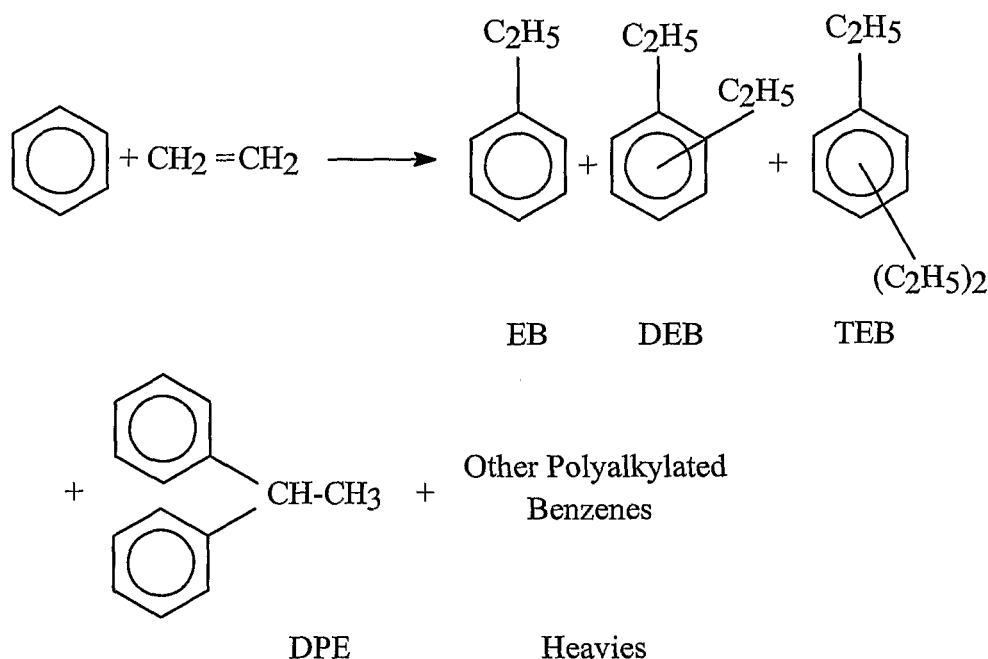


ALKYLATION PROCESS USING UZM-8 ZEOLITE

BACKGROUND OF THE INVENTION

[0001] This invention relates to a process for producing monoalkylated aromatic compounds by alkylation. Specifically, this invention relates to highly-selective alkylation and transalkylation to produce ethylbenzene, cumene, and butylbenzenes.

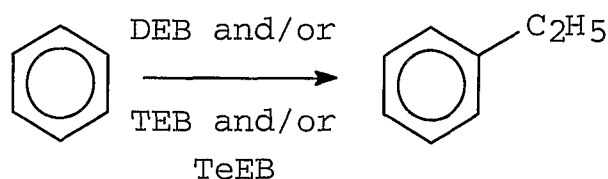
[0002] Alkylation of aromatic compounds with a C₂ to C₄ olefin and transalkylation of polyalkylaromatic compounds are two common reactions for producing monoalkylated aromatic compounds. Examples of these two reactions that are practiced industrially to produce ethylbenzene are the alkylation of benzene with ethylene and the transalkylation of benzene and a diethylbenzene. A simplified summary of the alkylation reaction and its common product and byproducts is given below:



[0003] In addition to these byproducts, the C₂-C₄ olefin can dimerize or oligomerize to form higher olefins, which can react with benzene to form alkylbenzenes having alkyl groups with 4 to 12 carbon atoms, such as butylbenzenes, hexylbenzenes, octylbenzenes, and dodecylbenzenes. These heavy alkylbenzenes can be further alkylated to form other heavy polyalkylated benzenes.

[0004] Although the formation of the diethylbenzene, triethylbenzene, and tetraethylbenzene (TeEB) isomers might, at first glance, be viewed as byproducts that represent a reduction in the efficient utilization of ethylene, in fact each can be readily transalkylated by benzene to produce ethylbenzene, as shown below:

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[0005] Combining alkylation and transalkylation can thus maximize ethylbenzene production. Such a combination can be carried out in a process having two reaction zones, one for alkylation and the other for transalkylation, or in a process having a single reaction zone in which alkylation and transalkylation both occur. In many cases, a single reaction zone is preferred over two reaction zones because of the savings in capital investment.

[0006] In contrast to diethylbenzene and triethylbenzene, 1-diphenylethane (1,1-DPE) can not be readily converted to ethylbenzene by alkylation or transalkylation. In fact, the byproduction of 1,1-DPE, as well as of the heavier polyethylated benzenes other than diethylbenzene and triethylbenzene, and of the butylbenzenes and octylbenzenes represents virtually all of the reduction in the ethylene utilization efficiency and a loss of benzene as well. As used herein, the term "heavies" refers to polyalkyl aromatics other than dialkyl and trialkyl and tetraalkyl aromatics where the alkyl group has the same number of carbon atoms as the feed olefin, to alkylaromatics formed from dimerized or oligomerized olefins such as butylbenzenes when the olefin is ethylene, and to other even heavier alkylation and transalkylation byproducts including diphenylalkanes (DPA) and alkylated diarylalkanes (DAAs), such as diphenylethanes (DPEs), alkylated diarylethanes (DAEs), diphenylpropane (DPP), and alkylated diarylpropanes (DAPs). The current minimum requirement for combination processes is that 1,1-DPE be not more than 1.0 wt-% relative to ethylbenzene. The formation of 1,1-DPE itself is assuming added importance and significance in view of the expectation in some areas of near-term minimum standards for the content of 1,1-DPE of not more than 0.5 wt-%.

[0007] In reaction zones where alkylation and transalkylation occur to produce a monoalkylated aromatic, a key operating variable is the molar ratio of aryl groups per alkyl group. The numerator of this ratio is the number of moles of aryl groups passing through the reaction zone during a specified period of time. The number of moles of aryl groups is the sum of all aryl groups, regardless of the compound in which the aryl group happens to be. In the context of ethylbenzene production, for example, one mole of benzene, one mole of ethylbenzene, and one mole of diethylbenzene each contribute one mole of aryl group to the sum of aryl groups. The denominator of this ratio is the number of moles of alkyl groups that have the same number of carbon atoms as that of the alkyl group on the desired monoalkylated aromatic and which pass through the reaction zone during the same specified period of time. The number of moles of alkyl groups is the sum of all alkyl and alkenyl groups with the same number of carbon atoms as that of the alkyl group on the desired monoalkylated aromatic, regardless of the compound in which the alkyl or alkyl group happens to be, except that paraffins are not included. In the context of ethylbenzene production, the number of moles of ethyl groups is the sum of all ethyl and ethenyl groups, regardless of the compound in which the ethyl or ethenyl group happens to be, except that paraffins, such as ethane, propane, n-butane, isobutane, pentanes, and higher paraffins are excluded from the computation of the number of moles of ethyl groups. For example, one mole of ethylene and one mole of ethylbenzene each contribute one mole of ethyl group to the sum of ethyl groups, whereas one mole of diethylbenzene contributes two moles of ethyl groups and one mole of triethylbenzene contributes three moles of ethyl groups. Butylbenzene and octylbenzene contribute no moles of ethyl groups.

[0008] In response to the hydrocarbon processing industry's demands for lower molar ratios of aryl groups per alkyl group and more efficient utilization of feed olefins, improved processes for the production of alkylbenzenes are sought.

SUMMARY OF THE INVENTION

[0009] A process has been discovered to significantly reduce the formation of diphenylalkanes and/or alkyl aromatics with alkyl groups corresponding to dimerized or oligomerized feed olefin in an alkylation process to produce alkyl aromatics, such as ethylbenzene and cumene, by alkylation using solid catalysts containing UZM-8 zeolite. The

process disclosed herein is particularly useful at a low molar ratio of aryl groups per C₂-C₄ alkyl group, such as 6 or less. The process passes a feed aromatic, a C₂-C₄ olefin, and an alkylated derivative of the feed aromatic to an alkylation catalyst bed containing UZM-8 zeolite. The concentration of the C₂-C₄ olefin based on the weight of the feed aromatic, the C₂-C₄ olefin, and the alkylated derivative of the feed aromatic passed to the UZM-8 zeolite bed can be relatively high, such as at most 17 wt-%, at most 10 wt-%, or at most 5 wt-%. The concentration of the C₂-C₄ olefin may preferably be positive and near zero in circumstances, but minimum concentrations of 0.1 wt-% or 1.5 wt-% may also be used. The alkylated derivative of the feed aromatic may have from one to six more C₂-C₄ alkyl groups than the feed aromatic, and preferably the alkylated derivative is a dialkylated or trialkylated derivative of the feed aromatic. The alkylated derivative may be introduced to the UZM-8 zeolite bed in any process stream, and preferably the stream is an aliquot recycled portion of the UZM-8 zeolite bed effluent. A relatively low amount of the aliquot portion of the effluent may be used, such as a ratio of the weight of the aliquot portion to the combined weight of the feed aromatic and the C₂-C₄ olefin of at least 0.1. Higher ratios of at least 1.0, 2.5, or 4.0 can also be used. This result using a solid catalyst comprising UZM-8 zeolite was surprising. UZM-8 zeolite has a unique layered structure, and its performance was not predictable from the prior art zeolites such as beta, omega, ZSM-5, PSH-3, MCM-22, MCM-36, MCM-49, and MCM-56.

[0010] An alkylation process of alkylating benzene with ethylene, propylene, or butene using effluent recycle shows a significant selectivity advantage if UZM-8 zeolite is used instead of a zeolite used in prior art processes. By using the process disclosed herein, ethylbenzene and cumene processes can now minimize formation of 1,1-DPE and/or other undesired byproducts even while operating profitably at low molar ratios of aryl groups per ethyl group.

