

A Method for Recovering the Regeneration Heat in a Process for Preparing Lower Olefins from Methanol

5 FIELD OF THE INVENTION

This invention relates to a method for recovering regeneration heat, specifically, to a method for recovering the heat carried by the high temperature catalyst after regeneration in a process for preparing lower olefins from methanol.

10 BACKGROUND

Lower olefins such as ethylene, propylene and the like are basic raw materials for chemical industry. Conventionally, ethylene and propylene is mainly derived from the steam cracking processes of hydrocarbon feedstock, such as naphtha, diesoline, hydro-cracked tail oil and the like. Recently, with the substantial increment of petroleum price, the production
15 cost for obtaining ethylene and propylene from the above raw materials is increasing continuously. Meanwhile, the conventional methods for producing ethylene and propylene mainly employ a cracking process using high temperature tubular furnace and the energy consumption is relatively high. These factors urge the development of new processes for the production of olefins.

20 The routes of preparing lower olefins from non-petroleum raw materials have drawn much attention recently, wherein a conversion route from coal or natural gas into methanol *via* synthesis gas, and then from methanol into lower olefins, has been of broad interest. The process for forming lower (C₂-C₄) olefins selectively on a molecular sieve catalyst from methanol (or the dimethyl ether formed by the dehydration of methanol firstly) is commonly
25 referred to as MTO process.

Recently, a fluid bed MTO process employing a continuous reaction-regeneration has been attracting much attention. The basic principle for this fluid bed process is as follows: Methanol, as feedstock, is mixed with a catalyst in a reactor to make the latter fluidized, and then it converts into a product mixture containing ethylene, propylene and the like at a
30 certain temperature, while the catalyst is deactivated partly or totally by carbon deposit

thereon after the reaction. The gaseous reaction product flows out of the reactor and enters into a fractionation plant, and the deactivated catalyst flows continuously out of the reactor, and enters into a regenerator to be regenerated, that is, removing the deposited carbon by combustion in an oxygen-containing atmosphere, and then returns to the reactor to contact with the reaction feedstock again.

In the above fluid bed process for continuous reaction-regeneration, the carbon deposited on the surface of the deactivated catalyst is removed in a manner of high temperature combustion. Generally, the temperature of the combustion reaction is above 600 °C and can be up to 700 °C or higher. If the loss of heat is not taken into account, the heat produced by the combustion of the deposited carbon will be transferred out of the regenerator in two manners: a part of the heat will be taken away by a high temperature regeneration flue gas, and the other part of the heat will be taken away by the high temperature regenerated catalyst.

On the one hand, the heat carried by the high temperature regeneration flue gas is generally recovered by the methods of producing steam or generating electricity or the like. For example, US Patent US20050238543 A1 disclosed a method for recovering heat from a regeneration flue gas, comprising reducing the temperature of the regeneration flue gas by several times of heat exchange and using the extracted heat for producing steam and the like.

On the other hand, the heat carried by the high temperature regenerated catalyst is often used for supplying heat for reaction. Conventionally, a fluid bed reaction process is generally used for the endothermic reactions such as the catalytic cracking of hydrocarbons or the like wherein a part of the reaction heat is provided by the high temperature catalyst regenerated in the process for regeneration. That is, the deactivated catalyst is regenerated and heated by combusting the deposited carbon in an oxygen-containing atmosphere within a regenerator, then returned to the reactor, and simultaneously, the heat is transferred from the former to the latter so as to provide at least a part of the reaction heat.

However, such a manner of heat utilization is not applicable to the reaction for converting methanol into olefins. Because the conversion of methanol into olefins is a strongly exothermal reaction and the catalyst is heated in the reactor, the stabilization of the reaction temperature can not be maintained unless the temperature of the catalyst entering

into the reactor is lower than that of the reaction bed. The preferable reaction temperature for the conversion of methanol into lower olefins is 350-600°C which is lower than the regeneration temperature for burning the carbon on the catalyst (600-700°C), that is, the temperature of the regenerated catalyst. Therefore, in order to meet the demand of the reaction for converting methanol into lower olefins, the high temperature regenerated catalyst has to release the heat obtained in the process for regeneration, and the temperature thereof has to be lowered before entering into the reactor.

DISCLOSURE OF THE INVENTION

10 An object of the invention is to provide a method for recovering the heat of a high temperature regenerated catalyst in a process for preparing lower olefins from methanol.

