conducted with respect to a raw material just fed into the beginning of the process, a second step may be carried out simultaneously with respect to an intermediate material resulting from treating the raw materials fed into the process at an earlier time in the first step. Preferably, the steps are conducted in the order described.

[0018] Unless otherwise indicated, all numbers indicating quantities in this disclosure are to be understood as being modified by the term “about” in all instances. It should also be understood that the precise numerical values used in the specification and claims constitute specific embodiments. Efforts have been made to ensure the accuracy of the data in the examples. However, it should be understood that any measured data inherently contains a certain level of error due to the limitation of the technique and/or equipment used for acquiring the measurement.

[0019] Certain embodiments and features are described herein using a set of numerical upper limits and a set of numerical lower limits. It should be appreciated that ranges including the combination of any two values, e.g., the combination of any lower value with any upper value, the combination of any two lower values, and/or the combination of any two upper values are contemplated unless otherwise indicated.

[0020] The indefinite article “a” or “an”, as used herein, means “at least one” unless specified to the contrary or the context clearly indicates otherwise. Thus, embodiments using “a reactor” or “a conversion zone” include embodiments where one, two or more reactors or conversion zones are used, unless specified to the contrary or the context clearly indicates that only one reactor or conversion zone is used.

[0021] The term “hydrocarbon” means (i) any compound consisting of hydrogen and carbon atoms or (ii) any mixture of two or more such compounds in (i). The term “C_n hydrocarbon,”

where n is a positive integer, means (i) any hydrocarbon compound comprising carbon atom(s) in its molecule at the total number of n, or (ii) any mixture of two or more such hydrocarbon compounds in (i). Thus, a C2 hydrocarbon can be ethane, ethylene, acetylene, or mixtures of at least two of these compounds at any proportion. A “Cm to Cn hydrocarbon” or “Cm-Cn hydrocarbon,” where m and n are positive integers and $m < n$, means any of Cm, Cm+1, Cm+2, ..., Cn-1, Cn hydrocarbons, or any mixtures of two or more thereof. Thus, a “C2 to C3 hydrocarbon” or “C2-C3 hydrocarbon” can be any of ethane, ethylene, acetylene, propane, propene, propyne, propadiene, cyclopropane, and any mixtures of two or more thereof at any proportion between and among the components. A “saturated C2-C3 hydrocarbon” can be ethane, propane, cyclopropane, or any mixture thereof of two or more thereof at any proportion. A “Cn+ hydrocarbon” means (i) any hydrocarbon compound comprising carbon atom(s) in its molecule at the total number of at least n, or (ii) any mixture of two or more such hydrocarbon compounds in (i). A “Cn- hydrocarbon” means (i) any hydrocarbon compound comprising carbon atoms in its molecule at the total number of at most n, or (ii) any mixture of two or more such hydrocarbon compounds in (i). A “Cm hydrocarbon stream” means a hydrocarbon stream consisting essentially of Cm hydrocarbon(s). A “Cm-Cn hydrocarbon stream” means a hydrocarbon stream consisting essentially of Cm-Cn hydrocarbon(s).

[0022] For the purposes of this disclosure, the nomenclature of elements is pursuant to the version of the Periodic Table of Elements (under the new notation) as provided in Hawley's Condensed Chemical Dictionary, 16th Ed., John Wiley & Sons, Inc., (2016), Appendix V. For example, a Group 2 element includes Mg, a Group 8 element includes Fe, a Group 9 element includes Co, a Group 10 element includes Ni, and a Group 13 element includes Al. The term “metalloid”, as used herein, refers to the following elements: B, Si, Ge, As, Sb, Te, and At. In this disclosure, when a given element is indicated as present, it can be present in the elemental state or as any chemical compound thereof, unless it is specified otherwise or clearly indicated otherwise by the context.

[0023] The term “alkane” means a saturated hydrocarbon. The term “cyclic alkane” means a saturated hydrocarbon comprising a cyclic carbon ring in the molecular structure thereof. An alkane can be linear, branched, or cyclic.

[0024] The term “aromatic” is to be understood in accordance with its art-recognized scope, which includes alkyl substituted and unsubstituted mono- and polynuclear compounds.

[0025] The term “rich” when used in phrases such as “X-rich” or “rich in X” means, with respect to an outgoing stream obtained from a device, e.g., a conversion zone, that the stream comprises material X at a concentration higher than in the feed material fed to the same device

from which the stream is derived. The term “lean” when used in phrases such as “X-lean” or “lean in X” means, with respect to an outgoing stream obtained from a device, e.g., a conversion zone, that the stream comprises material X at a concentration lower than in the feed material fed to the same device from which the stream is derived.

- 5 **[0026]** The term “mixed metal oxide” refers to a composition that includes oxygen atoms and at least two different metal atoms that are mixed on an atomic scale. For example, a “mixed Mg/Al metal oxide” has O, Mg, and Al atoms mixed on an atomic scale and is substantially the same as or identical to a composition obtained by calcining an Mg/Al hydrotalcite that has the general chemical formula $[Mg_{(1-x)}Al_x(OH)_2](A_x^{n-}) \cdot mH_2O$, where A is a counter anion of
- 10 a negative charge n, x is in a range of from > 0 to < 1 , and m is ≥ 0 . A material consisting of nm sized MgO particles and nm sized Al_2O_3 particles mixed together is not a mixed metal oxide because the Mg and Al atoms are not mixed on an atomic scale but are instead mixed on a nm scale.

- [0027]** The term “selectivity” refers to the production (on a carbon mole basis) of a specified
- 15 compound in a catalytic reaction. As an example, the phrase “an alkane hydrocarbon conversion reaction has a 100% selectivity for an olefin hydrocarbon” means that 100% of the alkane hydrocarbon (carbon mole basis) that is converted in the reaction is converted to the olefin hydrocarbon. When used in connection with a specified reactant, the term “conversion” means the amount of the reactant consumed in the reaction. For example, when the specified
- 20 reactant is propane, 100% conversion means 100% of the propane is consumed in the reaction. In another example, when the specified reactant is propane, if one mole of propane converts to one mole of methane and one mole of ethylene, the selectivity to methane is 33.3% and the selectivity to ethylene is 66.7%. Yield (carbon mole basis) is conversion times selectivity.

Hydrocarbon Upgrading and Catalyst Regeneration Process

- 25 **[0028]** The hydrocarbon-containing feed can be or can include, but is not limited to, one or more alkane hydrocarbons, e.g., C_2 - C_{16} linear or branched alkanes and/or C_4 - C_{16} cyclic alkanes, and/or one or more alkyl aromatic hydrocarbons, e.g., C_8 - C_{16} alkyl aromatics. In some embodiments, the hydrocarbon-containing feed can optionally include 0.1 vol% to 50 vol% of steam, based on a total volume of any C_2 - C_{16} alkanes and any C_8 - C_{16} alkyl aromatics in the
- 30 hydrocarbon-containing feed. In other embodiments, the hydrocarbon-containing feed can include < 0.1 vol% of steam or can be free of steam, based on the total volume of any C_2 - C_{16} alkanes and any C_8 - C_{16} alkyl aromatics in the hydrocarbon-containing feed. The hydrocarbon-containing feed can be contacted with a catalyst that includes a Group 10 element, e.g., Pt, and an inorganic support, to effect one or more of dehydrogenation, dehydroaromatization, and

dehydrocyclization of at least a portion of the hydrocarbon-containing feed to produce an at least partially deactivated catalyst that includes the Group 10 element, the inorganic support, and a contaminant, e.g., coke, and an effluent that can include one or more upgraded hydrocarbons and molecular hydrogen.

- 5 **[0029]** The one or more upgraded hydrocarbons can be or can include, but are not limited to, one or more dehydrogenated hydrocarbons, one or more dehydroaromatized hydrocarbons, one or more dehydrocyclized hydrocarbons, or a mixture thereof. The hydrocarbon-containing feed and catalyst can be contacted at a temperature in a range from 300°C to 900°C. In some embodiments, the hydrocarbon-containing feed and catalyst can be contacted for a time period
10 of ≤ 5 hours, ≤ 4 hours, or ≤ 3 hours, ≤ 1 hour, ≤ 0.5 hours, ≤ 0.1 hours, ≤ 3 minutes, ≤ 1 minute, ≤ 30 seconds, or ≤ 0.1 second. In some embodiments, the hydrocarbon-containing feed and catalyst can be contacted under a hydrocarbon partial pressure of at least 20 kPa-absolute, where the hydrocarbon partial pressure is the total partial pressure of any C₂-C₁₆ alkanes and any C₈-C₁₆ alkyl aromatics in the hydrocarbon-containing feed. The catalyst can include from
15 0.001 wt% to 6 wt% of the Group 10 element, e.g., Pt, based on the weight of the inorganic support.

- [0030]** A precursor catalyst can be obtained from the at least partially deactivated catalyst. In some embodiments, the at least partially deactivated catalyst can be provided directly as the precursor catalyst. In other embodiments, the precursor catalyst can be obtained by heating the
20 at least partially deactivated catalyst using a heating gas mixture that includes H₂O at a concentration of greater than 5 mol%, based on the total moles in the heating gas mixture to produce the precursor catalyst. In some embodiments, the heating gas mixture can be produced by combusting at least a portion of the contaminant, e.g., coke and/or residual hydrocarbon-containing feed, disposed on the at least partially deactivated catalyst with an oxidizing gas. In
25 some embodiments, the heating gas mixture can be produced by combusting a fuel with the oxidizing gas. In other embodiments, the heating gas mixture can be produced by combusting at least a portion of the contaminant disposed on the at least partially deactivated catalyst and the fuel with the oxidizing gas. In other embodiments, the heating gas mixture that can include H₂O at a concentration of greater than 5 mol% of H₂O can be provided with the H₂O, e.g.,
30 heated air having greater than 5 mol% of H₂O. The fuel can be or can include, but is not limited to at least one of H₂, CO, and a hydrocarbon. The oxidizing gas can be or can include, but is not limited to, O₂, O₃, CO, or any mixture thereof. In some embodiments, the heating gas mixture can contact the partially deactivated catalyst for a duration < 5 min, < 2 min, < 1 min, < 30 s, < 10 s, < 5 s, < 1 s, < 0.5 s, < 0.1 s.

[0031] An oxidative gas can be provided. The oxidative gas can include no greater than 5 mol% of H₂O, no greater than 4.5 mol% of H₂O, no greater than 4 mol% of H₂O, no greater than 3.5 mol% of H₂O, no greater than 3 mol% of H₂O, no greater than 2.5 mol% of H₂O, no greater than 2 mol% of H₂O, no greater than 1.7 mol% of H₂O, no greater than 1.5 mol% of H₂O, no greater than 1.3 mol% of H₂O, no greater than 1 mol% of H₂O, no greater than 0.7 mol% of H₂O, no greater than 0.5 mol% of H₂O, no greater than 0.3 mol% of H₂O, or no greater than 0.1 mol% of H₂O, based on the total moles in the oxidative gas. The precursor catalyst, whether the at least partially deactivated catalyst is provided directly as the precursor catalyst or the at least partially deactivated catalyst is heated using the heating gas mixture, can be contacted with the oxidative gas.

[0032] It has been surprisingly and unexpectedly discovered that contacting the precursor catalyst, whether the at least partially deactivated catalyst is provided directly as the precursor catalyst or the at least partially deactivated catalyst is heated using the heating gas mixture to produce the precursor catalyst, with the oxidative gas that includes no greater than 5 mol% of H₂O can significantly improve the activity and/or selectivity of the regenerated catalyst. Without wishing to be bound by theory, it is believed that an H₂O present in the oxidative gas may significantly reduce the effectiveness of Pt re-dispersion and hence the effectiveness of the regenerated catalyst.

[0033] The precursor catalyst can be contacted with the oxidative gas at an oxidizing temperature in a range of from 620°C, 650°C, 675°C, 700°C, or 750°C to 775°C, 800°C, 850°C, 900°C, 950°C, or 1,000°C to produce an oxidized precursor catalyst. The precursor catalyst can be contacted with the oxidative gas for a duration of at least 30 seconds, at least 1 minute, at least 5 minutes, at least 7 minutes, at least 10 minutes, at least 15 minutes, at least 20 minutes, at least 25 minutes, or at least 30 minutes to produce the oxidized precursor catalyst. In some embodiments, the precursor catalyst can be contacted with the oxidative gas for a duration in a range of from 30 seconds, 1 minute, 5 minutes, or 10 minutes to 30 minutes, 60 minutes, or 120 minutes to produce the oxidized precursor catalyst. In some embodiments, the precursor catalyst and the oxidative gas can be contacted with one another for a duration of ≤ 2 hours, ≤ 1 hour, ≤ 30 minutes, ≤ 10 minutes, ≤ 5 minutes, ≤ 1 min, ≤ 30 seconds, ≤ 10 seconds, ≤ 5 seconds, or ≤ 1 second to produce the oxidized precursor catalyst. For example, the precursor catalyst and oxidative gas can be contacted with one another for a duration in a range from 2 seconds to 2 hours to produce the oxidized precursor catalyst. In some embodiments, the precursor catalyst and oxidative gas can be contacted for a duration sufficient to remove ≥ 50

wt%, ≥ 75 wt%, or ≥ 90 wt% or > 99 wt% of the contaminant, e.g., coke disposed on the precursor catalyst.

[0034] The precursor catalyst and oxidative gas can be contacted with one another under an oxidative gas partial pressure in a range from 5 kPa-absolute, 10 kPa-absolute, 20 kPa-absolute, 50 kPa-absolute, 100 kPa-absolute, 300 kPa-absolute, 500 kPa-absolute, 750 kPa-absolute, or 1,000 kPa-absolute to 1,500 kPa-absolute, 2,500 kPa-absolute, 4,000 kPa-absolute, 5,000 kPa-absolute, 7,000 kPa-absolute, 8,500 kPa-absolute, or 10,000 kPa-absolute to produce the oxidized precursor catalyst. In some embodiments, the oxidative gas partial pressure during contact with the precursor catalyst can be in a range from 5 kPa-absolute, 10 kPa-absolute, 20 kPa-absolute, 50 kPa-absolute, 100 kPa-absolute, 150 kPa-absolute, 200 kPa-absolute, 250 kPa-absolute, or 300 kPa-absolute to 500 kPa-absolute, 600 kPa-absolute, 700 kPa-absolute, 800 kPa-absolute, 900 kPa-absolute, or 1,000 kPa-absolute to produce the oxidized precursor catalyst.

[0035] Without wishing to be bound by theory, it is believed that at least a portion of the Group 10 element, e.g., Pt, disposed on the precursor catalyst can be agglomerated as compared to the catalyst prior to contact with the hydrocarbon-containing feed. It is believed that during contact of the precursor catalyst with the oxidative gas, when at least a portion of the contaminant on the precursor catalyst can be combusted, at least a portion of the Group 10 element can be re-dispersed about the inorganic support. Re-dispersing at least a portion of the agglomerated Group 10 element can increase the activity and improve the stability of the catalyst over many cycles.

[0036] In some embodiments, the oxidative gas can be provided at a temperature below the oxidizing temperature and the oxidative gas can be pre-heated to a temperature higher than the temperature of the precursor catalyst before contacting the precursor catalyst with the oxidative gas at the oxidizing temperature. In some embodiments, the oxidative gas can be pre-heated by using a radiant/conductive heat source, a heat exchanger, or a combination thereof. In other embodiments, the oxidative gas, the precursor catalyst, or both the oxidative gas and the precursor catalyst can be heated by using a radiant/conductive heat source, a heat exchanger, or a combination thereof. In other words, the precursor catalyst and/or the oxidative gas can be heated separately and then contacted with one another at the oxidizing temperature or heated in the presence of one another to the oxidizing temperature. In some embodiments, the radiant/conductive heat source can be or can include one or more electric heating elements.

[0037] A regenerated catalyst can be obtained from the oxidized precursor catalyst. In some embodiments, the oxidized precursor catalyst can be provided directly as the regenerated

catalyst. In some embodiments, the oxidized precursor catalyst can optionally be contacted with a first stripping gas that can be free of O₂ to produce a stripped oxidized precursor catalyst and the regenerated catalyst can be obtained from the stripped oxidized precursor catalyst. The first stripping gas can be or can include, but is not limited to, CO, CO₂, N₂, a C₁-C₄ hydrocarbon, H₂O, He, Ne, Ar, or any mixture thereof. In some embodiments, the stripped oxidized precursor catalyst can be provided directly as the regenerated catalyst.

[0038] In some embodiments, at least a portion of the Group 10 element, e.g., Pt, in the oxidized precursor catalyst can be at a higher oxidized state as compared to the Group 10 element in the catalyst contacted with the hydrocarbon-containing feed and as compared to the Group 10 element in the at least partially deactivated catalyst. In some embodiments, the oxidized precursor catalyst or the stripped oxidized precursor catalyst can be contacted with a H₂-containing atmosphere to produce a reduced catalyst. In other embodiments, the oxidized precursor catalyst or the stripped oxidized precursor catalyst can be contacted with an atmosphere containing H₂, CO, CH₄, C₂H₆, C₃H₈, C₂H₄, C₃H₆, steam, or a mixture thereof to produce the reduced catalyst. In some embodiments, the atmosphere contacted with the oxidized precursor catalyst can also include an inert gas such as Ar, Ne, He, N₂, CO₂, H₂O or a mixture thereof. In such embodiments, at least a portion of the Group 10 element in the reduced catalyst can be reduced to a lower oxidation state, e.g., the elemental state, as compared to the Group 10 element in the oxidized precursor catalyst.

[0039] In some embodiments, the oxidized precursor catalyst or the stripped oxidized precursor catalyst can be contacted with the H₂-containing atmosphere or the atmosphere containing H₂, CO, CH₄, C₂H₆, C₃H₈, C₂H₄, C₃H₆, steam, or a mixture thereof at a temperature in a range from 400°C, 450°C, 500°C, 550°C, 600°C, 620°C, 650°C, or 670°C to 720°C, 750°C, 800°C, or 900°C. The oxidized precursor catalyst or the stripped oxidized precursor catalyst and the H₂-containing atmosphere or the atmosphere containing H₂, CO, CH₄, C₂H₆, C₃H₈, C₂H₄, C₃H₆, steam, or a mixture thereof can be contacted for a duration in a range from 0.01 seconds, 0.1 seconds, 1 second, 5 seconds, 10 seconds, 20 seconds, 30 seconds, or 1 minute to 10 minutes, 30 minutes, or 60 minutes. The oxidized precursor catalyst or the stripped oxidized precursor catalyst and the H₂-containing atmosphere or the atmosphere containing H₂, CO, CH₄, C₂H₆, C₃H₈, C₂H₄, C₃H₆, steam, or a mixture thereof can be contacted at a reducing agent partial pressure of 0.1 kPa-absolute, 1 kPa-absolute, 5 kPa-absolute, 10 kPa-absolute, 20 kPa-absolute, 50 kPa-absolute, or 100 kPa-absolute, 300 kPa-absolute, 500 kPa-absolute, 750 kPa-absolute, or 1,000 kPa-absolute to 1,500 kPa-absolute, 2,500 kPa-absolute, 4,000 kPa-absolute, 5,000 kPa-absolute, 7,000 kPa-absolute, 8,500 kPa-absolute, or 10,000 kPa-absolute, where the

reducing agent includes any H₂, CO, CH₄, C₂H₆, C₃H₈, C₂H₄, C₃H₆, and steam. In other embodiments, the reducing agent partial pressure can be in a range from 0.1 kPa-absolute, 1 kPa-absolute, 5 kPa-absolute, 10 kPa-absolute, 20 kPa-absolute, 50 kPa-absolute, 100 kPa-absolute, 150 kPa-absolute, 200 kPa-absolute, 250 kPa-absolute, or 300 kPa-absolute to 500 kPa-absolute, 600 kPa-absolute, 700 kPa-absolute, 800 kPa-absolute, 900 kPa-absolute, or 1,000 kPa-absolute to produce the regenerated catalyst, where the reducing agent includes any H₂, CO, CH₄, C₂H₆, C₃H₈, C₂H₄, C₃H₆, and steam.

[0040] In some embodiments, the oxidized precursor catalyst or the stripped oxidized precursor catalyst can be contacted with the H₂-containing atmosphere at a temperature higher than a use temperature of the regenerated catalyst. In such embodiment, the reduced catalyst can be cooled to the use temperature. In some embodiments, the reduced catalyst can be cooled to the use temperature in a duration no greater than 20 minutes, no greater than 15 minutes, no greater than 10 minutes, no greater than 7 minutes, no greater than 5 minutes, no greater than 2 minutes, no greater than 1 minute, no greater than 30 seconds, no greater than 10 seconds, no greater than 5 seconds, no greater than 2 seconds, no greater than 1 second, no greater than 0.1 seconds, no greater than 0.01 seconds, or no greater than 0.001 seconds. The use temperature of the catalyst is the temperature at which the hydrocarbon-containing feed or the additional quantity of the hydrocarbon-containing feed is contacted with the catalyst or the regenerated catalyst to effect one or more of dehydrogenation, dehydroaromatization, and dehydrocyclization of at least a portion of the hydrocarbon-containing feed to produce the at least partially deactivated catalyst that can include the Group 10 element, the inorganic support, and the contaminant and the effluent that can include the one or more upgraded hydrocarbons and molecular hydrogen.

