

"INTEGRATED PROCESS FOR AROMATICS PRODUCTION"

BACKGROUND OF THE INVENTION

[0001] This invention relates to a process for an aromatics complex which is a combination of process units that can be used to convert naphtha into basic
5 petrochemical intermediates of benzene, toluene, and xylene. Based on a metal catalyzed transalkylation process that handles unextracted toluene and heavier aromatics and an olefin saturation process, the improved flow scheme removes items of equipment and processing steps, such as a reformate splitter column and a heavy aromatics column, resulting in significant economic benefits when producing para-
10 xylene.

[0002] Most new aromatics complexes are designed to maximize the yield of benzene and para-xylene. Benzene is a versatile petrochemical building block used in many different products based on its derivation including ethylbenzene, cumene, and cyclohexane. Para-xylene is also an important building block, which is used almost
15 exclusively for the production of polyester fibers, resins, and films formed via terephthalic acid or dimethyl terephthalate intermediates. Accordingly, an aromatics complex may be configured in many different ways depending on the desired products, available feedstocks, and investment capital available. A wide range of options permits flexibility in varying the product slate balance of benzene and para-
20 xylene to meet downstream processing requirements.

[0003] A prior art aromatics complex flow scheme has been disclosed by Meyers in the *Handbook of Petroleum Refining Processes*, 2d. Edition in 1997 by McGraw-Hill.

[0004] US-A-3,996,305 to Berger discloses a fractionation scheme primarily directed to transalkylation of toluene and C₉ alkylaromatics in order to produce benzene and xylene. The transalkylation process is also combined with an aromatics extraction process. The fractionation scheme includes a single column with two
5 streams entering and with three streams exiting the column for integrated economic benefits.

[0005] US-A-4,341,914 to Berger discloses a transalkylation process with recycle of C₁₀ alkylaromatics in order to increase yield of xylenes from the process. The transalkylation process is also preferably integrated with a para-xylene separation
10 zone and a xylene isomerization zone operated as a continuous loop receiving mixed xylenes from the transalkylation zone feedstock and effluent fractionation zones.

[0006] US-A-4,642,406 to Schmidt discloses a high severity process for xylene production that employs a transalkylation zone that simultaneously performs as an isomerization zone over a nonmetal catalyst. High quality benzene is produced along
15 with a mixture of xylenes, which allows para-xylene to be separated by absorptive separation from the mixture with the isomer-depleted stream being passed back to the transalkylation zone.

[0007] US-A-5,417,844 to Boitiaux et al. discloses a process for the selective dehydrogenation of olefins in steam cracking petrol in the presence of a nickel
20 catalyst and is characterized in that prior to the use of the catalyst, a sulfur-containing organic compound is incorporated into the catalyst outside of the reactor prior to use.

[0008] US-A-5,658,453 to Russ et al. discloses an integrated reforming and olefin saturation process. The olefin saturation reaction uses a mixed vapor phase with

addition of hydrogen gas to a reformat liquid in contact with a refractory inorganic oxide containing preferably a platinum-group metal and optionally a metal modifier.

[0009] US-A-5,763,720 to Buchanan et al. discloses a transalkylation process for producing benzene and xylenes by contacting a C₉⁺ alkylaromatics with benzene
5 and/or toluene over a catalyst comprising a zeolite such as ZSM-12 and a hydrogenation noble metal such as platinum. Sulfur or stream is used to treat the catalyst.

[0010] US-A-5,847,256 to Ichioka et al. discloses a process for producing xylene from a feedstock containing C₉ alkylaromatics with the aid of a catalyst with a zeolite
10 that is preferably mordenite and with a metal that is preferably rhenium.

SUMMARY OF THE INVENTION

[0011] An aromatics complex flow scheme with an enabled transalkylation process requires stabilization of a transalkylation catalyst through the introduction of a metal function. Enabling a transalkylation process to handle both C₁₀
15 alkylaromatics and unextracted toluene permits the following flow scheme improvements to be realized. By using toluene without first passing it to an extraction unit, the flow scheme omits a reformat-splitter column. The concomitantly smaller capacity extraction unit is moved to the overhead of a benzene column. By only extracting benzene, simple extractive distillation is used, since a more expensive
20 combined liquid-liquid extraction method is only required for heavier contaminants. By using both C₉ and C₁₀ alkylaromatics in an enabled transalkylation unit, the flow scheme further omits a heavy aromatics column. A further enhancement to the flow scheme is accomplished through the elimination of clay treaters in favor of selective

olefin saturation at the exits of a reforming process unit or of an alkylaromatic isomerization process unit.

