**The Promoting Effect of Various Metals on the Activity of Molybdenum-Based Catalysts**

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**Abstract.** The study examined the enhancing influence of different metals on the performance of molybdenum-based catalysts in the catalytic aromatisation of propane-butane fractions and the kinetics of the process. The reaction was conducted in a differential reactor at 500-600 °C with a propane-butane fraction volume flow rate of 1000 h-1 and a pressure of 0.1 MPa. The reaction products were examined using gas chromatography on a "Kristallux-4000M" chromatograph. The tests revealed that the yield of aromatic hydrocarbons is enhanced in the presence of a catalyst exhibiting high catalytic activity, specifically 5.0% Mo, 2.0% Zn, and 2.0% Zr. The catalyst's catalytic activity was assessed by including metals into its composition as different compounds and evaluating its performance. In determining the ideal quantity of zinc, attention was paid not only to the maximal output of aromatic hydrocarbons but also to the sustained operational lifespan of the highly active catalyst.

**Keywords**: propane-butane fraction, low molecular weight hydrocarbons, aromatization, catalyst, promoter, kinetic model, activation energy.

##### INTRODUCTION

Modern oil refining and petrochemical processes are based on catalytic technologies. Catalytic reforming is considered one of the fundamental processes in oil refining, enabling the production of high-octane motor fuel components and individual aromatic hydrocarbons such as benzene, toluene, and xylene [1-3].

To meet the stringent requirements imposed on oil refining processes, the international market offers the latest catalytic technologies, including ultra-deep hydrotreating [4-8], processes of chain branching and hydrogen release, as well as hydrogen recycle [8-10]. The need to develop, upgrade, and phase out obsolete technologies has become paramount in oil and gas processing, driven by increasing demands for the quality of gasoline as a motor fuel, primarily its octane number.

For modified high-silica mesoporous aluminosilicates, the selectivity for aromatic hydrocarbons (AHs) in the n-hexane conversion process increases up to 60% [10-12]; however, the yield of key substances in the process decreases significantly. For a catalyst exhibiting good catalytic activity and selectivity in the catalytic aromatisation of n-hexane to aromatic hydrocarbons, the proportions of propane and methane are roughly 15% each, while that of ethane is around 5%. The mechanism of aromatization is a highly complex process. In addition to the viewpoints mentioned above, a mechanism for activating lower molecular weight saturated hydrocarbon molecules has been proposed, involving hydride ion abstraction by carbenium ions or initial dehydrogenation, which is determined by the nature of protonation [12-17].

Recently, due to increasingly stringent environmental requirements, reforming processes have been increasingly directed toward producing high-octane gasoline components with a high content of aromatic hydrocarbons (AHs) or iso-saturated hydrocarbons. In petroleum chemistry and industrial refining, high-silica mesoporous aluminosilicates serve as catalysts exhibiting significant catalytic activity and selectivity for hydrocarbon conversion, particularly in the catalytic aromatisation of linear normal alkanes, such as n-hexane, into aromatic hydrocarbons [18-24].

At present, reforming is advancing in two areas: the synthesis of aromatic hydrocarbons (as precursors for organic synthesis) and the generation of high-octane components for automotive fuels – isoparaffins. It is essential to enhance the reforming process by differentiating between aromatisation and isomerisation catalysts. The modification of alumozirconium-containing catalysts is a viable approach for acquiring highly active and selective aromatisation catalysts. Optimising the reforming process and developing novel catalysts is crucial for the synthesis of environmentally friendly motor fuel components, including isoparaffins.

##### EXPERIMENTAL PART

This research employed a high-silica zeolite (HSZ) derived from kaolin in the Pakhtachi district of Uzbekistan as the porous substrate. The catalyst with exceptional catalytic activity was synthesised as follows: 100g of HSZ was saturated with a 30% aqueous solution of (NH4)2MoO4, Zn(NO3)2, and ZrO(NO3)2 for 12 hours. Subsequently, the highly catalytically active catalyst was extracted from the solution and dried in a nitrogen stream at 350-400°C for 3 hours, subsequently shaped into granules measuring 5-7 mm in diameter.