[0041] The regenerated catalyst can be obtained from the reduced catalyst. In some embodiments, the reduced catalyst can be provided directly as the regenerated catalyst. In other embodiments, the reduced catalyst can be contacted with a second stripping gas to produce the regenerated catalyst. The second stripping gas can be or can include, but is not limited to, CO, CO₂, N₂, a C₁-C₄ hydrocarbon, H₂O, He, Ne, Ar, or any mixture thereof.

[0042] At least a portion of the regenerated catalyst, new or fresh catalyst, or a mixture thereof can be contacted with the additional quantity of the hydrocarbon-containing feed within the reaction or conversion zone to produce additional effluent and additional at least partially deactivated catalyst. The cycle time from the contacting the hydrocarbon-containing feed with the catalyst to the contacting the additional quantity of the hydrocarbon-containing feed with at least a portion of the regenerated catalyst and optionally with new or fresh catalyst can be ≤

5 hours, ≤ 4.5 hours, ≤ 4 hours, ≤ 3.5 hours, ≤ 3 hours, ≤ 2.5 hours, ≤ 2 hours, ≤ 1 hour, ≤ 0.5 hours, ≤ 0.2 hours, ≤ 0.1 hours, ≤ 0.05 hours, or ≤ 0.01 hours.

[0043] The first cycle begins upon contact of the catalyst with the hydrocarbon-containing feed, followed by contact with at least the oxidative gas to produce the oxidized precursor catalyst, which can be provided directly as the regenerated catalyst, or at least the oxidative gas and the optional reducing gas to produce the regenerated catalyst, and the first cycle ends upon contact of the regenerated catalyst with the additional quantity of the hydrocarbon-containing feed. If the first stripping gas and/or the second stripping gas or any other stripping gas(es) are utilized between flows of the hydrocarbon-containing feed and the oxidative gas, between the oxidative gas and the reducing gas (if used), between the oxidative gas and the additional quantity of the hydrocarbon-containing feed, and/or between the reducing gas (if used) and the additional quantity of the hydrocarbon-containing feed, the period of time such stripping gas(es) is/are utilized would be included in the period included in the cycle time. As such, the cycle time from the contacting the hydrocarbon-containing feed with the catalyst in step to the contacting the additional quantity of the hydrocarbon-containing feed with the regenerated catalyst can be ≤ 5 hours.

[0044] The catalyst that includes a Group 10 element, e.g., Pt, and the inorganic support can remain sufficiently active and stable after many cycles, e.g., at least 15, at least 20, at least 30, at least 40, at least 50, at least 60, at least 70, at least 100 cycles, at least 125 cycles, at least 150 cycles, at least 175 cycles, or at least 200 cycles with each cycle time lasting for ≤ 5 hours, ≤ 4 hours, ≤ 3 hours, ≤ 2 hours, ≤ 1 hour, ≤ 50 minutes, ≤ 45 minutes, ≤ 30 minutes, ≤ 15 minutes, ≤ 10 minutes, ≤ 5 minutes, ≤ 1 minute, ≤ 30 seconds, or ≤ 10 seconds. In some embodiments, the cycle time can be from 5 seconds, 30 seconds, 1 minute or 5 minutes to 10 minutes, 20 minutes, 30 minutes, 45 minutes, 50 minutes, 70 minutes, 2 hours, 3 hours, 4 hours, or 5 hours. In some embodiments, after the catalyst performance stabilizes (sometimes the first few cycles can have a relatively poor or a relatively good performance, but the performance can eventually stabilize), the process can produce a first upgraded hydrocarbon product yield, e.g., propylene when the hydrocarbon-containing feed includes propane, at an upgraded hydrocarbon selectivity, e.g., propylene, of $\geq 75\%$, $\geq 80\%$, $\geq 85\%$, or $\geq 90\%$, or $> 95\%$ when initially contacted with the hydrocarbon-containing feed, and can have a second upgraded hydrocarbon product yield upon completion of the last cycle (at least 15 cycles total) that can be at least 90%, at least 93%, at least 95%, at least 97%, at least 98%, at least 99%, at least 99.5%, or at least 100% of the first upgraded hydrocarbon product yield at an upgraded hydrocarbon selectivity, e.g., propylene, of $\geq 75\%$, $\geq 80\%$, $\geq 85\%$, or $\geq 90\%$, or $> 95\%$.

[0045] In some embodiments, when the hydrocarbon-containing feed includes propane and the upgraded hydrocarbon includes propylene, contacting the hydrocarbon-containing feed with the catalyst composition can produce a propylene yield of $\geq 48\%$, $\geq 49\%$, $\geq 50\%$, $\geq 51\%$, $\geq 52\%$, $\geq 53\%$, $\geq 54\%$, $\geq 55\%$, $\geq 56\%$, $\geq 57\%$, $\geq 58\%$, $\geq 59\%$, $\geq 60\%$, $\geq 61\%$, $\geq 62\%$, $\geq 63\%$, $\geq 64\%$, $\geq 65\%$, or $\geq 66\%$ at a propylene selectivity of $\geq 75\%$, $\geq 80\%$, $\geq 85\%$, $\geq 90\%$, $\geq 93\%$, or $\geq 95\%$. In some embodiments, when the hydrocarbon-containing feed includes propane and the upgraded hydrocarbon includes propylene, contacting the hydrocarbon-containing feed with the catalyst can produce a propylene yield of at least 48%, at least 49%, at least 50%, at least 51%, at least 52%, at least 53%, at least 55%, at least 57%, at least 60%, at least 62%, at least 63%, at least 64%, at least 65%, at or at least 66% at a propylene selectivity of at least 75%, at least 80%, at least 85%, at least 90%, or at least 95% for at least 15, at least 20, at least 30, at least 40, at least 50, at least 60, at least 70, at least 100 cycles, at least 125 cycles, at least 150 cycles, at least 175 cycles, or at least 200 cycles. In other embodiments, when the hydrocarbon-containing feed includes at least 70 vol%, at least 75 vol%, at least 80 vol%, at least 85 vol%, at least 90 vol%, or at least 95 vol% of propane, based on a total volume of the hydrocarbon-containing feed, is contacted under a propane partial pressure of at least 20 kPa-absolute, a propylene yield of at least 48%, at least 49%, at least 50%, at least 51%, at least 52%, at least 53%, at least 55%, at least 57%, at least 60%, at least 62%, at least 63%, at least 64%, at least 65%, or at least 66% at a propylene selectivity of at least 75%, at least 80%, at least 85%, at least 90%, or at least 95% can be obtained for at least 15, at least 20, at least 30, at least 40, at least 50, at least 60, at least 70, at least 100 cycles, at least 125 cycles, at least 150 cycles, at least 175 cycles, or at least 200 cycles. It is believed that the propylene yield can be further increased to at least 67%, at least 68%, at least 70%, at least 72%, at least 75%, at least 77%, at least 80%, or at least 82% at a propylene selectivity of at least 75%, at least 80%, at least 85%, at least 90%, or at least 95% for at least 15 cycles, at least 20, at least 30, at least 40, at least 50, at least 60, at least 70, at least 100 cycles, at least 125 cycles, at least 150 cycles, at least 175 cycles, or at least 200 cycles by further optimizing the composition of the support and/or adjusting one or more process conditions. In some embodiments, the propylene yield can be obtained when the catalyst is contacted with the hydrocarbon feed at a temperature of at least 620°C, at least 630°C, at least 640°C, at least 650°C, at least 655°C, at least 660°C, at least 670°C, at least 680°C, at least 690°C, at least 700°C, or at least 750°C for at least 15, at least 20, at least 30, at least 40, at least 50, at least 60, at least 70, at least 100 cycles, at least 125 cycles, at least 150 cycles, at least 175 cycles, or at least 200 cycles.

[0046] Systems suitable for carrying out the processes disclosed herein can include systems that are well-known in the art such as the fixed bed reactors disclosed in WO Publication No. WO2017078894; the fluidized riser reactors and/or downer reactors disclosed in U.S. Patent Nos. 3,888,762; 7,102,050; 7,195,741; 7,122,160; and 8,653,317; and U.S. Patent Application
 5 Publication Nos. 2004/0082824; 2008/0194891; and the reverse flow reactors disclosed in U.S. Patent No. 8,754,276; U.S. Patent Application Publication No. 2015/0065767; and WO Publication No. WO2013169461.

Catalyst

[0047] The catalyst can include 0.001 wt%, 0.002 wt%, 0.003 wt%, 0.004 wt%, 0.005 wt%,
 10 0.006 wt%, 0.007 wt%, 0.008 wt%, 0.009 wt%, 0.01 wt%, 0.015 wt%, 0.02 wt%, 0.025 wt%, 0.03 wt%, 0.035 wt%, 0.04 wt%, 0.045 wt%, 0.05 wt%, 0.055 wt%, 0.06 wt%, 0.065 wt%, 0.07 wt%, 0.075 wt%, 0.08 wt%, 0.085 wt%, 0.09 wt%, 0.095 wt%, 0.1 wt%, 0.2 wt%, 0.3 wt%, 0.4 wt%, 0.5 wt%, 0.6 wt%, 0.7 wt%, 0.8 wt%, 0.9 wt%, or 1 wt% to 2 wt%, 3 wt%, 4 wt%, 5 wt%, or 6 wt% of the Group 10 element disposed on the inorganic support, based on
 15 the weight of the inorganic support. In some embodiments, the catalyst composition can include ≤ 5.5 wt%, ≤ 4.5 wt%, ≤ 3.5 wt%, ≤ 2.5 wt%, ≤ 1.5 wt%, ≤ 1 wt%, ≤ 0.9 wt%, ≤ 0.8 wt%, ≤ 0.7 wt%, ≤ 0.6 wt%, ≤ 0.5 wt%, ≤ 0.4 wt%, ≤ 0.3 wt%, ≤ 0.2 wt%, ≤ 0.15 wt%, ≤ 0.1 wt%, ≤ 0.09 wt%, ≤ 0.08 wt%, ≤ 0.07 wt%, ≤ 0.06 wt%, ≤ 0.05 wt%, ≤ 0.04 wt%, ≤ 0.03 wt%, ≤ 0.02 wt%, ≤ 0.01 wt%, ≤ 0.009 wt%, ≤ 0.008 wt%, ≤ 0.007 wt%, ≤ 0.006 wt%, ≤ 0.005 wt%,
 20 ≤ 0.004 wt%, ≤ 0.003 wt%, or ≤ 0.002 wt% of the Group 10 element disposed on the inorganic support, based on the weight of the inorganic support. In some embodiments, the catalyst can include > 0.001 , > 0.003 wt%, > 0.005 wt%, > 0.007 , > 0.009 wt%, > 0.01 wt%, > 0.02 wt%, > 0.04 wt%, > 0.06 wt%, > 0.08 wt%, > 0.1 wt%, > 0.13 wt%, > 0.15 wt%, > 0.17 wt%, > 0.2 wt%, > 0.2 wt%, > 0.23 , > 0.25 wt%, > 0.27 wt%, or > 0.3 wt% and < 0.5 wt%, < 1 wt%, < 2
 25 wt%, < 3 wt%, < 4 wt%, < 5 wt%, or < 6 wt% of the Group 10 element disposed on the inorganic support, based on the weight of the inorganic support. In some embodiments, the Group 10 element can be or can include Ni, Pd, Pt, a combination thereof, or a mixture thereof. In at least one embodiment, the Group 10 element can be or can include Pt. If two or more Group 10 elements are disposed on the inorganic support, the catalyst can include 0.001 wt%,
 30 0.002 wt%, 0.003 wt%, 0.004 wt%, 0.005 wt%, 0.006 wt%, 0.007 wt%, 0.008 wt%, 0.009 wt%, 0.01 wt%, 0.015 wt%, 0.02 wt%, 0.025 wt%, 0.03 wt%, 0.035 wt%, 0.04 wt%, 0.045 wt%, 0.05 wt%, 0.055 wt%, 0.06 wt%, 0.065 wt%, 0.07 wt%, 0.075 wt%, 0.08 wt%, 0.085 wt%, 0.09 wt%, 0.095 wt%, 0.1 wt%, 0.2 wt%, 0.3 wt%, 0.4 wt%, 0.5 wt%, 0.6 wt%, 0.7 wt%, 0.8 wt%, 0.9 wt%, or 1 wt% to 2 wt%, 3 wt%, 4 wt%, 5 wt%, or 6 wt% of a combined amount of

the two or more Group 10 elements disposed on the inorganic support, based on the weight of the inorganic support. In some embodiments, an active component of the regenerated catalyst that can be capable of effecting one or more of dehydrogenation, dehydroaromatization, and dehydrocyclization of the hydrocarbon-containing feed that includes one or more of C₂-C₁₆ linear or branched alkanes, or one or more of C₄-C₁₆ cyclic alkanes, one or more C₈-C₁₆ alkyl aromatics, or a mixture thereof can include the group 10 element.

[0048] The inorganic support can be or can include, but is not limited to, one or more Group 2 elements, a combination thereof, or a mixture thereof. In some embodiments, the Group 2 element can be present in its elemental form. In other embodiments, the Group 2 element can be present in the form of a compound. For example, the Group 2 element can be present as an oxide, a phosphate, a halide, a halate, a sulfate, a sulfide, a borate, a nitride, a carbide, an aluminate, an aluminosilicate, a silicate, a carbonate, metaphosphate, a selenide, a tungstate, a molybdate, a chromite, a chromate, a dichromate, or a silicide. In some embodiments, a mixture of any two or more compounds that include the Group 2 element can be present in different forms. For example, a first compound can be an oxide and a second compound can be an aluminate where the first compound and the second compound include the same or different Group 2 element, with respect to one another.

[0049] The inorganic support can include ≥ 0.5 wt%, ≥ 1 wt%, ≥ 2 wt%, ≥ 3 wt%, ≥ 4 wt%, ≥ 5 wt%, ≥ 6 wt%, ≥ 7 wt%, ≥ 8 wt%, ≥ 9 wt%, ≥ 10 wt%, ≥ 11 wt%, ≥ 12 wt%, ≥ 13 wt%, ≥ 14 wt%, ≥ 15 wt%, ≥ 16 wt%, ≥ 17 wt%, ≥ 18 wt%, ≥ 19 wt%, ≥ 20 wt%, ≥ 21 wt%, ≥ 22 wt%, ≥ 23 wt%, ≥ 24 wt%, ≥ 25 wt%, ≥ 26 wt%, ≥ 27 wt%, ≥ 28 wt%, ≥ 29 wt%, ≥ 30 wt%, ≥ 35 wt%, ≥ 40 wt%, ≥ 45 wt%, ≥ 50 wt%, ≥ 55 wt%, ≥ 60 wt%, ≥ 65 wt%, ≥ 70 wt%, ≥ 75 wt%, ≥ 80 wt%, ≥ 85 wt%, or ≥ 90 wt% of the Group 2 element, based on the weight of the inorganic support. In some embodiments, the inorganic support can include the Group 2 element in a range of from 0.5 wt%, 1 wt%, 2 wt%, 2.5 wt%, 3 wt%, 5 wt%, 7 wt%, 10 wt%, 11 wt%, 13 wt%, 15 wt%, 17 wt%, 19 wt%, 21 wt%, 23 wt%, or 25 wt% to 30 wt%, 35 wt%, 40 wt%, 45 wt%, 50 wt%, 55 wt%, 60 wt%, 65 wt%, 70 wt%, 75 wt%, 80 wt%, 85 wt%, 90 wt%, or 92.34 wt% based on the weight of the inorganic support. In some embodiments, a molar ratio of the Group 2 element to the Group 10 element can be in a range from 0.24, 0.5, 1, 10, 50, 100, 300, 450, 600, 800, 1,000, 1,200, 1,500, 1,700, or 2,000 to 3,000, 3,500, 4,000, 4,500, 5,000, 5,500, 6,000, 6,500, 7,000, 7,500, 8,000, 8,500, 9,000, 9,500, 10,000, 15,000, 20,000, 25,000, 30,000, 35,000, 40,000, 45,000, 50,000, 55,000, 60,000, 65,000, 70,000, 75,000, 80,000, 85,000, 90,000, 95,000, 100,000, 200,000, 300,000, 400,000, 500,000, 600,000, 700,000, 800,000, or 900,000.

[0050] In some embodiments, the inorganic support can include the Group 2 element and Al and can be in the form of a mixed Group 2 element/Al metal oxide that has O, Mg, and Al atoms mixed on an atomic scale. In some embodiments the inorganic support can be or can include the Group 2 element and Al in the form of an oxide or one or more oxides of the Group 2 element and Al_2O_3 that can be mixed on a nm scale. In some embodiments, the inorganic support can be or can include an oxide of the Group 2 element, e.g., MgO , and Al_2O_3 mixed on a nm scale.

[0051] In some embodiments, the inorganic support can be or can include a first quantity of the Group 2 element and Al in the form of a mixed Group 2 element/Al metal oxide and a second quantity of the Group 2 element in the form of an oxide of the Group 2 element. In such embodiment, the mixed Group 2 element/Al metal oxide and the oxide of the Group 2 element can be mixed on the nm scale and the Group 2 element and Al in the mixed Group 2 element/Al metal oxide can be mixed on the atomic scale.

[0052] In other embodiments, the inorganic support can be or can include a first quantity of the Group 2 element and a first quantity of Al in the form of a mixed Group 2 element/Al metal oxide, a second quantity of the Group 2 element in the form of an oxide of the Group 2 element, and a second quantity of Al in the form of Al_2O_3 . In such embodiment, the mixed Group 2 element/Al metal oxide, the oxide of the Group 2 element, and the Al_2O_3 can be mixed on a nm scale and the Group 2 element and Al in the mixed Group 2 element/Al metal oxide can be mixed on the atomic scale.

[0053] In some embodiments, when the inorganic support includes the Group 2 element and Al, a weight ratio of the Group 2 element to the Al in the inorganic support can be in a range from 0.001, 0.005, 0.01, 0.05, 0.1, 0.15, 0.2, 0.3, 0.5, 0.7, or 1 to 3, 6, 12.5, 25, 50, 75, 100, 200, 300, 400, 500, 600, 700, 800, 900, or 1,000. In some embodiments, when the inorganic support includes Al, the inorganic support can include Al in a range from 0.5 wt%, 1 wt%, 1.5 wt%, 2 wt%, 2.1 wt%, 2.3 wt%, 2.5 wt%, 2.7 wt%, 3 wt%, 4 wt%, 5 wt%, 6 wt%, 7 wt%, 8 wt%, 9 wt%, 10 wt%, or 11 wt% to 15 wt%, 20 wt%, 25 wt%, 30 wt%, 40 wt%, 45 wt%, or 50 wt%, based on the weight of the inorganic support.

[0054] In some embodiments, the inorganic support can be or can include, but is not limited to, one or more of the following compounds: $\text{Mg}_w\text{Al}_2\text{O}_{3+w}$, where w is a positive number; $\text{Ca}_x\text{Al}_2\text{O}_{3+x}$, where x is a positive number; $\text{Sr}_y\text{Al}_2\text{O}_{3+y}$, where y is a positive number; $\text{Ba}_z\text{Al}_2\text{O}_{3+z}$, where z is a positive number. BeO ; MgO ; CaO ; BaO ; SrO ; BeCO_3 ; MgCO_3 ; CaCO_3 ; SrCO_3 ; BaCO_3 ; CaZrO_3 ; $\text{Ca}_7\text{ZrAl}_6\text{O}_{18}$; CaTiO_3 ; $\text{Ca}_7\text{Al}_6\text{O}_{18}$; $\text{Ca}_7\text{HfAl}_6\text{O}_{18}$; BaCeO_3 ; one or more magnesium chromates, one or more magnesium tungstates, one or more magnesium

molybdates, combinations thereof, and mixtures thereof. In some embodiments, the Group 2 element can include Mg and at least a portion of the Group 2 element can be in the form of MgO or a mixed oxide that includes MgO. In some embodiments, the inorganic support can be or can include, but is not limited to, a MgO-Al₂O₃ mixed metal oxide. In some embodiments, when the inorganic support is a MgO-Al₂O₃ mixed metal oxide, the inorganic support can have a molar ratio of Mg to Al equal to 20, 10, 5, 2, 1 to 0.5, 0.1, or 0.01.