[0011] Without limiting the process disclosed herein to any particular theory, a working hypothesis may in part explain the observed results. It is believed the unique layered structure of UZM-8 zeolite makes available a relatively large number of catalytically active sites near or on the surface of the UZM-8 zeolite. Once monoalkylation at an active site, the resulting monoalkylated aromatic can be readily removed from the surface. This reduces the formation of polyalkylated and heavier byproducts. Using effluent recycle with UZM-8 zeolite facilitates the transport of the monoalkylated aromatic from the active site. This working hypothesis explains the decreased formation of diarylalkanes and other byproducts when alkylating aromatics with

ethylene, propylene, butenes, and higher olefins. For example, with ethylene alkylating benzene less 1,1-DPE and butylbenzenes would form. Where the olefin is propylene for instance, less 2,2-diphenylpropane (2,2-DPP) and hexylbenzenes would form, and possibly less 1,1-diphenylpropane (1,1-DPP) too. In the case where the olefin is butene, less diphenylbutanes and butylbenzenes would form.

[0012] US-B-6,835,862 describes a process for producing a monoalkylated aromatic comprising passing an aromatic feedstock, an olefinic feedstock, and an aliquot portion of an effluent stream to an alkylation catalyst bed containing zeolite such as zeolite beta, omega, ZSM-5, PSH-3, MCM-22, MCM-36, MCM-49, and MCM-56. The ratio of the weight of the aliquot portion per unit time to the sum of the weights of the aromatic feedstock and the olefinic feedstock per said unit time is more than 2.5.

[0013] US-A-5,877,370 describes a reduction in the amount of 1,1-DPE formed in the production of ethylbenzene. The highest ratio of weight of recycle effluent per weight of fresh benzene is 3 which, at a phenyl/ethyl molar ratio of 5.0, corresponds to a ratio of weight of recycle effluent per weight of fresh feed (i.e., fresh benzene and fresh olefin) of 2.5.

[0014] Figures 1-4 are graphs showing the results of tests using reactor effluent recycle.

DETAILED DESCRIPTION

[0015] The process disclosed herein can be expected to be applicable generally to the alkylation of an alkylation substrate with an alkylation agent. The process disclosed herein is more specifically applicable to the production of an alkyl aromatic by alkylation of a feed aromatic with a feed olefin. Although benzene is the principal feed aromatic of interest, feed aromatics such as alkyl-substituted benzenes, condensed ring systems generally, and alkylated derivatives thereof may be used. Examples of such feed aromatics are toluene, ethylbenzene, propylbenzene, and so forth; xylene, mesitylene, methylethylbenzene, and so on; naphthalene, anthracene, phenanthrene, methylnaphthalene, dimethyl-naphthalene, and tetralin. More than one feed aromatic can be used. The feed aromatic may be introduced into an alkylation catalyst bed in one or more aromatic feed stream. Each aromatic feed stream may contain one or more feed aromatics. Besides the feed aromatic(s), an aromatic feed stream may contain non-aromatics, including but not limited to saturated and unsaturated cyclic hydrocarbons that have

the same, one more, or one less, number of carbon atoms as the feed aromatic. For example, an aromatic feed stream containing benzene may also contain cyclohexane, cycloheptane, cyclohexenes, or cycloheptenes, as well as methylated versions of any of these hydrocarbons, or mixtures thereof. The concentration of each feed aromatic in each aromatic feed stream may range from 0.01 to 100 wt-%.

[0016] Feed olefins containing from 2 to 6 carbon atoms are the principal alkylating agents contemplated for the process disclosed herein. Examples of such feed olefins include C₂-C₄ olefins, namely ethylene, propylene, butene-1, cis-butene-2, trans-butene-2, and iso-butene. However, feed olefins having from 2 to 20 carbon atoms may be used effectively in the process disclosed herein. More than one feed olefin may be used. The feed olefin may be introduced into an alkylation catalyst bed in one or more olefinic feed streams. Each olefinic feed stream may contain one or more feed olefins. In addition to the feed olefin(s), an olefinic feed stream may contain non-olefins, such as paraffins that have the same number of carbon atoms as the olefin. For example, a propylene-containing olefinic feed stream may also contain propane, while an olefinic feed stream containing ethylene may also contain ethane. The concentration of each feed olefin in each olefinic feed stream may range from 0.01 to 100 wt-%.

[0017] The most widely practiced hydrocarbon conversion processes to which the present invention is applicable are the catalytic alkylation of benzene with ethylene to produce ethylbenzene, the catalytic alkylation of benzene with propylene to produce cumene, and the catalytic alkylation of benzene with butene to produce butylbenzene. Although the discussion herein of the present invention will occasionally refer to catalytic cumene and butylbenzene reaction systems, the discussion mainly is in reference to its application to a catalytic ethylbenzene reaction system.

[0018] In practicing the process disclosed herein, a portion of the effluent of the alkylation reaction zone is reintroduced into the alkylation reaction zone. Unless otherwise noted in this specification, the term "portion" – when describing a process stream – refers to either an aliquot portion of the stream or a dissimilar fraction of the stream having a different composition than the total stream from which it was derived. An aliquot portion of the stream is a portion of the stream that has essentially the same composition as the stream from which it was derived. In some embodiments of the process disclosed herein, the reintroduced

portion of the alkylation reaction zone effluent may contain a transalkylation agent.

Consequently, in practicing these embodiments a transalkylation agent is introduced into the alkylation reaction zone with the alkylation effluent. In other embodiments, the

transalkylation agent may be introduced via a stream other than a portion of the alkylation

5 effluent stream. In theory, the transalkylation agent, if present, may be any compound that is

capable of transalkylating with the alkylation substrate (e.g., benzene), mixing with the

alkylating agent (e.g., ethylene), and decreasing the concentration of the alkylating agent at

and downstream of the alkylation agent injection point. In practice, however, the

transalkylation agent preferably has a number of characteristics that are consistent with the

10 process objective of producing high yields of high-purity product ethylbenzene. First, the

transalkylation agent should increase ethylbenzene yield by transalkylation, in addition to

increasing ethylbenzene yield by minimizing 1,1-DPE formation. Accordingly, a

polyethylbenzene, such as diethylbenzene, triethylbenzene, and so forth up to even

hexaethylbenzene, is preferred because each can transalkylate to ethylbenzene, regardless of

15 whether each is alkylated by ethylene. Because of the possibility of alkylation of the

polyethylbenzene by ethylene, however, the lighter polyethylbenzenes are more preferred over

the heavier polyethylbenzenes, with diethylbenzene being most preferred. More generally

when alkylating a feed aromatic with a C₂-C₄ olefin, the transalkylation agent is an alkylated

derivative of the feed aromatic having from one to six more C₂-C₄ alkyl groups than the feed

20 aromatic.

[0019] A second characteristic of the transalkylation agent is that the transalkylation agent

preferably decreases the molar ratio of aryl groups per alkyl groups in the alkylation reaction

zone. This is usually not a limiting characteristic, however, because if the transalkylation

agent has at least one aryl group and one alkyl group, then the transalkylation agent will

25 decrease the molar ratio of aryl groups per alkyl group if the ratio is greater than 1.

Transalkylation agents with two or more alkyl groups per aryl group will decrease the molar ratio of aryl groups per alkyl group if the ratio is greater than 0.5, and so on for

transalkylation agents with more alkyl groups per aryl group. Third, the transalkylation agent

preferably should not adversely affect the yield of the desired monoalkylated aromatic. For

30 example in the context of ethylbenzene production, toluene and cumene are not preferred,

because ethylene can alkylate toluene or cumene and produce byproducts that cannot be

converted readily to ethylbenzene by alkylation or transalkylation. Even though generally

present in the alkylation effluent, ethylbenzene is also not preferred, because ethylbenzene can shift the equilibrium of the reactions away from the formation of ethylbenzene and because ethylbenzene can react with ethylene to produce styrene and ultimately 1,1-DPE. Thus, it would be preferred to not recycle to the alkylation reaction zone a stream containing more than 75 wt-% of the desired monoalkylated aromatic, such as the ethylbenzene or cumene product stream produced by the ethylbenzene or cumene column of the product separation zone. Fourth, the transalkylation agent preferably should not adversely affect the purity of the product stream containing the desired monoalkylated aromatic. For example in the context of ethylbenzene production, xylenes are not preferred because they are relatively difficult to separate from ethylbenzene by distillation. Another reason that xylenes are not preferred is that they can adversely affect ethylbenzene yield by alkylating with ethylene.

[0020] In general, the transalkylation agent, when present, is preferably a compound that corresponds to the alkylation substrate alkylated with at least one more alkyl group corresponding to the alkylation agent than the number of alkyl groups on the desired product of alkylating the alkylation substrate with the alkylating agent. In the general case, the transalkylation agent, when present, is different from the desired product of alkylating the alkylation substrate with the alkylation agent. Where the aromatic is benzene and the olefin is ethylene, the transalkylation agent can generally be a polyethylbenzene, and suitable transalkylation agents include di-, tri-, and tetra-ethyl aromatic hydrocarbons such as diethylbenzene, triethylbenzene, diethylmethylbenzene, diethylpropylbenzene, etc. Diethylbenzenes are preferred. Where the aromatic is benzene and the olefin is propylene, the transalkylation agent can generally be a polypropylbenzene, and suitable transalkylation agents include di-, tri-, and tetra-propyl aromatic hydrocarbons such as diisopropylbenzene, triisopropylbenzene, diisopropylmethylbenzene, triisopropylmethylbenzene, etc. Diisopropylbenzenes are especially preferred transalkylation agents.

[0021] The ratio of the weight of the recycled portion of the effluent stream entering the alkylation catalyst bed per unit time to the sum of the weights entering the alkylation catalyst bed of the feed aromatic and the feed olefin per the unit time may be at least 0.1, at least 1.0, at least 2.5, at least 4.0, at least 7.0, or at least 10.0. This ratio is sometimes referred to herein as the effluent recycle ratio or R/FF.