The highly catalytically active catalyst (V=1.0 cm3) was positioned in a tubular quartz reactor with a diameter of 12 mm. Before the reaction, the highly catalytically active catalyst was heated in a helium stream to 750°C and maintained for 30 minutes, after which a propane-butane fraction was introduced at a volumetric flow rate of 1000 h-1.

The studies were performed in a flow-type catalytic unit utilising a fixed bed of a highly active catalyst (catalyst volume of 6 cm3), at temperatures ranging from 500 to 650°C, at atmospheric pressure (P=0.1 MPa), and with a propane-butane volumetric flow rate of 400-600 h-1.

The highly catalytically active catalyst (V=1.0 cm3) was positioned within a tubular quartz reactor with a diameter of 12 mm. Before the reaction, the highly catalytically active catalyst was heated in a helium stream to 750°C and maintained for 30 minutes, after which a gas mixture was introduced at a volumetric flow rate of 1000 h-1. The reaction products were examined via gas chromatography employing a "Crystallux-4000M" chromatograph equipped with a thermal conductivity detector. The separation of H2, O2, N2, CH4, and CO was conducted using a 2-meter-long column with a 3 mm inner diameter, filled with mesoporous high-silica zeolite 5A. The separation temperature was isothermal at 70°C, utilising argon as the carrier gas with a flow rate of 20 mL/min.

The separation of C2 hydrocarbons and CO2 was conducted using a column of 2 meters in length and 3 millimetres in inner diameter. The separation temperature was isothermal at 80°C, utilising helium as the carrier gas, with a flow rate of 20 mL/min. The separation of C3-C4 hydrocarbons was conducted using a 4 m long column with a 3 mm inner diameter, filled with alumina coated with Vaseline oil. The separation temperature was isothermal at 20°C, the carrier gas utilised was helium, and the gas flow rate was 20 mL/min. The original and resultant chemicals were analysed using the “Crystallux-4000M” chromatograph, employing a column (dimensions 3.5x0.5 cm) filled with fluorinated Al2O3.

The analysis conditions for the initial materials and gaseous reaction products were as follows: a mesoporous high-silica zeolite with a specific surface area of 257 m2/g was used as the adsorbent. The adsorbent was impregnated with 25% by mass of hexadecane (liquid phase). The column length was 6 m with an inner diameter of 3 mm.

Analysis of Hydrogen. The analysis of hydrogen and hydrocarbon components in the gaseous products was performed on a column packed with a mesoporous high-silica zeolite 5A.

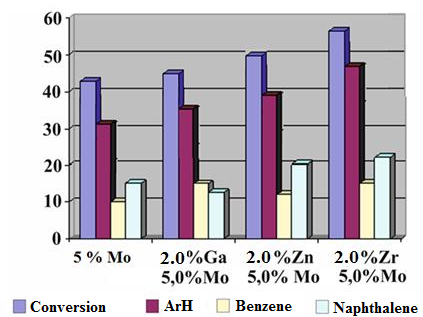
The mesoporous high-silica zeolite 5A adsorbs all hydrocarbons (except methane) at temperatures up to 100°C. Therefore, the chromatograms recorded signals specific to hydrogen and methane. The analysis conditions were: bridge current katharometer Im=100mA; thermostat temperature TT = 30°C, katharometer temperature TK = 40°C, carrier gas flow rate 30 mL/min.

Analysis of Liquid Products (Catalyzate). The analysis of liquid products was performed on a 50 m long, 3 mm inner diameter quartz capillary column coated with SE-30 liquid phase under the following optimal conditions: initial thermostat temperature T0 = 50°C; temperature ramp rate 5°C/min, hold time at maximum temperature 10 minutes, vaporizer temperature T =300°C, carrier gas (helium) flow rate 1 mL/min.

##### RESULTS AND DISCUSSION

We evaluated the promotional effects of several metals on a molybdenum-based catalyst exhibiting significant catalytic activity. The optimal results were achieved by incorporating zirconium into a highly active molybdenum-based catalyst.

The experimental findings are illustrated in Figure 1.



**FIGURE 1.** Molybdenum-based catalyst with high catalytic activity due to the promoting properties of various metals

Figure 1 illustrates that the high-silica mesoporous zeolite catalyst derived from bentonite, comprising 2.0% Zr and 5.0% Mo, exhibits significant catalytic activity.