[0055] The Mg_wAl₂O_{3+w}, where w is a positive number, if present as the inorganic support or as a component of the inorganic support can have a molar ratio of Mg to Al in a range from 0.5, 1, 2, 3, 4, or 5 to 6, 7, 8, 9, or 10. In some embodiments, the Mg_wAl₂O_{3+w} can include MgAl₂O₄, Mg₂Al₂O₅, or a mixture thereof. The Ca_xAl₂O_{3+x}, where x is a positive number, if present as the inorganic support or as a component of the inorganic support can have a molar ratio of Ca to Al in a range from 1:12, 1:4, 1:2, 2:3, 5:6, 1:1, 12:14, or 1.5:1. In some embodiments, the Ca_xAl₂O_{3+x} can include tricalcium aluminate, dodecacalcium hepta-aluminate, monocalcium aluminate, monocalcium dialuminate, monocalcium hexa-aluminate, dicalcium aluminate, pentacalcium trialuminate, tetracalcium trialuminate, or any mixture thereof. The Sr_yAl₂O_{3+y}, where y is a positive number, if present as the inorganic support or as a component of the inorganic support can have a molar ratio of Sr to Al in a range from 0.05, 0.3, or 0.6 to 0.9, 1.5, or 3. The Ba_zAl₂O_{3+z}, where z is a positive number, if present as the inorganic support or as a component of the inorganic support can have a molar ratio of Ba to Al 0.05, 0.3, or 0.6 to 0.9, 1.5, or 3.

[0056] In some embodiments, the inorganic support can also include one or more promoters disposed thereon. The promoter can be or can include, but is not limited to, Sn, Ag, Cu, a combination thereof, or a mixture thereof. In some embodiments, the promoter can be associated with the Group 10 element, e.g., Pt. For example, the promoter and the Group 10 element disposed on the inorganic support can form Group 10 element-promoter clusters that can be dispersed on the inorganic support. The promoter, if present, can improve the selectivity/activity/longevity of the catalyst for a given upgraded hydrocarbon. In some embodiments, the addition of the promoter can improve the propylene selectivity of the catalyst when the hydrocarbon-containing feed includes propane. The catalyst can include the promoter in an amount of 0.01 wt%, 0.1 wt%, 0.2 wt%, 0.3 wt%, 0.4 wt%, 0.5 wt%, 0.6 wt%, 0.7 wt%, 0.8 wt%, 0.9 wt%, or 1 wt% to 3 wt%, 5 wt%, 7 wt%, or 10 wt%, based on the weight of the inorganic support.

[0057] In some embodiments, the inorganic support can also include one or more alkali metal elements disposed thereon. The alkali metal element, if present, can be or can include, but is

not limited to, Li, Na, K, Rb, Cs, a combination thereof, or a mixture thereof. In at least some embodiments, the alkali metal element can be or can include K and/or Cs. The alkali metal element, if present, can improve the selectivity of the catalyst for a given upgraded hydrocarbon. The catalyst can include the alkali metal element in an amount 0.01 wt%, 0.1 wt%, 0.2 wt%, 5 0.3 wt%, 0.4 wt%, 0.5 wt%, 0.6 wt%, 0.7 wt%, 0.8 wt%, 0.9 wt%, or 1 wt% to 2 wt%, 3 wt%, 4 wt%, or 5 wt%, based on the weight of the inorganic support. In some embodiments, suitable catalyst can include those described in U.S. Patent Nos.; 5,073,662 and 6,313,063; U.S. Patent Application Publication Nos: 2011/0301392 and 2005/0003960; and European Patent Application Publication Nos.: EP0486993A1 and EP1073516A1.

10 **[0058]** In some embodiments, the inorganic support can also include, but is not limited to, at least one metal element and/or at least one metalloid element selected from Groups other than Group 2 and Group 10 and/or at least one compound thereof, where the at least one metal element and/or at least one metalloid element is not Li, Na, K, Rb, Cs, Sn, Ag, Cu. If the support also includes a compound that includes the metal element and/or metalloid element
15 selected from Groups other than Group 2 and Group 10, where the at least one metal element and/or at least one metalloid element is not Li, Na, K, Rb, Cs, Sn, Ag, or Cu, the compound can be present in the support as an oxide, a phosphate, a halide, a halate, a sulfate, a sulfide, a borate, a nitride, a carbide, an aluminate, an aluminosilicate, a silicate, a carbonate, metaphosphate, a selenide, a tungstate, a molybdate, a chromite, a chromate, a dichromate, or
20 a silicide. In some embodiments, suitable compounds that include the metal element and/or metalloid element selected from Groups other than Group 2 and Group 10, where the at least one metal element and/or at least one metalloid element is not Li, Na, K, Rb, Cs, Sn, Ag, or Cu, can be or can include, but are not limited to, one or more of the following: B_2O_3 , $AlBO_3$, Al_2O_3 , SiO_2 , SiC , Si_3N_4 , an aluminosilicate, zinc aluminate, ZnO , VO , V_2O_3 , VO_2 , V_2O_5 , Ga_2O_3 ,
25 In_2O_3 , Mn_2O_3 , Mn_3O_4 , MnO , one or more molybdenum oxides, one or more tungsten oxides, one or more zeolites, where s, t, u, and v are positive numbers and mixtures and combinations thereof.

[0059] The preparation of the inorganic support can be accomplished via any known process. For simplicity and ease of description, the preparation of a suitable inorganic support that
30 includes a mixed oxide of magnesium and aluminum ($Mg(Al)O$ or MgO/Al_2O_3) inorganic support will be described in more detail. Catalyst synthesis techniques are well-known and the following description is for illustrative purposes and not to be considered as limiting the synthesis of the inorganic support or the catalyst. In some embodiments, to make the MgO/Al_2O_3 mixed oxide inorganic support, Mg and Al precursors such as $Mg(NO_3)_2$ and

Al(NO₃)₃ can be mixed together, e.g., ball-milled, followed by calcination. In another embodiment, the two precursors can be dissolved in H₂O, stirred until dry (with heat optionally applied), followed by calcination to produce the inorganic support. In another embodiment, the two precursors can be dissolved in H₂O, followed by the addition of a base and a carbonate, e.g., NaOH/Na₂CO₃, to produce hydrotalcite, followed by calcination to produce the inorganic support. In another embodiment, a commercial ready MgO and Al₂O₃ may be mixed and ball-milled to produce the inorganic support. In another embodiment, the Mg(NO₃)₂ precursor can be dissolved in H₂O and the solution can be impregnated onto an existing inorganic support, e.g., an Al₂O₃ inorganic support, that can be dried and calcined to produce the inorganic support.

10 In another embodiment, Mg from Mg(NO₃)₂ can be loaded onto an existing Al₂O₃ inorganic support through ion adsorption, followed by liquid-solid separation, drying and calcination to produce the inorganic support. Without wishing to be bound by theory, it is believed that the inorganic support produced via any one of the above methods and/or other methods can include (i) the Mg and Al mixed together on the nm scale, (ii) the Mg and Al in the form of a mixed

15 Mg/Al metal oxide, or (iii) a combination of (i) and (ii).

[0060] Group 10 metals and any promoter and/or any alkali metal element may be loaded onto the mixed oxide inorganic support by any known technique. For example, one or more Group 10 element precursors, e.g., chloroplatinic acid, tetramineplatinum nitrate, and/or tetramineplatinum hydroxide, one or more promoter precursors (if used), e.g., a salt such as

20 SnCl₄ and/or AgNO₃, and one or more alkali metal element precursors (if used), e.g., KNO₃, KCl, and/or NaCl, can be dissolved in water. The solution can be impregnated onto the inorganic support, followed by drying and calcination. In some embodiments, the Group 10 element precursor and optionally the promoter precursor and/or the alkali metal element precursor can be loaded onto the inorganic support at the same time, or separately in a sequence

25 separated by drying and/or calcination steps. In other embodiments, the Group 10 element and, optionally the promoter and/or alkali metal element, can be loaded onto the inorganic support by chemical vapor deposition, where the precursors are volatilized and deposited onto the inorganic support, followed by calcination. In other embodiments, the Group 10 element precursor and, optionally, the promoter precursor and/or alkali metal precursor, can be loaded

30 onto the inorganic support through ion adsorption, followed by liquid-solid separation, drying and calcination. Optionally, the catalyst can also be synthesized using a one-pot synthesis method where the precursors of the inorganic support, the Group 10 metal active phase and the promoters are all mixed together, dry or wet, with or without any other additives to aid the synthesis, followed by drying and calcination.

[0061] Suitable processes that can be used to prepare the catalysts disclosed herein can include the processes described in U.S. Patent Nos. 4,788,371; 4,962,265; 5,922,925; 8,653,317; EP Patent No. EP0098622; Journal of Catalysis 94 (1985), pp. 547-557; and/or Applied Catalysis 54 (1989), pp. 79-90.

5 [0062] The as-synthesized catalyst, when examined under scanning electron microscope or transmission electron microscope, can appear as either primary particles, as agglomerates of primary particles, as aggregated primary particles, or a combination thereof. The primary particles in the as-synthesized catalyst, when examined under scanning electron microscope or transmission electron microscope, can have an average particle size, *e.g.*, a diameter when
10 spherical, in a range from 0.2 nm, 0.5 nm, 1 nm, 5 nm, 10 nm, 25 nm, 30 nm, 40 nm, 50 nm, 60 nm, 70 nm, 80 nm, 90 nm, 100 nm, 150 nm, 200 nm, 250 nm, 300 nm, 350 nm, 400 nm, 450 nm, or 500 nm to 1 μ m, 10 μ m, 25 μ m, 50 μ m, 100 μ m, 150 μ m, 200 μ m, 250 μ m, 300 μ m, 400 μ m, or 500 μ m. In some embodiments, the catalyst particles can have an average cross-sectional length of 0.2 nm to 500 μ m, 0.5 nm to 300 μ m, 1 nm to 200 μ m, 2 nm to 100 μ m, or
15 2 nm to 500 nm as measured by a transmission electron microscope.

[0063] The catalyst can have a surface area in a range from 0.1 m²/g, 1 m²/g, 10 m²/g, or 100 m²/g to 500 m²/g, 800 m²/g, 1,000 m²/g, or 1,500 m²/g. The surface area of the catalyst can be measured according to the Brunauer-Emmett-Teller (BET) method using adsorption-desorption of nitrogen (temperature of liquid nitrogen, 77 K) with a Micromeritics 3flex
20 instrument after degassing of the powders for 4 hrs at 350°C. More information regarding the method can be found, for example, in “Characterization of Porous Solids and Powders: Surface Area, Pore Size and Density,” S. Lowell et al., Springer, 2004.

[0064] In some embodiments, the inorganic support can be extruded or otherwise formed into any desired monolithic structure and the Group 10 element and any optional promoter
25 and/or alkali metal element can be disposed thereon. Suitable monolithic structures can be or can include, but are not limited to, structures having a plurality of substantially parallel internal passages such as those in the form of a ceramic honeycomb. In some embodiments, the support can be in the form of beads, spheres, rings, toroidal shapes, irregular shapes, rods, cylinders, flakes, films, cubes, polygonal geometric shapes, sheets, fibers, coils, helices, meshes, sintered
30 porous masses, granules, pellets, tablets, powders, particulates, extrudates, cloth or web form materials, honeycomb matrix monolith, including in comminuted or crushed forms, and the Group 10 element and any optional promoter and/or alkali metal element can be disposed thereon.

[0065] The as-synthesized catalyst can be formulated into one or more appropriate forms for different short cycle (≤ 5 hours) hydrocarbon upgrading processes. Alternatively, the support can be formulated into appropriate forms for different short cycle hydrocarbon upgrading processes, before the addition of the Group 10 element and, any optional promoter and/or alkali metal element. During formulation, one or more binders and/or additives can be added to the catalyst and/or support to improve the chemical/physical properties of the catalyst. For example, spray-dried catalyst particles having an average cross-sectional diameter in a range from 40 μm to 100 μm are typically used in an FCC type fluid-bed reactor. To make spray-dried catalyst, the support/catalyst needs to be made into a slurry with binder/additive in the slurry before spray-drying and calcination.

Hydrocarbon Upgrading Process

[0066] Returning to the hydrocarbon upgrading process, the hydrocarbon-containing feed and the catalyst and/or at least a portion of the regenerated catalyst can be contacted with one another within any suitable environment such as one or more reaction or conversion zones disposed within one or more reactors to produce the effluent and the at least partially deactivated catalyst. In some embodiments, the reaction or conversion zone can be disposed or otherwise located within one or more fixed bed reactors, one or more fluidized or moving bed reactors, one or more reverse flow reactors, or any combination thereof.

[0067] The hydrocarbon-containing feed and catalyst and/or at least a portion of the regenerated catalyst can be contacted at a temperature in a range from 300°C, 350°C, 400°C, 450°C, 500°C, 550°C, 600°C, 620°C, 650°C, 660°C, 670°C, 680°C, 690°C, or 700°C to 725°C, 750°C, 760°C, 780°C, 800°C, 825°C, 850°C, 875°C, or 900°C. In some embodiments, the hydrocarbon-containing feed and catalyst and/or at least a portion of the regenerated catalyst can be contacted at a temperature of at least 620°C, at least 650°C, at least 660°C, at least 670°C, at least 680°C, at least 690°C, or at least 700°C to 725°C, 750°C, 760°C, 780°C, 800°C, 825°C, 850°C, 875°C, or 900°C. The hydrocarbon-containing feed can be introduced into the reaction or conversion zone and contacted with the catalyst and/or at least a portion of the regenerated catalyst therein for a time period of ≤ 3 hours, ≤ 2.5 hours, ≤ 2 hours, ≤ 1.5 hours, ≤ 1 hour, ≤ 45 minutes, ≤ 30 minutes, ≤ 20 minutes, ≤ 10 minutes, ≤ 5 minutes, ≤ 1 minute, ≤ 30 seconds, ≤ 10 seconds, ≤ 5 seconds, or ≤ 1 second or ≤ 0.5 second. In some embodiments, the hydrocarbon-containing feed can be contacted with the catalyst and/or at least a portion of the regenerated catalyst for a time period in a range from 0.1 seconds, 0.5 seconds, 0.7 seconds, 1 second, 30 second, 1 minute, 5 minutes, or 10 minutes to 30 minutes, 50 minutes, 70 minutes, 1.5 hours, 2 hours, or 3 hours.

[0068] The hydrocarbon-containing feed and the catalyst and/or at least a portion of the regenerated catalyst can be contacted under a hydrocarbon partial pressure of at least 20 kPa-absolute, where the hydrocarbon partial pressure is the total partial pressure of any C₂-C₁₆ alkanes and any C₈-C₁₆ alkyl aromatics in the hydrocarbon-containing feed. In some
5 embodiments, the hydrocarbon partial pressure during contact of the hydrocarbon-containing feed and the catalyst and/or at least a portion of the regenerated catalyst can be in a range from 20 kPa-absolute, 50 kPa-absolute, 100 kPa-absolute, at least 150 kPa, at least 200 kPa 300 kPa-absolute, 500 kPa-absolute, 750 kPa-absolute, or 1,000 kPa-absolute to 1,500 kPa-absolute, 2,500 kPa-absolute, 4,000 kPa-absolute, 5,000 kPa-absolute, 7,000 kPa-absolute, 8,500 kPa-
10 absolute, or 10,000 kPa-absolute, where the hydrocarbon partial pressure is the total partial pressure of any C₂-C₁₆ alkanes and any C₈-C₁₆ alkyl aromatics in the hydrocarbon-containing feed. In other embodiments, the hydrocarbon partial pressure during contact of the hydrocarbon-containing feed and the catalyst and/or at least a portion of the regenerated catalyst can be in a range from 20 kPa-absolute, 50 kPa-absolute, 100 kPa-absolute, 150 kPa-
15 absolute, 200 kPa-absolute, 250 kPa-absolute, or 300 kPa-absolute to 500 kPa-absolute, 600 kPa-absolute, 700 kPa-absolute, 800 kPa-absolute, 900 kPa-absolute, or 1,000 kPa-absolute, where the hydrocarbon partial pressure is the total partial pressure of any C₂-C₁₆ alkanes and any C₈-C₁₆ alkyl aromatics in the hydrocarbon-containing feed.

[0069] In some embodiments, the hydrocarbon-containing feed can include at least 60 vol%,
20 at least 65 vol%, at least 70 vol%, at least 75 vol%, at least 80 vol%, at least 85 vol%, at least 90 vol%, at least 95 vol%, or at least 99 vol% of a single C₂-C₁₆ alkane, e.g., propane, based on a total volume of the hydrocarbon-containing feed. The hydrocarbon-containing feed and catalyst and/or at least a portion of the regenerated catalyst can be contacted under a single C₂-C₁₆ alkane, e.g., propane, pressure of at least 20 kPa-absolute, at least 50 kPa-absolute, at least
25 100 kPa-absolute, at least 150 kPa-absolute, at least 250 kPa-absolute, at least 300 kPa-absolute, at least 400 kPa-absolute, at least 500 kPa-absolute, or at least 1,000 kPa-absolute.

[0070] The hydrocarbon-containing feed can be contacted with the catalyst and/or at least a portion of the regenerated catalyst within the reaction or conversion zone at any weight hourly space velocity (WHSV) effective for carrying out the upgrading process. In some
30 embodiments, the WHSV can be 0.01 hr⁻¹, 0.1 hr⁻¹, 1 hr⁻¹, 2 hr⁻¹, 5 hr⁻¹, 10 hr⁻¹, 20 hr⁻¹, 30 hr⁻¹, or 50 hr⁻¹ to 100 hr⁻¹, 250 hr⁻¹, 500 hr⁻¹, or 1,000 hr⁻¹. In some embodiments, when the hydrocarbon upgrading process includes a fluidized or otherwise moving catalyst and/or moving regenerated catalyst, a ratio of the catalyst circulation mass flow rate to a combined amount of any C₂-C₁₆ alkanes and any C₈-C₁₆ alkyl aromatics mass flow rate can be in a range

from 1, 3, 5, 10, 15, 20, 25, 30, or 40 to 50, 60, 70, 80, 90, 100, 110, 125, or 150 on a weight to weight basis.

[0071] When the activity of the at least partially deactivated catalyst decreases below a desired minimum amount, the at least partially deactivated catalyst or at least a portion thereof can be subjected to the regeneration process described above to produce the regenerated catalyst. Regeneration of the at least partially deactivated catalyst can occur within the reaction or conversion zone or within a combustion zone that is separate and apart from the reaction or conversion zone, depending on the particular reactor configuration, to produce a regenerated catalyst. For example, regeneration of the catalyst can occur within the reaction or conversion zone when a fixed bed or reverse flow reactor is used, or within a separate combustion zone that can be separate and apart from the reaction or conversion zone when a fluidized bed reactor or other circulating or fluidized type reactor is used. Similarly, the optional reduction step can also occur within the reaction or conversion zone, within the combustion zone, and/or within a separate reduction zone. Accordingly, the hydrocarbon containing feed can be contacted with the catalyst to effect one or more of dehydrogenation, dehydroaromatization, and dehydrocyclization of at least a portion of the hydrocarbon-containing feed to produce a coked catalyst and a first effluent that includes the one or more upgraded hydrocarbons and molecular hydrogen in a cyclic type process such as those commonly employed in fixed bed and reverse flow reactors and/or a continuous type process commonly employed in fluidized bed reactors. The separation of the effluent that includes the upgraded hydrocarbon and molecular hydrogen from the coked catalyst, if needed, can be accomplished via one or more separators such as a cyclone separator. As noted above, the oxidative gas can be or can include, but is not limited to, O₂, O₃, CO₂, or a mixture thereof and can include no greater than 5 mol% of H₂O. In some embodiments, an amount of oxidative gas in excess of that needed to combust 100% of the contaminant, e.g., coke, disposed on the catalyst can be used to increase the rate of contaminant removal from the catalyst, so that the time needed for removal of the contaminant can be reduced and lead to an increased yield in the upgraded product produced within a given period of time.

Hydrocarbon-Containing Feed

[0072] The C₂-C₁₆ alkanes can be or can include, but are not limited to, ethane, propane, n-butane, isobutane, n-pentane, isopentane, n-hexane, 2-methylpentane, 3-methylpentane, 2,2-dimethylbutane, n-heptane, 2-methylhexane, 2,2,3-trimethylbutane, cyclopentane, cyclohexane, methylcyclopentane, ethylcyclopentane, n-propylcyclopentane, 1,3-dimethylcyclohexane, or a mixture thereof. For example, the hydrocarbon-containing feed can

include propane, which can be dehydrogenated to produce propylene, and/or isobutane, which can be dehydrogenated to produce isobutylene. In another example, the hydrocarbon-containing feed can include liquid petroleum gas (LP gas), which can be in the gaseous phase when contacted with the catalyst. In some embodiments, the hydrocarbon in the hydrocarbon-containing feed can be composed of substantially a single alkane such as propane. In some
5 embodiments, the hydrocarbon-containing feed can include ≥ 50 mol%, ≥ 75 mol%, ≥ 95 mol%, ≥ 98 mol%, or ≥ 99 mol% of a single C₂-C₁₆ alkane, e.g., propane, based on a total weight of all hydrocarbons in the hydrocarbon-containing feed. In some embodiments, the hydrocarbon-containing feed can include at least 50 vol%, at least 55 vol%, at least 60 vol%, at least 65
10 vol%, at least 70 vol%, at least 75 vol%, at least 80 vol%, at least 85 vol%, at least 90 vol%, at least 95 vol%, at least 97 vol%, or at least 99 vol% of a single C₂-C₁₆ alkane, e.g., propane, based on a total volume of the hydrocarbon-containing feed.