[0012] Another embodiment of the present invention comprises an apparatus that is based on the process steps, which efficiently converts naphtha into para-xylene.

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BRIEF DESCRIPTION OF THE DRAWING

[0013] The Figure shows an aromatics complex flow scheme of the present invention, which includes olefin saturation and a metal stabilized transalkylation catalyst.

DETAILED DESCRIPTION OF THE INVENTION

10 [0014] Feed to the complex may be naphtha, but can also be pygas, imported mixed xylene, or imported toluene. Naphtha fed to an aromatics complex is first hydrotreated to remove sulfur and nitrogen compounds to less than 0.5 wt-ppm before passing the treated naphtha on to a reforming unit 13. Naphtha hydrotreating occurs by contacting naphtha in a line 10 with a naphtha hydrotreating catalyst under naphtha
15 hydrotreating conditions in a unit 11. The naphtha hydrotreating catalyst is typically composed of a first component of cobalt oxide or nickel oxide, along with a second component of molybdenum oxide or tungsten oxide, and a third component inorganic oxide support, which is typically a high purity alumina. Generally good results are achieved when the cobalt oxide or nickel oxide component is in the range of 1 to 5
20 wt-% and the molybdenum oxide component is in the range of 6 to 25 wt-%. The alumina (or aluminum oxide) is set to balance the composition of the naphtha hydrotreating catalyst to sum all components up to 100 wt-%. One hydrotreating

catalyst for use in the present invention is disclosed in US-A-5,723,710, the teachings of which are incorporated herein by reference. Typical hydrotreating conditions include a liquid hourly space velocity (LHSV) from 1.0 to 5.0 hr⁻¹, a ratio of hydrogen to hydrocarbon (or naphtha feedstock) from 50 to 135 Nm³/m³, and a
5 pressure from 10 to 35 kg/cm².

[0015] In the reforming unit 13, paraffins and naphthenes are converted to aromatics. This is the only unit in the complex that actually creates aromatic rings. The other units in the complex separate the various aromatic components into individual products and convert various aromatic species into higher-value products.
10 The reforming unit 13 is usually designed to run at very high severity, equivalent to producing 100 to 106 Research Octane Number (RONC) gasoline reformate, in order to maximize the production of aromatics. This high severity operation also extinguishes virtually all non-aromatic impurities in the C₈⁺ fraction of reformate, and eliminates the need for extraction of the C₈ and C₉ aromatics.

15 [0016] In the reforming unit 13, hydrotreated naphtha from a line 12 is contacted with a reforming catalyst under reforming conditions. The reforming catalyst is typically composed of a first component platinum-group metal, a second component modifier metal, and a third component inorganic-oxide support, which is typically high purity alumina. Generally good results are achieved when the platinum-group
20 metal is in the range of 0.01 to 2.0 wt-% and the modifier metal component is in the range of 0.01 to 5 wt-%. The alumina is set to balance the composition of the naphtha hydrotreating catalyst to sum all components up to 100 wt-%. The platinum-group metal is selected from platinum, palladium, rhodium, ruthenium, osmium, and iridium. The preferred platinum-group metal component is platinum. The metal

modifiers may include rhenium, tin, germanium, lead, cobalt, nickel, indium, gallium, zinc, uranium, dysprosium, thallium, and mixtures thereof. One reforming catalyst for use in the present invention is disclosed in US-A-5,665,223, the teachings of which are incorporated herein by reference. Typical reforming conditions include a liquid
5 hourly space velocity from 1.0 to 5.0 hr⁻¹, a ratio of hydrogen to hydrocarbon from 1 to 10 moles of hydrogen per mole of hydrocarbon feed entering the reforming zone, and a pressure from 2.5 to 35 kg/cm². Hydrogen produced in the reforming unit 13 exits in a line 14.

[0017] The reformate product from the reforming unit 13 in a line 15 is sent to a
10 debutanizer zone 53, which typically comprises a debutanizer column 20 that strips off the light end hydrocarbons (butanes and lighter) in a line 21. The debutanizer zone 53 may also comprise at least one olefin saturation zone 16, which may be placed upstream or downstream from the debutanizer column 20. Moreover, streams from other units in the aromatics complex may also be sent to the debutanizer column 20
15 for stripping. These other units include the transalkylation zone, which sends a transalkylation stripper-overhead stream in a line 17, and the isomerization zone, which sends a deheptanizer overhead stream in a line 18. Both of these units are described in greater detail below.