To enhance the reaction yield, we included Zr and Zn metals into the chosen high-catalytic catalyst and conducted the reaction. The production of aromatic hydrocarbons was elevated with a high-catalytic catalyst including 5.0% Mo, 2.0% Zr, and 2.0% Zn.

The catalytic efficacy of a high-performance catalyst is contingent not only upon its qualitative and quantitative composition but also on the manner in which the active components are incorporated into the catalyst's formulation. Therefore, we introduced spirit in the form of various compounds into the composition of the high-catalytic catalyst and tested its catalytic activity. Initial tests were carried out by absorbing spirit acetate and spirit nitrate into YCC (high silica zeolite), with the spirit content of 2 and 5% (by mass) at a temperature of 500-650 °C. The studies lasted 120 minutes, and the volume flow rate of the initial gas mixture was adjusted between 1000 and 1200 h-1. At temperatures below 550 °C, the conversion of saturated hydrocarbons with a composition of C1-C4 decreases sharply, while at temperatures above 650 °C, coke formation increases and selectivity decreases. The results acquired are displayed in Tables 1 and 2.

The yield of aromatic hydrocarbons increases as the volumetric flow rate of the starting mixture diminishes. As the temperature rises, the conversion of saturated hydrocarbons including C1-C4 and the yield of aromatic hydrocarbons increase, while selectivity diminishes. Simultaneously, the concentration of benzene in the catalyst rises due to the hydrodealkylation of toluene, while the quantities of hydrogen and methane in the gas phase also escalate.

**TABLE 1.** Dependence of the aromatization process of C1-C4 saturated hydrocarbons on temperature, nature of promoter, and volume rate of raw material flow

|  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- |
| **A highly active catalyst based on High-Silica Zeolite (HSZ)** | **t, 0C** | **V, c-1** | **Conversion,%** | **Selective exposure,%** | **Product,%** |
| **2 % Zn**  **Zn(CH3COO)2** | 550 | 1200 | 32,0 | 30,6 | 9,8 |
| 1000 | 49,8 | 65,5 | 32,6 |
| 600 | 1200 | 43,0 | 55,3 | 23,8 |
| 1000 | 71,5 | 57,2 | 40,9 |
| 650 | 1200 | 76,7 | 40,0 | 30,7 |
| 1000 | 96,5 | 41,7 | 40,4 |
| **A highly active catalyst based on High-Silica Zeolite (HSZ)** | **t, 0C** | **V, c-1** | **Conversion,%** | **Selective exposure,%** | **Product,%** |
| **5 % Zn**  **Zn(CH3COO)2** | 600 | 1200 | 50,1 | 50,7 | 25,4 |
| 1000 | 69,3 | 60,7 | 42,1 |
| 650 | 1200 | 81,7 | 30,7 | 25,1 |
| **2 % Zn**  **Zn(NO3)2** | 550 | 1200 | 54,0 | 61,3 | 33,1 |
| 600 | 1200 | 66,7 | 52,3 | 34,9 |
| 1000 | 70,6 | 56,8 | 40,1 |
| 625 | 1200 | 84,3 | 48,0 | 40,5 |
| 1000 | 79,7 | 60,6 | 48,3 |
| 650 | 1200 | 90,1 | 43,6 | 39,3 |
| 1000 | 89,4 | 54,6 | 48,8 |
| **5 % Zn**  **Zn(NO3)2** | 550 | 1200 | 58,6 | 55,1 | 32,3 |
| 600 | 1200 | 75,6 | 52,6 | 39,8 |
| 1000 | 73,6 | 53,3 | 39,2 |
| 625 | 1200 | 90,8 | 43,2 | 39,3 |
| 1000 | 89,3 | 59,1 | 52,8 |
| **2% Zn**  **ZnO** | 550 | 1200 | 46,0 | 49,3 | 22,7 |
| 600 | 1000 | 90,3 | 32,3 | 29,2 |
| 1200 | 69,5 | 44,3 | 30,8 |
| 600 | 1200 | 73,5 | 52,8 | 38,8 |
| 625 | 1200 | 91,5 | 44,0 | 40,3 |