[0073] The C₈-C₁₆ alkyl aromatics can be or can include, but are not limited to, ethylbenzene, propylbenzenes, butylbenzenes, one or more ethyl toluenes, or a mixture thereof. In some
15 embodiments, the hydrocarbon-containing feed can include ≥ 50 mol%, ≥ 75 mol%, ≥ 95 mol%, ≥ 98 mol%, or ≥ 99 mol% of a single C₈-C₁₆ alkyl aromatic, e.g., ethylbenzene, based on a total weight of all hydrocarbons in the hydrocarbon-containing feed. In some embodiments, the ethylbenzene can be dehydrogenated to produce styrene. As such, in some embodiments, the processes disclosed herein can include propane dehydrogenation, butane dehydrogenation,
20 isobutane dehydrogenation, pentane dehydrogenation, pentane dehydrocyclization to cyclopentadiene, naphtha reforming, ethylbenzene dehydrogenation, ethyltoluenes dehydrogenation, and the like.

[0074] In some embodiments, the hydrocarbon-containing feed can be diluted, e.g., with one or more diluents such as one or more inert gases. Suitable inert gases can be or can include,
25 but are not limited to, Ar, Ne, He, N₂, CO₂, CH₄, or a mixture thereof. If the hydrocarbon containing-feed includes a diluent, the hydrocarbon-containing feed can include 0.1 vol%, 0.5 vol%, 1 vol%, or 2 vol% to 3 vol%, 8 vol%, 16 vol%, or 32 vol% of the diluent, based on a total volume of any C₂-C₁₆ alkanes and any C₈-C₁₆ alkyl aromatics in the hydrocarbon-containing feed.

30 **[0075]** In some embodiments, the hydrocarbon-containing feed can also include H₂. In some embodiments, when the hydrocarbon-containing feed includes H₂, a molar ratio of the H₂ to a combined amount of any C₂-C₁₆ alkane and any C₈-C₁₆ alkyl aromatic can be in a range from 0.1, 0.3, 0.5, 0.7, or 1 to 2, 3, 4, 5, 6, 7, 8, 9, or 10.

[0076] In some embodiments, the hydrocarbon-containing feed can be substantially free of any steam, *e.g.*, < 0.1 vol% of steam, based on a total volume of any C₂-C₁₆ alkanes and any C₈-C₁₆ alkyl aromatics in the hydrocarbon-containing feed. In other embodiments, the hydrocarbon-containing feed can include steam. For example, the hydrocarbon-containing feed can include 0.1 vol%, 0.3 vol%, 0.5 vol%, 0.7 vol%, 1 vol%, 3 vol%, or 5 vol% to 10 vol%, 15 vol%, 20 vol%, 25 vol%, 30 vol%, 35 vol%, 40 vol%, 45 vol%, or 50 vol% of steam, based on a total volume of any C₂-C₁₆ alkanes and any C₈-C₁₆ alkyl aromatics in the hydrocarbon-containing feed. In other embodiments, the hydrocarbon-containing feed can include ≤ 50 vol%, ≤ 45 vol%, ≤ 40 vol%, ≤ 35 vol%, ≤ 30 vol%, ≤ 25 vol%, ≤ 20 vol%, or ≤ 15 vol% of steam, based on a total volume of any C₂-C₁₆ alkanes and any C₈-C₁₆ alkyl aromatics in the hydrocarbon-containing feed. In other embodiments, the hydrocarbon-containing feed can include at least 1 vol%, at least 3 vol%, at least 5 vol%, at least 10 vol%, at least 15 vol%, at least 20 vol%, at least 25 vol%, or at least 30 vol% of steam, based on a total volume of any C₂-C₁₆ alkanes and any C₈-C₁₆ alkyl aromatics in the hydrocarbon-containing feed.

[0077] In some embodiments, the hydrocarbon-containing feed can include sulfur. For example, the hydrocarbon-containing feed can include sulfur in a range from 0.5 ppm, 1 ppm, 5 ppm, 10 ppm, 20 ppm, 30 ppm, 40 ppm, 50 ppm, 60 ppm, 70 ppm, or 80 ppm to 100 ppm, 150 ppm, 200 ppm, 300 ppm, 400 ppm, or 500 ppm. In other embodiments, the hydrocarbon-containing feed can include sulfur in a range from 1 ppm to 10 ppm, 10 ppm to 20 ppm, 20 ppm to 50 ppm, 50 ppm to 100 ppm, or 100 ppm to 500 ppm. The sulfur, if present in the hydrocarbon-containing feed, can be or can include, but is not limited to, H₂S, dimethyl disulfide, as one or more mercaptans, or any mixture thereof.

[0078] The hydrocarbon feed can be substantially free or free of molecular oxygen. In some embodiments, the hydrocarbon feed can include ≤ 5 mol%, ≤ 3 mol%, or ≤ 1 mol% of molecular oxygen (O₂). It is believed that providing a hydrocarbon feed substantially-free of molecular oxygen substantially prevents oxidative coupling reactions that would otherwise consume at least a portion of the alkane and/or the alkyl aromatic in the hydrocarbon feed.

Recovery and Use of the Upgraded Hydrocarbons

[0079] The upgraded hydrocarbon can include at least one upgraded hydrocarbon, *e.g.*, an olefin, water, unreacted hydrocarbons, molecular hydrogen, etc. The upgraded hydrocarbon can be recovered or otherwise obtained via any convenient process, *e.g.*, by one or more conventional processes. One such process can include cooling and/or compressing the effluent to condense at least a portion of any water and any heavy hydrocarbon that may be present, leaving the olefin and any unreacted alkane or alkyl aromatic primarily in the vapor phase.

Olefin and unreacted alkane or alkyl aromatic hydrocarbons can then be removed from the reaction product in one or more separator drums. For example, one or more splitters or distillation columns can be used to separate the dehydrogenated product from the unreacted hydrocarbon feed.

- 5 **[0080]** In some embodiments, a recovered olefin, *e.g.*, propylene, can be used for producing polymer, *e.g.*, recovered propylene can be polymerized to produce polymer having segments or units derived from the recovered propylene such as polypropylene, ethylene-propylene copolymer, etc. Recovered isobutene can be used, *e.g.*, for producing one or more of: an oxygenate such as methyl tert-butyl ether, fuel additives such as diisobutene, synthetic
10 elastomeric polymer such as butyl rubber, etc.

Examples:

[0081] The foregoing discussion can be further described with reference to the following non-limiting examples. Catalysts 1-27 and a comparative catalyst were prepared according to the following procedures.

- 15 **[0082]** Catalyst 1: The catalyst was prepared according to the following procedure: Set aside 2.3 g PURALOX® MG 70/170 (Sasol), which was a MgO-Al₂O₃ mixed metal oxide that was obtained by calcining hydrotalcite. The mixed metal oxide contained 70 wt% MgO and 30 wt% Al₂O₃. The BET surface area was 170 m²/g according to Sasol. Tin (IV) chloride pentahydrate (0.103 g) (Acros Organics), chloroplatinic acid hexahydrate (0.0184 g) (BioXtra), and
20 deionized water (2.2 mL) were mixed in a small glass vial to make a solution. The PURALOX® MG 70/170 support was impregnated with the solution. The impregnated material was dried at 110°C for 6 hours, and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.3 wt% Pt and 1.5 wt% Sn.

- [0083]** Catalyst 2: The catalyst was prepared according to the following procedure: Set aside
25 3 g PURALOX® MG 80/150 (Sasol), which was a MgO-Al₂O₃ mixed metal oxide obtained by calcining hydrotalcite. The mixed metal oxide contained 80 wt% MgO and 20 wt% Al₂O₃. The BET surface area was 150 m²/g according to Sasol. Tin (IV) chloride pentahydrate (0.134 g) (Acros Organics), chloroplatinic acid hexahydrate (0.024 g) (BioXtra), and deionized water (2.25 mL) were mixed in a small glass vial to make a solution. The PURALOX® MG 80/150
30 support was impregnated with the solution. The impregnated material was sitting in a closed container at room temperature for 24 h before it was dried at 110°C for 6 hours, and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.3 wt% Pt and 1.5 wt% Sn.

- [0084] Catalyst 3: The catalyst was a Pt-based, Sn-containing catalyst supported on an Mg/Al mixed oxide support. Elemental analysis showed that the catalyst contained 0.48 wt% Pt, 1.25 wt% Sn, 67.93 wt% of Mg, and 29.23 wt% of Al, based on the total weight of the metal elements, with an Mg to Al molar ratio of about 2.58.
- 5 [0085] Catalyst 4: The catalyst was prepared according to the following procedure: Set aside 3 g PURALOX® MG 30/260 (Sasol), which was a MgO-Al₂O₃ mixed metal oxide obtained by calcining hydrotalcite. The mixed metal oxide contained 30 wt% MgO and 70 wt% Al₂O₃. The BET surface area was 260 m²/g according to Sasol. Tin (IV) chloride pentahydrate (0.134 g) (Acros Organics), chloroplatinic acid hexahydrate (0.024 g) (BioXtra), and deionized water
10 (3.15 mL) were mixed in a small glass vial to make a solution. The PURALOX® MG 30/260 support was impregnated with the solution. The impregnated material was stored in a closed container at room temperature for 120 h before it was dried at 110°C for 6 hours, and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.3 wt% Pt and 1.5 wt% Sn.
- 15 [0086] Catalyst 5: The catalyst was prepared according to the following procedure: Set aside 3 g PURALOX® MG 30/70 (Sasol), which was a MgO-Al₂O₃ mixed metal oxide obtained by calcining hydrotalcite. The mixed metal oxide contained 30 wt% MgO and 70 wt% Al₂O₃. The BET surface area was 70 m²/g according to Sasol. Puralox MG30/70 contained more MgAl₂O₄ spinel phase than Puralox MG30/260. Tin (IV) chloride pentahydrate (0.134 g) (Acros
20 Organics), chloroplatinic acid hexahydrate (0.024 g) (BioXtra), and deionized water (2.4 mL) were mixed in a small glass vial to make a solution. The PURALOX® MG 30/70 support was impregnated with the solution. The impregnated material was stored in a closed container at room temperature for 120 h before it was dried at 110°C for 6 hours, and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.3 wt% Pt and 1.5 wt% Sn.
- 25 [0087] Catalyst 6: The catalyst was prepared according to the following procedure: Set aside 2.3 g PURALOX® MG 28/100 (Sasol), which was primarily a MgAl₂O₄ spinel. The BET surface area was 100 m²/g according to Sasol. Tin (IV) chloride pentahydrate (0.103 g) (Acros Organics), chloroplatinic acid hexahydrate (0.0184 g) (BioXtra), and deionized water (2.4 mL) were mixed in a small glass vial to make a solution. The PURALOX® MG 28/100 support
30 was impregnated with the solution. The impregnated material was dried at 110°C for 6 hours, and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.3 wt% Pt and 1.5 wt% Sn.
- [0088] Catalyst 7: The catalyst was prepared according to the following procedure: Set aside 3.5 g PURAL® MG 70 (Sasol), which was a Mg/Al-based hydrotalcite that, upon activation at

550°C for 3 h, form a MgO-Al₂O₃ mixed metal oxide that contains 70 wt% MgO and 30 wt% Al₂O₃. Tin (IV) chloride pentahydrate (0.103 g) (Acros Organics), chloroplatinic acid hexahydrate (0.0184 g) (BioXtra), and deionized water (2.6 mL) were mixed in a small glass vial to make a solution. The PURAL 70 support was impregnated with the solution. The
5 impregnated material was stored in a closed container at room temperature for 24 h before it was dried at 110°C for 6 hours, and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.3 wt% Pt and 1.5 wt% Sn.

[0089] Catalyst 8: The catalyst was prepared according to the following procedure: Set aside 2.5 g MgO (50 nm, Sigma Aldrich). Tin (IV) chloride pentahydrate (0.0158 g) (Acros
10 Organics), chloroplatinic acid hexahydrate (0.0314 g) (BioXtra), and deionized water (1 mL) were mixed in a small glass vial to make a solution. The MgO support was impregnated with the solution. The impregnated material was dried at 110°C for 6 hours, and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.5 wt% Pt and 0.25 wt% Sn.

[0090] Catalyst 9: 3 g of alumina (Sigma Aldrich), 1.93 g of magnesium nitrate hexahydrate
15 (Sigma Aldrich) and 2.06 g of deionized water were mixed and stirred on a hot plate set at 60°C until the mixture was dry. The mixture was further heated to 110°C for 6 hours, and calcined at 800°C for 12 hours, all in air. The solid support obtained contained 91 wt% Al₂O₃ and 9 wt% MgO. Tin (IV) chloride pentahydrate (0.134 g) (Acros Organics), Chloroplatinic acid hexahydrate (0.0244 g) (BioXtra), and deionized water (1 mL) were mixed in a small glass vial
20 to make a solution. The solution was impregnated onto the solid support mentioned above. The impregnated material was dried at 110°C for 6 hours, and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.3 wt% Pt and 1.5 wt% Sn.

[0091] Catalyst 10: The catalyst was prepared according to the following procedure: Set aside 20 g PURALOX® MG 70/170 (Sasol), which was a MgO-Al₂O₃ mixed metal oxide
25 obtained by calcining hydrotalcite. The mixed metal oxide contained 70 wt% MgO and 30 wt% Al₂O₃. The BET surface area was 170 m²/g according to Sasol. An appropriate amount of silver nitrate, tetraammineplatinum(II) nitrate, deionized water were mixed to form a solution. The PURALOX® MG 70/170 support was impregnated with the solution. The impregnated material was stored in a closed container at room temperature for 1 h before it was dried at
30 120°C overnight, and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.3 wt% Pt and 1.5 wt% Ag.

[0092] Catalyst 11: The catalyst was prepared according to the following procedure: Set aside 20 g PURALOX® MG 70/170 (Sasol), which was a MgO-Al₂O₃ mixed metal oxide obtained by calcining hydrotalcite. The mixed metal oxide contained 70 wt% MgO and 30 wt%

Al₂O₃. The BET surface area was 170 m²/g according to Sasol. An appropriate amount of cupric nitrate trihydrate, tetraammineplatinum(II) nitrate, and deionized water were mixed to form a solution. The PURALOX® MG 70/170 support was impregnated with the solution. The impregnated material was stored in a closed container at room temperature for 1 h before
5 it was dried at 120°C overnight, and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.3 wt% Pt and 1.5 wt% Cu.

[0093] Catalyst 12: The catalyst was prepared according to the following procedure: Set aside 20 g PURALOX® MG 70/170 (Sasol), which was a MgO-Al₂O₃ mixed metal oxide obtained by calcining hydrotalcite. The mixed metal oxide contained 70 wt% MgO and 30 wt%
10 Al₂O₃. The BET surface area was 170 m²/g according to Sasol. An appropriate amount of gallium(III) nitrate, tetraammineplatinum(II) nitrate, and deionized water were mixed to form a solution. The PURALOX® MG 70/170 support was impregnated with the solution. The impregnated material was stored in a closed container at room temperature for 1 h before it was dried at 120°C overnight, and calcined at 800°C for 12 hours, all in air. The final product
15 contained nominally 0.3 wt% Pt and 1.5 wt% Ga.

[0094] Catalyst 13: The catalyst was prepared according to the following procedure: Set aside 2.3 g PURALOX® MG 80/150 (Sasol), which was a MgO-Al₂O₃ mixed metal oxide obtained by calcining hydrotalcite. The mixed metal oxide contained 80 wt% MgO and 20 wt% Al₂O₃. The BET surface area was 150 m²/g according to Sasol. Tin (IV) chloride pentahydrate
20 (0.054 g) (Acros Organics), chloroplatinic acid hexahydrate (0.0184 g) (BioXtra), and deionized water (1.725 mL) were mixed in a small glass vial to make a solution. The PURALOX® MG 80/150 support was impregnated with the solution. The impregnated material was dried at 110°C for 6 hours, and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.3 wt% Pt and 0.75 wt% Sn.

[0095] Catalyst 14: The catalyst was prepared according to the following procedure: Set aside 2.3 g PURALOX® MG 80/150 (Sasol), which was a MgO-Al₂O₃ mixed metal oxide obtained by calcining hydrotalcite. The mixed metal oxide contained 80 wt% MgO and 20 wt% Al₂O₃. The BET surface area was 150 m²/g according to Sasol. Tin (IV) chloride pentahydrate
25 (0.103 g) (Acros Organics), chloroplatinic acid hexahydrate (0.0184 g) (BioXtra), and deionized water (1.725 mL) were mixed in a small glass vial to make a solution. The PURALOX® MG 80/150 support was impregnated with the solution. The impregnated material was dried at 110°C for 6 hours, and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.3 wt% Pt and 1.5 wt% Sn.
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[0096] Catalyst 15: The catalyst was prepared according to the following procedure: Set aside 2.3 g PURALOX® MG 80/150 (Sasol), which was a MgO-Al₂O₃ mixed metal oxide obtained by calcining hydrotalcite. The mixed metal oxide contained 80 wt% MgO and 20 wt% Al₂O₃. The BET surface area was 150 m²/g according to Sasol. Tin (IV) chloride pentahydrate (0.214 g) (Acros Organics), chloroplatinic acid hexahydrate (0.0184 g) (BioXtra), and deionized water (1.725 mL) were mixed in a small glass vial to make a solution. The PURALOX® MG 80/150 support was impregnated with the solution. The impregnated material was dried at 110°C for 6 hours, and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.3 wt% Pt and 3.0 wt% Sn.

[0097] Catalyst 16: The catalyst was prepared according to the following procedure: Set aside 5 g PURALOX® MG 80/150 (Sasol), which is an MgO-Al₂O₃ mixed metal oxide obtained by calcining hydrotalcite. The mixed metal oxide contained 80 wt% MgO and 20 wt% Al₂O₃. The BET surface area was 150 m²/g according to Sasol. Tin (IV) chloride pentahydrate (0.100 g) (Acros Organics), chloroplatinic acid hexahydrate (0.04 g) (BioXtra), and deionized water (3.75 mL) were mixed in a small glass vial to make a solution. The PURALOX® MG 80/150 support was impregnated with the solution. The impregnated material was dried at 110°C for 6 hours, and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.3 wt% Pt and 0.67 wt% Sn.

[0098] Catalyst 17: KNO₃ (0.00812 g) was mixed with deionized water (0.7 mL) in a small glass vial to make a solution. Catalyst 16 (1.5 g) was impregnated with the solution. The impregnated material was dried at 110°C for 6 hours, and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.3 wt% Pt, 0.67 wt% Sn and 0.21 wt% of K.

[0099] Catalyst 18: The catalyst was prepared according to the following procedure: Set aside 2.3 g PURALOX® MG 80/150 (Sasol), which was a MgO-Al₂O₃ mixed metal oxide obtained by calcining hydrotalcite. The mixed metal oxide contained 80 wt% MgO and 20 wt% Al₂O₃. The BET surface area was 150 m²/g according to Sasol. Tin (IV) chloride pentahydrate (0.103 g) (Acros Organics), chloroplatinic acid hexahydrate (0.0307 g) (BioXtra), and deionized water (1.725 mL) were mixed in a small glass vial to make a solution. The PURALOX® MG 80/150 support was impregnated with the solution. The impregnated material was dried at 100 °C for 10 hours, and calcined at 800 °C for 15 hours, all in air. The final product contained nominally 0.5 wt% Pt and 1.5 wt% Sn.

[0100] Catalyst 19: The catalyst was prepared according to the following procedure: Set aside 2.3 g PURALOX® MG 80/150 (Sasol), which was a MgO-Al₂O₃ mixed metal oxide obtained by calcining hydrotalcite. The mixed metal oxide contained 80 wt% MgO and 20 wt%

Al₂O₃. The BET surface area was 150 m²/g according to Sasol. Tin (IV) chloride pentahydrate (0.103 g) (Acros Organics), chloroplatinic acid hexahydrate (0.0187 g) (BioXtra), and deionized water (1.725 mL) were mixed in a small glass vial to make a solution. The PURALOX® MG 80/150 support was impregnated with the solution. The impregnated material was dried at 100°C for 10 hours, and calcined at 800°C for 15 hours, all in air. The final product contained nominally 0.3 wt% Pt and 1.5 wt% Sn.

[0101] Catalyst 20: The catalyst was prepared according to the following procedure: Set aside 2.3 g PURALOX® MG 80/150 (Sasol), which was a MgO-Al₂O₃ mixed metal oxide obtained by calcining hydrotalcite. The mixed metal oxide contained 80 wt% MgO and 20 wt% Al₂O₃. The BET surface area was 150 m²/g according to Sasol. The PURALOX® MG 80/150 support was transferred to a 200 mL beaker filled with 80 mL of deionized water. The mixture was stirred to form slurry A. Tin (II) chloride dihydrate (0.0663 g) (Sigma Aldrich) and fuming HCl (2 mL) were mixed to form solution A. Chloroplatinic acid hexahydrate (0.0307 g) (BioXtra) was mixed with deionized water (10 mL) to form solution B. Solution A and solution B were mixed to form solution C which had an amber color. Solution C was added drop-wise to slurry A and was stirred for 45 minutes. The resulting mixture was filtered to form a filter cake that was washed with ample deionized water three times. Both the supernatant and the wastewater from washing the filter cake were colorless. The filter cake was dried at 100°C for 10 hours, and calcined at 800°C for 15 hours, all in air. The final product contained nominally 0.5 wt% Pt and 1.5 wt% Sn.