The inventors completed the invention through an intensive research.

In an aspect of the invention, a method for recovering heat from the high temperature regenerated catalyst in a process for preparing lower olefins from methanol is provided, which comprises the steps of: leading the high temperature regenerated catalyst firstly into a heat absorption zone of cracking reaction, and then into a methanol conversion reactor, wherein

hydrocarbons are introduced into said heat absorption zone of cracking reaction and contacted with said high temperature regenerated catalyst to occur a cracking reaction; and
20 the temperature of said high temperature regenerated catalyst ranges from 500 to 800 °C.

In a preferable aspect of the invention, the catalyst used is an aluminosilicate zeolite or/and a phosphosilicate molecular sieve catalyst and an element-modified product thereof, having a micropore diameter of 0.3-0.6 nm.

25 In another preferable aspect of the invention, the matrix material of the catalyst is one or more of silica, alumina or clay.

In a preferable aspect of the invention, the temperature of the heat absorption zone of cracking reaction is 400-700°C.

In another preferable aspect of the invention, the difference in temperature of the catalyst between the inlet and the outlet of the heat absorption zone of cracking reaction is
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50-300°C.

In a preferable aspect of the invention, the hydrocarbons introduced into the heat absorption zone of cracking reaction are the C₄ or higher products in the reaction of methanol to olefins, or/and other hydrocarbons of C₄-C₂₀.

5 In another preferable aspect of the invention, the hydrocarbons introduced into the heat absorption zone of cracking reaction are naphtha, gasoline, gas condensate, diesoline, hydrogenation tail oil or/and kerosene.

In a preferable aspect of the invention, the product produced in the heat absorption zone of cracking reaction is joined into the product obtained from the preparation of lower olefins from methanol.

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In another preferable aspect of the invention, said heat absorption zone of cracking reaction is an independent reaction section in the methanol conversion unit.

In a preferable aspect of the invention, said heat absorption zone of cracking reaction is a riser of the methanol conversion reactor.

15 By using the method involved in the invention, a part of the heat carried by the regenerated catalyst in the process for preparing lower olefins from methanol can be recovered, and the temperature of the regenerated catalyst can be adjusted to meet the temperature demand of the reaction for preparing olefins from methanol.

20 SPECIFIC MODE OF CARRYING OUT THIS INVENTION

In order to realize the above object, the method provided by the present invention for recovering the regeneration heat in the process for preparing lower olefins from methanol, is that, the high temperature regenerated catalyst is led into a heat absorption zone of cracking reaction, in which the heat carried by the catalyst is absorbed through a catalytic cracking reaction of hydrocarbons, and then the catalyst is led into a methanol conversion reactor when the temperature thereof is lowered.

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In said method, the catalyst used is an aluminosilicate zeolite or/and a phosphosilicate molecular sieve catalyst and an element-modified product thereof, having a micropore diameter of 0.3-0.6 nm.

30 In said method, the matrix material of the catalyst is one or more of silica, alumina or

clay.

In said method, the temperature of the heat absorption zone of cracking reaction is 400-700°C.

5 In said method, the difference in temperature of the catalyst between the inlet and the outlet of the heat absorption zone of cracking reaction is 50-300°C.

In said method, the hydrocarbons introduced into the heat absorption zone of cracking reaction are the C₄ or higher products in the reaction of methanol to olefins, or/and other hydrocarbons of C₄-C₂₀.

10 In said method, the hydrocarbons introduced into the heat absorption zone of cracking reaction are naphtha, gasoline, gas condensate, diesoline, hydrogenation tail oil or/and kerosene.

In said method, the product containing ethylene and propylene produced in the heat absorption zone of cracking reaction is joined into the product obtained from methanol conversion.

15 According to the invention, a part of the regeneration heat in the process for preparing lower olefins from methanol can be recovered using the endothermic hydrocarbon cracking reaction. The invention is characterized by that, in a reaction-regeneration fluid bed process for preparing lower olefins from methanol with a feature of continuous reaction, the high temperature regenerated catalyst, before being contacted with methanol, is led firstly into a
20 heat absorption zone of cracking reaction, in which the high temperature regenerated catalyst is contacted with the introduced C₄-C₂₀ hydrocarbons so that the heat carried by the catalyst is absorbed using the hydrocarbon cracking reaction, and then the catalyst is led into a methanol conversion reactor when its temperature is lowered to meet the temperature demand of methanol conversion. Simultaneously, the lower olefins produced by the cracking
25 reaction can be added into the olefin product formed from methanol.