**TABLE 2.** Dependence of the composition of reaction gases on temperature, nature of the promoter, and volumetric flow rate of the feedstock during the aromatization of saturated hydrocarbons with C1-C4 composition

|  |  |  |  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| **A highly active catalyst based on High-Silica Zeolite (HSZ)** | **T, 0C** | **V, c-1** | **Composition of the gases produced, %** | | | | | | |
| **H2** | **CH4** | **C2H6** | **C2H4** | **C3H8** | **C3H6** | **∑C4** |
| **2 % Zn**  **Zn(CH3COO)2** | 550 | 1200 | 28,9 | 4,5 | 2,0 | 2,3 | 40,0 | 3,7 | 18,6 |
| 1000 | 46,4 | 12,8 | 8,5 | 1,6 | 24,9 | 2,5 | 3,4 |
| 600 | 1200 | 50,5 | 9,6 | 5,8 | 2,7 | 21,8 | 5,3 | 4,3 |
| 1000 | 49,3 | 18,2 | 12,4 | 2,2 | 13,1 | 3,3 | 1,1 |
| 625 | 1200 | 51,6 | 17,9 | 11,3 | 3,2 | 10,2 | 4,6 | 1,2 |
| 1000 | 51,0 | 6,6 | 19,0 | 16,5 | 1,6 | 3,8 | 1,5 |
| **5 % Zn**  **Zn(CH3COO)2** | 600 | 1200 | 40,1 | 14,2 | 6,6 | 4,0 | 25,5 | 4,8 | 4,9 |
| 1000 | 41,7 | 22,4 | 12,0 | 3,9 | 14,1 | 4,4 | 1,6 |
| **2%Zn**  **Zn(NO3)2** | 550 | 1200 | 46,8 | 15,1 | 9,1 | 1,7 | 25,1 | 1,8 | 0,3 |
| 575 | 1200 | 53,9 | 16,8 | 8,9 | 2,0 | 16,0 |  | 2,3 |
| 1000 | 53,3 | 19,2 | 7,8 | 2,8 | 14,4 | 1,8 | 0,9 |
| 600 | 1200 | 51,6 | 24,5 | 14,1 | 1,6 | 6,6 | 1,5 | 0,1 |
| 1000 | 58,1 | 18,4 | 8,8 | 3,4 | 9,1 | 1,7 | 0,6 |
| 625 | 1200 | 61,3 | 21,2 | 10,0 | 2,1 | 4,2 |  | 1,3 |
| 1000 | 56,3 | 24,6 | 10,2 | 2,7 | 4,7 | 1,6 |  |
| **5%Zn**  **Zn(NO3)2** | 550 | 1200 | 46,9 | 15,1 | 10,8 | 1,3 | 23,0 |  | 3,1 |
| 575 | 1200 | 49,1 | 21,0 | 14,7 | 1,3 | 12,1 |  | 2,0 |
| 1000 | 47,8 | 21,6 | 14,5 | 1,4 | 13,0 |  | 1,6 |
| 600 | 1200 | 47,7 | 28,6 | 17,3 | 1,5 | 3,9 |  | 1,1 |
| 1000 | 52,4 | 24,8 | 15,0 | 1,4 | 5,3 |  | 1,2 |
| **5%Zn**  **ZnO** | 550 | 1200 | 39,9 | 12,4 | 4,5 | 3,0 | 31,7 | 2,6 | 6,0 |
| 575 | 1200 | 52,8 | 17,1 | 8,0 | 2,1 | 17,5 | 2,2 | 0,5 |
| 1000 | 54,3 | 24,0 | 11,8 | 1,6 | 6,6 | 1,3 | 0,3 |
| 600  625 | 1200 | 57,7 | 17,9 | 8,1 | 2,4 | 11,0 | 2,1 | 0,9 |
| 1200 | 57,5 | 22,6 | 9,7 | 2,1 | 3,8 | 1,4 | 3,0 |

In determining the ideal zinc quantity, both the greatest output of aromatic hydrocarbons and the sustained operating lifespan of the highly active catalyst were taken into account.