[0102] Catalyst 21: A 1/1 (wt/wt) Mg/Al mixed metal oxide support was made by dissolving Mg(NO₃)₂•6H₂O (44.58 g) (Sigma-Aldrich) and Al(NO₃)₃•9H₂O (51.55 g) (Sigma-Aldrich) in deionized water (100 g). The solution was stirred at 70°C to evaporate water until solids started to form. The material was then calcined at 200°C for 4 h and then at 800°C for 4 h, all in air. The material was finally ball milled for 2 h in an agate cup at 500 rpm to obtain the support.

[0103] In a glass vial, chloroplatinic acid hexahydrate (0.1280 g) (Sigma Aldrich) and Tin (II) chloride dihydrate (0.0998 g) (Fluka) were dissolved in deionized water (10.4960 g) to obtain a dark orange clear solution. The Mg/Al mixed metal oxide support (13.3279 g) was transferred to a 50 mL plastic bottle, to which the Pt/Sn solution (9.8957 g) was added. The material was mixed using a small lab shaker until a homogeneous light orange colored powder was obtained. The metal-impregnated material was calcined at 120°C for 4 h and then at 800°C for 12 h, all in air. The final product contained nominally 0.33 wt% Pt and 0.33 wt% Sn.

[0104] Catalyst 22: A 0.5/1 (wt/wt) Mg/Al mixed metal oxide support was made by dissolving Mg(NO₃)₂•6H₂O (31.89 g) (Sigma-Aldrich) and Al(NO₃)₃•9H₂O (73.64 g) (Sigma-

Aldrich) in deionized water (25.47 g). The solution was stirred at 70°C to evaporate water until solids started to form. The material was placed in an oven at 120°C for 1 h to further evaporate water. The dried material was calcined at 200°C for 4 h and then at 800°C for 4 h, all in air. The calcined material was finally ball milled for 2 h in an agate cup at 500 rpm to obtain the support.

[0105] In a glass vial, chloroplatinic acid hexahydrate (0.1727 g) (Sigma Aldrich) and tin(II) chloride dihydrate (0.1313 g) (Fluka) was added to deionized water (10.7014 g) to obtain a dark orange clear solution. The Mg/Al mixed metal oxide support (14.2533 g) was transferred to a 50 mL plastic bottle, to which the Pt/Sn solution (8.0550 g) was added. The material was mixed using a small lab shaker until a homogeneous light orange colored powder was obtained. The metal-impregnated material was calcined at 120°C for 4 h and then at 800 °C for 12 h, all in air. The final product contained nominally 0.33 wt% Pt and 0.33 wt% Sn.

[0106] Catalyst 23: A 2/1 (wt/wt) Mg/Zr mixed metal oxide support was made by dissolving zirconium(IV) oxynitrate hydrate (3.69 g) (Sigma-Aldrich) in deionized water (40 g) at 50°C. $\text{Mg}(\text{NO}_3)_2 \cdot 6\text{H}_2\text{O}$ (12.55 g) (Sigma-Aldrich) was added to the solution. The mixture was stirred at 70°C to evaporate water until solids started to form. The material was then calcined at 200°C for 4 h and then at 800°C for 4 h, all in air, to obtain the support.

[0107] In a glass vial, chloroplatinic acid hexahydrate (0.0212 g) (Sigma Aldrich) and tin(IV) chloride pentahydrate (0.0318 g) (Sigma Aldrich) were dissolved in deionized water (0.6647 g) to obtain a dark orange clear solution. The Mg/Zr mixed metal oxide support (2.0412 g) was transferred to a 50 mL plastic bottle, to which the Pt/Sn solution (0.6007 g) was added. The material was mixed using a small lab shaker until a homogeneous light orange colored powder was obtained. The metal-impregnated catalyst was dried at 120°C for 4 h, and then calcined at 800°C for 12 h, all in air. The final product contained nominally 0.33 wt% Pt and 0.33 wt% Sn.

[0108] Catalyst 24: To make a 2/1 (wt/wt) Mg/Ti mixed metal oxide support, in an agate ball mill cup, 8.04 g MgO (Acros), 4.02 g TiO_2 (Anatase, Sigma Aldrich) and 10 g H_2O were added together with 3 agate balls. The material was ball milled for 70 h at 500 rpm. The lid was sealed to reduce water evaporation while ball milling. After ball milling, a thick paste was obtained. The paste was calcined at 200 °C for 4 h and then at 800 °C for 4 h to obtain the support.

[0109] In a glass vial, 0.0288 g chloroplatinic acid hexahydrate (Sigma Aldrich) and 0.0336 g tin(IV) chloride pentahydrate (Sigma Aldrich) were dissolved in 1.6007 g H_2O to obtain a dark orange clear solution. 2.7321 g of the Mg/Ti mixed metal oxide support was transferred

to a 50 mL plastic bottle, followed by dropwise addition of 1.4080 g of the Pt/Sn solution made above. The material was mixed using a small lab shaker until a homogeneous light orange colored powder was obtained. The metal-impregnated catalyst was dried at 120 °C for 4 h, and then calcined at 800°C for 12 h. The final product contained nominally 0.33 wt% Pt and 0.33 wt% Sn.

[0110] Catalyst 25: A 2/1 (wt/wt) Mg/Si mixed metal oxide was made in an agate ball mill cup by adding MgO (8.04 g) (Acros), SiO₂ (4.03 g) (Sigma-Aldrich), deionized water (10 g) and, 3 agate balls. The material was ball milled for 70 h at 500 rpm. The lid was sealed to reduce water evaporation during ball milling. After ball milling, a thick paste was obtained. The paste was calcined at 200°C for 4 h and then at 800°C for 4 h, all in air, to obtain the support.

[0111] In a glass vial, chloroplatinic acid hexahydrate (0.0296 g) (Sigma Aldrich) and tin (IV) chloride pentahydrate (0.0365 g) (Sigma Aldrich) were dissolved in deionized water (1.4319 g) to obtain a dark orange clear solution. The 2/1 Mg/Si mixed metal oxide support (2.8178 g) was added to a 50 mL plastic bottle, followed by dropwise addition of the Pt/Sn solution (1.2704 g). The material was mixed using a small lab shaker until a homogeneous light orange colored powder was obtained. The metal-impregnated catalyst was dried at 120°C for 4 h, and then calcined at 800°C for 12 h, all in air. The final product contained nominally 0.33 wt% Pt and 0.33 wt% Sn.

[0112] Catalyst 26: A Mg/Ca/Sn/Al mixed metal oxide support was made by combining and thoroughly mixing MgO (23.4 g) (Aldrich), SnO₂ (1.56 g) (Aldrich), calcium aluminate cement (12.8 g) (Almatis), and Al₂O₃ (60.5 g) (Versal 300). An aqueous acetic acid solution was made by adding 10.24 mL of acetic acid to 100 mL deionized water. A 60.4 mL aliquot was taken out of the solution and an additional 0.5 mL of acetic acid was added to the aliquot. The aliquot was added to the powder in 5 mL increments, with two minutes of stirring between additions, until it was used up. Deionized water was subsequently added to the mixture until a slurry was formed. The slurry was then calcined at 300°C for 20 h and then at 843°C for 5 h, all in air, to obtain the support. 0.3 wt% Pt and an additional 0.3 wt% of Sn was impregnated onto the support made above, yielding the final catalyst.

[0113] Catalyst 27: The catalyst was prepared according to the following procedure: Set aside 20 g PURALOX® MG 70/170 (Sasol), which was a MgO-Al₂O₃ mixed metal oxide obtained by calcining hydrotalcite. The mixed metal oxide contained 70 wt% MgO and 30 wt% Al₂O₃. The BET surface area was 170 m²/g according to Sasol. An appropriate amount of tin(II) chloride dehydrate and deionized water were mixed to form a solution. The

PURALOX® MG 70/170 support was impregnated with the solution. The impregnated material was stored in a closed container at room temperature for 1 h before it was dried at 120°C overnight. An appropriate amount of tetraammineplatinum(II) nitrate and deionized water were mixed to form a solution. The support impregnated with Sn was further
 5 impregnated with the Pt solution. The impregnated material was sitting in a closed container at room temperature for 1 h before it was dried at 120°C overnight and calcined at 800°C for 12 hours, all in air. The final product contained nominally 0.3 wt% Pt and 1.5 wt% Sn.

[0114] Catalyst compositions 28-41 were prepared according to the following procedure. For each catalyst composition PURALOX® MG 80/150 (3 grams) (Sasol), which was a mixed
 10 Mg/Al metal oxide that contained 80 wt% of MgO and 20 wt% of Al₂O₃ and had a surface area of 150 m²/g, was calcined under air at 550°C for 3 hours to form a support. Solutions that contained a proper amount of tin (IV) chloride pentahydrate when used to make the catalyst composition (Acros Organics) and/or chloroplatinic acid when used to make the catalyst composition (Sigma Aldrich), and 1.8 ml of deionized water were prepared in small glass vials.
 15 The calcined PURALOX® MG 80/150 supports (2.3 grams) for each catalyst composition were impregnated with the corresponding solution. The impregnated materials were allowed to equilibrate in a closed container at room temperature (RT) for 24 hours, dried at 110°C for 6 hours, and calcined at 800°C for 12 hours. Table 1 shows the nominal Pt and Sn content of each catalyst composition based on the weight of the support.

Table 1		
Catalyst	Pt (wt%)	Sn (wt%)
28	0.4	1
29	0.3	1
30	0.2	1
31	0.1	1
32	0.05	1
33	0.025	1
34	0.0125	1
35	0	1
36	0.1	0.5
37	0.1	1
38	0.1	2
39	0.0125	0
40	0.0125	0.5
41	0.0125	2

[0115] Comparative catalyst 1: In a graduated cylinder, SnCl_2 (0.048 g) (Aldrich), chloroplatinic acid, 8% solution, (0.79 g) (Aldrich), and remainder HCl (1.2 M) (Acculute) were combined to make a dark solution of 5.6 mL. The solution was added to theta-alumina (10 g) and stirred for 15 minutes. The catalyst was allowed to rest for 1 hr. The catalyst was placed in a muffle furnace and ramped at $3^\circ\text{C}/\text{min}$ to 120°C , held for 2 hours at 120°C , and then the catalyst was ramped at $3^\circ\text{C}/\text{min}$ to 550°C , which was maintained for 2 hours, all in air. The catalyst was then cooled to room temperature.

[0116] In a graduated cylinder, KNO_3 (0.258 g) (Aldrich) was dissolved in deionized water to yield 5.6 mL of solution. The solution was added to the Pt-Sn catalyst and stirred for 15 minutes. The catalyst was allowed to rest for 1 hr. The catalyst was placed in a muffle furnace and ramped at $3^\circ\text{C}/\text{min}$ to 120°C , held for 2 hours at 120°C , and then the catalyst was ramped at $3^\circ\text{C}/\text{min}$ to 550°C , which was maintained for 2 hours, all in air. The catalyst was then cooled to room temperature. The final product contained nominally 0.3 wt% Pt, 0.3 wt% Sn, and 1.0 wt% K.

15 Examples using the catalysts described above.

[0117] Fixed bed experiments were conducted at approximately 100 kPa-absolute. A gas chromatograph (GC) was used to measure the composition of the reactor effluents. The concentrations of each component in the reactor effluents were then used to calculate the C_3H_6 yield and selectivity. The C_3H_6 yield and selectivity, as reported in these examples, were calculated on the carbon mole basis.

[0118] In each example, a certain amount of the catalyst "Mcat" was mixed with an appropriate amount of quartz diluent and loaded into a quartz reactor. The amount of diluent was determined so that the catalyst bed (catalyst + diluent) overlaps with the isothermal zone of the quartz reactor and the catalyst bed is largely isothermal during operation. The dead volume of the reactor was filled with quartz chips/rods.

[0119] The concentration of each component in the reactor effluent was used to calculate the C_3H_6 yield and selectivity. The C_3H_6 yield and the selectivity at the beginning of t_{rxn} and at the end of t_{rxn} is denoted as Y_{ini} , Y_{end} , S_{ini} , and S_{end} , respectively, and reported as percentages in the data tables below.

30 [0120] Example 1 - Effect of exposing the spent catalyst to steam at 800°C before oxidation. Two options (Option A and Option B) were used in step 1 to evaluate the effect of exposing the spent catalyst to steam, referred to as Case 1A and Case 1B, respectively. The process steps were as follows: 1. (Option A/Case 1A) - The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature T_{oxi} . After that, 46.6 sccm of He with

5.1 sccm of steam was passed through the reaction zone for 5 min. 1. (Option B/Case 1B) - The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature T_{oxi} . After that 46.6 sccm of He was passed through the reaction zone for 5 min. 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. After t_{oxi} , an inert was passed through the reaction zone and the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}). 5. A H_2 containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H_2 containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 670 °C. 7. A hydrocarbon-containing (HC_{gas}) feed that included 81 vol% C_3H_8 , 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. As shown in Table 2 below, comparing Case 1A that used Option A to Case 1B that used Option B shows that when the spent catalyst was exposed to steam at 800°C before oxidation, the performance of the catalyst was not affected.

Table 2		
	Case 1A	Case 1B
Catalyst	1	1
M _{cat} (g)	0.3	0.3
F _{rxn} (sccm)	17.6	17.6
H _{gas}	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar
F _{red} (sccm)	46.6	46.6
T _{red} (°C)	800	800
t _{red} (min)	0.05	0.05
O _{gas}	Dry air	Dry air
F _{oxi} (sccm)	83.9	83.9
T _{oxi} (°C)	800	800
t _{oxi} (min)	10	10
Performance	Y _{ini}	65.3
	Y _{end}	59.2
	S _{ini}	95.5
	S _{end}	96.5

[0121] Example 2 - Effect of exposing the spent catalyst to steam at 670°C before oxidation. Two options (Option A and Option B) were used in step 8 to evaluate the effect of exposing the spent catalyst to steam referred to as Case 2A and Case 2B, respectively. The process steps were as follows:

1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature T_{oxi}.
2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone.
3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst.
4. After t_{oxi}, an inert was passed through the reaction zone and the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}).
5. A H₂ containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H₂ containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}).
6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 670°C.
7. A hydrocarbon-containing (HC_{gas}) feed that included 81 vol% C₃H₈, 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone

at 670 °C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. 8. (Option A) 46.6 sccm of He with 5.1 sccm of steam was passed through the reaction zone for 5 min. 8. (Option B) 46.6 sccm of He was passed through the reaction zone for 5 min. The above process steps were repeated in cycles until stable performance was obtained. As shown in Table 3 below, comparing Case 2A that used Option A to Case 2B that used Option B shows that when the spent catalyst was exposed to steam at 670°C before oxidation, the performance of the catalyst was not affected.

Table 3		
	Case 2A	Case 2B
Catalyst	1	1
M _{cat} (g)	0.3	0.3
F _{rxn} (sccm)	17.6	17.6
H _{gas}	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar
F _{red} (sccm)	46.6	46.6
T _{red} (°C)	800	800
t _{red} (min)	0.05	0.05
O _{gas}	Dry air	Dry air
F _{oxi} (sccm)	83.9	83.9
T _{oxi} (°C)	800	800
t _{oxi} (min)	10	10
Performance	Y _{ini}	65.7
	Y _{end}	63.5
	S _{ini}	94.1
	S _{end}	95.9

[0122] Example 3A - Preferred oxidation temperature/duration. The process steps were as follows: 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature T_{oxi}. 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. After t_{oxi}, an inert was passed through the reaction zone and the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}). 5. A H₂ containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert gas was passed through the reaction zone. This was then followed by flowing the H₂ containing gas through the reaction zone at

T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 670°C. 7. A hydrocarbon-containing (HCgas) feed that included 81 vol% C_3H_8 , 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670 °C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained.

Table 4A		
Catalyst	1	1
M_{cat} (g)	0.3	0.3
F_{rxn} (sccm)	17.6	17.6
Hgas	10% H_2 ; 90% Ar	10% H_2 ; 90% Ar
F_{red} (sccm)	14.0	14.0
T_{red} (°C)	670	670
t_{red} (min)	0.05	0.05
Ogas	Dry air	Dry air
F_{oxi} (sccm)	37.3	37.3
T_{oxi} (°C)	800	800
t_{oxi} (min)	10	30
Performance	Y_{ini}	62.7
	Y_{end}	54.5
	S_{ini}	93.7
	S_{end}	95.3

10 **[0123]** Example 3B - Preferred oxidation temperature/duration. The process steps were as follows: 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature T_{oxi} . 2. An oxygen containing gas (Ogas) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. After t_{oxi} , an inert was passed through the reaction zone and the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}). 5. A H_2 containing gas (Hgas) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H_2 containing gas through the reaction zone at

T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 670°C. 7. A hydrocarbon-containing (HCgas) feed that included 81 vol% C_3H_8 , 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. Tables 4A, 4B-1, and 4B-2 show that a longer oxidation duration resulted in a more effective regeneration of the catalysts. Tables 4B-3 shows that a higher oxidation temperature of 800°C or 850°C was a more effective oxidation temperature than 750°C. For example, the time that was needed to achieve a similar yield/selectivity using an oxidizing temperature of 750°C was three (3) times greater than was needed at 800°C.

Table 4B-1			
Catalyst		1	1
M_{cat} (g)		0.3	0.3
F_{rxn} (sccm)		19.7	19.7
H_{gas}		10% H_2 ; 90% Ar	10% H_2 ; 90% Ar
F_{red} (sccm)		46.6	46.6
T_{red} (°C)		800	800
t_{red} (min)		0.05	0.05
O_{gas}		Dry air	Dry air
F_{oxi} (sccm)		83.9	83.9
T_{oxi} (°C)		800	800
t_{oxi} (min)		10	5
Performance	Y_{ini}	65.7	65.1
	Y_{end}	58.9	53.8
	S_{ini}	95.6	95.6
	S_{end}	96.5	96.8

Table 4B-2				
Catalyst	2	2	2	2
M _{cat} (g)	0.3	0.3	0.3	0.3
F _{rxn} (sccm)	17.6	17.6	17.6	17.6
H _{gas}	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar
F _{red} (sccm)	46.6	46.6	46.6	46.6
T _{red} (°C)	800	800	800	850
t _{red} (min)	0.05	0.05	0.05	0.05
O _{gas}	Dry air	Dry air	Dry air	Dry air
F _{oxi} (sccm)	83.9	83.9	83.9	83.9
T _{oxi} (°C)	800	850	850	850
t _{oxi} (min)	10	10	5	5
Performance	Y _{ini}	65.2	65.5	64.5
	Y _{end}	60.8	60.8	58.7
	S _{ini}	95.0	95.2	95.4
	S _{end}	96.3	96.5	96.6

Table 4B-3			
Catalyst	3	3	3
M _{cat} (g)	0.225	0.225	0.225
F _{rxn} (sccm)	17.6	17.6	17.6
H _{gas}	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar
F _{red} (sccm)	46.6	46.6	46.6
T _{red} (°C)	800	800	800
t _{red} (min)	0.05	0.05	0.05
O _{gas}	Dry air	Dry air	Dry air
F _{oxi} (sccm)	83.9	83.9	83.9
T _{oxi} (°C)	800	750	800
t _{oxi} (min)	10	30	10
Performance	Y _{ini}	64.9	65.0
	Y _{end}	61.1	61.5
	S _{ini}	94.7	94.3
	S _{end}	95.8	95.5

[0124] Example 4 - effect of O₂ partial pressure and the presence of CO₂. The process steps were as follows: 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature T_{oxi}. 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was

passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. The system was flushed with an inert gas. During this process, the temperature of the reaction zone was changed to T_{red} . 5. A H_2 containing gas (Hgas) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H_2 containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 670°C . 7. A hydrocarbon-containing (HCgas) feed that included 81 vol% C_3H_8 , 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 5 shows the impact of the partial pressure of O_2 to oxidation and the presence of CO_2 to oxidation.