[0022] As used herein, aromatic feedstock entering or passing to the alkylation catalyst bed means all of the feed aromatic entering or passing to the alkylation catalyst bed in streams other than an aliquot portion of the effluent stream that enters or passes to the alkylation catalyst bed. As used herein, olefinic feedstock entering or passing to the alkylation catalyst bed means all of the feed olefin entering or passing to the alkylation catalyst bed in streams other than the portion of the effluent stream that enters or passes to the alkylation catalyst bed.

[0023] Streams that enter or pass to the alkylation catalyst bed, but which are not an aliquot portion of the effluent stream include but are not limited to the aromatic feed stream, olefinic feed stream, and any other feed stream to the alkylation catalyst bed. Such streams also include streams that have a substantially different composition from the effluent stream but which are produced by separating the effluent stream. For example, a portion of the alkylation effluent stream is typically passed to a separation zone or a product recovery zone to recover monoalkylated aromatic from that portion of the effluent stream. As a result of this separation, one or more streams that have a composition substantially different from the alkylation effluent stream are recovered. These recovered streams include the product monoalkylated aromatic stream, a stream comprising light ends, a recycle stream comprising the feed aromatic, one or more streams comprising polyalkyl aromatics (e.g., dialkyl aromatics and trialkyl aromatics), and a stream comprising heavies.

[0024] Of these recovered streams, it is most common for some or all of the feed aromatic recycle stream to be passed to the alkylation catalyst bed. However, some or all of the other recovered streams may also be passed to the alkylation catalyst bed. Any feed aromatic thus passed to the alkylation catalyst bed in some or all of these recovered streams is deemed to be included in the aromatic feedstock entering or passing to the alkylation catalyst bed for purposes of the process disclosed herein. Likewise, any feed olefin passed to the alkylation catalyst bed in some or all of these recovered streams is included in the olefinic feedstock for purposes of the invention. However, any feed aromatic or feed olefin in the aliquot portion of the alkylation catalyst bed effluent that is passed to the alkylation catalyst bed is not counted as aromatic feedstock or olefinic feedstock, respectively, passing to or entering the alkylation catalyst bed, for purposes of the process disclosed herein.

[0025] In some embodiments, the ratio of the weight of the olefinic feedstock entering the alkylation catalyst bed per unit time to the sum of the weights of compounds entering the alkylation catalyst bed per the same unit time, multiplied by 100, is generally at most 17 wt-%, at most 10 wt-%, at most 7 wt-%, at most 5 wt-%, or at most 3 wt-%. This ratio is sometimes referred to herein as the feed olefin ratio. The alkylation conditions comprise a maximum feed olefin concentration based on the weight of the feed aromatic, the C₂-C₄ olefin, and an alkylated derivative of the feed aromatic having from one to six more C₂-C₄ alkyl groups than the feed aromatic entering the alkylation catalyst bed of at most 17 wt-%, at most 10 wt-%, at most 7 wt-%, at most 5 wt-%, or at most 3 wt-%.

[0026] The aromatic feed stream, the olefinic feed stream, and the aliquot portion of the effluent stream are preferably combined upstream of the alkylation catalyst bed to form a combined feed stream having preferably a homogeneous mixture and a uniform composition and temperature. If one or more other streams also pass to the alkylation catalyst bed reaction zone, then preferably the other stream or streams mix so that the combined feed stream is formed from all entering streams. This helps to ensure that the feed olefin ratio and/or the maximum feed olefin concentration at alkylation conditions are minimized. Combining these streams may occur batch-wise, or on a non-continuous basis, or preferably on a continuous basis.

[0027] The alkylation reaction zone can comprise one or more alkylation catalyst beds and/or one or more alkylation catalyst reactors, and each reactor may contain one or more alkylation catalyst beds. A common configuration of an alkylation zone employs two alkylation reactors, each of which has two alkylation catalyst beds. The number of reactors is generally less than eight, and the number of catalyst beds in a given reactor is generally less than six.

[0028] Alkylation conditions for the process disclosed herein include a molar ratio of aryl groups per alkyl group of generally from 25 to 1. The molar ratio may be less than 1, and it is believed that the molar ratio may be 0.75 or lower. The molar ratio of aryl groups per ethyl group (or per propyl group, in cumene production) is preferably at least 1.2. The molar ratio is preferably at most 6, and more preferably at most 3.

[0029] In general, for a given molar ratio of alkylation substrate per alkylation agent, especially an olefinic alkylation agent, the greater the molar ratio of aryl groups to alkyl

groups in the feed stream, the less is the rise in temperature in the reaction zone that occurs as a result of the alkylation reactions. The alkylation reactions are considered to be moderately exothermic. Although the reactor may have indirect heat exchange means to remove the heat as it is produced, the reactor is preferably adiabatic, and so the outlet temperature of the effluent stream is higher than the inlet temperature of the reactants. An increase in R/FF, as well as an increase in the molar ratio of aryl groups to alkyl groups in the feed stream, increases the quantity of aryl groups available to act as a heat sink in the reaction zone and thus decreases the temperature rise in the reaction zone. The appropriate reaction temperature may be generally from 60°C to the critical temperature of the alkylation substrate, which may be 475°C or even higher. The inlet temperature in the reaction zone is generally from 60 to 260°C, and preferably from 100 to 250°C. Although the temperature rise that occurs in the reaction zone could be from 10 to 190°C depending on the total mass flows in the reactor, the temperature rise is generally from 5 to 60°C, and preferably from 5 to 50°C.

[0030] Minimizing the temperature rise in the reaction zone helps prevent high reactor outlet temperatures, which cause undesirable side reactions such as cracking of hydrocarbons to occur. High reaction temperatures can also cause vaporization of benzene and the desired monoalkylaromatic (e.g. ethylbenzene or cumene) in the reaction zone. The temperature rise in the reaction zone can be controlled by withdrawing an effluent stream from the reaction zone, cooling a portion of the effluent stream, and recycling the cooled portion of the effluent stream to the reaction zone. Although recycling reactor effluent to the reaction zone in this manner may be disadvantageous for some reaction zones, it is not disadvantageous for the process disclosed herein because recycling reactor effluent to the reaction zone does not significantly alter the product distribution when the catalyst is UZM-8 zeolite. A significant alteration in the product distribution is a change in the concentration of any of the products in the reactor effluent stream of more than 0.5 wt-%. A significant alteration in the product distribution does not occur because at the reaction conditions UZM-8 zeolite is such an active promoter of the alkylation reaction between benzene and ethylene and of the transalkylation reaction between benzene and diethylbenzene that these reactions proceed to an extent of at least 80% and generally more than 90% of the way to equilibrium. The analogous reactions between benzene and propylene and between benzene and diisopropylbenzene also proceed to these high extents. Thus, recycling reactor effluent to the reaction zone does not interfere in a

significant way with the extent of the alkylation or transalkylation reactions, and recycling reactor effluent may be employed for the purpose of controlling reaction zone temperatures.

[0031] Alkylation is preferably performed in the liquid phase. Consequently, reaction pressure needs to be sufficiently high to ensure at least a partial liquid phase. Where ethylene is the olefin, the pressure range for the reactions is usually from 1379 to 6985 kPa(g), and more commonly from 3103 to 4137 kPa(g). Preferably, the reaction conditions are sufficient to maintain benzene in a liquid phase and are supercritical conditions for ethylene. Pressure is not a critical variable in the success of the process disclosed herein, however, and the only criterion is that the pressure be sufficiently great to ensure at least partial liquid phase. For olefins other than ethylene, the process disclosed herein may be practiced generally at a pressure of from 345 to 6985 kPa(g).

[0032] The weight hourly space velocity (WHSV) of the feed olefin may range from 0.01 to 8.0 hr^{-1} . As used herein, weight hourly space velocity of a component means the weight flow rate of the component per hour divided by the catalyst weight, where the weight flow rate of the component per hour and the catalyst weight are in the same weight units. The WHSV of aromatics, including benzene and a polyalkylaromatic having at least two C_2^+ groups, if any, is generally from 0.3 to 480 hr^{-1} . In a preferred embodiment, in which the polyalkyl aromatic is a diethylbenzene or a triethylbenzene, the molar ratio of benzene per ethylene is from 1.5:1 to 6:1, the WHSV of ethylene is from 0.1 to 6.0 hr^{-1} , and the WHSV of aromatics including benzene and the polyethylbenzenes is from 0.5 to 70 hr^{-1} .

[0033] In the context of ethylbenzene production, the principal reaction that occurs in the reaction zone is the alkylation of the benzene by ethylene to produce ethylbenzene. In addition, other reactions can occur in the reaction zone. For example, benzene can transalkylate with a polyethylbenzene to produce ethylbenzene. Also, polyethylbenzene can be alkylated with ethylene. The reactor effluent stream thus contains ethylbenzene and may also contain unreacted polyethylbenzene or a byproduct of an alkylation side reaction involving the polyethylbenzene or a byproduct of a transalkylation side reaction involving the polyethylbenzene. Although the extent to which other reactions form byproducts is diminished by the practice of the process disclosed herein, the reactor effluent stream usually contains the byproducts of these side reactions. The reactor effluent stream may also contain unreacted benzene, as well as a byproduct of an alkylation side reaction involving benzene or

a byproduct of a transalkylation side reaction involving benzene. In addition, the reactor effluent stream may contain unreacted ethylene, but the concentration of unreacted ethylene is likely to be insignificant because benzene is usually present at least in a stoichiometric proportion. Although it is not common for the feed stream to contain C₁ to C₃ paraffins in addition to ethylene, if ethane is present in the feed stream then the reactor effluent stream may also contain unreacted ethane.

[0034] The alkylation effluent stream contains preferably less than 1.0 wt-% diarylalkane, more preferably less than 0.5 wt-% diarylalkane, and even more preferably less than 0.2 wt-% diarylalkane, relative to the monoalkylated aromatic. These low yields of diarylalkane relative to the monoalkylated aromatic can be attained using an alkylation catalyst bed without using a separate transalkylation catalyst bed that is distinct from the alkylation catalyst bed.