The invention provides a method for recovering a part of the heat in the process for preparing lower olefins from methanol as follows: in the process for preparing lower olefins from methanol using a fluid bed process, methanol feedstock is mixed with a catalyst in a reactor to fluidize the latter and converted into a product mixture containing ethylene,
30 propylene and other hydrocarbons at a certain temperature; the catalyst is deactivated partly

or totally by the carbon deposit thereon after the reaction; the gaseous reaction product flows out of the reactor and enters a separation apparatus, while the deactivated catalyst continuously flows out of the reactor and enters a regenerator to be regenerated; the deactivated catalyst, before being entered into the regenerator, is passed through a stripping
5 apparatus to remove the hydrocarbons remained on the catalyst with an inert gas such as steam or the like, and then burned in a oxygen-containing atmosphere in the regenerator to remove the deposited carbon; the combustion of the deposited carbon releases heat wherein a part of this heat is taken away by a regeneration flue gas, while the other part is taken away by the regenerated catalyst; the catalyst heated to 600-700°C in the regenerator, after
10 the residual oxygen therein is removed by a inert gas such as steam or the like, is led into a heat absorption zone of cracking reaction which can be either an independent dense phase reaction section, or a riser which conveys the catalyst into the methanol conversion reactor simultaneously. In this heat absorption zone of cracking reaction, the hydrocarbon feedstock contacts with the regenerated catalyst to occur a cracking reaction so as to absorb the heat
15 carried by the catalyst. The temperature of the reaction endothermic zone is 400-700°C, and after the catalyst is flowed out of the above heat absorption zone of reaction, the temperature thereof is 50-300°C lower than that at the inlet of the heat absorption zone of reaction so that it attains the temperature required by the reaction for converting methanol into olefins, and then the catalyst is entered into a reactor for converting methanol into olefins. The
20 product containing ethylene and propylene produced in the above heat absorption zone of cracking reaction can be joined into the product of the methanol conversion.

The catalyst described above contains a aluminosilicate zeolite or/and a phosphosilicate molecular sieve catalyst with a micropore diameter of 0.3-0.6 nm, such as ZSM-5, ZSM-11, SAPO-34, SAPO-11 or the like, and an element modified product thereof.
25 The catalyst described above also contains a matrix material which is one or more of silica, alumina or clay. The hydrocarbons used in the above heat absorption zone of reaction have a carbon number between 4-20, which can be either a C₄ or higher product in the reaction for producing olefins from methanol, or other hydrocarbons of C₄-C₂₀, including naphtha, gasoline, gas condensate, diesoline, hydrogenation tail oil and kerosene or the like.

SPECIFIC EXAMPLES

The invention is described in detail below by way of examples, but the invention is not limited thereto.

5 Example 1:

The cracking of 2-butylene on SAPO-34 molecular sieve catalyst. The preparation process for the catalyst was as follows: SAPO-34 (Dalian Institute of Chemical Physics, the ratio of silicon to aluminum was 0.2), clay, alumina sol and silica sol (all purchased from Zhejiang Yuda Chemical Engineering Co. Ltd.) were mixed and dispersed into water
10 to form a slurry, which was then spray-dried to form microspheres with a particle diameter distribution of 20-100 microns. The microspheres described above, after calcination at 600 °C for 4 h, became the catalyst used in this example. The content of SAPO-34 in the catalyst was 30% by weight. The reaction was performed in a fluid bed microreactor with diameter of 20 mm under the reaction conditions of: a catalyst charge of 10 g, a feeding
15 weight hourly space velocity (WHSV) for 2-butylene (Fushun Petrochemical Co., purity of 98%, the ratio of *cis*- and *trans*-2-butylene being 1) of 1.0 hr⁻¹ and a reaction pressure of 0.1 MPa. The reaction product was analyzed by using a Varian CP-3800 gas chromatography with a Pona column and a FID. The sampling time was 2 min.

The reaction heat was obtained from a calculation based on the thermodynamic
20 constants of the raw material and the product, wherein the thermodynamic data of C₅ was the average value of 10 kinds of C₅ hydrocarbons including paraffins, olefins, dialkenes and naphthene isomers, and that of C₆ was the average value of 12 kinds of C₆ hydrocarbons including paraffins, olefins, dialkenes and naphthene isomers. Each substance was presumed to be an ideal gas under the reaction condition.