Table 5				
Catalyst		1	1	1
M_{cat} (g)		0.3	0.3	0.3
F_{rxn} (sccm)		17.6	17.6	17.6
Hgas		10% H_2 ; 90% Ar	10% H_2 ; 90% Ar	10% H_2 ; 90% Ar
F_{red} (sccm)		46.6	46.6	46.6
T_{red} ($^\circ\text{C}$)		800	800	800
t_{red} (min)		0.05	0.05	0.05
Ogas		Dry air	10% O_2 ; 90% He	10% O_2 ; 10% CO_2 ; 80% He
F_{oxi} (sccm)		83.9	83.9	83.9
T_{oxi} ($^\circ\text{C}$)		800	800	800
t_{oxi} (min)		10	10	10
Performance	Y_{ini}	64.9	62.0	61.8
	Y_{end}	60.4	55.8	55.3
	S_{ini}	95.6	96.3	96.3
	S_{end}	96.4	96.6	96.6

[0125] Example 5 - Effect of steam during oxidation. The process steps were as follows: 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation

temperature T_{oxi} . 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. After t_{oxi} , an inert was passed through the reaction zone and the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}). 5. The system was flushed with an inert gas. 6. A H_2 containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H_2 containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 7. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 670°C . 8. A hydrocarbon-containing (HC_{gas}) feed that included 81 vol% C_3H_8 , 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 6 shows that the presence of more than 10 vol% of steam in air during oxidation yielded an even more deactivated catalyst after regeneration. The more steam present in air during oxidation, the lower the activity. On the other hand, if the moist air was switched to dry air after 2 min of oxidation, the catalyst was effectively regenerated.

Table 6				
Catalyst	1	1	1	1
M _{cat} (g)	0.3	0.3	0.3	0.3
F _{rxn} (sccm)	17.6	17.6	17.6	17.6
H _{gas}	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar
F _{red} (sccm)	46.6	46.6	46.6	46.6
T _{red} (°C)	750	750	750	750
t _{red} (min)	0.05	0.05	0.05	0.05
O _{gas}	Dry air	95.7% Air; 4.3% H ₂ O	90% Air; 10% H ₂ O	90% Air; 10% H ₂ O; Then Dry air
F _{oxi} (sccm)	46.6	48.7	51.7	51.7; Then 46.6
T _{oxi} (°C)	800	800	800	800
t _{oxi} (min)	10	10	10	2; Then 8
Performance	Y _{ini}	65.1	61.2	56.8
	Y _{end}	61.1	56.5	51.3
	S _{ini}	95.2	95.5	95.2
	S _{end}	95.9	96.0	95.4

[0126] Example 6A - Effect of steam during oxidation. 1. The system was flushed with an inert gas. 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. The reaction zone was heated to an oxidation temperature T_{oxi}. 3. The oxygen containing gas was then

5 passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. After t_{oxi}, the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}) while maintaining the oxygen-containing gas flow. 4. The system was flushed with an inert gas. 5. A H₂ containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction

10 zone. This was then followed by flowing the H₂ containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 670°C. 7. A hydrocarbon-containing (HC_{gas}) feed that included 81 vol% C₃H₈, 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the

15 reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in

cycles until stable performance was obtained. Table 7A shows the impact of the presence of steam during oxidation. The shorter the duration of contact between the catalyst and steam (1 vs 3 min) was during oxidation, the easier it was to restore the performance of the catalyst through subsequent oxidation by dry air.

Table 7A				
Catalyst	1	1	1	1
M _{cat} (g)	0.3	0.3	0.3	0.3
F _{rxn} (sccm)	17.6	17.6	17.6	17.6
H _{gas}	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar
F _{red} (sccm)	14.0	14.0	14.0	14.0
T _{red} (°C)	750	750	750	750
t _{red} (min)	0.05	0.05	0.05	0.05
O _{gas}	Dry air	90% Air; 10% H ₂ O; Then Dry air	90% Air; 10% H ₂ O; Then Dry air	90% Air; 10% H ₂ O; Then Dry air
F _{oxi} (sccm)	37.3	41.4; Then 37.3	41.4; Then 37.3	41.4; Then 37.3
T _{oxi} (°C)	800	800	800	800
t _{oxi} (min)	10	1; Then 10	3; Then 10	3; Then 10
Performance	Y _{ini}	66.2	65.3	64.9
	Y _{end}	62.3	61.6	60.6
	S _{ini}	94.5	94.0	94.2
	S _{end}	95.7	95.3	95.4

- 5 **[0127]** Example 6B - Effect of steam during oxidation. 1. The system was flushed with an inert gas. 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. The reaction zone was heated to an oxidation temperature T_{oxi}. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. After
- 10 t_{oxi}, the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}) while maintaining the oxygen-containing gas flow. 4. The system was flushed with an inert gas. 5. A H₂ containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H₂ containing gas through the reaction zone at
- 15 T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 655 °C. 7. A hydrocarbon-containing (HC_{gas}) feed that included 81 vol% C₃H₈, 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the

reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 655°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 7B shows that if a two-step oxidation scheme where moist air oxidation of 1 min was immediately followed by dry air oxidation of 10 min, was employed, the negative effect of moist air to oxidation was negligible. FIG. 1 shows that the catalyst can be effectively regenerated by using such a two-step oxidation scheme for 80+ cycles.

Table 7B			
Catalyst	3		3
M _{cat} (g)	0.225		0.225
F _{rxn} (sccm)	17.6		17.6
H _{gas}	10% H ₂ ; 90% Ar		10% H ₂ ; 90% Ar
F _{red} (sccm)	46.6		46.6
T _{red} (°C)	800		800
t _{red} (min)	0.05		0.05
O _{gas}	Dry air		90% Air; 10% H ₂ O; Then Dry air
F _{oxi} (sccm)	83.9		93.2; Then 83.9
T _{oxi} (°C)	800		800
t _{oxi} (min)	10		1; Then 10
Performance	Y _{ini}	62.4	62.4
	Y _{end}	59.3	59.8
	S _{ini}	94.9	94.5
	S _{end}	95.8	96.1

10 [0128] Example 7 - Effect of reduction temperature. 1. The system was flushed with an inert gas. 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone. During this process, the temperature of the reaction zone was changed to an oxidation temperature (T_{oxi}) and it was kept at T_{oxi} for a certain period of time (t_{oxi}) to oxidize the catalyst. After t_{oxi}, the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}). 4. The system was flushed with an inert gas. 5. A H₂ containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H₂ containing gas through the reaction

- zone at T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 670°C. 7. A hydrocarbon-containing (HCgas) feed that included 81 vol% C_3H_8 , 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 8 shows that reduction can be carried out at a variety of temperatures. A reduction duration as short as 0.05 min can be used.

Table 8				
Catalyst	1	1	1	1
M_{cat} (g)	0.3	0.3	0.3	0.3
F_{rxn} (sccm)	35.2	35.2	35.2	35.2
Hgas	10% H_2 , ; 90% Ar	10% H_2 , ; 90% Ar	10% H_2 , ; 90% Ar	10% H_2 , ; 90% Ar
F_{red} (sccm)	14.0	14.0	14.0	14.0
T_{red} (°C)	800	750	670	670
t_{red} (min)	0.05	0.05	0.05	0.05
Ogas	Dry air	Dry air	Dry air	Dry air
F_{oxi} (sccm)	83.9	83.9	83.9	83.9
T_{oxi} (°C)	800	800	800	800
t_{oxi} (min)	10	10	10	10
Performance	Y_{ini}	59.7	62.9	61.4
	Y_{end}	53.5	56.4	53.1
	S_{ini}	96.6	96.2	96.3
	S_{end}	97.7	97.5	97.5

- [0129] Example 8 - Effect of steam during reduction. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature T_{oxi} . 2. An oxygen containing gas (Ogas) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. After t_{oxi} , an inert was passed through the reaction zone and the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}). 5. A H_2 containing gas (Hgas) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing

the H₂ containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 670 °C. 7. A hydrocarbon-containing (HCgas) feed that included 81 vol% C₃H₈, 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670 °C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 9 shows that the negative impact of steam to the effectiveness of reduction increases with the amount of steam in the reduction gas.

Table 9				
Catalyst	1	1	1	1
M _{cat} (g)	0.3	0.3	0.3	0.3
F _{rxn} (sccm)	17.6	17.6	17.6	17.6
H _{gas}	10% H ₂ ; 90% Ar	9.6% H ₂ ; 86.1% Ar; 4.3% H ₂ O	9% H ₂ ; 81% Ar; 10% H ₂ O	
F _{red} (sccm)	46.6	48.7	51.7	
T _{red} (°C)	800	800	800	
t _{red} (min)	0.2	0.2	0.2	
O _{gas}	Dry air	Dry air	Dry air	
F _{oxi} (sccm)	83.9	83.9	83.9	
T _{oxi} (°C)	800	800	800	
t _{oxi} (min)	10	10	10	
Performance	Y _{ini}	66.1	66.2	65.2
	Y _{end}	61.5	61.8	60.6
	S _{ini}	95.8	95.6	95.9
	S _{end}	96.4	96.5	96.6

[0130] Example 9 - Effect of hydrocarbon during reduction. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature T_{oxi}. 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. After t_{oxi}, an inert was passed through the reaction zone and the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}). 5. A H₂ containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain

period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H_2 containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of $670^\circ C$. 7. A hydrocarbon-containing (HCgas) feed that included 81 vol% C_3H_8 , 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at $670^\circ C$ for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 10 shows the negative impact of trace hydrocarbon to the effectiveness of reduction.

Table 10		
Catalyst	1	1
M_{cat} (g)	0.3	0.3
F_{rxn} (sccm)	17.6	17.6
Hgas	9% H_2 ; 81% Ar; 10% H_2O ; 40 ppm C_3H_8	9% H_2 ; 81% Ar; 10% H_2O ; 110 ppm C_3H_8
F_{red} (sccm)	46.6	46.6
T_{red} ($^\circ C$)	800	800
t_{red} (min)	0.2	0.2
Ogas	Dry air	Dry air
F_{oxi} (sccm)	83.9	83.9
T_{oxi} ($^\circ C$)	800	800
t_{oxi} (min)	10	10
Performance	Y_{ini}	64.4
	Y_{end}	61.1
	S_{ini}	95.8
	S_{end}	96.5

[0131] Example 10 - Reduction of wet solid. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature T_{oxi} . 2. An oxygen containing gas (Ogas) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. After t_{oxi} , an inert was passed through the reaction zone and the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}). Step 5 was carried out in one of the

following three options to evaluate the effectiveness of reducing a wet solid. 5. (Option 1/Case 5A) – A H_2 containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for 1.5 min, while 83.9 sccm of He was passed through the reaction zone. This was then followed by flowing the H_2 containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 5. (Option 2/Case 5B) – A H_2 containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for 1.5 min, while 83.9 sccm of He with 9.2 sccm of steam was passed through the reaction zone. This was then followed by flowing the H_2 containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 5. (Option 3/Case 5C) – A H_2 containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for 3.0 min. Meanwhile, 83.9 sccm of He with 9.2 sccm of steam was passed through the reaction zone during the initial 1.5 min, 89.3 sccm of He was passed through the reaction zone during the subsequent 1.5 min. This was then followed by flowing the H_2 containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 670°C. 7. A hydrocarbon-containing (HCgas) feed that included 81 vol% C_3H_8 , 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670 °C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 11 shows the results for Cases 5A-5C. Comparing case 5A to 5B shows that the reduction of a oxidized catalyst with adsorbed H_2O is not effective. Comparing case 5B to 5C suggests that adsorbed H_2O may be removed by a dry gas purge. The re-dried catalyst can then be effectively reduced by H_2 .

Table 11				
		Case 5A	Case 5B	Case 5C
Catalyst		1	1	1
M _{cat} (g)		0.3	0.3	0.3
F _{rxn} (sccm)		17.6	17.6	17.6
H _{gas}		10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar
F _{red} (sccm)		46.6	46.6	46.6
T _{red} (°C)		800	800	800
t _{red} (min)		0.05	0.05	0.05
O _{gas}		Dry air	Dry air	Dry air
F _{oxi} (sccm)		83.9	83.9	83.9
T _{oxi} (°C)		800	800	800
t _{oxi} (min)		10	10	10
Performance	Y _{ini}	65.0	59.8	63.3
	Y _{end}	62.0	56.1	60.0
	S _{ini}	96.1	96.6	96.5
	S _{end}	96.5	96.5	96.9

[0132] Example 11 - effect of steam to reduced catalyst. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature T_{oxi}. 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then

5 passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. The system was flushed with an inert gas. During this process, the temperature of the reaction zone was changed to T_{red}. 5. A H₂ containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H₂ containing gas through

10 the reaction zone at T_{red} for a certain period of time (t_{red}). 6. After t_{red}, 111.8 sccm of He, with or without 12.3 sccm of additional steam, was passed through the reaction zone while the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 670°C. 7. A hydrocarbon-containing (HC_{gas}) feed that included 81 vol% C₃H₈, 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for

15 a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670 °C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable

performance was obtained. Table 12 shows the impact of the presence of steam to a reduced catalyst.

Table 12		
Catalyst	1	1
M _{cat} (g)	0.3	0.3
F _{rxn} (sccm)	17.6	17.6
H _{gas}	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar
F _{red} (sccm)	46.6	46.6
T _{red} (°C)	800	800
t _{red} (min)	0.05	0.05
Gas used during step 6	111.8 sccm He	111.8 sccm He + 12.3 sccm H ₂ O
O _{gas}	Dry air	Dry air
F _{oxi} (sccm)	83.9	83.9
T _{oxi} (°C)	800	800
t _{oxi} (min)	10	10
Performance	Y _{ini}	65.2
	Y _{end}	59.2
	S _{ini}	95.4
	S _{end}	96.4

[0133] Example 12 - Effect of cooling after reduction. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature of 800°C. 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. The system was flushed with an inert gas. During this process, the temperature of the reaction zone was kept at 800°C. 5. A H₂ containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H₂ containing gas through the reaction zone at 800 °C for a certain period of time (t_{red}). 6. The reaction zone was passed with He of various flow rates (F_{he}). During this process, the temperature of the reaction zone was reduced from 800 °C to a reaction temperature of 670 °C. A higher F_{he} resulted in a faster cooling rate (R_c), defined as the temperature drop during the 1st min of cooling. 7. A hydrocarbon-containing (HC_{gas}) feed that included 81 vol% C₃H₈, 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-

- containing feed was then passed through the reaction zone at 670 °C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 13 shows that after reduction, it can be desirable to cool down the reduced catalyst in a short period of time to preserve high activity.

Table 13				
Catalyst	1	1	1	1
M _{cat} (g)	0.3	0.3	0.3	0.3
F _{rxn} (sccm)	17.6	17.6	17.6	17.6
H _{gas}	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar
F _{red} (sccm)	46.6	46.6	46.6	46.6
t _{red} (min)	0.05	0.05	0.05	0.05
F _{he} (sccm)	14.0	46.6	111.8	111.8
R _c (°C/min)	16.6	27.3	30.8	30.8
O _{gas}	Dry air	Dry air	Dry air	Dry air
F _{oxi} (sccm)	46.6	46.6	46.6	46.6
t _{oxi} (min)	10	10	10	10
Performance	Y _{ini}	58.2	60.5	64.7
	Y _{end}	53.6	55.7	56.7
	S _{ini}	96.0	95.9	95.6
	S _{end}	96.5	96.0	96.4

- [0134]** Example 13 - Effect of H₂ partial pressure during reduction. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature of 800°C. 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. The system was flushed with an inert gas. During this process, the temperature of the reaction zone was kept at 800°C. 5. A H₂ containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H₂ containing gas through the reaction zone at 800°C for a certain period of time (t_{red}). 6. The reaction zone was passed with He. During this process, the temperature of the reaction zone was reduced from 800°C to a reaction temperature of 670°C. 7. A hydrocarbon-containing (HC_{gas}) feed that included 81 vol% C₃H₈, 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time,

while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained.

- 5 Table 14 shows that there is little difference between 10% vs 40% H₂.

Table 14		
Catalyst	1	1
M _{cat} (g)	0.3	0.3
F _{rxn} (sccm)	17.6	17.6
H _{gas}	10% H ₂ ; 90% Ar	40% H ₂ ; 90% Ar
F _{red} (sccm)	46.6	46.6
t _{red} (min)	0.05	0.05
O _{gas}	Dry air	Dry air
F _{oxi} (sccm)	83.9	83.9
t _{oxi} (min)	10	10
Performance	Y _{ini}	65.4
	Y _{end}	60.9
	S _{ini}	96.0
	S _{end}	96.9

- [0135] Example 14A - Effect of H₂ reduction duration. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature of 800°C. 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. The system was flushed with an inert gas. During this process, the temperature of the reaction zone was kept at 800°C. 5. A H₂ containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H₂ containing gas through the reaction zone at 800°C for a certain period of time (t_{red}). 6. The reaction zone was passed with He. During this process, the temperature of the reaction zone was reduced from 800°C to a reaction temperature of 670°C. 7. A hydrocarbon-containing (HC_{gas}) feed that included 81 vol% C₃H₈, 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as

the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 15A shows that there is little difference between the reduction durations in terms of the performance of the catalyst.

Table 15A				
Catalyst	1	1	1	1
M _{cat} (g)	0.3	0.3	0.3	0.3
F _{rxn} (sccm)	17.6	17.6	17.6	17.6
Hgas	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar
F _{red} (sccm)	46.6	46.6	46.6	46.6
t _{red} (min)	0.02	0.05	0.1	0.2
Ogas	Dry air	Dry air	Dry air	Dry air
F _{oxi} (sccm)	83.9	83.9	83.9	83.9
t _{oxi} (min)	10	10	10	10
Perfor- mance	Y _{ini}	64.5	65.3	64.7
	Y _{end}	59.5	61.9	62.1
	S _{ini}	95.3	95.3	95.1
	S _{end}	96.2	96.3	96.2

- 5 **[0136]** Example 14B - Effect of H₂ reduction duration. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature of 800°C. 2. An oxygen containing gas (Ogas) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. The system was flushed with an inert gas. During this process, the temperature of the reaction zone was kept at 800°C. 5. A H₂ containing gas (Hgas) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H₂ containing gas through the reaction zone at 800°C for a certain period of time (t_{red}). 6. The reaction zone was passed with He. During this process, the temperature of the reaction zone was reduced from 800°C to a reaction temperature of 655°C. 7. A hydrocarbon-containing (HCgas) feed that included 81 vol% C₃H₈, 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 655°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above

process steps were repeated in cycles until stable performance was obtained. Table 15B shows that without catalyst reduction, the propylene yield of the oxidized catalyst (53.7%) was even lower than that of the deactivated catalyst (61.3%).

Table 15B		
Catalyst	27	27
M _{cat} (g)	0.3	0.3
F _{rxn} (sccm)	17.6	17.6
H _{gas}	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar
F _{red} (sccm)	46.6	46.6
t _{red} (min)	0	0.05
O _{gas}	90% Air; 10% H ₂ O; Then Dry air	90% Air; 10% H ₂ O; Then Dry air
F _{oxi} (sccm)	93.2; Then 83.9	93.2; Then 83.9
t _{oxi} (min)	1; Then 10	1; Then 10
Performance	Y _{ini}	53.7
	Y _{end}	63.3
	S _{ini}	37.9
	S _{end}	61.3
		93.1
		94.5
		95.9
		96.1

- [0137] Example 15 - Effect of exposing the spent catalyst to inert at 800°C. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature of 800°C. Step 2 was carried out in one of the following two options to evaluate the exposing the spent catalyst to inert at 800°C. 2. (Option 1/Case 15A) An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone for 1 min, while He was passed through the reaction zone for 1 min. 2. (Option 2/Case 15B) An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone for 3 min, while He was passed through the reaction zone for 7 min. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. After t_{oxi}, an inert was passed through the reaction zone and the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}). 5. A H₂ containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H₂ containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 670°C. 7. A hydrocarbon-containing (HC_{gas}) feed that included 81 vol% C₃H₈, 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time,

while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained.

- 5 Table 16 shows that there is little difference in the performance of the catalyst for the two cases.

Table 16		
	Case 15A	Case 15B
Catalyst	1	1
M _{cat} (g)	0.3	0.3
F _{rxn} (sccm)	17.6	17.6
Hgas	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar
F _{red} (sccm)	46.6	46.6
t _{red} (min)	0.05	0.05
Ogas	Dry air	Dry air
F _{oxi} (sccm)	83.9	83.9
t _{oxi} (min)	10	10
Performance	Y _{ini}	64.8
	Y _{end}	60.9
	S _{ini}	95.3
	S _{end}	96.2

- [0138] Example 16 - n-Butane dehydrogenation. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature of 800°C. 2. An oxygen containing gas (Ogas) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then
- 10 passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. After t_{oxi}, an inert was passed through the reaction zone and the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}). 5. A H₂ containing gas (Hgas) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing
- 15 the H₂ containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 635°C. 7. A hydrocarbon-containing (HCgas) feed that included 89 vol% n-butane and 11 vol% steam at a flow rate (F_{rxn}) was passed
- 20 through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the

reaction zone at 635°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. The catalyst showed no deactivation after 30+ cycles. No internal standard was used during this experiment, therefore, to calculate selectivity and yield, the species that were not analyzed by GC (C_5 and C_{5+} , coke) were assumed to be negligible. The absence of a large amount of coke made during reaction was confirmed by the low amounts of CO/CO₂ made during oxidation. Yield/selectivity is defined on the basis of the molar flow rate of all major linear C₄ species, including 1-butene, cis-2-butene, trans-2-butene, 1,3-butadiene, made during the reaction.