[0035] In practicing the process disclosed herein, the reactor effluent stream can be separated into at least two aliquot portions, in order that an aliquot portion can be recycled and passed to the alkylation reaction zone.

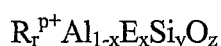
[0036] When one aliquot portion of the alkylation effluent is recycled to and introduced into the alkylation reaction zone, at least one other aliquot portion of the alkylation effluent usually passes to a separation zone for recovering the monoalkylated aromatic. The separation zone generally comprises a benzene fractionation column in order to recycle unreacted benzene to the alkylation zone, and an ethylbenzene fractionation column in order to recover ethylbenzene as product from the heavier polyalkylbenzenes. A polyalkylbenzene fractionation column may also be used in order to separate diethylbenzenes and triethylbenzenes from the other heavier polyalkylbenzenes, particularly where the polyalkylbenzene that is present in the feed stream is a diethylbenzene or a triethylbenzene. The separation zone generally does not comprise a deethanizer unless the concentrations of unreacted ethylene, ethane, or light C₃-minus paraffins in the reactor effluent are high enough to justify their being separated from the reactor effluent stream.

[0037] Thus, in addition to producing a fraction comprising the monoalkylated aromatic, the separation zone may also produce one or more other fractions of the alkylation effluent from the aliquot portion of the alkylation effluent. Accordingly, in addition to recycling an aliquot portion of the alkylation effluent to the alkylation reaction zone, some or all of at least

one of these other fractions recovered from the separation zone can also pass to the alkylation reaction zone. These other recovered fractions can comprise polyethylbenzenes, which in turn can be recycled to the alkylation reaction zone as transalkylation agents. In a commercial ethylbenzene process, several process streams produced by the separation zone can be used to supply such polyethylbenzenes to the alkylation reaction zone.

[0038] The catalyst for the process disclosed herein contains one or more members of the family of aluminosilicate and substituted aluminosilicate zeolites designated UZM-8.

US-B-6,756,030 describes UZM-8 and its preparation. Briefly, UZM-8 zeolites are prepared in an alkali-free reaction medium in which only one or more organoammonium species are used as structure directing agents. In this case, the microporous crystalline zeolite (UZM-8) has a composition in the as-synthesized form and on an anhydrous basis expressed by the empirical formula:



where R is at least one organoammonium cation selected from the group consisting of protonated amines, protonated diamines, quaternary ammonium ions, diquaternary ammonium ions, protonated alkanolamines and quaternized alkanolammonium ions. Preferred organoammonium cations are those that are non-cyclic or those that do not contain a cyclic group as one substituent. Of these those that contain at least two methyl groups as substituents are especially preferred. Examples of preferred cations include without limitation DEDMA, ETMA, HM and mixtures thereof. The ratio of R to (Al + E) is represented by "r" which varies from 0.05 to 5. The value of "p" which is the weighted average valence of R varies from 1 to 2. The ratio of Si to (Al + E) is represented by "y" which varies from 6.5 to 35. E is an element which is tetrahedrally coordinated, is present in the framework and is selected from the group consisting of gallium, iron, chromium, indium and boron. The mole fraction of E is represented by "x" and has a value from 0 to 0.5, while "z" is the mole ratio of O to (Al + E) and is given by the equation

$$z = (r \cdot p + 3 + 4 \cdot y)/2.$$

[0039] The UZM-8 zeolites can be prepared using both organoammonium cations and alkali and/or alkaline earth cations as structure directing agents. As in the alkali-free case above, the same organoammonium cations can be used here. Alkali or alkaline earth cations

are observed to speed up the crystallization of UZM-8, often when present in amounts less than 0.05 M⁺/Si. For the alkali and/or alkaline earth metal containing systems, the microporous crystalline zeolite (UZM-8) has a composition in the as-synthesized form and on an anhydrous basis expressed by the empirical formula:



where M is at least one exchangeable cation and is selected from the group consisting of alkali and alkaline earth metals. Specific examples of the M cations include but are not limited to lithium, sodium, potassium, rubidium, cesium, calcium, strontium, barium and mixtures thereof. Preferred R cations include without limitation DEDMA, ETMA, HM and mixtures thereof.

The value of "m" which is the ratio of M to (Al + E) varies from 0.01 to 2. The value of "n" which is the weighted average valence of M varies from 1 to 2. The ratio of R to (Al + E) is represented by "r" which varies from 0.05 to 5. The value of "p" which is the weighted average valence of R varies from 1 to 2. The ratio of Si to (Al + E) is represented by "y" which varies from 6.5 to 35. E is an element which is tetrahedrally coordinated, is present in the framework and is selected from the group consisting of gallium, iron, chromium, indium and boron. The mole fraction of E is represented by "x" and has a value from 0 to 0.5, while "z" is the mole ratio of O to (Al + E) and is given by the equation

$$z = (m \cdot n + r \cdot p + 3 + 4 \cdot y) / 2$$

where M is only one metal, then the weighted average valence is the valence of that one metal, i.e. +1 or +2. However, when more than one M metal is present, the total amount of

$$M_m^{n+} = M_{m1}^{(n1)+} + M_{m2}^{(n2)+} + M_{m3}^{(n3)+} + \dots$$

and the weighted average valence "n" is given by the equation:

$$n = \frac{m1 \cdot n1 + m2 \cdot n2 + m3 \cdot n3 + \dots}{m1 + m2 + m3 + \dots}$$

[0040] Similarly when only one R organic cation is present, the weighted average valence is the valence of the single R cation, i.e., +1 or +2. When more than one R cation is present, the total amount of R is given by the equation.

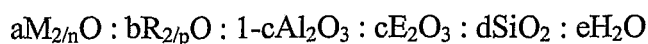
$$R_r^{p+} = R_{r1}^{(p1)+} + R_{r2}^{(p2)+} + R_{r3}^{(p3)+}$$

and the weighted average valence "p" is given by the equation

$$p = \frac{p_1 \bullet r_1 + p_2 \bullet r_2 + p_3 \bullet r_3 + \dots}{r_1 + r_2 + r_3 + \dots}$$

[0041] The microporous crystalline zeolites used in the process disclosed herein are prepared by a hydrothermal crystallization of a reaction mixture prepared by combining reactive sources of R, aluminum, silicon and optionally M and E. The sources of aluminum include but are not limited to aluminum alkoxides, precipitated aluminas, aluminum metal, sodium aluminate, organoammonium aluminates, aluminum salts and alumina sols. Specific examples of aluminum alkoxides include, but are not limited to aluminum ortho sec-butoxide and aluminum ortho isopropoxide. Sources of silica include but are not limited to tetraethylorthosilicate, colloidal silica, precipitated silica, alkali silicates and organoammonium silicates. A special reagent consisting of an organoammonium aluminosilicate solution can also serve as the simultaneous source of Al, Si, and R. Sources of the E elements include but are not limited to alkali borates, boric acid, precipitated gallium oxyhydroxide, gallium sulfate, ferric sulfate, ferric chloride, chromium nitrate and indium chloride. Sources of the M metals include the halide salts, nitrate salts, acetate salts, and hydroxides of the respective alkali or alkaline earth metals. R can be introduced as an organoammonium cation or an amine. When R is a quaternary ammonium cation or a quaternized alkanolammonium cation, the sources include but are not limited the hydroxide, chloride, bromide, iodide and fluoride compounds. Specific examples include without limitation DEDMA hydroxide, ETMA hydroxide, tetramethylammonium hydroxide, tetraethylammonium hydroxide, hexamethonium bromide, tetrapropylammonium hydroxide, methyltriethylammonium hydroxide, tetramethylammonium chloride and choline chloride. R may also be introduced as an amine, diamine, or alkanolamine that subsequently hydrolyzes to form an organoammonium cation. Specific non-limiting examples are N,N,N',N'-tetramethyl-1,6-hexanediamine, triethylamine, and triethanolamine. Preferred sources of R without limitation are ETMAOH, DEDMAOH, and hexamethonium dihydroxide (HM(OH)₂).

[0042] The reaction mixture containing reactive sources of the desired components can be described in terms of molar ratios of the oxides by the formula:



where "a" varies from 0 to 25, "b" varies from 1.5 to 80, "c" varies from 0 to 1.0, "d" varies from 10 to 100, and "e" varies from 100 to 15000. If alkoxides are used, it is preferred to include a distillation or evaporative step to remove the alcohol hydrolysis products. The reaction mixture is now reacted at a temperature of 85°C to 225°C and preferably from 125°C to 150°C for a period of 1 day to 28 days and preferably for a time of 5 days to 14 days in a sealed reaction vessel under autogenous pressure. After crystallization is complete, the solid product is isolated from the heterogeneous mixture by means such as filtration or centrifugation, and then washed with deionized water and dried in air at ambient temperature up to 100°C.

[0043] The UZM-8 aluminosilicate zeolite, which is obtained from the above-described process, is characterized by an x-ray diffraction pattern, having at least the *d*-spacings and relative intensities set forth in Table A below

Table A
d-Spacings and Relative Intensities for as-synthesized UZM-8

2- θ	<i>d</i> (Å)	I/I ₀ %
6.40 - 6.90	13.80 - 12.80	w - s
6.95 - 7.42	12.70 - 11.90	m - s
8.33 - 9.11	10.60 - 9.70	w - vs
19.62 - 20.49	4.52 - 4.33	m - vs
21.93 - 22.84	4.05 - 3.89	m - vs
24.71 - 25.35	3.60 - 3.51	w - m
25.73 - 26.35	3.46 - 3.38	m - vs

[0044] The UZM-8 compositions are stable to at least 600°C (and usually at least 700°C). The characteristic diffraction lines associated with typical calcined UZM-8 samples are shown below in table B. The as-synthesized form of UZM-8 is expandable with organic cations, indicating a layered structure.