[0018] The olefin saturation zone 16 may consist of the well-known clay treating
20 means or other means to treat residual olefin contaminants. Preferably, the olefin saturation zone 16 comprises an olefin saturation catalyst operating under olefin saturation conditions. Catalytic olefin saturation is also a flow scheme enabler. If a catalytic process is used, the olefins are converted to useful products. For example, a C₆ olefin can be converted to benzene (if the olefin is cyclic), a C₇ to toluene, and a

C₈ to xylene. If clay is used, then the olefin will be polymerized, often to C₁₁+, which is not very useful for an aromatics complex. Thus, the catalytic olefin saturation helps improve the economics of the flow scheme. Suitable olefin saturation catalysts in the present invention contain elemental nickel or a platinum-group component preferably supported on an inorganic oxide support, which is typically alumina. In the case where the elemental nickel is present on a support, the nickel is preferably present in an amount from 2 to 40 wt-% of the total catalyst weight. One catalyst for use in the present invention is disclosed in US-A-5,658,453, the teachings of which are incorporated herein by reference. Typical olefin saturation conditions include a temperature from 20° to 200°C, a pressure from 5 to 70 kg/cm² and a stoichiometric ratio of hydrogen to olefins from 1:1 to 5:1. Olefin treated reformate is shown in a line 19.

[0019] The debutanized reformate comprising aromatics in a line 22 is combined with a transalkylation stripper-bottoms stream in a line 24 and sent to a benzene-toluene (BT) fractionation zone 54 via a line 23. The BT fractionation zone 54 generally comprises at least one column, and usually comprises a benzene column 25 and a toluene column 31. However, the benzene column 25 may be eliminated in favor of a transalkylation stripper column, with a stabilizer section sufficient to produce a suitable benzene stream. The BT fractionation zone 54 produces a benzene-enriched stream in a line 26, a toluene-enriched stream in a line 32, and a xylene-plus-enriched stream in a line 33. Typically, the benzene-enriched stream in line 26 is produced from the overhead of the benzene column 25, with the bottom of the benzene column 25 being sent via a line 30 to feed the toluene column 31. The toluene-enriched stream in line 32 is produced from the overhead of the toluene

column 31 and sent to a transalkylation unit 36, with the bottom of the toluene column 31 producing the xylene-plus-enriched stream in line 33. The xylene-plus-enriched stream in line 33 from the bottom of the toluene column 31 is sent to a xylene recovery section 55 of the aromatics complex described below.

5 [0020] The benzene-enriched stream in line 26 is sent to an extractive distillation zone 27 which produces a high purity benzene product stream in a line 29 and rejects a by-product raffinate stream in a line 28. The raffinate stream may be blended into gasoline, used as feedstock for an ethylene plant, or converted into additional benzene by recycling to the reforming unit 13. The use of extractive distillation instead of
10 liquid-liquid extraction or combined liquid-liquid extraction/extractive distillation processes results in an economic improvement.

[0021] Extractive distillation is a technique for separating mixtures of components having nearly equal volatility and having nearly the same boiling point. It is difficult to separate the components of such mixtures by conventional fractional distillation. In
15 extractive distillation, a solvent is introduced into a main distillation column above the entry point of the hydrocarbon-containing fluid mixture that is to be separated. The solvent affects the volatility of the hydrocarbon-containing fluid component boiling at a higher temperature differently than the hydrocarbon-containing fluid component boiling at a lower temperature sufficiently to facilitate the separation of
20 the various hydrocarbon-containing fluid components by distillation and such solvent exits with the bottoms fraction. Suitable solvents include tetrahydrothiophene 1,1-dioxide (or sulfolane), diethylene glycol, triethylene glycol, or tetraethylene glycol. The raffinate stream in line 28 comprising nonaromatic compounds exits overhead of the main distillation column, while the bottoms fraction containing solvent and

benzene exits below. Often the raffinate will be sent to a wash column in order to be contacted with water and thus remove any residual dissolved solvent. The bottoms stream from the main distillation column is sent to a solvent recovery column, where benzene is recovered overhead and the solvent is recovered from the bottom and
5 passed back to the main distillation column. The recovery of high purity benzene in the line 29 from extractive-distillation typically exceeds 99 wt-%.

