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PROCESS FOR PRODUCING BTX FROM A MIXED HYDROCARBON SOURCE USING COKING

The present invention relates to a process for producing BTX comprising coking,
5 aromatic ring opening and BTX recovery. Furthermore, the present invention relates to a process installation to convert a coker feedstream into BTX comprising a coker unit, an aromatic ring opening unit and a BTX recovery unit.

It has been previously described that chemical grade BTX can be produced from a mixed feedstream comprising C5-C12 hydrocarbons by contacting said feedstream in
10 the presence of hydrogen with a catalyst having hydrocracking/hydrodesulphurisation activity; see e.g. WO 2013/182534 A1.

A major drawback of the process of WO 2013/182534 A1 is that it is not particularly suitable to convert relatively heavy mixed hydrocarbon feedstreams, such as coker gasoil, to BTX.

15 It was an object of the present invention to provide a process for producing BTX from a mixed hydrocarbon stream having an improved yield of high-value petrochemical products, such as BTX.

The solution to the above problem is achieved by providing the embodiments as described herein below and as characterized in the claims. Accordingly, the present
20 invention provides a process for producing BTX comprising:

- (a) subjecting a coker feedstream comprising heavy hydrocarbons to coking to produce coker naphtha and coker gasoil
- (b) subjecting coker gasoil to aromatic ring opening to produce BTX; and
- (c) recovering BTX from coker naphtha.

25 In the context of the present invention, it was surprisingly found that the yield of high-value petrochemical products, such as BTX can be improved by using the improved process as described herein.

In the process of the present invention, any hydrocarbon composition that is suitable as a feed for coking can be used. The coker feedstream preferably comprises resid,
30 more preferably vacuum residue. However, also crude oil, such as extra heavy crude oil can be used as a coker feedstream.

Preferably, the coker feedstream comprises hydrocarbons having a boiling point of 350 °C or more.

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The terms naphtha, gasoil and resid are used herein having their generally accepted meaning in the field of petroleum refinery processes; see Alfke et al. (2007) Oil Refining, Ullmann's Encyclopedia of Industrial Chemistry and Speight (2005) Petroleum Refinery Processes, Kirk-Othmer Encyclopedia of Chemical Technology. In
5 this respect, it is to be noted that there may be overlap between the different crude oil fractions due to the complex mixture of the hydrocarbon compounds comprised in the crude oil and the technical limits to the crude oil distillation process. Preferably, the term "naphtha" as used herein relates to the petroleum fraction obtained by crude oil distillation having a boiling point range of about 20-200 °C, more
10 preferably of about 30-190 °C. Preferably, light naphtha is the fraction having a boiling point range of about 20-100 °C, more preferably of about 30-90 °C. Heavy naphtha preferably has a boiling point range of about 80-200 °C, more preferably of about 90-190 °C. Preferably, the term "kerosene" as used herein relates to the petroleum fraction obtained by crude oil distillation having a boiling point range of
15 about 180-270 °C, more preferably of about 190-260 °C. Preferably, the term "gasoil" as used herein relates to the petroleum fraction obtained by crude oil distillation having a boiling point range of about 250-360 °C, more preferably of about 260-350 °C. Preferably, the term "resid" as used herein relates to the petroleum fraction obtained by crude oil distillation having a boiling point of more
20 than about 340 °C, more preferably of more than about 350 °C. Preferably, the resid is further fractioned, e.g. using a vacuum distillation unit, to separate the resid into a vacuum gas oil fraction and vacuum residue fraction.

The process of the present invention involves coking, which comprises subjecting the coker feedstream to coking conditions. The process conditions useful in coking, also
25 described herein as "coking conditions", can be easily determined by the person skilled in the art; see e.g. Alfke et al. (2007) loc. cit.

The term "coking" is used herein in its generally accepted sense and thus may be defined as a (non-catalytic) process to convert heavy hydrocarbon feedstream, which preferably is selected from the group consisting of atmospheric resid and
30 vacuum resid feed, into a gaseous hydrocarbon product comprising methane and C2-C4 hydrocarbons, coker naphtha, coker gas oil and petroleum coke by heating the feed to its thermal cracking temperature; see Alfke et al. (2007) Oil Refining, Ullmann's Encyclopedia of Industrial Chemistry; US 4,547,284 and US 20070108036. The C2-C4 hydrocarbons fraction produced by coking is a mixture of
35 paraffins and olefins. As used herein, the term "coker naphtha" relates to the light-distillate produced by coking that is relatively rich in mono-aromatic hydrocarbons.

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As used herein, the term "coker gasoil" relates to the middle-distillate, and optionally also the heavy-distillate, produced by coking that is relatively rich in aromatic hydrocarbons having two or more condensed aromatic rings. One form of coking is "delayed coking" which comprises introducing the heavy hydrocarbon feedstream to a fractionator where cracked vapors are condensed. The fractionator bottom product is subsequently heated in a furnace to a temperature of 450-550 °C, and the cracked furnace effluent flows through one of the coke drums in which coke is being formed and deposited. The cracked vapors from the coke drum may be separated further in a fractionator. The coke drums are alternately in use to allow coke removal. A further form of coking is "fluidized coking", which, in contrast to the delayed coking process, allows continuous operation. Fluidized coking comprises performing the cracking reaction in reactor in a fluid bed of coke particles into which the heavy hydrocarbon feedstream is injected. Coke fines are removed from the cracked vapors in cyclone separators before fractionation. The coke formed in the reactor may flow continuously to a heater, where it is heated to a temperature of 550-700 °C by partial combustion in a fluid bed, from where the net coke production is withdrawn. Another part of the heated coke particles is returned to the reactor to provide process heat.

Preferably, the coking comprises subjecting the coker feedstream to coking conditions, wherein the coking conditions comprise a temperature of 450-700 °C and a pressure of 50-800 kPa absolute.

The coker naphtha produced in the process of the present invention is relatively rich in olefins and diolefins. Preferably, said olefins and diolefins are separated from other hydrocarbons comprised in the coker naphtha by extraction; see e.g. US 7,019,188. The accordingly separated olefins may be subjected to aromatization.