25 The reaction result of the conversion of 2-butylene on the molecular sieve catalyst and the reaction heat were shown in Table 1. The reaction temperature was 500°C. As shown by the data of this condition in the table, the selectivity of ethylene and propylene in the reaction product was 78.1%, and under the condition of this product distribution, the reaction heat was 471 kJ/Kg.

Table 1: Cracking reaction products of 2-butylene and reaction heat in Example 1 (500°C)

Product distribution (wt %)	CH ₄	C ₂ H ₄	C ₂ H ₆	C ₃ H ₆	C ₃ H ₈	C ₅	C ₆ ⁺
	0.62	18.40	0.25	59.70	10.70	10.01	0.31
ΔH(J/Kmol)	2.26*10 ⁷						
Reaction heat (KJ/Kg)	4.71*10 ²						

C₆⁺ represented the product of C₆ or higher.

Example 2:

- 5 Except that the reaction temperature was 550°C, the procedure described in Example 1 was repeated.

The reaction result of the conversion of 2-butylene on the molecular sieve catalyst and the reaction heat were shown in Table 2. As shown by the data of this condition in the table, the selectivity of ethylene and propylene in the reaction product was 82.21%, and under the
10 condition of this product distribution, the reaction heat was 704 kJ/Kg.

Table 2: Cracking reaction products of 2-butylene and reaction heat in Example 2 (550°C)

Product distribution (wt %)	CH ₄	C ₂ H ₄	C ₂ H ₆	C ₃ H ₆	C ₃ H ₈	C ₅	C ₆ ⁺
	1.01	27.30	0.44	54.91	7.81	5.39	3.14
ΔH(J/Kmol)	3.38*10 ⁷						
Reaction heat (KJ/Kg)	7.04*10 ²						

C₆⁺ represented the product of C₆ or higher.

15 Example 3:

Except that the reaction temperature was 600°C, the procedure described in Example 1 was repeated.

The reaction result of the conversion of 2-butylene on the molecular sieve catalyst and the reaction heat were shown in Table 3. As shown by the data of this condition in the table,
20 the selectivity of ethylene and propylene in the reaction product was 80.97%, and under the

condition of this product distribution, the reaction heat was 825 kJ/Kg.

Table 3: Cracking reaction products of 2-butylene and reaction heat in Example 3 (600°C)

Product distribution (wt %)	CH ₄	C ₂ H ₄	C ₂ H ₆	C ₃ H ₆	C ₃ H ₈	C ₅	C ₆ ⁺
	4.16	44.03	1.20	36.94	6.40	1.69	5.57
$\Delta H(\text{J/Kmol})$	3.96*10 ⁷						
Reaction heat (KJ/Kg)	8.25*10 ²						

C₆⁺ represented the product of C₆ or higher.

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Example 4:

The cracking of kerosene on ZSM-5 molecular sieve catalyst. The preparation steps of the catalyst and the reaction operation were the same as those in Example 1 except that SAPO-34 molecular sieve catalyst therein was replaced by ZSM-5 (The molecular sieve plant of Nankai University, the ratio of silicon to aluminum was 50) and the reaction feedstock was replaced by kerosene (No.3 aviation kerosene, Qilu Petrochemical Co.). The calculating method of reaction heat was the same as that in Example 1 and the combustion enthalpy of kerosene was taken as -7513 kJ·Kg⁻¹.

The reaction result of kerosene conversion on the molecular sieve catalyst and the reaction heat were shown in Table 4. The reaction temperature was 550 °C. Other thermodynamic functions of the kerosene were in terms of n-dodecane. As shown by the data of this condition in the table, the selectivity of ethylene and propylene in the reaction product was 28.91%, and under the condition of this product distribution, the reaction heat was 2238 kJ/Kg.

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Table 4: Cracking reaction products of kerosene and reaction heat in Example 4 (550°C)

Product distribution (wt %)	CH ₄	C ₂ H ₄	C ₂ H ₆	C ₃ H ₆	C ₃ H ₈	C ₄	C ₅ ⁺
	1.70	10.61	2.78	18.30	11.22	22.36	33.03
Reaction heat (KJ/Kg)	22.38*10 ²						

C_5^+ represented the product of C_5 or higher.

Example 5:

Except that the reaction temperature was 600°C , the procedure described in Example 4 was repeated.