Table B
d-Spacings and Relative Intensity for Calcined UZM-8

2- θ	d(Å)	I/I ₀ %
4.05 – 4.60	21.80 – 19.19	w - m
7.00 – 7.55	12.62 – 11.70	m - vs
8.55 – 9.15	10.33 – 9.66	w - vs
12.55 – 13.15	7.05 – 6.73	w
14.30 – 14.90	6.19 – 5.94	m - vs
19.55 – 20.35	4.54 – 4.36	w - m
22.35 – 23.10	3.97 – 3.85	m - vs
24.95 – 25.85	3.57 – 3.44	w - m
25.95 – 26.75	3.43 – 3.33	m - s

[0045] An aspect of the UZM-8 synthesis that contributes to some of its unique properties is that it can be synthesized from a homogenous solution. In this chemistry, soluble aluminosilicate precursors condense during digestion to form extremely small crystallites that have a great deal of external surface area and short diffusion paths within the pores of the crystallites. This can affect both adsorption and catalytic properties of the material.

[0046] As-synthesized, the UZM-8 material will contain some of the charge balancing cations in its pores. In the case of syntheses from alkali or alkaline earth metal-containing reaction mixtures, some of these cations may be exchangeable cations that can be exchanged for other cations. In the case of organoammonium cations, they can be removed by heating under controlled conditions. In the cases where UZM-8 is prepared in an alkali-free system, the organoammonium cations are best removed by controlled calcination, thus generating the acid form of the zeolite without any intervening ion-exchange steps. The controlled calcination conditions include the calcination conditions described herein below for the composite catalyst, and it may sometimes be possible desirable to perform the controlled calcination of the zeolite after the zeolite has been combined with a binder. On the other hand, it may sometimes be possible to remove a portion of the organoammonium via ion exchange. In a special case of ion exchange, the ammonium form of UZM-8 may be

generated via calcination of the organoammonium form of UZM-8 in an ammonia atmosphere.

[0047] The catalyst used in the process disclosed herein preferably contains calcined UZM-8. Calcination of as-synthesized UZM-8 effects changes such as in the x-ray diffraction pattern. The UZM-8 zeolite used in the catalyst used in the process disclosed herein contains preferably less than 0.1 wt-%, more preferably less than 0.05 wt-%, and even more preferably less than 0.02 wt-% of alkali and alkaline earth metals.

[0048] For use in the process disclosed herein, the zeolite preferably is mixed with a binder for convenient formation of catalyst particles in a proportion of 5 to 100 mass % zeolite and 0 to 95 mass-% binder, with the zeolite preferably comprising from 10 to 90 mass-% of the composite. The binder should preferably be porous, have a surface area of 5 to 800 m²/g, and be relatively refractory to the conditions utilized in the hydrocarbon conversion process. Non-limiting examples of binders are aluminas, titania, zirconia, zinc oxide, magnesia, boria, silica-alumina, silica-magnesia, chromia-alumina, alumina-boria, silica-zirconia, etc.; silica, silica gel, and clays. Preferred binders are amorphous silica and alumina, including gamma-, eta-, and theta-alumina, with gamma- and eta-alumina being especially preferred.

[0049] The zeolite with or without a binder can be formed into various shapes such as pills, pellets, extrudates, spheres, etc. Preferred shapes are extrudates and spheres.

Extrudates are prepared by conventional means which involves mixing of zeolite either before or after adding metallic components, with the binder and a suitable peptizing agent to form a homogeneous dough or thick paste having the correct moisture content to allow for the formation of extrudates with acceptable integrity to withstand direct calcination. The dough then is extruded through a die to give the shaped extrudate. A multitude of different extrudate shapes are possible, including, but not limited to, cylinders, cloverleaf, dumbbell and symmetrical and asymmetrical polylobates. It is also within the scope of this invention that the extrudates may be further shaped to any desired form, such as spheres. Spheres can also be prepared by the well known oil-drop method which is described in US-A-2,620,314.

[0050] The catalyst composite is dried at a temperature of from 100° to 320°C for a period of from 2 to 24 or more hours and, usually, calcined at a temperature of from 400° to 650°C in an air atmosphere for a period of from 1 to 20 hours. The calcining in air may be

preceded by heating the catalyst composite in nitrogen to the temperature range for calcination and holding the catalyst composite in that temperature range for from 1 to 10 hours. A catalyst composite used in the process disclosed herein preferably has an x-ray diffraction pattern having at least the d-spacings and relative intensities set forth in Table B.

5 **[0051]** The binder used in the catalyst composite for the process disclosed herein preferably contains less alkali and alkaline earth metals than the UZM-8 zeolite used in the catalyst composite, and more preferably contains little or no alkali and alkaline earth metals. Therefore, the catalyst composite has a content of alkali and alkaline earth metals of less than that of the UZM-8 zeolite used in forming the catalyst composite, owing to the binder
10 effectively lowering the alkali and alkaline earth metals content of the catalyst composite as a whole.

15 **[0052]** In any of the following nine embodiments of the process disclosed herein, any of the feed olefins disclosed herein may be used, such as ethylene, propylene, or butenes. The following embodiments describe the use of benzene as the feed aromatic, but any of the feed aromatics may be used. Also, in any of these embodiments, a fractionation column may be used to recover feed aromatic, such as benzene, as an overhead stream from the alkylation reactor effluent for recycle to the alkylation reactor.

20 **[0053]** In one embodiment, not only an aliquot portion of an alkylation reactor effluent, but also a polyalkylbenzene stream is recycled to the alkylation reactor which contains a UZM-8 zeolite catalyst. The polyalkylbenzene stream typically contains only dialkylbenzenes and trialkylbenzenes, contains a relatively low concentration of alkylbenzene, and does not contain a high concentration of the heaviest polyalkylbenzenes that are produced in the alkylation reactors. The polyalkylbenzene stream is typically recovered as an overhead stream from a polyalkylbenzene fractionation column.

25 **[0054]** In another embodiment, two alkylation reactors are in a series flow arrangement and an aliquot portion of the second alkylation reactor effluent stream is recycled to the first alkylation reactor. Each alkylation reactor contains a UZM-8 zeolite catalyst. At least in theory, the aliquot portion of the second alkylation reactor effluent stream can be recycled at a rate that is limited only by economic considerations. Except for the aliquot portion of the
30 alkylation reactor effluent stream that is passed downstream to product separation facilities, the second alkylation reactor effluent stream is available for recycle in what amounts to an

unlimited quantity. Unlike other streams in the process, the second alkylation reactor effluent stream can be recycled to the alkylation reactors without interfering with the extent to which the alkylation and transalkylation reactions proceed.

[0055] In another embodiment, two alkylation reactors are in a series flow arrangement and an alkylbenzene column bottom stream is recycled along with an aliquot portion of the second alkylation reactor effluent stream to the first reactor. The bottom stream of the alkylbenzene column contains a relatively low concentration of alkylbenzene. Each alkylation reactor contains a UZM-8 zeolite catalyst.

[0056] In another embodiment, an olefin stream is admixed with a benzene stream and with a cooled aliquot portion of alkylation reactor effluent. The combined stream enters an alkylation reactor and contacts a solid alkylation catalyst comprising UZM-8 zeolite to form alkylbenzene. The reactor effluent stream divides into two aliquot portions. One aliquot portion passes to another alkylation reactor or a product separation zone. The other aliquot portion of the reactor effluent stream is cooled by exchanging heat indirectly with boiler feed water to produce steam, and then combines to form the alkylation reactor feed stream.

[0057] In another embodiment, two alkylation reactors are in a series flow arrangement, and each reactor contains two beds of UZM-8 zeolite catalyst that are also in a series flow arrangement. Feed aromatic is charged to the first bed of the first alkylation reactor and feed olefin is charged to each of the beds. The effluent of the first alkylation reactor is cooled in a heat exchanger and flows to the second alkylation reactor. The effluent of the second alkylation reactor divides into two aliquot portions. One aliquot portion flows to a downstream alkylation reactor or to a product separation zone. The other aliquot portion is cooled in a heat exchanger and divides into two aliquot portions. One aliquot flows to the first alkylation reactor and the other aliquot portion flows to the second alkylation reactor.

[0058] In another embodiment, two alkylation reactors are in a series flow arrangement, each alkylation reactor contains a pair of beds of UZM-8 zeolite catalyst, and each of the two beds in a pair are also in a series flow arrangement. Feed aromatic is charged to the first bed of the first alkylation reactor and feed olefin is charged to each of the beds. The effluent of the first alkylation reactor is cooled in a heat exchanger and flows to the second alkylation reactor. The effluent of the second alkylation reactor divides into two aliquot portions. One aliquot portion flows to a downstream alkylation reactor or to a product separation zone. The

other aliquot portion is cooled in a heat exchanger and divides into two aliquot portions. One aliquot portion flows to the first bed of the first alkylation reactor and the other aliquot portion flows to the first bed of the second alkylation reactor.

[0059] In another embodiment, a transalkylation reactor and an alkylation reactor are in a

series flow arrangement. Polyalkylbenzenes and benzene enter any suitable transalkylation reactor and contact any suitable transalkylation catalyst, such as the beta-containing catalysts described in US-B-6,835,862, to form alkylbenzene. See US-A-4,008,290, US-A-4,774,377 and US-A-4,891,458, for example, for suitable transalkylation reactors. The transalkylation reactor effluent is optionally heated or cooled and flows to the alkylation reactor. The alkylation reactor contains two beds of UZM-8 zeolite catalyst in a series flow arrangement, and feed olefin is charged to each bed. The effluent of the alkylation reactor divides into two aliquot portions. One aliquot portion flows to a downstream alkylation reactor or to a product separation zone, and the other aliquot portion is cooled in a heat exchanger and flows to the first bed of the alkylation reactor.

[0060] In another embodiment, an alkylation reactor contains two catalyst beds in a series flow arrangement. Feed olefin and feed aromatic are charged to the first bed, which can be any suitable alkylation reactor containing any suitable alkylation catalyst, such as those described in US-A-4,008,290, US-A-4,774,377, US-A-4,891,458, and US-B-6,835,862. The effluent of the first bed flows to the second bed, which contains a UZM-8 zeolite catalyst.