The term "alkane" or "alkanes" is used herein having its established meaning and accordingly describes acyclic branched or unbranched hydrocarbons having the general formula C_nH_{2n+2} , and therefore consisting entirely of hydrogen atoms and saturated carbon atoms; see e.g. IUPAC. Compendium of Chemical Terminology, 2nd ed. (1997). The term "alkanes" accordingly describes unbranched alkanes ("normal-paraffins" or "n-paraffins" or "n-alkanes") and branched alkanes ("iso-paraffins" or "iso-alkanes") but excludes naphthenes (cycloalkanes).

The term "aromatic hydrocarbons" or "aromatics" is very well known in the art. Accordingly, the term "aromatic hydrocarbon" relates to cyclically conjugated hydrocarbon with a stability (due to delocalization) that is significantly greater than

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that of a hypothetical localized structure (e.g. Kekulé structure). The most common method for determining aromaticity of a given hydrocarbon is the observation of diatropicity in the ^1H NMR spectrum, for example the presence of chemical shifts in the range of from 7.2 to 7.3 ppm for benzene ring protons.

- 5 The terms “naphthenic hydrocarbons” or “naphthenes” or “cycloalkanes” is used herein having its established meaning and accordingly describes saturated cyclic hydrocarbons.

- The term “olefin” is used herein having its well-established meaning. Accordingly, olefin relates to an unsaturated hydrocarbon compound containing at least one
10 carbon-carbon double bond. Preferably, the term “olefins” relates to a mixture comprising two or more of ethylene, propylene, butadiene, butylene-1, isobutylene, isoprene and cyclopentadiene.

- The term “LPG” as used herein refers to the well-established acronym for the term “liquefied petroleum gas”. LPG generally consists of a blend of C2-C4 hydrocarbons
15 i.e. a mixture of ethane, propane and butanes and, depending on the source, also ethylene, propylene and butylenes.

- As used herein, the term “C# hydrocarbons”, wherein “#” is a positive integer, is meant to describe all hydrocarbons having # carbon atoms. Moreover, the term “C# + hydrocarbons” is meant to describe all hydrocarbon molecules having # or
20 more carbon atoms. Accordingly, the term “C5+ hydrocarbons” is meant to describe a mixture of hydrocarbons having 5 or more carbon atoms. The term “C5+ alkanes” accordingly relates to alkanes having 5 or more carbon atoms.

- The terms light-distillate, middle-distillate and heavy-distillate are used herein having their generally accepted meaning in the field of petroleum refinery processes;
25 see Speight, J. G. (2005) loc.cit. In this respect, it is to be noted that there may be overlap between different distillation fractions due to the complex mixture of the hydrocarbon compounds comprised in the product stream produced by refinery unit operations and the technical limits to the distillation process used to separate the different fractions. Preferably, a “light-distillate” is a hydrocarbon distillate obtained
30 in a refinery unit process having a boiling point range of about 20-200 °C, more preferably of about 30-190 °C. The “light-distillate” is often relatively rich in aromatic hydrocarbons having one aromatic ring. Preferably, a “middle-distillate” is a hydrocarbon distillate obtained in a refinery unit process having a boiling point range of about 180-360 °C, more preferably of about 190-350 °C. The “middle-distillate” is

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relatively rich in aromatic hydrocarbons having two aromatic rings. Preferably, a "heavy-distillate" is a hydrocarbon distillate obtained in a refinery unit process having a boiling point of more than about 340 °C, more preferably of more than about 350 °C. The "heavy-distillate" is relatively rich in hydrocarbons having more than 2 aromatic rings. Accordingly, a refinery or petrochemical process-derived distillate is obtained as the result of a chemical conversion followed by a fractionation, e.g. by distillation or by extraction, which is in contrast to a crude oil fraction.

The process of the present invention involves aromatic ring opening, which comprises contacting the coker gasoil in the presence of hydrogen with an aromatic ring opening catalyst under aromatic ring opening conditions. The process conditions useful in aromatic ring opening, also described herein as "aromatic ring opening conditions", can be easily determined by the person skilled in the art; see e.g. e.g. US3256176, US4789457 and US 7,513,988.

The term "aromatic ring opening" is used herein in its generally accepted sense and thus may be defined as a process to convert a hydrocarbon feed that is relatively rich in hydrocarbons having condensed aromatic rings, such as coker gasoil, to produce a product stream comprising a light-distillate that is relatively rich in BTX (ARO-derived gasoline) and preferably LPG. Such an aromatic ring opening process (ARO process) is for instance described in US3256176 and US4789457. Such processes may comprise of either a single fixed bed catalytic reactor or two such reactors in series together with one or more fractionation units to separate desired products from unconverted material and may also incorporate the ability to recycle unconverted material to one or both of the reactors. Reactors may be operated at a temperature of 200-600 °C, preferably 300-400 °C, a pressure of 3-35 MPa, preferably 5 to 20MPa together with 5-20 wt-% of hydrogen (in relation to the hydrocarbon feedstock), wherein said hydrogen may flow co-current with the hydrocarbon feedstock or counter current to the direction of flow of the hydrocarbon feedstock, in the presence of a dual functional catalyst active for both hydrogenation-dehydrogenation and ring cleavage, wherein said aromatic ring saturation and ring cleavage may be performed. Catalysts used in such processes comprise one or more elements selected from the group consisting of Pd, Rh, Ru, Ir, Os, Cu, Co, Ni, Pt, Fe, Zn, Ga, In, Mo, W and V in metallic or metal sulphide form supported on an acidic solid such as alumina, silica, alumina-silica and zeolites. In this respect, it is to be noted that the term "supported on" as used herein includes any conventional way to provide a catalyst which combines one or more elements