The reaction result of kerosene conversion on the molecular sieve catalyst and the reaction heat were shown in Table 5. The combustion enthalpy of kerosene was taken as $-7513 \text{ kJ}\cdot\text{Kg}^{-1}$ and other thermodynamic functions of kerosene were in terms of n-dodecane. As shown by the data of this condition in the table, the selectivity of ethylene and propylene in the reaction product was 36.97%, and under the condition of this product distribution, the reaction heat was 2821 kJ/Kg .

Table 5: Cracking reaction products of kerosene and reaction heat in Example 5 (600°C)

Product distribution (wt %)	CH_4	C_2H_4	C_2H_6	C_3H_6	C_3H_8	C_4	C_5^+
	5.36	14.80	5.12	22.17	4.61	15.16	32.78
Reaction heat (KJ/Kg)	28.21×10^2						

C_5^+ represented the product of C_5 or higher.

Example 6:

The reaction result of gasoline conversion on ZSM-5 sieve catalyst and the reaction heat were shown in Table 6. The preparation steps of the catalyst and reaction operation were the same as those in Example 4 except that the reaction feedstock was replaced by gasoline (a catalytic gasoline of Fushun Petrochemical Co., the olefin content was 40%). The reaction temperature was 640°C . The thermodynamic functions of gasoline were in terms of the average value of various amylenes. As shown by the data of this condition in the table, the yield of ethylene and propylene in the reaction product was 44.01%, and under the condition of this product distribution, the reaction heat was 361 kJ/Kg .

Table 6: Cracking reaction products of gasoline and reaction heat in Example 6 (640°C)

Product yield (wt %)	CH ₄	C ₂ H ₄	C ₂ H ₆	C ₃ H ₆	C ₃ H ₈	C ₄	C ₅ ⁺
	2.52	17.22	1.96	26.79	2.18	10.99	38.34
Reaction heat (KJ/Kg)	3.61*10 ²						

C₅⁺ represented the product of C₅ or higher.

Example 7:

Except that the reaction temperature was 610°C, the procedure described in Example 6
5 was repeated.

The reaction result of the conversion of gasoline on sieve catalyst and the reaction heat
were shown in Table 7. The thermodynamic functions of gasoline were in terms of the
average value of various amylenes. As shown by the data of this condition in the table, the
yield of ethylene and propylene in the reaction product was 39.68%, and under the condition
10 of this product distribution, the reaction heat was 367 kJ/Kg.

Table 7: Cracking reaction products of gasoline and reaction heat in Example 7 (610°C)

Product yield (wt %)	CH ₄	C ₂ H ₄	C ₂ H ₆	C ₃ H ₆	C ₃ H ₈	C ₄	C ₅ ⁺
	2.28	16.63	1.98	23.05	2.85	9.45	43.76
Reaction heat (KJ/Kg)	3.67*10 ²						

C₅⁺ represented the product of C₅ or higher.

15 Example 8:

In terms of 300 days ordinary operation on average per year, a unit for producing
olefins from methanol with a methanol conversion capacity of 600 thousand tons/year has a
methanol treating capacity of 2000 tons per day. This unit applies a reaction-regeneration
fluid bed process, mainly comprising a methanol conversion reactor, a catalyst regenerator
20 and a heat absorption cracking reactor, wherein a manner of fluid bed was adopted, on a
route conveying catalyst from the generator to the methanol conversion reactor. The high
temperature regenerated catalyst was passed through the above heat absorption cracking
reactor and then led into the methanol conversion reactor.

SAPO-34 fluid catalyst (the preparation steps of the catalyst were the same as that in

Example 1) was used and the recycle amount of the catalyst was 2000 tons/day. The temperature of the regenerated catalyst (after stripping) was 650°C and the catalyst before being entered into the methanol conversion reactor was required to be reduced to a temperature of 430°C.

5 In the heat absorption cracking reactor described above, the catalytic cracking of 2-butylene (for example derived from Example 1) was utilized to absorb the heat carried by the regenerated catalyst and make its temperature lowered to 550°C. The heat capacity of the catalyst was 840J/ (Kg·K), and the heat released by the catalyst during the temperature decreasing was 1.68×10^8 KJ/day. The catalyst was flowed out of the heat absorption
10 cracking reactor described above, and before it was entered into the methanol conversion reactor, its temperature was reduced to be 430°C by the heat dispersion of the line and the loss of the heat dispersion was 2.02×10^8 KJ/day. In order to control the heat dispersion loss of the line, the catalyst conveying line was needed to be subjected to an adequate insulation.