Table 17 shows that the catalyst is effective in n-butane dehydrogenation.

Table 17		
Catalyst	1	
M _{cat} (g)	0.3	
F _{rxn} (sccm)	18.5	
H _{gas}	10% H ₂ ; 90% Ar	
F _{red} (sccm)	46.6	
t _{red} (min)	0.05	
O _{gas}	Dry air	
F _{oxi} (sccm)	83.9	
t _{oxi} (min)	15	
Performance	Y _{ini}	66.2
	Y _{end}	57.7
	S _{ini}	95.6
	S _{end}	96.2

[0139] Example 17 - Isobutane dehydrogenation. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature of 800°C. 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. After t_{oxi}, an inert was passed through the reaction zone and the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}). 5. A H₂ containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H₂ containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone

was changed from T_{red} to a reaction temperature of 670°C. 7. A hydrocarbon-containing (HCgas) feed that included 89 vol% iso-butane and 11 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. The catalyst showed no deactivation after 30+ cycles. No internal standard was used during this experiment, therefore, to calculate selectivity and yield, the species that were not analyzed by GC (C_5 and C_{5+} , coke) were assumed to be negligible. The absence of a large amount of coke made during reaction was confirmed by the low amounts of CO/CO₂ made during oxidation. Yield/selectivity is defined on the basis of the molar flow rate of isobutene made during the reaction. Table 18 shows that the catalyst is effective in isobutane dehydrogenation. FIG. 2 shows that isobutane dehydrogenation was stable over 30+ cycles on this catalyst, despite of the high temperature used during reaction, reduction and oxidation.

Table 18		
Catalyst		1
M_{cat} (g)		0.3
F_{rxn} (sccm)		18.1
Hgas		10% H ₂ ; 90% Ar
F_{red} (sccm)		46.6
t_{red} (min)		0.05
Ogas		Dry air
F_{oxi} (sccm)		83.9
t_{oxi} (min)		20
Performance	Y_{ini}	77.3
	Y_{end}	70.5
	S_{ini}	92.1
	S_{end}	94.0

[0140] Example 18 - Ethane dehydrogenation. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature of 800°C. 2. An oxygen containing gas (Ogas) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. After t_{oxi} , an inert was passed through the reaction zone and the temperature within the reaction

zone was changed from T_{oxi} to a reduction temperature (T_{red}). 5. A H_2 containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H_2 containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 670°C . 7. A hydrocarbon-containing (HC_{gas}) feed that included 81 vol% C_2H_6 , 9 vol% Ar and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. The catalyst showed no deactivation after 30+ cycles. Table 19 shows that the catalyst is effective in ethane dehydrogenation. When the catalyst in the reaction zone was replaced by quartz, the C_2H_4 yield measured was less than 1 Cmol%, indicating that homogeneous reactions were not significant at the testing conditions.

Table 19	
Catalyst	3
M_{cat} (g)	0.225
F_{rxn} (sccm)	50.6
H_{gas}	10% H_2 ; 90% Ar
F_{red} (sccm)	46.6
t_{red} (min)	0.05
O_{gas}	Dry air
F_{oxi} (sccm)	83.9
t_{oxi} (min)	10
Performance	Y_{ini} 29.8
	Y_{end} 27.7
	S_{ini} 92.6
	S_{end} 95.9

[0141] Example 19 - Effect of support. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature of 800°C . 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. The system

was flushed with an inert gas. During this process, the temperature of the reaction zone was kept at 800°C. 5. A H₂ containing gas (Hgas) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H₂ containing gas through the reaction zone at 800°C for a certain period of time (t_{red}). 6. The reaction zone was passed with He. During this process, the temperature of the reaction zone was reduced from 800°C to a reaction temperature of 670°C. 7. A hydrocarbon-containing (HCgas) feed that included 81 vol% C₃H₈, 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 20 shows that highly active/selective/stable catalysts can be made with various Mg-containing catalyst supports.

Table 20				
Catalyst	4	5	6	7
M _{cat} (g)	0.3	0.3	0.3	0.3
F _{rxn} (sccm)	17.6	17.6	17.6	17.6
Hgas	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar
F _{red} (sccm)	46.6	46.6	46.6	46.6
t _{red} (min)	0.05	0.05	0.05	0.05
Ogas	Dry air	Dry air	Dry air	Dry air
F _{oxi} (sccm)	83.9	83.9	83.9	83.9
t _{oxi} (min)	10	10	10	10
Performance	Y _{ini}	60.8	55.6	52.3
	Y _{end}	52.2	45.9	42.6
	S _{ini}	93.4	94.0	93.6
	S _{end}	94.7	94.2	93.9

15 [0142] FIG. 3 shows that the performance of catalyst 6 was stable over 30+ cycles despite the high temperature used during reaction, reduction and oxidation.

[0143] Example 20 - Effect of support. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature of 800°C. 2. An oxygen containing gas (Ogas) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. The

system was flushed with an inert gas. During this process, the temperature of the reaction zone was cooled down to 620°C. 5. A H₂ containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H₂ containing gas through the reaction zone at 620°C for a certain period of time (t_{red}). 6. The reaction zone was passed with an inert gas. During this process, the temperature of the reaction zone was maintained at 620°C. 7. A hydrocarbon-containing (HC_{gas}) feed that included 81 vol% C₃H₈, 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 620°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 21 shows that highly active/selective/stable catalysts can be made with various Mg-containing catalyst supports.

Table 21		
Catalyst	8	9
M _{cat} (g)	0.3	0.3
F _{rxn} (sccm)	5.2	5.2
H _{gas}	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar
F _{red} (sccm)	46.6	46.6
t _{red} (min)	1.0	1.0
O _{gas}	Dry air	Dry air
F _{oxi} (sccm)	83.9	83.9
t _{oxi} (min)	15	15
Performance	Y _{ini}	49.2
	Y _{end}	39.5
	S _{ini}	95.5
	S _{end}	96.4

15 [0144] FIG. 4 shows that the performance of catalyst 8 was stable over 20+ cycles despite of the high temperature used during reaction, reduction and oxidation. FIG. 5 shows that the performance of catalyst 9 was stable over 30+ cycles on this catalyst, despite of the high temperature used during reaction, reduction and oxidation.

[0145] Example 21 - Effect of metallic promoters. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature of 800°C. 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone,

while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. After t_{oxi} , an inert was passed through the reaction zone and the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}). 5. A H_2 containing gas (Hgas) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H_2 containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 670°C . 7. A hydrocarbon-containing (HCgas) feed that included 81 vol% C_3H_8 , 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 22 shows that metals other than Sn may be used together with Pt for dehydrogenation.

Table 22				
Catalyst		10	11	12
M_{cat} (g)		0.3	0.3	0.3
F_{rxn} (sccm)		17.6	17.6	17.6
Hgas		10% H_2 ; 90% Ar	10% H_2 ; 90% Ar	10% H_2 ; 90% Ar
F_{red} (sccm)		46.6	46.6	46.6
t_{red} (min)		0.05	0.05	0.05
Ogas		Dry air	Dry air	Dry air
F_{oxi} (sccm)		83.9	83.9	83.9
t_{oxi} (min)		10	10	10
Performance	Y_{ini}	24.3	41.0	49.9
	Y_{end}	9.1	27.7	35.3
	S_{ini}	63.2	82.0	83.1
	S_{end}	67.9	85.7	84.9

[0146] Example 22 - Effect of Sn level. 1. An oxygen containing gas (Ogas) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone kept at 670°C , while an inert was passed through the reaction zone. 2. The oxygen containing gas was then passed through the reaction zone while the temperature of the reaction zone was ramped up to 800°C . The oxygen containing gas kept on flowing through the reaction zone for a certain period of time (t_{oxi}) to

oxidize the catalyst. After that, the reaction zone was cooled down to 670°C in the oxygen containing gas. 3. The system was flushed with an inert gas. During this process, the temperature of the reaction zone was maintained at 670°C. 4. A H₂ containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H₂ containing gas through the reaction zone at 670°C for a certain period of time (t_{red}). 5. The reaction zone was passed with an inert gas. During this process, the temperature of the reaction zone was maintained at 670°C. 6. A hydrocarbon-containing (HC_{gas}) feed that included 81 vol% C₃H₈, 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 23 shows that the Sn loading on the catalyst can be varied.

Table 23				
Catalyst	13	14	15	
M _{cat} (g)	0.3	0.3	0.3	
F _{rxn} (sccm)	17.6	17.6	17.6	
H _{gas}	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	
F _{red} (sccm)	14.0	14.0	14.0	
t _{red} (min)	1	1	1	
O _{gas}	Dry air	Dry air	Dry air	
F _{oxi} (sccm)	14.0	14.0	14.0	
t _{oxi} (min)	30	30	30	
Performance	Y _{ini}	65.7	67.0	60.6
	Y _{end}	61.3	64.5	52.7
	S _{ini}	93.3	94.4	95.7
	S _{end}	95.6	95.7	96.3

[0147] Example 23 - Effect of alkali metal additive. 1. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone kept at 670°C, while an inert was passed through the reaction zone. 2. The oxygen containing gas was then passed through the reaction zone while the temperature of the reaction zone was ramped up to 800 °C. 3. The oxygen containing gas kept on flowing through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. After that, the reaction zone was cooled down to 670 °C in

the oxygen containing gas. 3. The system was flushed with an inert gas. During this process, the temperature of the reaction zone was maintained at 670°C. 4. A H₂ containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H₂ containing gas through the reaction zone at 670°C for a certain period of time (t_{red}). 5. The reaction zone was passed with an inert gas. During this process, the temperature of the reaction zone was maintained at 670°C. 6. A hydrocarbon-containing (HC_{gas}) feed that included 81 vol% C₃H₈, 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 24 shows that alkali metals such as K may be added to the catalyst.

Table 24			
Catalyst		16	17
M _{cat} (g)		0.3	0.3
F _{rxn} (sccm)		17.6	17.6
H _{gas}		10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar
F _{red} (sccm)		14.0	14.0
t _{red} (min)		1	1
O _{gas}		Dry air	Dry air
F _{oxi} (sccm)		14.0	14.0
t _{oxi} (min)		30	30
Performance	Y _{ini}	61.5	62.2
	Y _{end}	50.9	53.7
	S _{ini}	94.1	95.5
	S _{end}	95.2	96.2

15 **[0148]** Example 24 - Effect of Pt level and synthesis method. 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature of 800°C. 2. An oxygen containing gas (O_{gas}) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. After t_{oxi}, an inert was passed through the reaction zone and the temperature within the reaction zone was changed from T_{oxi} to a reduction temperature (T_{red}). 5. A H₂ containing

gas (Hgas) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H_2 containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 6. The system was flushed with an inert. During this process, the temperature of the reaction zone was changed from T_{red} to a reaction temperature of 670°C. 7. A hydrocarbon-containing (HCgas) feed that included 81 vol% C_3H_8 , 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670 °C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 25 shows that the Pt loading on the catalyst and the method of synthesis can be varied.

Table 25				
Catalyst		18	19	20
M_{cat} (g)		0.18	0.3	0.3
F_{rxn} (sccm)		17.6	17.6	17.6
Hgas		10% H_2 ; 90% Ar	10% H_2 ; 90% Ar	10% H_2 ; 90% Ar
F_{red} (sccm)		46.6	46.6	46.6
t_{red} (min)		0.05	0.05	0.05
Ogas		Dry air	Dry air	Dry air
F_{oxi} (sccm)		83.9	83.9	83.9
t_{oxi} (min)		10	10	10
Performance	Y_{ini}	63.4	66.1	58.5
	Y_{end}	58.8	62.3	51.6
	S_{ini}	96.0	95.1	96.2
	S_{end}	96.8	96.2	96.7

[0149] Example 25 - Effect of support. 1. An oxygen containing gas (Ogas) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone kept at 620°C, while an inert was passed through the reaction zone. 2. The oxygen containing gas was then passed through the reaction zone while the temperature of the reaction zone was ramped up to 800°C. The oxygen containing gas kept on flowing through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. After that, the reaction zone was cooled down to 620 °C in the oxygen containing gas. 3. The system was flushed with an inert gas. During this process, the temperature of the reaction zone was maintained at 620°C. 4. A H_2 containing gas (Hgas) at

a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H_2 containing gas through the reaction zone at $620^\circ C$ for a certain period of time (t_{red}). 5. The reaction zone was passed with an inert gas. During this process, the temperature of the reaction zone was maintained at $620^\circ C$. 6. A hydrocarbon-containing (HCgas) feed that included 81 vol% C_3H_8 , 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at $620^\circ C$ for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained. Table 26 shows that catalysts with good activity, selectivity and high temperature stability can be made with various Mg-containing catalyst supports.

Table 26		
Catalyst	21	22
M_{cat} (g)	0.0825	0.0825
F_{rxn} (sccm)	10.4	10.4
Hgas	10% H_2 ; 90% Ar	10% H_2 ; 90% Ar
F_{red} (sccm)	14.0	14.0
t_{red} (min)	1	1
Ogas	10% O_2 ; 90% He	10% O_2 ; 90% He
F_{oxi} (sccm)	14.0	14.0
t_{oxi} (min)	30	30
Performance	Y_{ini}	36.9
	Y_{end}	29.9
	S_{ini}	95.1
	S_{end}	96.2

[0150] Example 26 - Effect of support. 1. The system was flushed with an inert gas while the reaction zone was changed to an oxidation temperature of T_{oxi} . 2. An oxygen containing gas (Ogas) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. The system was flushed with an inert gas. During this process, the temperature of the reaction zone was changed to T_{red} . 5. A H_2 containing gas (Hgas) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the

- reaction zone. This was then followed by flowing the H₂ containing gas through the reaction zone at T_{red} for a certain period of time (t_{red}). 6. The reaction zone was passed with an inert gas. During this process, the temperature of the reaction zone was changed to T_{rxn}. 7. A hydrocarbon-containing (HCgas) feed that included 81 vol% C₃H₈, 9 vol% inert (Ar or Kr) and 10 vol% steam at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at T_{rxn} for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. The above process steps were repeated in cycles until stable performance was obtained.

Table 27					
Catalyst	23	24	25	26	
M _{cat} (g)	0.3	0.3	0.3	0.3	
F _{rxn} (sccm)	5.2	17.6	17.6	17.6	
T _{rxn} (°C)	620	620	620	640	
Hgas	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	10% H ₂ ; 90% Ar	
F _{red} (sccm)	46.6	46.6	46.6	46.6	
T _{red} (°C)	620	620	620	800	
t _{red} (min)	1	1	1	0.05	
Ogas	Dry air	Dry air	Dry air	Dry air	
F _{oxi} (sccm)	83.9	83.9	83.9	83.9	
T _{oxi} (°C)	700	800	800	800	
t _{oxi} (min)	15	15	15	15	
Perfor- mance	Y _{ini}	23.1	36.0	38.3	42.5
	Y _{end}	17.9	30.1	31.4	34.0
	S _{ini}	84.1	88.2	84.0	95.1
	S _{end}	88.3	94.7	92.8	95.1

- [0151] Table 27 above shows that catalysts with good activity, selectivity and high temperature stability can be made with various Mg-containing catalyst supports. FIG. 6 shows that the performance of catalyst 24 was stable over 20+ cycles, despite of the high temperature used during reaction, reduction and oxidation.
- 15 [0152] Comparative Example 1: 1. The system was flushed with an inert gas while the reaction zone was heated to an oxidation temperature of T_{oxi}. 2. An oxygen containing gas (Ogas) at a flow rate (F_{oxi}) was passed through the by-pass of the reaction zone, while an inert was passed through the reaction zone. 3. The oxygen containing gas was then passed through

the reaction zone for a certain period of time (t_{oxi}) to oxidize the catalyst. 4. The system was flushed with an inert gas. During this process, the temperature of the reaction zone was cooled down to 620°C. 5. A H_2 containing gas (H_{gas}) at a flow rate (F_{red}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. This was then followed by flowing the H_2 containing gas through the reaction zone at 620°C for a certain period of time (t_{red}). 6. The reaction zone was passed with an inert gas. During this process, the temperature of the reaction zone was maintained at 620°C. 7. A hydrocarbon-containing (HC_{gas}) feed that included 90 vol% C_3H_8 , 10 vol% inert (Ar or Kr) at a flow rate (F_{rxn}) was passed through the by-pass of the reaction zone for a certain period of time, while an inert was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 620°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone. Table 28 shows further details of the testing conditions of the comparative example. FIG. 7 shows that the performance of comparative catalyst 1 kept on deactivating, even though the oxidation temperature (620°C) was much lower than the other examples.

Table 28	
Catalyst	Comparative catalyst 1
M_{cat} (g)	0.3
F_{rxn} (sccm)	15.8
H_{gas}	10% H_2 ; 90% Ar
F_{red} (sccm)	46.6
t_{red} (min)	1
O_{gas}	Dry air
F_{oxi} (sccm)	83.9
T_{oxi} (°C)	620
t_{oxi} (min)	30

[0153] Example 27 - Fixed bed experiments were conducted at approximately 100 kPa-absolute that used catalysts 28-41. A gas chromatograph (GC) was used to measure the composition of the reactor effluents. The concentrations of each component in the reactor effluents were then used to calculate the C_3H_6 yield and selectivity. The C_3H_6 yield and selectivity, as reported in these examples, were calculated on the carbon mole basis.

[0154] In each example, 0.3 g of the catalyst composition was mixed with an appropriate amount of quartz diluent and loaded into a quartz reactor. The amount of diluent was determined so that the catalyst bed (catalyst + diluent) overlapped with the isothermal zone of

the quartz reactor and the catalyst bed was largely isothermal during operation. The dead volume of the reactor was filled with quartz chips/rods.

[0155] The C_3H_6 yield and the selectivity at the beginning of t_{rxn} and at the end of t_{rxn} is denoted as Y_{ini} , Y_{end} , S_{ini} , and S_{end} , respectively, and reported as percentages in Tables 29 and 30 below for catalysts 28-35.

[0156] The process steps for catalysts 28-35 were as follows: 1. The system was flushed with an inert gas. 2. Dry air at a flow rate of 83.9 sccm was passed through a by-pass of the reaction zone, while an inert was passed through the reaction zone. The reaction zone was heated to a regeneration temperature of 800°C. 3. Dry air at a flow rate of 83.9 sccm was then passed through the reaction zone for 10 min to regenerate the catalyst. 4. The system was flushed with an inert gas. 5. A H_2 containing gas with 10 vol% H_2 and 90 vol % Ar at a flow rate of 46.6 sccm was passed through the by-pass of the reaction zone for a certain period of time, while an inert gas was passed through the reaction zone. This is then followed by flowing the H_2 containing gas through the reaction zone at 800°C for 3 seconds. 6. The system was flushed with an inert gas. During this process, the temperature of the reaction zone was changed from 800°C to a reaction temperature of 670°C. 7. A hydrocarbon-containing (HCgas) feed that included 81 vol% of C_3H_8 , 9 vol% of inert gas (Ar or Kr) and 10 vol% of steam at a flow rate of 35.2 sccm was passed through the by-pass of the reaction zone for a certain period of time, while an inert gas was passed through the reaction zone. The hydrocarbon-containing feed was then passed through the reaction zone at 670°C for 10 min. GC sampling of the reaction effluent started as soon as the feed was switched from the by-pass of the reaction zone to the reaction zone.

[0157] The above process steps were repeated in cycles until stable performance was obtained. Tables 29 and 30 show that Catalyst 33 that contained only 0.025 wt% of Pt and 1 wt% of Sn had both a similar yield and a similar selectivity as compared to Catalyst 28 that contained 0.4 wt% of Pt and 1 wt% of Sn, which was surprising and unexpected. Catalyst 35 that did not include any Pt did not show an appreciable propylene yield.

Table 29					
		Catalyst 28	Catalyst 29	Catalyst 30	Catalyst 31
Performance	Y _{ini}	61.7	61.7	60.7	63.7
	Y _{end}	55.2	55.7	54.2	56.7
	S _{ini}	97.3	97.2	97.0	97.1
	S _{end}	98.1	98.0	97.7	98.3

Table 30					
		Catalyst 32	Catalyst 33	Catalyst 34	Catalyst 35
Performance	Y _{ini}	62.4	62.0	56.7	2.0
	Y _{end}	57.2	54.6	45.7	1.7
	S _{ini}	96.7	97.3	96.9	64.2
	S _{end}	97.7	98.0	97.6	49.5

- [0158] Catalysts 36-41 were also tested using the same process steps 1-7 described above with regard to catalysts 28-35. Table 31 shows that the level of Sn should not be too low or too high for optimal propylene yield for the catalyst compositions that included 0.1 wt% of Pt based on the weight of the support.