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with a catalytic support. By adapting either single or in combination the catalyst composition, operating temperature, operating space velocity and/or hydrogen partial pressure, the process can be steered towards full saturation and subsequent cleavage of all rings or towards keeping one aromatic ring unsaturated and subsequent cleavage of all but one ring. In the latter case, the ARO process produces a light-distillate ("ARO-gasoline") which is relatively rich in hydrocarbon compounds having one aromatic and or naphthenic ring. In the context of the present invention, it is preferred to use an aromatic ring opening process that is optimized to keep one aromatic or naphthenic ring intact and thus to produce a light-distillate which is relatively rich in hydrocarbon compounds having one aromatic or naphthenic ring. A further aromatic ring opening process (ARO process) is described in US 7,513,988. Accordingly, the ARO process may comprise aromatic ring saturation at a temperature of 100-500 °C, preferably 200-500 °C, more preferably 300-500 °C, a pressure of 2-10 MPa together with 1-30 wt-%, preferably 5-30 wt-% of hydrogen (in relation to the hydrocarbon feedstock) in the presence of an aromatic hydrogenation catalyst and ring cleavage at a temperature of 200-600 °C, preferably 300-400 °C, a pressure of 1-12 MPa together with 1-20 wt-% of hydrogen (in relation to the hydrocarbon feedstock) in the presence of a ring cleavage catalyst, wherein said aromatic ring saturation and ring cleavage may be performed in one reactor or in two consecutive reactors. The aromatic hydrogenation catalyst may be a conventional hydrogenation/hydrotreating catalyst such as a catalyst comprising a mixture of Ni, W and Mo on a refractory support, typically alumina. The ring cleavage catalyst comprises a transition metal or metal sulphide component and a support. Preferably the catalyst comprises one or more elements selected from the group consisting of Pd, Rh, Ru, Ir, Os, Cu, Co, Ni, Pt, Fe, Zn, Ga, In, Mo, W and V in metallic or metal sulphide form supported on an acidic solid such as alumina, silica, alumina-silica and zeolites. In this respect, it is to be noted that the term "supported on" as used herein includes any conventional way of to provide a catalyst which combines one or more elements with a catalyst support. By adapting either single or in combination the catalyst composition, operating temperature, operating space velocity and/or hydrogen partial pressure, the process can be steered towards full saturation and subsequent cleavage of all rings or towards keeping one aromatic ring unsaturated and subsequent cleavage of all but one ring. In the latter case, the ARO process produces a light-distillate ("ARO-gasoline") which is relatively rich in hydrocarbon compounds having one aromatic ring. In the context of the present invention, it is preferred to use an aromatic ring opening process that is optimized to keep one aromatic ring intact and thus to

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produce a light-distillate which is relatively rich in hydrocarbon compounds having one aromatic ring.

Preferably, the aromatic ring opening comprises contacting the coker gasoil in the presence of hydrogen with an aromatic ring opening catalyst under aromatic ring opening conditions, wherein the aromatic ring opening catalyst comprises a
5 transition metal or metal sulphide component and a support, preferably comprising one or more elements selected from the group consisting of Pd, Rh, Ru, Ir, Os, Cu, Co, Ni, Pt, Fe, Zn, Ga, In, Mo, W and V in metallic or metal sulphide form supported on an acidic solid, preferably selected from the group consisting of alumina, silica,
10 alumina-silica and zeolites and wherein the aromatic ring opening conditions comprise a temperature of 100-600 °C, a pressure of 1-12 MPa. Preferably, the aromatic ring opening conditions further comprise the presence and the presence of 5-30 wt-% of hydrogen (in relation to the hydrocarbon feedstock).

Preferably, the aromatic ring opening catalyst comprises an aromatic hydrogenation
15 catalyst comprising one or more elements selected from the group consisting of Ni, W and Mo on a refractory support, preferably alumina; and a ring cleavage catalyst comprising a transition metal or metal sulphide component and a support, preferably comprising one or more elements selected from the group consisting of Pd, Rh, Ru,
20 Ir, Os, Cu, Co, Ni, Pt, Fe, Zn, Ga, In, Mo, W and V in metallic or metal sulphide form supported on an acidic solid, preferably selected from the group consisting of alumina, silica, alumina-silica and zeolites, and wherein the conditions for aromatic hydrogenation comprise a temperature of 100-500 °C, preferably 200-500 °C, more preferably 300-500 °C, a pressure of 2-10 MPa and the presence of 1-30 wt-%, preferably 5-30 wt-%, of hydrogen (in relation to the hydrocarbon feedstock) and
25 wherein the ring cleavage comprises a temperature of 200-600 °C, preferably 300-400 °C, a pressure of 1-12 MPa and the presence of 5-20 wt-% of hydrogen (in relation to the hydrocarbon feedstock).

The process of the present invention involves recovery of BTX from a mixed hydrocarbon stream comprising aromatic hydrocarbons, such as coker naphtha. Any
30 conventional means for separating BTX from a mixed hydrocarbons stream may be used to recover the BTX. One such suitable means for BTX recovery involves conventional solvent extraction. The coker naphtha and the light-distillate may be subjected to "gasoline treatment" prior to solvent extraction. As used herein, the term "gasoline treatment" or "gasoline hydrotreatment" relates to a process wherein
35 an unsaturated and aromatics-rich hydrocarbon feedstream, such as coker naphtha,

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is selectively hydrotreated so that the carbon-carbon double bonds of the olefins and di-olefins comprised in said feedstream are hydrogenated; see also US 3,556,983. Conventionally, a gasoline treatment unit may include a first-stage process to improve the stability of the aromatics-rich hydrocarbon stream by selectively
5 hydrogenating diolefins and alkenyl compounds thus making it suitable for further processing in a second stage. The first stage hydrogenation reaction is carried out using a hydrogenation catalyst commonly comprising Ni and/or Pd, with or without promoters, supported on alumina in a fixed-bed reactor. The first stage
10 hydrogenation is commonly performed in the liquid phase comprising a process inlet temperature of 200 °C or less, preferably of 30-100 °C. In a second stage, the first-stage hydrotreated aromatics-rich hydrocarbon stream may be further processed to prepare a feedstock suitable for aromatics recovery by selectively hydrogenating the olefins and removing sulfur via hydrodesulfurization. In the second stage
15 hydrogenation a hydrogenation catalyst is commonly used comprising elements selected from the group consisting of Ni, Mo, Co, W and Pt, with or without promoters, supported on alumina in a fixed-bed reactor, wherein the catalyst is in a sulfide form. The process conditions generally comprise a process temperature of 200-400 °C, preferably of 250-350 °C and a pressure of 1-3.5 MPa, preferably 2-3.5 MPa gauge. The aromatics-rich product produced by gasoline treatment is then
20 further subject to BTX recovery using conventional solvent extraction. In case the aromatics-rich hydrocarbon mixture that is to be subjected to the gasoline treatment is low in diolefins and alkenyl compounds, the aromatics-rich hydrocarbon stream can be directly subjected to the second stage hydrogenation or even directly subjected to aromatics extraction. Preferably, the gasoline treatment unit is a
25 hydrocracking unit as described herein below that is suitable for converting a feedstream that is rich in aromatic hydrocarbons having one aromatic ring into purified BTX.