In the heat absorption cracking reactor, 2-butylene was pre-heated to 200°C for
15 feeding, and the reaction temperature was 550°C. 2-butylene has a heat capacity of $C_p=1456$ J/ (Kg·K). According to Example 2, the reaction heat for the catalytic cracking of 2-butylene was taken as 704 KJ/Kg. The heat absorption of 2-butylene per ton from feeding to completing the reaction was 12.14×10^5 KJ/Kg and the single-pass conversion of 2-butylene was 75%, a feeding amount of 184 tons/day for 2-butylene can make the catalyst
20 temperature reduced from 650°C to 550°C.

In the reaction process described above, the heat recoverable by the catalytic cracking of 2-butylene comprises 45% of the heat carried by the regenerated catalyst, and by applying the recovered heat to the catalytic cracking of 2-butylene, the productions of ethylene and propylene can be increased by 37.8 tons/day and 76.0 tons/day, respectively.

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Comparative example 1:

In terms of 300 days ordinary operation on average per year, a unit for producing olefins from methanol with a methanol conversion capacity of 600 thousand tons/year has a methanol treating capacity of 2000 tons per day. A reaction-regeneration fluid bed process
30 similar to that of Example 8, mainly included a methanol conversion reactor and a catalyst

regenerator, was utilized by this conversion unit, except that no heat absorption zone of cracking reaction was disposed and the catalyst was flowed directly from the regenerator to the methanol conversion reactor through a conveying route.

SAPO-34 fluid catalyst, the preparation steps of which was the same as those in
5 Example 1, was used and the recycle amount of the catalyst was 2000 tons/day. The temperature of the regenerated catalyst (after stripping) was 650°C and was required to be reduced to 430°C before being entered into the methanol conversion reactor.

When the catalyst temperature was reduced to be 430°C totally through the heat dispersion of the conveying line and the heat capacity of the catalyst was 840 J/(Kg·K), the
10 loss of heat dispersion during the temperature decreasing was 3.70×10^8 KJ/day.

We claim:

1. A method for recovering heat from the high temperature regenerated catalyst in a process for preparing lower olefins from methanol, comprising the steps of: leading the high
5 temperature regenerated catalyst firstly into a heat absorption zone of cracking reaction, and then into a methanol conversion reactor, wherein

hydrocarbons are introduced into said heat absorption zone of cracking reaction and contacted with said high temperature regenerated catalyst to occur a cracking reaction, wherein the heat carried by the high temperature regenerated catalyst is absorbed through the catalytic cracking
10 reaction of hydrocarbons, and wherein 1 kg of the hydrocarbons absorbs 361-2821 KJ of the heat; and

the temperature of said high temperature regenerated catalyst ranges from 500 to 800°C.

2. The method as claimed in claim 1, wherein the catalyst used is an aluminosilicate
15 zeolite or/and a phosphosilicate molecular sieve catalyst, and an element-modified product thereof, having a micropore diameter of 0.3-0.6 nm.

3. The method as claimed in claim 1, wherein the matrix material of the catalyst is one or
20 more of silica, alumina or clay.

4. The method as claimed in claim 1, wherein the temperature of the heat absorption zone of cracking reaction is 400-700°C.

5. The method as claimed in claim 1, wherein the difference in temperature of the
25 catalyst between the inlet and the outlet of the heat absorption zone of cracking reaction is 50-300°C.

6. The method as claimed in claim 1, wherein the hydrocarbons introduced into the heat absorption zone of cracking reaction are the C₄ or higher products in the reaction of
30 methanol to olefins, or/and other hydrocarbons of C₄-C₂₀.

7. The method as claimed in claim 1, wherein the hydrocarbons introduced into the

heat absorption zone of cracking reaction are naphtha, gasoline, gas condensate, diesoline, hydrogenation tail oil or/and kerosene.

8. The method as claimed in claim 1, wherein the product produced in the heat
5 absorption zone of cracking reaction is joined into the product obtained from the preparation of lower olefins from methanol.

9. The method as claimed in claim 1, wherein said heat absorption zone of cracking
reaction is an independent reaction section in the methanol conversion unit.

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10. The method as claimed in claim 1, wherein said heat absorption zone of cracking
reaction is a riser of the methanol conversion reactor.