**TABLE 3.** The dependence of the catalyst composition on temperature and the nature of the promoter in the aromatization process of saturated hydrocarbons with a C1-C4 composition, V=1000h-1

|  |  |  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- | --- | --- |
| **A highly active catalyst based on High-Silica Zeolite (HSZ)** | **t, 0C** | **Catalyst composition, % mass** | | | | | | |
|  |  | ∑C6 | C6H6 | C7H9 | ∑C8H10 | ∑C9H12 | C10H8 | C11+ |
| **2 % Zn**  **Zn(CH3COO)2** | 550 | 1,3 | 25,7 | 46,6 | 21,6 | 0,4 | 2,3 | 2,1 |
| 600 | 2,0 | 35,6 | 44,4 | 12,5 | 0,7 | 2,8 | 2,1 |
| 625 | 1,3 | 25,7 | 46,6 | 21,6 | 0,4 | 2,3 | 2,1 |
| **2%Zn**  **Zn(NO3)2** | 575 | - | 37,3 | 39,9 | 11,9 | 5,9 | 3,4 | 1,6 |
| 600 | - | 37,4 | 40,0 | 0,8 | 17,0 | 3,4 | 1,6 |
| 625 | - | 39,3 | 37,7 | 10,0 | 7,5 | 3,8 | 1,8 |
| **5%Zn**  **Zn(NO3)2** | 600 | - | 32,1 | 35,9 | 11,2 | 0,2 | 11,8 | 8,8 |

As the zinc concentration rises from 2% to 5%, the quantity of naphthenes in the catalyst correspondingly increases (see Table 3). The findings derived from the zinc acetate solution in water were inferior to those obtained with zinc nitrate.

Table 4 presents information regarding the influence of zirconium in mesoporous high-silica zeolite on propane conversion and the selectivity of reaction product formation.

**TABLE 4.** Effect of zirconium oxide content in mesoporous high-silica zeolite on propane conversion and   
selectivity of reaction product formation

|  |  |  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- | --- | --- |
| **A highly active catalyst** | **T, 0C** | **X, %** | **A,%** | **Selective influence on the formation of products, %** | | | | |
| **H2** | **CH4** | **Alkanes,C2-C5** | **Alkenes C2-C4** | **Arenes** |
| **1 %Zr – HSZ** | 500 | 21 | 0,9 | 1,3 | 21,1 | 62,5 | 5,8 | 4,4 |
| 550 | 43 | 2,7 | 1,5 | 35,3 | 43,2 | 13,7 | 6,4 |
| 575 | 75 | 16,2 | 2,0 | 38,3 | 30,0 | 13,2 | 21,6 |
| 600 | 93.2 | 34,0 | 2,4 | 33,9 | 15,8 | 14,8 | 31,6 |
| **2 %Zr- HSZ** | 500 | 16 | 0,8 | 0,5 | 19,8 | 57,1 | 17,4 | 5,2 |
| 550 | 33 | 3,1 | 1,1 | 29,3 | 38,8 | 21,6 | 9,2 |
| 575 | 58 | 4,4 | 1,6 | 38,0 | 26,8 | 25,9 | 7,7 |
| 600 | 88 | 24,9 | 2,0 | 32,8 | 16,1 | 21,0 | 27,2 |
| **3 % Zr- HSZ** | 500 | 8 | 0,3 | 0,3 | 19,2 | 50,9 | 25,1 | 4,5 |
| 550 | 21 | 0,7 | 0,9 | 27,3 | 39,6 | 28,9 | 3,2 |
| 575 | 42 | 1,8 | 1,5 | 34,9 | 26,1 | 33,1 | 4,3 |
| 600 | 72 | 5,0 | 2,6 | 40,9 | 18,7 | 30,9 | 7,0 |
| **4 % Zr- HSZ** | 500 | 3 | 0,2 | 0,2 | 21,3 | 30,7 | 41,2 | 6,6 |
| 550 | 11 | 0,2 | 0,5 | 24,4 | 27,2 | 45,8 | 2,1 |
| 575 | 28 | 1,3 | 0,9 | 27,7 | 20,4 | 46,4 | 4,6 |
| 600 | 54 | 3,9 | 1,5 | 30,3 | 13,1 | 47,9 | 7,2 |

An elevation in zirconium oxide content to 2% inside the highly active catalyst results in a reduction of its overall activity, as shown by the degree of propane conversion, as well as its aromatisation activity. At a reaction temperature of 600 °C, the propane conversion rate and the yield of aromatic hydrocarbons for the 2% Zr-HSZ catalyst are 88% and 24.9%, respectively.