The product produced in the process of the present invention is BTX. The term "BTX" as used herein relates to a mixture of benzene, toluene and xylenes. Preferably, the
30 product produced in the process of the present invention comprises further useful aromatic hydrocarbons such as ethylbenzene. Accordingly, the present invention preferably provides a process for producing a mixture of benzene, toluene xylenes and ethylbenzene ("BTXE"). The product as produced may be a physical mixture of the different aromatic hydrocarbons or may be directly subjected to further
35 separation, e.g. by distillation, to provide different purified product streams. Such purified product stream may include a benzene product stream, a toluene product stream, a xylene product stream and/or an ethylbenzene product stream.

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Preferably, the aromatic ring opening further produces light-distillate and wherein the BTX is recovered from said light-distillate. Preferably, the BTX produced by aromatic ring opening is comprised in the light-distillate. In this embodiment, the BTX comprised in the light-distillate is separated from the other hydrocarbons
5 comprised in said light-distillate by the BTX recovery.

Preferably the BTX is recovered from the coker naphtha and/or from the light-distillate by subjecting said coker naphtha and/or light-distillate to hydrocracking. By selecting hydrocracking for the BTX recovery, the BTX yield of of the process of the present invention can be improved since mono-aromatic hydrocarbons other than
10 BTX can be converted into BTX by hydrocracking.

Preferably, coker naphtha is hydrotreated before subjecting to hydrocracking to saturate all olefins and diolefins. By removing the olefins and diolefins in the coker naphtha, the exotherm during hydrocracking can be better controlled, thus improving operability. More preferably, the olefins and diolefins are separated from
15 the coker naphtha using conventional methods such as described in US 7,019,188 and WO 01/59033 A1. Preferably, the olefins and diolefins, which were separated from the coker naphtha, are subjected to aromatization, thereby improving the BTX yield of the process of the present invention.

The process of the present invention may involve hydrocracking, which comprises
20 contacting the coker naphtha and preferably the light-distillate in the presence of hydrogen with a hydrocracking catalyst under hydrocracking conditions. The process conditions useful hydrocracking, also described herein as "hydrocracking conditions", can be easily determined by the person skilled in the art; see Alfke et al. (2007) loc.cit. Preferably, the coker naphtha is subjected to gasoline hydrotreatment as
25 described herein above before subjecting to hydrocracking. Preferably, the C9+ hydrocarbons comprised in the hydrocracked product stream are recycled to either the either hydrocracker or, preferably, to aromatic ring opening.

The term "hydrocracking" is used herein in its generally accepted sense and thus may be defined as a catalytic cracking process assisted by the presence of an
30 elevated partial pressure of hydrogen; see e.g. Alfke et al. (2007) loc.cit. The products of this process are saturated hydrocarbons and, depending on the reaction conditions such as temperature, pressure and space velocity and catalyst activity, aromatic hydrocarbons including BTX. The process conditions used for hydrocracking generally includes a process temperature of 200-600 °C, elevated pressures of 0.2-
35 20 MPa, space velocities between 0.1-20 h⁻¹. Hydrocracking reactions proceed

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through a bifunctional mechanism which requires an acid function, which provides for the cracking and isomerization and which provides breaking and/or rearrangement of the carbon-carbon bonds comprised in the hydrocarbon compounds comprised in the feed, and a hydrogenation function. Many catalysts
5 used for the hydrocracking process are formed by combining various transition metals, or metal sulfides with the solid support such as alumina, silica, alumina-silica, magnesia and zeolites.

Preferably the BTX is recovered from the coker naphtha and/or from the light-distillate by subjecting said coker naphtha and/or light-distillate to gasoline
10 hydrocracking. As used herein, the term "gasoline hydrocracking" or "GHC" refers to a hydrocracking process that is particularly suitable for converting a complex hydrocarbon feed that is relatively rich in aromatic hydrocarbon compounds -such as coker naphtha- to LPG and BTX, wherein said process is optimized to keep one aromatic ring intact of the aromatics comprised in the GHC feedstream, but to
15 remove most of the side-chains from said aromatic ring. Accordingly, the main product produced by gasoline hydrocracking is BTX and the process can be optimized to provide chemicals-grade BTX. Preferably, the hydrocarbon feed that is subject to gasoline hydrocracking further comprises light-distillate. More preferably, the hydrocarbon feed that is subjected to gasoline hydrocracking preferably does not
20 comprise more than 1 wt-% of hydrocarbons having more than one aromatic ring. Preferably, the gasoline hydrocracking conditions include a temperature of 300-580 °C, more preferably of 400-580 °C and even more preferably of 430-530 °C. Lower temperatures must be avoided since hydrogenation of the aromatic ring becomes favourable, unless a specifically adapted hydrocracking catalyst is employed. For
25 instance, in case the catalyst comprises a further element that reduces the hydrogenation activity of the catalyst, such as tin, lead or bismuth, lower temperatures may be selected for gasoline hydrocracking; see e.g. WO 02/44306 A1 and WO 2007/055488. In case the reaction temperature is too high, the yield of LPG's (especially propane and butanes) declines and the yield of methane rises. As
30 the catalyst activity may decline over the lifetime of the catalyst, it is advantageous to increase the reactor temperature gradually over the life time of the catalyst to maintain the hydrocracking conversion rate. This means that the optimum temperature at the start of an operating cycle preferably is at the lower end of the hydrocracking temperature range. The optimum reactor temperature will rise as the
35 catalyst deactivates so that at the end of a cycle (shortly before the catalyst is replaced or regenerated) the temperature preferably is selected at the higher end of the hydrocracking temperature range.