Feed olefin is charged to the second bed. The effluent of the second bed is withdrawn from the alkylation reactor and divides into two aliquot portions. One aliquot portion flows to a downstream alkylation reactor or to a product separation zone, and the other aliquot portion is cooled in a heat exchanger and flows to the second bed of the alkylation reactor.

[0061] In another embodiment, two alkylation reactors are in a series flow arrangement, each alkylation reactor contains a pair of beds of UZM-8 zeolite catalyst, and each of the two beds in a pair are also in a series flow arrangement. Feed aromatic is charged to the first bed of the first alkylation reactor and feed olefin is charged to each of the beds. The effluent of the first alkylation reactor is cooled in a heat exchanger and flows to the second alkylation reactor. The effluent of the second alkylation reactor flows to a fractionation column, such as a deethanizer or a depropanizer or a debutanizer to remove lighter hydrocarbons from the second alkylation reactor effluent. These lighter hydrocarbons can be ethane and lighter

compounds when producing ethylbenzene, propane and lighter compounds when producing cumene, and butane and lighter compounds when producing butylbenzenes. The bottom stream of this fractionation column divides into two aliquot portions. One aliquot portion flows to a downstream alkylation reactor or to a product separation zone. The other aliquot portion is optionally cooled in a heat exchanger and divides into two aliquot portions. One aliquot portion flows to the first bed of the first alkylation reactor and the other aliquot portion flows to the first bed of the second alkylation reactor.

EXAMPLES

[0062] The following abbreviations will be used in the examples:

Al (Osec-Bu)₃ — aluminum tri-sec-butoxide

DEMAOH — diethyldimethylammonium hydroxide

ETMAOH — ethyltrimethylammonium hydroxide

TMABr — tetramethylammonium bromide

[0063] In the examples that follow, ethylene conversion is defined as the difference between the ethylene expressed in wt-% of the stream entering the reactor and the ethylene expressed in wt-% of the stream exiting the reactor divided by the ethylene expressed in wt-% of the stream entering the reactor. The selectivity of a component is defined as the difference between the number of carbon atoms in molecules of the component exiting the reactor and the number of carbon atoms in molecules of the component entering the reactor, divided by the total number of carbon atoms in molecules of ethylene converted and molecules of benzene converted, and multiplied by 100. The total of the selectivities of ethylbenzene (EB), diethylbenzene (DEB), triethylbenzene (TEB), and tetraethylbenzene (TeEB) accounts for the total selectivity to EB that would be produced if all the diethylbenzene, triethylbenzene, and tetraethylbenzene in the net reactor effluent were transalkylated to EB in a typical transalkylation zone and subsequently recovered. The total of the selectivities of butylbenzene (C₄-Bz), butyl-ethylbenzene (C₄-EB), and butyl-diethylbenzene (C₄-DEB) and the total of the selectivities of diphenylethane (DPE) and ethyldiphenylethane (EDPE) account for undesirable byproducts that cannot be readily transalkylated to EB in a typical transalkylation zone and therefore typically represent a loss of EB production.

EXAMPLE 1

[0064] A fresh alkylation catalyst comprising 70 wt-% zeolite beta and 30 wt-% alumina binder was prepared and is designated as Catalyst A. The zeolite beta for Catalyst A was prepared in substantially the same manner as described in US-A-5,522,984.

5 [0065] An aluminosilicate reaction mixture was prepared in the following manner. A 7329.73 g portion of DEDMAOH (20% aq) was added to a tank. A 804.38 g portion of Al (Osec-Bu)₃ (95%+) was added to the tank, and the resulting solution was thoroughly mixed for 45 min. A 2000 g quantity of deionized water was then added to the solution, followed by the addition of a 2526.96 g portion of precipitated silica (Ultrasil™ VN SP3, 89% SiO₂). Next, a
10 solution of 126.69 g of NaOH dissolved in 212.25 g of deionized water was prepared and added to the reaction mixture and the reaction mixture was thoroughly mixed for 30 min. The reaction mixture was then transferred to a 19-L stirred reactor. The tank was rinsed with 1000 g of deionized water and the rinse was transferred to the reactor and mixed into the reaction mixture. The reaction mixture was heated in 3 hr to 150°C and digested at 150°C for 290 hr. A solid
15 product was collected by filtering, washed with deionized water, and dried at 50°C. The isolated product was identified as UZM-8 by powder x-ray diffraction analysis. Elemental analysis revealed the composition of the isolated product to consist of the elemental mole ratios of Si/Al=11.77, Na/Al=0.26, N/Al=2.03, and C/N=3.04. The isolated product was ammonium ion-exchanged using an ion exchange solution of 1 part by weight of NH₄NO₃, 10 parts by
20 weight of deionized water, and 1 part by weight of the isolated product at 75°C for 3 hr, and the solids were collected by filtering. The ammonium ion exchange and filtration was repeated two more times, and the triple ammonium ion-exchanged material was washed with deionized water and dried at 50°C. A sample of the dried material was calcined by heating to 540°C and holding at that temperature for 2 hr in the presence of flowing nitrogen, and then
25 switching to flowing air and holding at that temperature for 14 hr. Thereafter the BET surface area was found to be 481 m²/g and the micropore volume was 0.14 cc/g. Another sample of the dried material was then formulated into a catalyst comprising 80 wt-% UZM-8 and 20 wt-% alumina. The extrusion was done using HNO₃-peptized Al₂O₃ as a binder and 3.0 wt-% based on the weight of the UZM-8 and the alumina of Solka-Floc™ powdered cellulose (BW-
30 40; International Fiber Corp., North Tonawanda, NY, USA) as an extrusion aid to obtain 1.6 mm (1/16 in) diameter extrudates. The extrudates were activated in a muffle oven by heating

to 538°C and holding at that temperature for 1 hr in the presence of flowing nitrogen, and then switching to flowing air and holding at that temperature for 15 hr. This catalyst is designated as Catalyst B.

[0066] An aluminosilicate reaction mixture was prepared in the following manner. In a beaker, a 32.77 g portion of Al (Osec-Bu)₃ (95%+) was dissolved in 538.39 g of DEDMAOH (20% aq) and thoroughly mixed for 10 min. A 500 g quantity of deionized water was added to the beaker. A 175 g portion of precipitated silica (Ultrasil™ VN SP3, 89% SiO₂) was slowly added, and the reaction mixture was thoroughly mixed for 10 min. In a separate beaker, a 6.58 g portion of NaBr and a 23.86 g portion of TMABr were dissolved in 123 g of deionized water, the resulting solution was added to the reaction mixture and then the reaction mixture was thoroughly mixed for 10 min. Next, a 14 g (on a volatile free basis) portion of finely ground UZM-8 seed was added to the reaction mixture, and the reaction mixture was thoroughly mixed for 20 min. The reaction mixture was then transferred to a 2-L stirred reactor and heated to 150°C in 2 hr and digested at 150°C for 166 hr. A solid product was collected by centrifugation, washed with deionized water, and dried in air. The isolated product was identified as UZM-8 by powder x-ray diffraction analysis. Elemental analysis revealed the composition of the isolated product to consist of the elemental mole ratios of Si/Al=15.69 and Na/Al=0.74. The isolated product was ammonium ion-exchanged, washed, and dried in the manner described for Catalyst B. One sample of the dried material was calcined in the manner described for Catalyst B, and thereafter for this sample the BET surface area was found to be 413 m²/g and the micropore volume was 0.146 cc/g. Another sample of the dried material was formulated into a catalyst comprising 80 wt-% UZM-8 and 20 wt-% alumina and activated in the manner described for Catalyst B, except that 0.5 wt-% based on the weight of the UZM-8 and the binder of Methocel™ methylcellulose (A4M; Dow Chemical Co., Midland, MI, USA) was used as the extrusion aid. This catalyst is designated as Catalyst C.

[0067] An aluminosilicate reaction mixture was prepared in the following manner. In a baffled tank, a 6 g (on a volatile free basis) portion of dry UZM-8 seed was added to 706.2 g of deionized water and thoroughly mixed. In a separate beaker, a 38.3 g portion of liquid sodium aluminate, a 108.5 g portion of ETMAOH (20% aq), and a 4 g portion of a 50% NaOH solution were mixed, and then added to the tank. Next, 137.2 grams of precipitated silica (Ultrasil™ VN SP3, 89% SiO₂) was added in 5 minutes to the tank, and the reaction

mixture was thoroughly mixed for 20 minutes. The reaction mixture was then transferred to a 2-L stirred reactor. The reaction mixture was then heated in 2 hr to 150°C and digested at 150°C for 165 hr. A solid product was collected by centrifugation, washed with deionized water, and dried at 50°C. The isolated product was identified as UZM-8 by powder x-ray diffraction analysis. Elemental analysis revealed the composition of the isolated product to consist of the elemental mole ratios of Si/Al=10.22, Na/Al=0.49, N/Al=0.74, and C/N=5.21. The isolated product was ammonium ion-exchanged, washed, and dried in the manner described for Catalyst B. One sample of the dried material was calcined in the manner described for Catalyst B, and thereafter for this sample the BET surface area was found to be 505 m²/g and the micropore volume was 0.134 cc/g. Another sample of the dried material was then formulated into a catalyst and activated in the manner described for Catalyst B, except the catalyst comprised 70 wt-% UZM-8 and 30 wt-% alumina. This catalyst is designated as Catalyst D.