[0022] The toluene-enriched stream in line 32 is usually blended with a stream in a line 41 rich in C₉ and C₁₀ alkylaromatics produced by a xylene column 39 and charged via a line 34 to the transalkylation unit 36 for production of additional
10 xylenes and benzene. In the transalkylation unit 36, the feed is contacted with a transalkylation catalyst under transalkylation conditions. The preferred catalyst is a metal stabilized transalkylation catalyst. Such catalyst comprises a zeolite component, a metal component, and an inorganic oxide component. The zeolite component typically is either a pentasil zeolite, which include the structures of MFI, MBL, MTW,
15 MTT and FER (IUPAC Commission on Zeolite Nomenclature), a beta zeolite, or a mordenite. Preferably it is mordenite zeolite. The metal component typically is a noble metal or base metal. The noble metal is a platinum-group metal is selected from platinum, palladium, rhodium, ruthenium, osmium, and iridium. The base metal is selected from the group consisting of rhenium, tin, germanium, lead, cobalt, nickel,
20 indium, gallium, zinc, uranium, dysprosium, thallium, and mixtures thereof. The base metal may be combined with another base metal, or with a noble metal. Preferably the metal component comprises rhenium. Suitable metal amounts in the transalkylation catalyst range from 0.01 to 10 wt-%, with the range from 0.1 to 3 wt-% being preferred, and the range from 0.1 to 1 wt-% being highly preferred. Suitable zeolite

amounts in the catalyst range from 1 to 99 wt-%, preferably between 10 to 90 wt-%, and more preferably between 25 to 75 wt-%. The balance of the catalyst is composed of inorganic oxide binder, preferably alumina. One transalkylation catalyst for use in the present invention is disclosed in US-A-5,847,256, the teachings of which are
5 incorporated herein by reference

[0023] Conditions employed in the transalkylation zone normally include a temperature of from 200° to 540°C. The transalkylation zone is operated at moderately elevated pressures broadly ranging from 1 to 60 kg/cm². The transalkylation reaction can be effected over a wide range of space velocities, with
10 higher space velocities effecting a higher ratio of para-xylene at the expense of conversion. Liquid hourly space velocity generally is in the range of from 0.1 to 20 hr⁻¹. The feedstock is preferably transalkylated in the vapor phase and in the presence of hydrogen supplied via a line 35. If transalkylated in the liquid phase, then the presence of hydrogen is optional. If present, free hydrogen is associated with the
15 feedstock and recycled hydrocarbons in an amount of 0.1 moles per mole of alkylaromatics up to 10 moles per mole of alkylaromatic. This ratio of hydrogen to alkylaromatic is also referred to as hydrogen to hydrocarbon ratio.

[0024] The effluent from the transalkylation zone 36 is sent to a transalkylation unit stripper column 52 to remove light ends, then sent to the BT fractionation zone
20 54 through the lines 24 and 23. There the benzene product is recovered, and the xylenes are fractionated out and sent to the xylene recovery section 55 via the xylene plus enriched stream in line 33. The overhead material from the transalkylation stripper column 52 is normally recycled back via the line 17 to the reforming unit debutanizer for recovery of residual benzene. Alternatively, a stabilizer section or

column is placed on or after the transalkylation unit stripper column 52. This transalkylation stabilizer section can produce a benzene-enriched stream suitable for extractive distillation, and eliminate the need for a separate benzene column in the BT fractionation section. Such a stabilizer or stripper column from the transalkylation unit is thus encompassed in the definition of the BT fractionation zone 54 when the separate benzene column is eliminated. The transalkylation unit stripper column 52 can also accept treated product from the olefin saturation zone or overhead from the alkylaromatic isomerization deheptanizer column that would normally be recycled back to the reforming unit debutanizer column 20.