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Preferably, the gasoline hydrocracking of a hydrocarbon feedstream is performed at a pressure of 0.3-5 MPa gauge, more preferably at a pressure of 0.6-3 MPa gauge, particularly preferably at a pressure of 1-2 MPa gauge and most preferably at a pressure of 1.2-1.6 MPa gauge. By increasing reactor pressure, conversion of C5+ non-aromatics can be increased, but this also increases the yield of methane and the hydrogenation of aromatic rings to cyclohexane species which can be cracked to LPG species. This results in a reduction in aromatic yield as the pressure is increased and, as some cyclohexane and its isomer methylcyclopentane, are not fully hydrocracked, there is an optimum in the purity of the resultant benzene at a pressure of 1.2-1.6 MPa.

Preferably, gasoline hydrocracking of a hydrocarbon feedstream is performed at a Weight Hourly Space Velocity (WHSV) of 0.1-20 h⁻¹, more preferably at a Weight Hourly Space Velocity of 0.2-15 h⁻¹ and most preferably at a Weight Hourly Space Velocity of 0.4-10 h⁻¹. When the space velocity is too high, not all BTX co-boiling paraffin components are hydrocracked, so it will not be possible to achieve BTX specification by simple distillation of the reactor product. At too low space velocity the yield of methane rises at the expense of propane and butane. By selecting the optimal Weight Hourly Space Velocity, it was surprisingly found that sufficiently complete reaction of the benzene co-boilers is achieved to produce on spec BTX without the need for a liquid recycle.

Preferably, hydrocracking comprises contacting the coker naphtha and preferably the light-distillate in the presence of hydrogen with a hydrocracking catalyst under hydrocracking conditions, wherein, wherein the hydrocracking catalyst comprises 0.1-1 wt-% hydrogenation metal in relation to the total catalyst weight and a zeolite having a pore size of 5-8 Å and a silica (SiO₂) to alumina (Al₂O₃) molar ratio of 5-200 and wherein the hydrocracking conditions comprise a temperature of 400-580 °C, a pressure of 300-5000 kPa gauge and a Weight Hourly Space Velocity (WHSV) of 0.1-20 h⁻¹. The hydrogenation metal preferably is at least one element selected from Group 10 of the periodic table of Elements, most preferably Pt. The zeolite preferably is MFI. Preferably a temperature of 420-550 °C, a pressure of 600-3000 kPa gauge and a Weight Hourly Space Velocity of 0.2-15 h⁻¹ and more preferably a temperature of 430-530 °C, a pressure of 1000-2000 kPa gauge and a Weight Hourly Space Velocity of 0.4-10 h⁻¹ is used.

One advantage of selecting this specific hydrocracking catalyst as described herein above is that no desulphurization of the feed to the hydrocracking is required.

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Accordingly, preferred gasoline hydrocracking conditions thus include a temperature of 400-580 °C, a pressure of 0.3-5 MPa gauge and a Weight Hourly Space Velocity of 0.1-20 h⁻¹. More preferred gasoline hydrocracking conditions include a temperature of 420-550 °C, a pressure of 0.6-3 MPa gauge and a Weight Hourly Space Velocity of 0.2-15 h⁻¹. Particularly preferred gasoline hydrocracking conditions include a temperature of 430-530 °C, a pressure of 1-2 MPa gauge and a Weight Hourly Space Velocity of 0.4-10 h⁻¹.

Preferably, the aromatic ring opening and preferably the hydrocracking further produce LPG and wherein said LPG is subjected to aromatization to produce BTX.

- 10 The process of the present invention may involve aromatization, which comprises contacting LPG with an aromatization catalyst under aromatization conditions. The process conditions useful for aromatization, also described herein as "aromatization conditions", can be easily determined by the person skilled in the art; see Encyclopaedia of Hydrocarbons (2006) Vol II, Chapter 10.6, p. 591-614.
- 15 By subjecting some or all of the LPG produced by hydrocracking to aromatization, the aromatics yield of the integrated process can be improved. In addition thereto, hydrogen is produced by said aromatization, which can be used as a feed for the hydrogen consuming processes such as the aromatic ring opening and/or the aromatics recovery.
- 20 The term "aromatization" is used herein in its generally accepted sense and thus may be defined as a process to convert aliphatic hydrocarbons to aromatic hydrocarbons. There are many aromatization technologies described in the prior art using C3-C8 aliphatic hydrocarbons as raw material; see e.g. US 4,056,575; US 4,157,356; US 4,180,689; Micropor. Mesopor. Mater 21, 439; WO 2004/013095 A2
- 25 and WO 2005/08515 A1. Accordingly, the aromatization catalyst may comprise a zeolite, preferably selected from the group consisting of ZSM-5 and zeolite L and may further comprising one or more elements selected from the group consisting of Ga, Zn, Ge and Pt. In case the feed mainly comprises C3-C5 aliphatic hydrocarbons, an acidic zeolite is preferred. As used herein, the term "acidic zeolite" relates to a
- 30 zeolite in its default, protonic form. In case the feed mainly comprises C6-C8 hydrocarbons a non-acidic zeolite preferred. As used herein, the term "non-acidic zeolite" relates to a zeolite that is base-exchanged, preferably with an alkali metal or alkaline earth metals such as cesium, potassium, sodium, rubidium, barium, calcium, magnesium and mixtures thereof, to reduce acidity. Base-exchange may take place
- 35 during synthesis of the zeolite with an alkali metal or alkaline earth metal being

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added as a component of the reaction mixture or may take place with a crystalline zeolite before or after deposition of a noble metal. The zeolite is base-exchanged to the extent that most or all of the cations associated with aluminum are alkali metal or alkaline earth metal. An example of a monovalent base:aluminum molar ratio in the zeolite after base exchange is at least about 0.9. Preferably, the catalyst is
5 selected from the group consisting of HZSM-5 (wherein HZSM-5 describes ZSM-5 in its protonic form), Ga/HZSM-5, Zn/HZSM-5 and Pt/GeHZSM-5. The aromatization conditions may comprise a temperature of 400-600 °C, preferably 450-550 °C, more preferably 480-520 °C a pressure of 100-1000 kPa gauge, preferably 200-500 kPa gauge, and a Weight Hourly Space Velocity (WHSV) of 0.1-20 h⁻¹, preferably of 0.4-4
10 h⁻¹.