Table 31					
		Catalyst 36	Catalyst 31	Catalyst 37	Catalyst 38
		0.5 wt% Sn	1 wt% Sn	1 wt% Sn	2 wt% Sn
Performance	Y _{ini}	58.4	63.7	63.4	56.5
	Y _{end}	49.5	56.7	55.5	47.7
	S _{ini}	96.9	97.1	97.2	97.8
	S _{end}	97.6	98.3	98.1	98.2

- [0159] Table 32 shows that the level of Sn should not be too high or too low for optimal propylene yield for the catalyst compositions that included 0.0125 wt% of Pt based on the weight of the support.

Table 32					
		Catalyst 39	Catalyst 40	Catalyst 34	Catalyst 41
		0 wt% Sn	0.5 wt% Sn	1 wt% Sn	2 wt% Sn
Performance	Y _{ini}	2.6	44	56.7	55.4
	Y _{end}	1.7	24.4	45.7	44.1
	S _{ini}	63.9	96.7	96.9	96.8
	S _{end}	61.1	95.6	97.6	97.6

[0160] Catalyst 33 that contained only 0.025 wt% of Pt and 1 wt% of Sn was also subjected to a longevity test using the same process steps 1-7 described above with regard to catalysts 28 to 35, except a flow rate of 17.6 sccm was used instead of 35.2 sccm in step 7. FIG. 8 shows that catalyst 33 maintained performance for 204 cycles (x-axis is time, y-axis is C₃H₆ yield and selectivity to C₃H₆, both in carbon mole %).

Listing of Embodiments

[0161] This disclosure further includes the following non-limiting embodiments.

[0162] A1. A process for regenerating an at least partially deactivated catalyst comprising a Group 10 element, an inorganic support, and a contaminant, wherein the Group 10 element has a concentration in the range of from 0.001 wt% to 6 wt%, based on the weight of the inorganic support, and the process comprises: (I) obtaining a precursor catalyst from the at least partially deactivated catalyst; (II) providing an oxidative gas comprising no greater than 5 mol% of H₂O, based on the total moles in the oxidative gas; (III) contacting the precursor catalyst at an oxidizing temperature in a range of from 620°C to 1,000°C with the oxidative gas for a duration of at least 30 seconds, preferably at least 1 minute, preferably at least 5 minutes, to produce an oxidized precursor catalyst; and (IV) obtaining a regenerated catalyst from the oxidized precursor catalyst.

[0163] A2. The process of A1, wherein the Group 10 element comprises Pt, and wherein the inorganic support comprises at least 0.5 wt% of a Group 2 element, based on the weight of the inorganic support.

[0164] A3. The process of A2, wherein: the Group 2 element comprises Mg, and at least a portion of the Group 2 element is in the form of MgO or a mixed oxide comprising MgO.

[0165] A4. The process of any one of A1 to A3, wherein the at least partially deactivated catalyst further comprises up to 10 wt% of a promoter, based on the weight of the inorganic

support, and wherein the promoter comprises one or more of the following elements: Sn, Ag, Cu, a combination thereof, or a mixture thereof.

5 [0166] A5. The process of any one of A1 to A4, wherein the at least partially deactivated catalyst further comprises up to 5 wt% an alkali metal element disposed on the inorganic support, and wherein the alkali metal element comprises at least one of: Li, Na, K, Rb, and Cs.

[0167] A6. The process of any one of A1 to A5, wherein an active component of the regenerated catalyst that is capable of effecting one or more of dehydrogenation, dehydroaromatization, and dehydrocyclization of a hydrocarbon-containing feed comprising one or more of C₂-C₁₆ linear or branched alkanes, or one or more of C₄-C₁₆ cyclic alkanes, one
10 or more C₈-C₁₆ alkyl aromatics, or a mixture thereof comprises the Group 10 element.

[0168] A7. The process of any one of A1 to A6, wherein step (I) comprises: heating the at least partially deactivated catalyst using a heating gas mixture comprising H₂O at a concentration of greater than 5 mol%, based on the total moles in the heating gas mixture to produce the precursor catalyst.

15 [0169] A8. The process of A7, wherein the heating gas mixture is produced by combusting a fuel with an oxidizing gas, and wherein the fuel comprises at least one of H₂, CO, and a hydrocarbon, and the oxidizing gas comprises O₂.

[0170] A9. The process of any one of A1 to A6, wherein, in step (I), the at least partially deactivated catalyst is provided directly as the precursor catalyst.

20 [0171] A10. The process of any one of A1 to A9, wherein step (II) comprises: (IIa) providing the oxidative gas at a temperature below the oxidizing temperature; and (IIb) pre-heating the oxidative gas to a temperature higher than the temperature of the precursor catalyst before the contacting in step (III).

[0172] A11. The process of any one of A1 to A10, further comprising: (V) heating the
25 oxidative gas or the precursor catalyst during step (III) by using a radiant heat source, a heat exchanger, or a combination thereof.

[0173] A12. The process of any one of A1 to A11, wherein step (IV) comprises: (IVa) contacting the oxidized precursor catalyst with a first stripping gas free of O₂ to produce a stripped oxidized precursor catalyst; and (IVb) obtaining the regenerated catalyst from the
30 stripped oxidized precursor catalyst.

[0174] A13. The process of any one of A1 to A12, wherein step (IV) comprises: (IVc) contacting the oxidized precursor catalyst or the stripped oxidized precursor catalyst with a H₂-

containing atmosphere to produce a reduced catalyst; and (IVd) obtaining the regenerated catalyst from the reduced catalyst.

[0175] A14. The process of A13, wherein step (IVd) comprises: (IVd-1) contacting the reduced catalyst with a second stripping gas to produce the regenerated catalyst.

- 5 **[0176]** A15. The process of A13 or A14, wherein step (IVc) is carried out at a temperature of the oxidized precursor catalyst higher than a use temperature of the regenerated catalyst, and step (IVd) further comprises: (IVd-2) cooling the reduced catalyst or the regenerated catalyst to the use temperature in a duration no greater than 10 minutes, no greater than 5 minutes, no greater than 1 minute, no greater than 30 seconds, no greater than 10 seconds, no greater than
10 5 seconds, no greater than 1 second, no greater than 0.5 seconds, no greater than 0.1 seconds, no greater than 0.01 seconds, or no greater than 0.001 seconds.

- [0177]** A16. A dehydrogenation process using the regenerated catalyst produced by a process of any one of A1 to A15, the dehydrogenation process comprising: (VI) contacting a hydrocarbon-containing feed with the regenerated catalyst to effect one or more of
15 dehydrogenation, dehydroaromatization, and dehydrocyclization of at least a portion of the hydrocarbon-containing feed to produce the at least partially deactivated catalyst comprising the Group 10 element, the inorganic support, and the contaminant and an effluent comprising one or more upgraded hydrocarbons and molecular hydrogen, wherein the hydrocarbon feed comprises one or more of C₂-C₁₆ linear or branched alkanes, one or more of C₄-C₁₆ cyclic
20 alkanes, one or more of C₈-C₁₆ alkyl aromatic hydrocarbons, or a mixture thereof; and (VII) repeating steps (I) through (IV), wherein, in step (III) additional oxidized precursor catalyst is produced, and wherein, in step (IV), additional regenerated catalyst is obtained from the additional oxidized precursor catalyst; and (VIII) contacting an additional quantity of the hydrocarbon-containing feed with at least a portion of the additional regenerated catalyst to
25 produce additional at least partially deactivated catalyst and additional effluent.

[0178] A17. The dehydrogenation process of A16, wherein a cycle time from the contacting the hydrocarbon-containing feed with the regenerated catalyst in step (VI) to the contacting the additional quantity of the hydrocarbon-containing feed with the additional regenerated catalyst in step (VIII) is ≤ 5 hours.

- 30 **[0179]** B1. A process for upgrading a hydrocarbon, comprising: (I) contacting a hydrocarbon-containing feed with a catalyst comprising a Group 10 element and an inorganic support to effect one or more of dehydrogenation, dehydroaromatization, and dehydrocyclization of at least a portion of the hydrocarbon-containing feed to produce an at least partially deactivated catalyst comprising the Group 10 element, the inorganic support, and

a contaminant and an effluent comprising one or more upgraded hydrocarbons and molecular hydrogen, wherein: the hydrocarbon-containing feed comprises one or more of C₂-C₁₆ linear or branched alkanes, or one or more of C₄-C₁₆ cyclic alkanes, one or more C₈-C₁₆ alkyl aromatics, or a mixture thereof; the Group 10 element has a concentration in the range of from 5 0.001 wt% to 6 wt%, based on the weight of the inorganic support; the hydrocarbon-containing feed and the catalyst are contacted at a temperature in a range from 300°C to 900°C; and the one or more upgraded hydrocarbons comprise at least one of a dehydrogenated hydrocarbon, a dehydroaromatized hydrocarbon, and a dehydrocyclized hydrocarbon; (II) obtaining a precursor catalyst from the at least partially deactivated catalyst; (III) providing an oxidative 10 gas comprising no greater than 2 mol% of H₂O, based on the total moles in the oxidative gas; (IV) contacting the precursor catalyst at an oxidizing temperature in a range of from 620°C to 1,000°C with the oxidative gas for a duration of at least 30 seconds, preferably at least 1 minute, preferably at least 5 minutes, to produce an oxidized precursor catalyst; (V) obtaining a regenerated catalyst from the oxidized precursor catalyst; and (VI) contacting an additional 15 quantity of the hydrocarbon-containing feed with at least a portion of the regenerated catalyst to produce additional at least partially deactivated catalyst and additional effluent.

[0180] B2. The process of B1, wherein: the Group 10 element comprises Pt, the inorganic support comprises at least 0.5 wt% of a Group 2 element, based on the weight of the inorganic support, the catalyst optionally further comprises up to 10 wt% of a promoter, based on the 20 weight of the inorganic support, wherein the promoter, if present, comprises one or more of the following elements: Sn, Ag, Cu, a combination thereof, or a mixture thereof, the catalyst optionally further comprises up to 5 wt% an alkali metal element, and the alkali metal element, if present, comprises at least one of: Li, Na, K, Rb, and Cs.

[0181] B3. The process of B1 or B2, wherein step (II) comprises: heating the at least partially 25 deactivated catalyst using a heating gas mixture comprising H₂O at a concentration of greater than 5 mol%, based on the total moles in the heating gas mixture to produce the precursor catalyst.

[0182] B4. The process of B3, wherein the heating gas mixture is produced by combusting a fuel with an oxidizing gas, and wherein the fuel comprises at least one of H₂, CO, and a 30 hydrocarbon, and the oxidizing gas comprises O₂.

[0183] B5. The process of any one of B1 to B4, wherein, in step (II), the at least partially deactivated catalyst is provided directly as the precursor catalyst.

[0184] B6. The process of any one of B1 to B5, wherein step (III) comprises: (IIIa) providing the oxidative gas at a temperature below the oxidizing temperature; and (IIIb) pre-heating the

oxidative gas to a temperature higher than the temperature of the precursor catalyst before the contacting in step (IV).

5 **[0185]** B7. The process of any one of B1 to B6, further comprising:(VI) heating the oxidative gas or the precursor catalyst during step (IV) by using a radiant heat source, a heat exchanger, or a combination thereof.

[0186] B8. The process of any one of B1 to B7, wherein a cycle time from the contacting the hydrocarbon-containing feed with the catalyst in step (I) to the contacting the additional quantity of the hydrocarbon-containing feed with the regenerated catalyst in step (VI) is ≤ 5 hours.

10 **[0187]** Various terms have been defined above. To the extent a term used in a claim is not defined above, it should be given the broadest definition persons in the pertinent art have given that term as reflected in at least one printed publication or issued patent. Furthermore, all patents, test procedures, and other documents cited in this application are fully incorporated by reference to the extent such disclosure is not inconsistent with this application and for all
15 jurisdictions in which such incorporation is permitted.

While the foregoing is directed to embodiments of the present invention, other and further embodiments of the invention may be devised without departing from the basic scope thereof, and the scope thereof is determined by the claims that follow.

CLAIMS:

What is claimed is:

1. A process for regenerating an at least partially deactivated catalyst comprising a Group
5 10 element, an inorganic support, and a contaminant, wherein the Group 10 element has a concentration in the range of from 0.001 wt% to 6 wt%, based on the weight of the inorganic support, and the process comprises:
 - (I) heating the at least partially deactivated catalyst using a heating gas mixture comprising H₂O at a concentration of greater than 5 mol%, based on the total moles in the
10 heating gas mixture to produce a precursor catalyst;
 - (II) providing an oxidative gas comprising no greater than 5 mol% of H₂O, based on the total moles in the oxidative gas;
 - (III) contacting the precursor catalyst at an oxidizing temperature in a range of from 620°C to 1,000°C with the oxidative gas for a duration of at least 30 seconds, preferably at
15 least 1 minute, preferably at least 5 minutes, to produce an oxidized precursor catalyst; and
 - (IV) obtaining a regenerated catalyst from the oxidized precursor catalyst.
2. The process of claim 1, wherein the heating gas mixture is produced by combusting a
20 fuel with an oxidizing gas.
3. The process of claim 2, wherein the fuel comprises at least one of H₂, CO, and a hydrocarbon, and the oxidizing gas comprises O₂.
4. The process of any one of claims 1 to 3, wherein the Group 10 element comprises Pt,
25 and wherein the inorganic support comprises at least 0.5 wt% of a Group 2 element, based on the weight of the inorganic support.
5. The process of claim 4, wherein:
the Group 2 element comprises Mg, and
30 at least a portion of the Group 2 element is in the form of MgO or a mixed oxide comprising MgO.
6. The process of any one of the preceding claims, wherein the at least partially deactivated catalyst further comprises up to 10 wt% of a promoter, based on the weight of the

inorganic support, and wherein the promoter comprises one or more of the following elements: Sn, Ag, Cu, a combination thereof, or a mixture thereof.

7. The process of any one of the preceding claims, wherein the at least partially deactivated catalyst further comprises up to 5 wt% an alkali metal element disposed on the inorganic support, and wherein the alkali metal element comprises at least one of: Li, Na, K, Rb, and Cs.

8. The process of any one of the preceding claims, wherein an active component of the regenerated catalyst that is capable of effecting one or more of dehydrogenation, dehydroaromatization, and dehydrocyclization of a hydrocarbon-containing feed comprising one or more of C₂-C₁₆ linear or branched alkanes, or one or more of C₄-C₁₆ cyclic alkanes, one or more C₈-C₁₆ alkyl aromatics, or a mixture thereof comprises the Group 10 element.

9. The process of any one of the preceding claims, wherein step (II) comprises: (IIa) providing the oxidative gas at a temperature below the oxidizing temperature; and (IIb) pre-heating the oxidative gas to a temperature higher than the temperature of the precursor catalyst before the contacting in step (III).

10. The process of any one of the preceding claims, further comprising: (V) heating the oxidative gas, the precursor catalyst, or both, during step (III) by using a radiant/conductive heat source, a heat exchanger, or a combination thereof.

11. The process of any one of the preceding claims, wherein step (IV) comprises: (IVa) contacting the oxidized precursor catalyst with a first stripping gas free of O₂ to produce a stripped oxidized precursor catalyst; and (IVb) obtaining the regenerated catalyst from the stripped oxidized precursor catalyst.

12. The process of any one of the preceding claims, wherein step (IV) comprises: (IVc) contacting the oxidized precursor catalyst or the stripped oxidized precursor catalyst with a H₂-containing atmosphere to produce a reduced catalyst; and (IVd) obtaining the regenerated catalyst from the reduced catalyst.

13. The process of claim 12, wherein step (IVd) comprises:

(IVd-1) contacting the reduced catalyst with a second stripping gas to produce the regenerated catalyst.

14. The process of claim 12 or claim 13, wherein step (IVc) is carried out at a temperature of the oxidized precursor catalyst higher than a use temperature of the regenerated catalyst, and step (IVd) further comprises:

(IVd-2) cooling the reduced catalyst or the regenerated catalyst to the use temperature in a duration no greater than 10 minutes, no greater than 5 minutes, no greater than 1 minute, no greater than 30 seconds, no greater than 10 seconds, no greater than 5 seconds, no greater than 1 second, no greater than 0.5 seconds, no greater than 0.1 seconds, no greater than 0.01 seconds, or no greater than 0.001 seconds.

15. A dehydrogenation process using the regenerated catalyst produced by a process of any one of the preceding claims, the dehydrogenation process comprising:

(VI) contacting a hydrocarbon-containing feed with the regenerated catalyst to effect one or more of dehydrogenation, dehydroaromatization, and dehydrocyclization of at least a portion of the hydrocarbon-containing feed to produce the at least partially deactivated catalyst comprising the Group 10 element, the inorganic support, and the contaminant and an effluent comprising one or more upgraded hydrocarbons and molecular hydrogen, wherein the hydrocarbon feed comprises one or more of C₂-C₁₆ linear or branched alkanes, one or more of C₄-C₁₆ cyclic alkanes, one or more of C₈-C₁₆ alkyl aromatic hydrocarbons, or a mixture thereof; and

(VII) repeating steps (I) through (IV), wherein, in step (III) additional oxidized precursor catalyst is produced, and wherein, in step (IV), additional regenerated catalyst is obtained from the additional oxidized precursor catalyst; and

(VIII) contacting an additional quantity of the hydrocarbon-containing feed with at least a portion of the additional regenerated catalyst to produce additional at least partially deactivated catalyst and additional effluent.

16. The dehydrogenation process of claim 15, wherein a cycle time from the contacting the hydrocarbon-containing feed with the regenerated catalyst in step (VI) to the contacting the additional quantity of the hydrocarbon-containing feed with the additional regenerated catalyst in step (VIII) is ≤ 5 hours.

17. A process for upgrading a hydrocarbon, comprising:

(I) contacting a hydrocarbon-containing feed with a catalyst comprising a Group 10 element and an inorganic support to effect one or more of dehydrogenation, dehydroaromatization, and dehydrocyclization of at least a portion of the hydrocarbon-containing feed to produce an at least partially deactivated catalyst comprising the Group 10 element, the inorganic support, and a contaminant and an effluent comprising one or more upgraded hydrocarbons and molecular hydrogen, wherein:

the hydrocarbon-containing feed comprises one or more of C₂-C₁₆ linear or branched alkanes, or one or more of C₄-C₁₆ cyclic alkanes, one or more C₈-C₁₆ alkyl aromatics, or a mixture thereof;

the Group 10 element has a concentration in the range of from 0.001 wt% to 6 wt%, based on the weight of the inorganic support;

the hydrocarbon-containing feed and the catalyst are contacted at a temperature in a range from 300°C to 900°C; and

the one or more upgraded hydrocarbons comprise at least one of a dehydrogenated hydrocarbon, a dehydroaromatized hydrocarbon, and a dehydrocyclized hydrocarbon;

(II) heating the at least partially deactivated catalyst using a heating gas mixture comprising H₂O at a concentration of greater than 5 mol%, based on the total moles in the heating gas mixture to produce a precursor catalyst;

(III) providing an oxidative gas comprising no greater than 2 mol% of H₂O, based on the total moles in the oxidative gas;

(IV) contacting the precursor catalyst at an oxidizing temperature in a range of from 620°C to 1,000°C with the oxidative gas for a duration of at least 30 seconds, preferably at least 1 minute, preferably at least 5 minutes, to produce an oxidized precursor catalyst;

(V) obtaining a regenerated catalyst from the oxidized precursor catalyst; and

(VI) contacting an additional quantity of the hydrocarbon-containing feed with at least a portion of the regenerated catalyst to produce additional at least partially deactivated catalyst and additional effluent.

18. The process of claim 17, wherein the heating gas mixture is produced by combusting a fuel with an oxidizing gas.

19. The process of claim 18, wherein the fuel comprises at least one of H₂, CO, and a hydrocarbon, and the oxidizing gas comprises O₂.
20. The process of any one of claims 17 to 19, wherein:
5 the Group 10 element comprises Pt, and
the inorganic support comprises at least 0.5 wt% of a Group 2 element, based on the weight of the inorganic support.
21. The process of any one of claims 17 to 20, wherein the catalyst further comprises up to
10 10 wt% of a promoter, based on the weight of the inorganic support, and wherein the promoter comprises one or more of the following elements: Sn, Ag, Cu, a combination thereof, or a mixture thereof.
22. The process of any one of claims 17 to 21, wherein the catalyst further comprises up to
15 5 wt% an alkali metal element, and wherein the alkali metal element comprises at least one of: Li, Na, K, Rb, and Cs.
23. The process of any one of claims 17 to 22, wherein step (III) comprises:
(IIIa) providing the oxidative gas at a temperature below the oxidizing temperature; and
20 (IIIb) pre-heating the oxidative gas to a temperature higher than the temperature of the precursor catalyst before the contacting in step (IV).
24. The process of any one of claims 17 to 23, further comprising:
(VII) heating the oxidative gas, the precursor catalyst, or both, during step (IV) by using
25 a radiant/conductive heat source, a heat exchanger, or a combination thereof.
25. The process of any one of claims 17 to 24, wherein a cycle time from the contacting the hydrocarbon-containing feed with the catalyst in step (I) to the contacting the additional quantity of the hydrocarbon-containing feed with the regenerated catalyst in step (VI) is ≤ 5
30 hours.