10 [0025] As noted above, the xylene-plus-enriched stream in line 33 from the bottom of the toluene column 31 is sent to the xylene recovery section 55 of the aromatics complex. This section of the aromatics complex comprises at least one xylene column 39, and generally will further include a process unit for separation of at least one xylene isomer, which is typically the para-xylene product from the aromatics complex. Preferably such a para-xylene separation zone 43 is operated in conjunction with an isomerization unit 51 for isomerization of the remaining alkylaromatic compounds back to an equilibrium or near equilibrium mixture containing para-xylene, which can be recycled around again for further recovery in a loop-wise fashion. Accordingly, the xylene-plus-enriched stream in line 33, which may be blended with a recycle stream in a line 38 to form a stream in a line 37, is charged to a xylene column 39. The xylene column 39 is designed to rerun a feed stream in a line 40 to the para-xylene separation zone 43 down to very low levels of C₉ alkylaromatics (A₉) concentration. A₉ compounds may build up in a desorbent circulation loop within the para-xylene separation zone 43, so it is more efficient to

remove this material upstream in the xylene column 39. The overhead feed stream in the line 40 from the xylene column 39 is charged directly to the para-xylene separation zone 43.

[0026] Material from the lower part of the xylene column 39 is withdrawn as a stream rich in C₉ and C₁₀ alkylaromatics via the line 41, which is then sent to the transalkylation zone 36 for production of additional xylenes and benzene. The stream in line 41 taken as a sidecut stream on the xylene column (which eliminates a heavy aromatics column) is really enabled by the metal stabilized transalkylation catalyst. A separate column doing a rigorous split to keep coke precursors such as methyl indan or naphthalene out of the stream is no longer needed because the metal stabilized transalkylation catalyst can handle them. Any remaining C₁₁⁺ material is rejected from the bottom of the xylene column 39 via a line 42. Another embodiment is to just send the whole xylene column bottoms stream to the transalkylation unit instead of the sidecut stream.

[0027] Alternatively, if ortho-xylene is to be produced in the complex, the xylene column is designed to make a split between meta and ortho-xylene and drop a targeted amount of orthoxylene to the bottoms. The xylene column **bottoms** is then sent to an ortho-xylene column (not shown) where high purity ortho-xylene product is recovered overhead. Material from the lower part of the ortho-xylene column is withdrawn as a stream rich in C₉ and C₁₀ alkylaromatics then sent to the transalkylation unit. Any remaining C₁₁⁺ material is rejected from the bottom of the ortho-xylene column.

[0028] The para-xylene separation zone 43 may be based on a fractional crystallization process or an adsorptive separation process, both of which are well known in the art, and preferably is based on the adsorptive separation process. Such

adsorptive separation can recover over 99 wt-% pure para-xylene in a line 44 at high recovery per pass. Any residual toluene in the feed to the separation unit is extracted along with the para-xylene, fractionated out in a finishing column within the unit, and then optionally recycled to the transalkylation unit stripper column 52. Thus, the
5 raffinate from the para-xylene separation zone 43 is almost entirely depleted of para-xylene, to a level usually of less than 1 wt-%. The raffinate is sent via a line 45 to the alkylaromatics isomerization unit 51, where additional para-xylene is produced by reestablishing an equilibrium or near-equilibrium distribution of xylene isomers. Any ethylbenzene in the para-xylene separation unit raffinate is either converted to
10 additional xylenes or converted to benzene by dealkylation, depending upon the type of isomerization catalyst used.

[0029] In the alkylaromatic isomerization unit 51, the raffinate stream in line 45 is contacted with an isomerization catalyst under isomerization conditions. The isomerization catalyst is typically composed of a molecular sieve component, a metal
15 component, and an inorganic oxide component. Selection of the molecular sieve component allows control over the catalyst performance between ethylbenzene isomerization and ethylbenzene dealkylation depending on overall demand for benzene. Consequently, the molecular sieve may be either a zeolitic aluminosilicate or a non-zeolitic molecular sieve. The zeolitic aluminosilicate (or zeolite) component
20 typically is either a pentasil zeolite, which include the structures of MFI, MEL, MTW, MTT and FER (IUPAC Commission on Zeolite Nomenclature), a beta zeolite, or a mordenite. The non-zeolitic molecular sieve is typically one or more of the AEL framework types, especially SAPO-11, or one or more of the ATO framework types, especially MAPSO-31, according to the "Atlas of Zeolite Structure Types"