[0068] An aluminosilicate reaction mixture was prepared in the following manner. In a beaker, a 84.88 g portion of Al (Osec-Bu)₃ (95%+) was added to a 386.73 g portion of DEDMAOH (20% aq) with vigorous mixing. A 266.66 g portion of precipitated silica (Ultrasil™ VN SP3, 89% SiO₂) was added to the reaction mixture. A quantity of 600 g of deionized water was added to the reaction mixture. Then a solution containing 13.37 g of NaOH dissolved in 48 g of deionized water was added to the reaction mixture with mixing. The reaction mixture was thoroughly mixed for 20 minutes. A 140 g portion of slurry UZM-8 seed was added to the reaction mixture and was thoroughly mixed for 20 minutes. A slurry UZM-8 seed is a reaction mixture used to make UZM-8 zeolite after digestion and prior to isolation by filtering, centrifugation, or other means. The reaction mixture was then transferred to a 2-L stirred reactor, heated to 150°C in 2 hr, and then digested at 150°C for 216 hr. A solid product was collected by filtering, washed with deionized water, and dried at 50°C. The isolated product was identified as UZM-8 by powder x-ray diffraction analysis. Elemental analysis revealed the composition of the isolated product to consist of the elemental mole ratios of Si/Al=11.00, Na/Al=0.47, N/Al=1.17, and C/N=5.53. The isolated product was ammonium ion-exchanged, washed, and dried in the manner described for Catalyst B. One sample of the dried material was calcined in the manner described for Catalyst B, and thereafter for this sample the BET surface area was found to be 493 m²/g and the micropore volume was 0.13

cc/g. Another sample of the dried material was formulated into a catalyst and activated in the manner described for Catalyst D. This catalyst is designated as Catalyst E.

[0069] An aluminosilicate reaction mixture was prepared in the following manner. In a tank, a 2115.55 g portion of DEDMAOH (20% aq) was added to 10 kg deionized water. A
5 781.88 g portion of liquid sodium aluminate was added slowly to the reaction mixture, and the reaction mixture was thoroughly mixed for 20 min. A 2877.01 g portion of precipitated silica (Ultrasil™ VN SP3, 89% SiO₂) was slowly added to the reaction mixture and thoroughly mixed for 20 min. A 1400 g portion of slurry UZM-8 seed was added to the reaction mixture and the reaction mixture was mixed thoroughly for an additional 20 minutes. The reaction
10 mixture was then transferred to a 19-L stirred reactor. The tank was rinsed with 525 g of deionized water and the rinse was transferred to the reactor and mixed into the reaction mixture. The reaction mixture was heated to 150°C in 6 hours and digested at 150°C for 138 hours. A solid product was collected by centrifugation, washed with deionized water, and dried at 50°C. The isolated product was identified as UZM-8 by powder x-ray diffraction analysis. Elemental
15 analysis revealed the composition of the isolated product to consist of the elemental mole ratios of Si/Al=10.46, Na/Al=0.49, N/Al=0.59, and C/N=6.22. The isolated product was ammonium ion-exchanged, washed, and dried in the manner described for Catalyst B. One sample of the dried material was calcined in the manner described for Catalyst B, and thereafter for this sample the BET surface area was found to be 530 m²/g and the micropore volume was 0.135
20 cc/g. Another sample of the dried material was formulated into a catalyst and activated in the manner described for Catalyst D. This catalyst is designated as Catalyst F.

EXAMPLE 2

[0070] The experimental procedure used in the tests for Example 1 was as follows. A volume of the catalyst to be tested was loaded into a cylindrical reactor. The reactor was
25 equipped with a thermocouple in a thermowell located to measure temperatures at distances along the length of the fixed catalyst bed. Dry benzene was passed through the reactor at 260°C and at a benzene LHSV of 6.7 hr⁻¹ for 24 hours.

[0071] Subsequently, the flow of fresh benzene was adjusted and the reactor inlet temperature was lowered to a temperature 50°C below the desired distance average bed
30 temperature (DABT) for the initial testing conditions. As used herein, DABT means the

temperature calculated by plotting the catalyst bed temperature versus distance along the catalyst bed, computing the area under the curve from the inlet to the outlet of the catalyst bed, and dividing the area by the length of the catalyst bed. Fresh ethylene was introduced into the reactor. Then a portion of the reactor effluent was recycled so that a combined feed of the fresh benzene, the fresh ethylene, and the recycled reactor effluent flowed to the reactor. The reactor inlet temperature was adjusted to maintain the desired DABT while the reactor effluent was sampled and analyzed. Then the reactor inlet temperature and/or the amount of recycled reactor effluent was adjusted, and the reactor effluent was sampled again. This process was repeated until measurements and samples were obtained at all of the desired DABTs and effluent recycle ratios (R/FF). It is believed that minimal catalyst deactivation occurred over the duration that the performance of each catalyst was measured.

[0072] The ethylene WHSV was 0.76 hr^{-1} for the tests of Catalyst A and 0.90 to 0.92 hr^{-1} for the tests of Catalysts B and C. The molar ratio of fresh benzene to fresh ethylene for these tests was 2. The R/FF for these tests was 4 or 8. Because the molar ratio of aryl groups per ethyl group is essentially the same in the combined reactor feed stream and the total reactor effluent stream, the molar ratio of aryl groups per ethyl group is not significantly affected by recycling any portion of the reactor effluent stream.

[0073] Results are shown in Figures 1-4. Additional results of tests done at DABTs from 214 to 218°C and from 239 to 244°C are shown in Table 1. Each result in Figures 1-4 and in Table 1 is an average from measurements and/or analyses at each test condition. Figure 1 is a graph of total selectivity to (EB + DEB + TEB + TeEB) versus selectivity to EB and shows that, in comparison with a process using Catalyst A, a process using Catalyst B or C has a higher total selectivity to (EB + DEB + TEB + TeEB) at a given EB selectivity. Figure 2 is a graph of total selectivity to (C₄-Bz + C₄-EB + C₄-DEB) versus DABT and shows that a process using Catalyst B or C has a lower total selectivity to (C₄-Bz + C₄-EB + C₄-DEB) at a given DABT than a process using Catalyst A. Figure 3 is a graph of total selectivity to (DPE + EDPE) versus DABT and shows that a process using Catalyst B or C has a lower total selectivity to (DPE + EDPE) at a given DABT than a process using Catalyst A. These results indicate that an alkylation-transalkylation combination process using Catalysts B or C for alkylation would produce more EB than a process using Catalyst A.

[0074] Tests 3 and 5 for Catalyst B in Table 1 show that despite nearly doubling the R/FF from 4.1 to 8.1 the selectivity to EB and each of the total selectivities to (EB + DEB + TEB + TeEB), (C₄-Bz + C₄-EB + C₄-DEB), and (DPE + EDPE) remained nearly the same.

[0075] At the same DABT and R/FF, a change in Si/Al₂ molar ratio of the UZM-8 zeolite from 31.38 (Catalyst C) to 23.54 (Catalyst B) provided essentially the same or higher selectivity to EB and total selectivity to (EB + DEB + TEB + TeEB). At 4.1-4.3 R/FF, tests 4 and 7 show this at 217-218°C DABT while tests 3 and 6 show this at 244°C DABT. In addition, the total selectivity to (C₄-Bz + C₄-EB + C₄-DEB) decreased.

EXAMPLE 3

[0076] The experimental procedure used in the tests for Example 3 was the same as for Example 2, except that for the tests for Example 3 the catalyst was Catalyst D, the ethylene WHSV was 0.89 hr⁻¹, and the molar ratio of fresh benzene to fresh ethylene was 1.8. In Example 3, the R/FF was 4.3 during tests 1 and 2. Subsequently, the reactor effluent recycle was stopped during test 3. It is believed that minimal catalyst deactivation occurred over the duration that the tests were performed.

[0077] The results shown in Table 2 are averages from measurements and/or analyses at each test condition. Table 2 shows that at nearly the same ethylene conversion recycling reactor effluent increased the EB selectivity and the total selectivity to (EB + DEB + TEB + TeEB), and decreased the total selectivities to (C₄-Bz + C₄-EB + C₄-DEB) and (DPE + EDPE). In addition, recycling reactor effluent decreased by at least 66% the selectivities to non-aromatics and to compounds heavier than DPE and EDPE.

EXAMPLE 4

[0078] A volume of the catalyst to be tested was loaded into the cylindrical reactor described in Example 1. Dry benzene was passed through the reactor at 260°C and at a benzene LHSV of 6.7 hr⁻¹ for 24 hours. Subsequently, the flow of fresh benzene was adjusted and the reactor inlet temperature was lowered to a temperature 50°C below the desired DABT for the initial testing conditions. Fresh ethylene was introduced into the reactor. The molar ratio of fresh benzene to fresh ethylene for these tests was 2. Then a portion of the reactor effluent was recycled so that a combined feed of the fresh benzene, the fresh ethylene, and the

recycled reactor effluent flowed to the reactor. The reactor inlet temperature was adjusted to maintain the desired DABT. The temperature within the catalyst bed rose as the incoming feed contacted the catalyst due to the exothermic nature of the reaction. At times during a period (e.g., 100 hours) at test conditions, temperature profiles (bed temperature versus distance through the bed) were plotted. The rate of catalyst deactivation was taken to be the rate of progression of these temperature profiles through the bed. The position of each temperature profile was defined by the end of the active zone, which was a measure of the end of the temperature rise in the temperature profile. On a temperature profile, the end of the active zone was the distance in the bed at the intersection of the linear extrapolation of the linear part of the temperature rise and a horizontal line at the maximum bed temperature. After the rate of deactivation was determined at one set of test conditions, the rates of fresh ethylene and fresh benzene were increased (higher LHSV) in order to accelerate the rate of deactivation.

[0079] The results are shown in Table 3. At the low WHSV conditions and at comparable times on stream, the deactivation rate of Catalyst E was one-tenth that of Catalyst A. After the ethylene WHSV was increased, the deactivation rate of Catalyst A decreased only slightly and the end of the active zone continued to move toward the outlet of the catalyst bed as Catalyst A continued to deactivate. During comparable times on stream, the deactivation rate of Catalyst E also decreased and remained much less than that of Catalyst A. Indeed, at the higher WHSV conditions the deactivation rate of Catalyst E decreased by so much that the end of the active zone stopped moving toward the outlet of the catalyst bed and instead moved toward the inlet of the catalyst bed. This is evidenced by the negative slope (i.e., -0.0004) of the plot of the position of the end of the active catalyst zone versus time on stream. This in turn indicates that the using Catalyst E with reactor effluent recycle in an alkylation process would permit increased ethylene throughput and therefore increased alkylate production for longer periods of time than would use of Catalyst A.