Preferably, the aromatization comprises contacting the LPG with an aromatization catalyst under aromatization conditions, wherein the aromatization catalyst comprises a zeolite selected from the group consisting of ZSM-5 and zeolite L,
15 optionally further comprising one or more elements selected from the group consisting of Ga, Zn, Ge and Pt and wherein the aromatization conditions comprise a temperature of 400-600 °C, preferably 450-550 °C, more preferably 480-520 °C a pressure of 100-1000 kPa gauge, preferably 200-500 kPa gauge, and a Weight Hourly Space Velocity (WHSV) of 0.1-20 h⁻¹, preferably of 0.4-4 h⁻¹.

20 Preferably, the coking further produces LPG and wherein said LPG produced by coking is subjected to aromatization to produce BTX.

Preferably, only part of the LPG produced in the process of the present invention (e.g. produced by one or more selected from the group consisting of aromatic ring opening, hydrocracking and coking) is subjected to aromatization to produce BTX.
25 The part of the LPG that is not subjected to aromatization may be subjected to olefins synthesis, e.g. by subjecting to pyrolysis or, preferably, to dehydrogenation.

Preferably, the LPG produced by hydrocracking and aromatic ring opening is subjected to a first aromatization that is optimized towards aromatization of paraffinic hydrocarbons. Preferably, said first aromatization preferably comprises the
30 aromatization conditions comprising a temperature of 450-550 °C, preferably 480-520 °C, a pressure of 100-1000 kPa gauge, preferably 200-500 kPa gauge, and a Weight Hourly Space Velocity (WHSV) of 0.1-7 h⁻¹, preferably of 0.4-2 h⁻¹.

Preferably, the LPG produced by coking is subjected to a second aromatization that is optimized towards aromatization of olefinic hydrocarbons. Preferably, said second
35 aromatization preferably comprises the aromatization conditions comprising a

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temperature of 400-600 °C, preferably 450-550 °C, more preferably 480-520 °C, a pressure of 100-1000 kPa gauge, preferably 200-700 kPa gauge, and a Weight Hourly Space Velocity (WHSV) of 1-20 h⁻¹, preferably of 2-4 h⁻¹.

It was found that the aromatic hydrocarbon product made from olefinic feeds may
5 comprise less benzene and more xylenes and C9+ aromatics than the liquid product resulting from paraffinic feeds. A similar effect may be observed when the process pressure is increased. It was found that olefinic aromatization feeds are suitable for higher pressure operation when compared to an aromatization process using paraffinic hydrocarbon feeds, which results in a higher conversion. With respect to
10 paraffinic feed and low pressure process, the detrimental effect of pressure on aromatics selectivity may be offset by the improved aromatic selectivities for olefinic aromatization feeds.

Preferably, propylene and/or butylenes are separated from the LPG produced by coking before subjecting to aromatization.

15 Means and methods for separating propylene and/or butylenes from mixed C2-C4 hydrocarbon streams are well known in the art and may involve distillation and/or extraction; see Ullmann's Encyclopedia of Industrial Chemistry, Vol. 6, Chapter "Butadiene", 388-390 and Vol. 13, Chapter "Ethylene", p. 512.

Preferably, some or all of the C2 hydrocarbons are separated from LPG produced in
20 the process of the present invention before subjecting said LPG to aromatization.

Preferably, the LPG produced by hydrocracking and aromatic ring opening is subjected to a first aromatization that is optimized towards aromatization of paraffinic hydrocarbons. Preferably, said first aromatization preferably comprises the aromatization conditions comprising a temperature of 450-550 °C, preferably 480-
25 520 °C, a pressure of 100-1000 kPa gauge, preferably 200-500 kPa gauge, and a Weight Hourly Space Velocity (WHSV) of 0.5-7 h⁻¹, preferably of 1-5 h⁻¹. Preferably the LPG produced by coking is subjected to a second aromatization that is optimized towards aromatization of olefinic hydrocarbons. Preferably, said second aromatization preferably comprises the aromatization conditions comprising a
30 temperature of 400-600 °C, preferably 450-550 °C, more preferably 480-520 °C, a pressure of 100-1000 kPa gauge, preferably 200-700 kPa gauge, and a Weight Hourly Space Velocity (WHSV) of 1-20 h⁻¹, preferably of 2-4 h⁻¹.

It was found that the aromatic hydrocarbon product made from olefinic feeds may comprise less benzene and more xylenes and C9+ aromatics than the liquid product

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resulting from paraffinic feeds. A similar effect may be observed when the process pressure is increased. It was found that olefinic aromatization feeds are suitable for higher pressure operation when compared to an aromatization process using paraffinic hydrocarbon feeds, which results in a higher conversion. With respect to
5 paraffinic feed and low pressure process, the detrimental effect of pressure on aromatics selectivity may be offset by the improved aromatic selectivities for olefinic aromatization feeds.

Preferably, one or more of the group consisting of the coking, the hydrocracking and the aromatic ring opening, and optionally the aromatization, further produce
10 methane and wherein said methane is used as fuel gas to provide process heat. Preferably, said fuel gas may be used to provide process heat to the hydrocracking, aromatic ring opening and/or aromatization. Process heat for coking preferably is provided by petroleum coke produced by coking.

Preferably, the aromatization further produces hydrogen and wherein said hydrogen
15 is used in the hydrocracking and/or the aromatic ring opening.

A representative process flow scheme illustrating particular embodiments for carrying out the process of the present invention is described in Figure 1. Figure 1 is to be understood to present an illustration of the invention and/or the principles involved.

20 In a further aspect, the present invention also relates to a process installation suitable for performing the process of the invention. This process installation and the process as performed in said process installation is particularly presented in figure 1 (Fig. 1).

Accordingly, the present invention provides a process installation for producing BTX comprising a coker unit (2) comprising an inlet for a coker feedstream (1) and an
25 outlet for coker naphtha (5) and an outlet for coker gasoil (6);
an aromatic ring opening unit (10) comprising an inlet for coker gasoil (6) and an outlet for BTX (19); and
a BTX recovery unit (9) comprising an inlet for coker naphtha (5) and an outlet for
30 BTX (16).