(Butterworth-Heinemann, Boston, Mass., 3rd ed. 1992). The metal component typically is a noble metal component, and may include an optional base metal modifier component in addition to the noble metal or in place of the noble metal. The noble metal is a platinum-group metal is selected from platinum, palladium, rhodium, ruthenium, osmium, and iridium. The base metal is selected from the group consisting of rhenium, tin, germanium, lead, cobalt, nickel, indium, gallium, zinc, uranium, dysprosium, thallium, and mixtures thereof. The base metal may be combined with another base metal, or with a noble metal. Suitable total metal amounts in the isomerization catalyst range from 0.01 to 10 wt-%, with the range from 0.1 to 3 wt-% preferred. Suitable zeolite amounts in the catalyst range from 1 to 99 wt-%, preferably between 10 to 90 wt-%, and more preferably between 25 to 75 wt-%. The balance of the catalyst is composed of inorganic oxide binder, typically alumina. One isomerization catalyst for use in the present invention is disclosed in US-A-4,899,012, the teachings of which are incorporated herein by reference.

15 [0030] Typical isomerization conditions include a temperature in the range from 0° to 600°C and a pressure from atmospheric to 50 kg/cm³. The liquid hourly hydrocarbon space velocity of the feedstock relative to the volume of catalyst is from 0.1 to 30 hr⁻¹. The hydrocarbon contacts the catalyst in admixture with a gaseous hydrogen-containing stream in a line 46 at a hydrogen-to-hydrocarbon mole ratio of from 0.5:1 to 15:1 or more, and preferably a ratio of from 0.5 to 10. If liquid phase conditions are used for isomerization, then no hydrogen is added to the unit.

20 [0031] The effluent from the isomerization unit 51 is sent via a line 47 to a deheptanizer column 48. A bottoms stream in a line 49 from the deheptanizer 48 is treated to remove olefins, if necessary, in an olefin saturation unit 50 with the olefin

saturation methods described above. An alternative is to put the olefin saturation unit 50 after the isomerization unit 51 and use the deheptanizer column 48 to remove residual hydrogen. If the catalyst used in the isomerization unit 51 is the ethylbenzene dealkylation type, then olefin saturation may not be required at all.

5 [0032] The deheptanizer bottoms stream in line 49, after olefin treatment, is then recycled back to the xylene column 39 via the line 38. In this way, all the C₈ aromatics are continually recycled within the xylenes recovery section of the complex until they exit the aromatics complex as para-xylene, benzene, or optionally ortho-xylene. The overhead from the deheptanizer is normally recycled back via the line 18
10 to the reforming unit debutanizer for recovery of residual benzene. Alternatively, the overhead liquid is recycled back to the transalkylation unit stripper column 52.

[0033] Accordingly, the aromatics complex of the present invention displays excellent economic benefits. These improvements result in an aromatics complex with savings on inside battery limits curve costs and an improvement on the return on
15 investment in such a complex.

WHAT IS CLAIMED IS:

1. A process for producing benzene and para-xylene from a catalytic reformat stream comprising the steps of:
 - (a) optionally providing the reformat stream to a first olefin saturation zone,
5 wherein said stream is contacted with an olefin saturation catalyst under olefin saturation conditions to produce a reformat stream treated for olefins;
 - (b) combining the reformat stream with at least a portion of a transalkylation product stream to form a blended stream;
 - 10 (c) separating the blended stream of step (b) in a benzene-toluene fractionation zone to produce a benzene-enriched stream, a toluene-enriched stream, and a xylene-plus-enriched stream;
 - (d) passing the benzene-enriched stream to an extractive-distillation zone to produce a raffinate stream and a benzene product which is recovered as a
15 product stream from said process;
 - (e) combining the toluene-enriched stream with a stream from a xylene column rich in C₉ and C₁₀ alkylaromatics;
 - (f) passing the combined streams of step (e) to a transalkylation zone, wherein said streams are contacted with a metal-stabilized transalkylation catalyst
20 under transalkylation conditions to produce the transalkylation product stream of step (b);
 - (g) separating the xylene-plus-enriched stream of step (c) in a xylene fractionation zone to produce an overhead xylene stream and the stream from a xylene column rich in C₉ and C₁₀ alkylaromatics of step (e); and

(h) passing the overhead xylene stream to a para-xylene separation zone, wherein para-xylene is concentrated into a para-xylene enriched product stream which is recovered as a product stream of said process.