EXAMPLE 5

[0080] The experimental procedure used in the tests for Example 5 was the same as for Example 2, except that Catalyst F was used, the olefin was propylene, propylene WHSV was 1.07 hr⁻¹, and the molar ratio of fresh benzene to fresh propylene was 2.9. The R/FF was 4.9,

7.2, and 11.4, and the propylene concentration in the combined feed was 2.7 wt-%, 1.9 wt-%, and 1.3 wt-%, respectively. The molar ratio of aryl groups per propyl group is essentially the same in the combined reactor feed stream and the total reactor effluent stream, and therefore the molar ratio of aryl groups per propyl group is not significantly affected by recycling any portion of the reactor effluent stream. It is believed that minimal catalyst deactivation occurred over the duration that the tests were performed.

[0081] The results shown in Figure 4 are averages from measurements and/or analyses at each test condition. Figure 4 shows that recycling more reactor effluent increased the total selectivity to cumene (IPB), dipropylbenzene (DPB), and tripropylbenzene (TPB).

EXAMPLE 6

[0082] The experimental procedure used in the tests for Example 6 was the same as for Example 4, except that Catalyst E was used, the olefin was propylene, and the molar ratio of fresh benzene to fresh propylene was 2.4.

[0083] The results are shown in Table 4. At the low WHSV conditions and at comparable times on stream, the deactivation rate of Catalyst E was 38% of that of Catalyst A. After the ethylene WHSV was increased, the deactivation rate of Catalyst A increased by a factor of 5 while during comparable times on stream the deactivation rate of Catalyst E decreased.

Table 1

Catalyst	A		B			C	
Test	1	2	3	4	5	6	7
Recycle / fresh feed ratio, wt/wt	4.3	4.3	4.1	4.1	8.1	4.3	4.3
Olefin concentration in combined feed, wt-%	2.9	2.9	3.0	3.0	1.7	2.9	2.9
Inlet temperature, °C	193	168	194	172	198	195	170
DABT, °C	243	214	244	218	239	244	217
Ethylene WHSV, hr ⁻¹	0.76	0.76	0.90	0.90	0.90	0.92	0.92
Ethylene conversion, mol-%	99.83	99.92	99.92	99.88	99.88	99.91	99.63
Selectivity, mol-%							
EB	75.70	71.15	75.61	76.76	75.90	75.11	75.77
EB + DEB + TEB + TeEB	98.55	99.26	99.64	99.75	99.64	99.65	99.65
C ₄ -Bz + C ₄ -EB + C ₄ -DEB	0.43	0.21	0.06	0.06	0.06	0.10	0.14
DPE + EDPE	0.16	0.08	0.05	0.01	0.05	0.02	0.01

Table 2

Catalyst	D		
Test	1	2	3
Recycle / fresh feed ratio, wt/wt	4.3	4.3	0
Olefin concentration in combined feed, wt-%	3.2	3.2	17.1
Inlet temperature, °C	208	163	226
Temperature rise, °C	45	44	37
Ethylene conversion, mol-%	99.7	99.7	99.9
Selectivity, mol-%			
EB	73.1	73.1	60.9
EB + DEB + TEB + TeEB	99.55	99.61	97.76
C ₄ -Bz + C ₄ -EB + C ₄ -DEB	0.06	0.13	0.65
DPE + EDPE	0.05	0.02	0.08

Table 3

Catalyst	A	E
Low WHSV Conditions		
Recycle / fresh feed ratio, wt/wt	4.3	4.2
Olefin concentration in combined feed, wt-%	2.9	2.9
Inlet temperature, °C	168	168
DABT, °C	215	215
Ethylene WHSV, hr ⁻¹	0.77	0.84
Deactivation rate, distance / unit time	0.003	0.0003
High WHSV Conditions		
Recycle / fresh feed ratio, wt/wt	2.9	2.8
Olefin concentration in combined feed, wt-%	4.0	4.0
Inlet temperature, °C	172	174
Ethylene WHSV, hr ⁻¹	1.12	1.26
Deactivation rate, distance / unit time	0.002	-0.0004

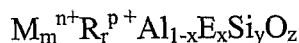
Table 4

Catalyst	A	E
Low WHSV Conditions		
Recycle / fresh feed ratio, wt/wt	4.8	4.9
Olefin concentration in combined feed, wt-%	3.2	3.1
Inlet temperature, °C	140	141
DABT, °C	161	161
Ethylene WHSV, hr ⁻¹	0.91	1.07
Deactivation rate, distance / unit time	0.0016	0.0006
High WHSV Conditions		
Recycle / fresh feed ratio, wt/wt	3.0	3.0
Olefin concentration in combined feed, wt-%	4.7	4.7
Inlet temperature, °C	141	142
Ethylene WHSV, hr ⁻¹	1.88	2.03
Deactivation rate, distance / unit time	0.0085	0.0005

CLAIMS:

1. An alkylation process for producing a monoalkylated aromatic, the process comprising:

a) passing an aromatic feedstock comprising a feed aromatic, an olefinic feedstock comprising a C₂-C₄ olefin, and an additional stream comprising an alkylated derivative of the feed aromatic having from one to six more C₂-C₄ alkyl groups than the feed aromatic, to an alkylation catalyst bed containing a solid catalyst, wherein the solid catalyst comprises a microporous crystalline zeolite having a layered framework of at least AlO₂ and SiO₂ tetrahedral units and a composition on an as-synthesized and anhydrous basis expressed by an empirical formula of:



where M is at least one exchangeable cation, "m" is the mole ratio of M to (Al + E) and varies from 0 to 2.0, R is at least one organoammonium cation selected from the group consisting of quaternary ammonium cations, diquaternary ammonium cations, protonated amines, protonated diamines, protonated alkanoamines and quaternized alkanolammonium cations, "r" is the mole ratio of R to (Al + E) and has a value of 0.05 to 5.0, "n" is the weighted average valence of M and has a value of 1 to 2, "p" is the weighted average valence of R and has a value of 1 to 2, E is an element selected from the group consisting of gallium, iron, boron, chromium, indium and mixtures thereof, "x" is the mole fraction of E and has a value from 0 to 1.0, "y" is the mole ratio of Si to (Al + E) and varies from 6.5 to 35 and "z" is the mole ratio of O to (Al + E) and has a value determined by the equation:

$$z = (m \cdot n + r \cdot p + 3 + 4 \cdot y)/2$$

and is characterized in that it has the x-ray diffraction pattern having at least the d spacings and intensities set forth in Table A:

Table A

2- θ	d(Å)	I/I ₀ %
6.40 - 6.90	13.80 - 12.80	w - s
6.95 - 7.42	12.70 - 11.90	m - s
8.33 - 9.11	10.60 - 9.70	w - vs
19.62 - 20.49	4.52 - 4.33	m - vs
21.93 - 22.84	4.05 - 3.89	m - vs
24.71 - 25.35	3.60 - 3.51	w - m
25.73 - 26.35	3.46 - 3.38	m - vs

b) alkylating the feed aromatic with the C₂-C₄ olefin in the alkylation catalyst bed in the presence of the solid catalyst at alkylation conditions to form a monoalkylated aromatic, wherein the monoalkylated aromatic has one more C₂-C₄ alkyl group than the feed aromatic; and

c) withdrawing an effluent stream comprising the monoalkylated aromatic from the alkylation catalyst bed.

2. The process of claim 1 wherein the alkylation conditions comprise a concentration of the C₂-C₄ olefin based on the weight of the feed aromatic, the C₂-C₄ olefin, and the alkylated derivative of the feed aromatic passed to the alkylation catalyst bed in Step (a) of at most 17 wt-%.

3. The process of claim 1 or 2 wherein the alkylated derivative of the feed aromatic has two more C₂-C₄ alkyl groups than the feed aromatic.

4. The process of claim 1, 2, or 3 further characterized in that the additional stream comprises an aliquot portion of the effluent stream and the alkylation conditions comprise a weight ratio of the aliquot portion of the effluent stream to the feed aromatic and the C₂-C₄ olefin passed to the alkylation catalyst bed in Step (a) of at least 0.1.

5. The process of any of claims 1, 2, 3, or 4 further characterized in that the alkylation conditions comprise a molar ratio of aryl groups to the C₂-C₄ alkyl group of at most 6.

6. The process of any of claims 1, 2, 3, 4, or 5 wherein the feed aromatic is selected from the group consisting of benzene, naphthalene, anthracene, tetralin, phenanthrene, and alkylated derivatives thereof.

7. The process of any of claims 1, 2, 3, 4, 5, or 6 wherein the feed aromatic comprises benzene, the feed olefin comprises ethylene, and the monoalkylated aromatic comprises ethylbenzene.

8. The process of any of claims 1, 2, 3, 4, 5, 6, or 7 wherein the microporous crystalline zeolite is ion-exchanged and contains less than 0.1 wt-% of alkali and alkaline earth metals.

9. The process of any of claims 1, 2, 3, 4, 5, 6, 7, or 8 wherein the microporous crystalline zeolite is calcined and is characterized in that it has the x-ray diffraction pattern having at least the d spacings and intensities set forth in Table B:

Table B

2-θ	d(Å)	I/I ₀ %
4.05 – 4.60	21.80 – 19.19	w - m
7.00 – 7.55	12.62 – 11.70	m - vs
8.55 – 9.15	10.33 – 9.66	w - vs
12.55 – 13.15	7.05 – 6.73	w
14.30 – 14.90	6.19 – 5.94	m - vs
19.55 – 20.35	4.54 – 4.36	w - m
22.35 – 23.10	3.97 – 3.85	m - vs
24.95 – 25.85	3.57 – 3.44	w - m
25.95 – 26.75	3.43 – 3.33	m - s

10. The process of any of claims 1, 2, 3, 4, 5, 6, 7, 8, or 9 wherein the microporous crystalline zeolite is UZM-8.