This aspect of the present invention is presented in figure 1 (Fig. 1).

As used herein, the term "an inlet for X" or "an outlet of X", wherein "X" is a given hydrocarbon fraction or the like relates to an inlet or outlet for a stream comprising said hydrocarbon fraction or the like. In case of an outlet for X is directly connected

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to a downstream refinery unit comprising an inlet for X, said direct connection may comprise further units such as heat exchangers, separation and/or purification units to remove undesired compounds comprised in said stream and the like.

If, in the context of the present invention, a unit is fed with more than one feed
5 stream, said feedstreams may be combined to form one single inlet into the unit or may form separate inlets to the unit.

The aromatic ring opening unit (10) preferably further has an outlet for light-distillate (17) which is fed to the BTX recovery unit (9). The BTX produced in the aromatic ring opening unit (10) may be separated from the light-distillate to form an
10 outlet for BTX (19). Preferably, the BTX produced in the aromatic ring opening unit (10) is comprised in the light-distillate (17) and is separated from said light-distillate in the BTX recovery unit (9).

The coker unit (2) preferably further has an outlet for fuel gas (3) and/or an outlet for LPG (4). Furthermore, the coker unit (2) preferably has an outlet for coke (7).
15 The aromatic ring opening unit (10) preferably further has an outlet for fuel gas (18) and/or an outlet for LPG (20). The BTX recovery unit (9) preferably further comprises an outlet for fuel gas (14) and/or an outlet for LPG (15).

Preferably, the process installation of the present invention further comprises an aromatization unit (8) comprising an inlet for LPG (4) and an outlet for BTX produced
20 by aromatization (22).

The LPG fed to the aromatization unit (8) is preferably produced by the coker unit (2), but may also be produced by other units such as the aromatic ring opening unit (10) and/or the BTX recovery unit (9). The aromatization unit (8) preferably further comprises an outlet for fuel gas (13) and/or an outlet for LPG (21). Preferably, the
25 aromatization unit (8) further comprises an outlet for hydrogen that is fed to the aromatic ring opening unit (12) and/or an outlet for hydrogen that is fed to the BTX recovery unit (11).

The following numeral references are used in Figure 1:

- 1 coker feedstream
- 30 2 coker unit
- 3 fuel gas produced by coking
- 4 LPG produced by coking
- 5 coker naphtha
- 6 coker gasoil

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- 7 coke
- 8 aromatization unit
- 9 BTX recovery unit
- 10 aromatic ring opening unit
- 5 11 hydrogen produced by aromatization that is fed to BTX recovery
- 12 hydrogen produced by aromatization that is fed to aromatic ring opening
- 13 fuel gas produced by aromatization
- 14 fuel gas produced by BTX recovery
- 15 LPG produced by BTX recovery
- 10 16 BTX produced by BTX recovery
- 17 light-distillate produced by aromatic ring opening
- 18 fuel gas produced by aromatic ring opening
- 19 BTX produced by aromatic ring opening
- 20 LPG produced by aromatic ring opening
- 15 21 LPG produced by aromatization
- 22 BTX produced by aromatization

It is noted that the invention relates to all possible combinations of features described herein, particularly features recited in the claims.

It is further noted that the term 'comprising' does not exclude the presence of other
20 elements. However, it is also to be understood that a description on a product comprising certain components also discloses a product consisting of these components. Similarly, it is also to be understood that a description on a process comprising certain steps also discloses a process consisting of these steps.

The present invention will now be more fully described by the following non-limiting
25 Examples.

Example 1

The experimental data as provided herein were obtained by flowsheet modelling in Aspen Plus. For the delayed coker, product yields and compositions are based on experimental data obtained from literature. For the aromatic ring opening followed
30 by gasoline hydrocracking a reaction scheme has been used in which all multi aromatic compounds were converted into BTX and LPG and all naphthenic and paraffinic compounds were converted to LPG.

In Example 1, Urals vacuum residue is sent to a delayed coker. This unit produces a gaseous stream, a light-distillate cut, a middle-distillate cut and coke. The light-

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distillate cut consisting of light naphtha and heavy naphtha (properties shown in Table 1) is further upgraded in the gasoline hydrocracker into a BTXE-rich stream and a non-aromatic stream. The middle-distillate consisting of light coker gas oil and heavy coker gas oil (properties shown in Table 1) is upgraded in the aromatic ring opening unit under conditions keeping 1 aromatic ring intact. The aromatic-rich product obtained in the latter unit is sent to the gasoline hydrocracker to improve the purity of the BTXE contained in that stream. The results are provided in Table 2 as provided herein below.

The products that are generated are divided into petrochemicals (olefins and BTXE, which is an acronym for BTX + ethylbenzene) and other products (hydrogen, methane, heavy fractions comprising C9 and heavier aromatic compounds and coke).

For Example 1 the BTXE yield is 35.2 wt-% of the total feed.

Example 2

Example 2 is identical to the Example 1 except for the following:

C3 and C4 hydrocarbons generated in different units of the overall complex are fed into an aromatization unit where BTXE (product), C9+ aromatics and gases are produced. Different yield patterns due to variations in feedstock composition (e.g. olefinic content) were obtained from literature and applied in the model to determine the battery-limit product slate (Table 2).

The hydrogen generated by the aromatization unit (hydrogen-producing unit) can be subsequently used in the hydrogen-consuming units (gasoline hydrocracker and aromatic ring opening).

For Example 2 the BTXE yield is 47.1 wt-% of the total feed.