2. The process of claim 1 wherein the reformate stream is produced from a naphtha feedstock by:

providing a naphtha feedstock stream to a hydrotreating zone, that contacts the feedstock with a hydrotreating catalyst under hydrotreating conditions to produce a hydrotreated naphtha stream; the hydrotreated naphtha stream is passed to a reforming zone and contacted with a reforming catalyst under reforming conditions to produce a catalytic reformate stream which comprises aromatic components and light-end hydrocarbons; the catalytic reformate stream is combined with a transalkylation stripper-overhead-liquid stream and the combined streams are separated in a debutanizer fractionation zone that produces a light-end hydrocarbon product stream which is removed from the process and a debutanized aromatic stream as the reformate stream; a transalkylation stripper-bottoms stream provides the at least a portion of the transalkylation zone product stream in the blended stream; the transalkylation product stream is separated in a transalkylation stripper fractionation zone to produce the transalkylation overhead-stripper-liquid stream and the transalkylation stripper-bottoms stream; the xylene-plus-enriched stream is combined with a deheptanizer bottoms stream and the combined streams are separated in the xylene fractionation zone which additionally produces a bottoms stream rich in C₁₁ alkylaromatics which is removed from the process;

the overhead-xylene-splitter stream is passed to the para-xylene separation zone which additionally produces a paraxylene separate zone effluent stream comprising metaxylene; the paraxylene separate zone effluent stream is passed to a xylene isomerization zone and contacted with an isomerization catalyst
5 under isomerization conditions to produce an isomerization product; the isomerization product is separated in a deheptanizer fractionation zone comprising at least one deheptanizer column and optionally at least one olefin saturation zone; and the deheptanizer fractionation zone produces the deheptanizer bottoms stream.

10 3. The process of claim 2 wherein the deheptanizer zone additionally produces an over head liquid stream and the catalytic reformat stream and the transalkylation stripper-overhead-liquid stream are combined with the overhead-liquid stream from the deheptanizer.

4. The process of any of claims 1 through 3 wherein the olefin saturation
15 catalyst comprises a nickel or a platinum-group component, and an inorganic oxide component and wherein the olefin saturation conditions comprise a temperature from 20° to 200°C, a pressure from 5 to 70 kg/cm² and a stoichiometric ratio of hydrogen to olefins from 1:1 to 5:1.

5. The process of any of claims 1 through 3 wherein the metal-stabilized
20 transalkylation catalyst comprises a zeolite component, a metal component, and an inorganic oxide component and wherein the metal component is preferably selected from the group consisting of platinum, palladium, rhodium, ruthenium, osmium, and iridium, rhenium, tin, germanium, lead, cobalt, nickel, indium, gallium, zinc, uranium,

dysprosium, thallium, and mixtures thereof and the selected metal group component is preferably rhenium.

6. The process of claim 5 wherein the zeolite component is selected from the group consisting of a pentasil zeolite, a beta zeolite, a mordenite zeolite, or mixtures thereof and wherein the zeolite component is preferably mordenite and more preferably the inorganic oxide component is alumina.

7. The process of any of claims 1 through 3 wherein the transalkylation conditions comprise a temperature from 200° to 540°C, a pressure from 1 to 60 kg/cm², and a liquid hourly space velocity from 0.1 to 20 hr⁻¹.

8. The process of claims 2 or 3 wherein the hydrotreating catalyst comprises a component of cobalt oxide or nickel oxide, a component of molybdenum oxide or tungsten oxide, and a component of inorganic oxide support and wherein the hydrotreating conditions comprise a liquid hourly space velocity from 1.0 to 5.0 hr⁻¹, a ratio of hydrogen to naphtha feedstock from 50 to 135 Nm³/m³, and a pressure from 10 to 35 kg/cm².

9. The process of claims 2 or 3 wherein the reforming conditions comprise a liquid hourly space velocity from 1.0 to 5.0 hr⁻¹, a ratio of hydrogen to hydrocarbon from 1 to 10 moles of hydrogen per mole of naphtha, and a pressure from 2.5 to 35 kg/cm².

10. The process of claims 2 or 3 wherein the debutanizer fractionation zone comprises at least one stabilizer column and at least one olefin saturation zone and preferably the olefin saturation zone comprises a nickel and alumina catalyst.

11. The process of claims 2 or 3 wherein the isomerization catalyst comprises a molecular sieve component, a metal component, and an inorganic oxide component

and wherein the isomerization conditions comprises a temperature in the range from 0° to 600°C, a pressure from atmospheric to 50 kg/cm³, and a liquid hourly hydrocarbon space velocity from 0.1 to 30 hr⁻¹.