Table 1. Properties of delayed coker naphthas and gas oils

FRACTION	BOILING RANGE	SPECIFIC GRAVITY (kg/ L)	PONA* (wt-%)
Light Naphtha	C5-82°C	0.6702	48/45/6/1
Heavy Naphtha	82-177°C	0.7569	36/32/14/18
Light Coker Gas Oil	177-343°C	0.8535	29/22/21/28
Heavy Coker Gas Oil	343°C and heavier	0.9568	26/19/9/46

* PONA stands for paraffinic/olefinic/naphthenic and aromatic content, respectively

Table 2. Battery-limit product slates

PRODUCTS	Example 1	Example 2
	wt-% of feed	wt-% of feed
H ₂ *	0.0%	0.8%
CH ₄	0.9%	4.1%
Ethylene	0.7%	0.7%
Ethane	6.3%	9.5%
Propylene	2.6%	0.1%
Propane	18.8%	8.6%
1-butene	1.6%	0.1%
i-butene	0.3%	0.0%
n-butane	4.9%	0.0%
i-butane	1.2%	0.0%
GASES	37.2%	23.8%
Benzene	8.8%	12.1%
Toluene	13.2%	18.9%
Xylenes	10.7%	12.1%
EB	2.5%	3.9%
BTXE	35.2%	47.1%
C9 AROMATICS	0.5%	2.0%
COKE	27.1%	27.1%

- 5 * Hydrogen amounts shown in Table 1 represent hydrogen produced in the system and not battery-limit product slate.

CLAIMS

1. Process for producing BTX comprising:
 - (a) subjecting a coker feedstream comprising heavy hydrocarbons to coking to
5 produce coker naphtha and coker gasoil;
 - (b) subjecting coker gasoil to aromatic ring opening to produce BTX; and
 - (c) recovering BTX from coker naphtha, wherein the coking further produces LPG wherein said LPG produced by coking is subjected to aromatization to produce BTX.
2. The process according to claim 1, wherein the aromatic ring opening further
10 produces light-distillate and wherein the BTX is recovered from said light-distillate.
3. The process according to claim 1 or 2, wherein the BTX is recovered from the coker naphtha and/or from the light-distillate by subjecting said coker naphtha and/or light-distillate to hydrocracking.
4. The process according to any one of claims 1-3, wherein the aromatic ring opening
15 and preferably the hydrocracking further produce LPG and wherein said LPG is subjected to aromatization to produce BTX.
5. The process according to any one of claims 1-4, wherein propylene and/or butylenes are separated from the LPG produced by coking before subjecting to aromatization.
6. The process according to any one of claims 1-5, wherein said coking comprises
20 subjecting the coker feedstream to coking conditions, wherein the coking conditions comprise a temperature of 450-700 °C and a pressure of 50-800 kPa absolute.
7. The process according to any one of claims 3-6, wherein said hydrocracking comprises contacting the coker naphtha and preferably the light-distillate in the presence of hydrogen with a hydrocracking catalyst under hydrocracking conditions, wherein
25 the hydrocracking catalyst comprises 0.1-1 wt.% hydrogenation metal in relation to the total catalyst weight and a zeolite having a pore size of 5-8 Å and a silica (SiO₂) to alumina (Al₂O₃) molar ratio of 5-200 and wherein the hydrocracking conditions comprise a temperature of 400-580 °C, a pressure of 300-5000 kPa gauge and a Weight Hourly Space Velocity (WHSV) of 0.1-20 h⁻¹.
8. The process according to any one of claims 1-7, wherein said aromatic ring opening
30 comprises contacting the coker gasoil in the presence of hydrogen with an aromatic ring opening catalyst under aromatic ring opening conditions, wherein the aromatic ring opening catalyst comprises a transition metal or metal sulphide component and a support, preferably comprising one or more elements selected from the group

consisting of Pd, Rh, Ru, Ir, Os, Cu, Co, Ni, Pt, Fe, Zn, Ga, In, Mo, W and V in metallic or metal sulphide form supported on an acidic solid, preferably selected from the group consisting of alumina, silica, alumina-silica and zeolites and wherein

the aromatic ring opening conditions comprise a temperature of 100-600 °C, a pressure of 1-12 MPa.

9. The process according to claim 8, wherein the aromatic ring opening catalyst comprises an aromatic hydrogenation catalyst comprising one or more elements selected from the group consisting of Ni, W and Mo on a refractory support; and a ring cleavage catalyst comprising a transition metal or metal sulphide component and a support and

wherein the conditions for aromatic hydrogenation comprise a temperature of 100-500 °C, a pressure of 2-10 MPa and the presence of 1-30 wt.% of hydrogen (in relation to the hydrocarbon feedstock) and wherein the ring cleavage comprises a temperature of 200-600 °C, a pressure of 1-12 MPa and the presence of 1-20 wt.% of hydrogen (in relation to the hydrocarbon feedstock).

10. The process according to any one of claims 4-9, wherein the aromatization comprises contacting the LPG with an aromatization catalyst under aromatization conditions, wherein the aromatization catalyst comprises a zeolite selected from the group consisting of ZSM-5 and zeolite L, optionally further comprising one or more elements selected from the group consisting of Ga, Zn, Ge and Pt and wherein

the aromatization conditions comprise a temperature of 400-600 °C, a pressure of 100-1000 kPa gauge and a Weight Hourly Space Velocity (WHSV) of 0.1-20 h⁻¹.

11. The process according to any one of claims 4-10, wherein the LPG produced by hydrocracking and aromatic ring opening is subjected to a first aromatization that is optimized towards aromatization of paraffinic hydrocarbons, wherein said first aromatization preferably comprises the aromatization conditions comprising a temperature of 400-600 °C, a pressure of 100-1000 kPa gauge and a Weight Hourly Space Velocity (WHSV) of 0.5-7 h⁻¹; and/or wherein

the LPG produced by coking is subjected to a second aromatization that is optimized towards aromatization of olefinic hydrocarbons, wherein said second aromatization preferably comprises the aromatization conditions comprising a temperature of 400-600 °C, a pressure of 100-1000 kPa gauge and a Weight Hourly Space Velocity (WHSV) of 1-20 h⁻¹.

12. The process according to any one of claims 1-11, wherein one or more of the group consisting of the coking, the hydrocracking and the aromatic ring opening, and optionally the

aromatization, further produce methane and wherein said methane is used as fuel gas to provide process heat.

13. The process according to any one of claims 1-12, wherein the coker feedstream comprises hydrocarbons having a boiling point of 350 °C or more.

5 14. The process according to any one of claims 4-13, wherein the aromatization further produces hydrogen and wherein said hydrogen is used in the hydrocracking and/or the aromatic ring opening.