An increase in zirconium oxide content to 3% or greater in the highly active catalyst results in a further decline in its catalytic activity during the propane aromatisation process. The 4% Zr-HSZ catalyst has a selectivity of merely 7.2% for aromatic hydrocarbon production at 600 °C, with a propane conversion rate of 54%.

Table 5 presents the primary indicators of the propane aromatisation reaction utilising extremely active catalysts with the formula (MoO3)x∙(ZnO)y∙(ZrO2)z, synthesised through various ways.

**TABLE 5.** The main indicators of the propane aromatization reaction in the presence of highly active catalysts with the composition (MoO3)x∙(ZnO)y∙(ZrO2)z, prepared by different methods (0.1 MPa; Vpropane = 1000 h-1; T = 550oC)

|  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- |
| **Catalytic conversion of propane using catalysts with high catalytic activity, %** | **BApY, %** | **SApY,%** | **, %** | **, %** | **%** | **%** |
| **Impregnation** | | | | | | |
| **58,5** | 23,9 | 46,4 | 17,8 | 39,9 | 4,8 | 7,2 |
| **59,3** | 24,6 | 47,2 | 18,9 | 38,4 | 5,1 | 9,9 |
| **58,2** | 27,0 | 45,6 | 23,8 | 41,4 | 4,2 | 6,5 |
| **48,1** | 18,1 | 46,8 | 13,1 | 34,4 | 4,8 | 12,8 |
| **42,8** | 17,2 | 51,7 | 8,1 | 23,8 | 6,0 | 18,0 |
| **solid-phase modification** | | | | | | |
| **58,6** | 24,6 | 41,6 | 26,6 | 45,7 | 3,8 | 6,5 |
| **56,9** | 23,1 | 48,9 | 15,6 | 33,5 | 5,6 | 9,5 |
| **57,9** | 25,6 | 47,8 | 19,8 | 38,4 | 3,5 | 6,8 |
| **48,0** | 21,3 | 46,3 | 16,2 | 36,2 | 5,0 | 11,2 |
| **42,2** | 17,4 | 47,5 | 11,8 | 30,5 | 5,6 | 14,5 |

Table 5 illustrates that the primary parameters—the catalytic conversion of propane (K), the yield of aromatic hydrocarbons (B), and the catalyst selectivity (S)—exhibit similarities for catalysts synthesised via impregnation and solid-phase modification techniques. This suggests that catalysts produced by both processes yield roughly an equivalent quantity of active sites.

A bentonite-derived catalyst comprising 1.0% Zr and 5.0% Mo demonstrated significant catalytic efficacy. Subsequently, changing the zirconium amount in the Mo-based catalyst from 0.25% to 2% yielded optimal results at a zirconium concentration of 1.0%. The incorporation of Zr and Zn metals into the chosen catalyst was examined to improve the reaction yield. The incorporation of promoters and zirconium oxides into the Zr–Mo catalyst system enhanced the catalytic activity of the catalysts.

The gaseous byproducts of aromatisation primarily include hydrogen and methane, whilst the liquid byproducts consist of benzene and toluene. The addition of a minimal quantity of zirconium (Zr/Zn = 0.05) results in a nearly twofold reduction in the concentration of naphthalene and its homologues. The Zr-supported catalyst produced the most substantial effect. The inclusion of these metals resulted in a more than threefold rise in the quantity of condensed aromatic hydrocarbons (ArH), but a Zr/Zn ratio of 0.1 led to a fourfold drop (from 20.6% in the initial Zn-HSZ to 5.1% in the Zr-supported catalyst).

The results obtained are displayed in Tables 6 and 7.

**TABLE 6**. The influence of zirconium content on the main parameters of the aromatization process of liquefied hydrocarbon gases at 600°C over a highly active (MoO3)x∙(ZnO)y/HSZ catalyst

|  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- |
| **Zr/Zn ratio, mole** | **Zr content, % mass** | **Conversion,%** | **Selective influence, %** | **Product,%** | | |
| **ArH** | **CH4** | **C10+** |
| 0 | 0 | 90,3 | 57,9 | 53,8 | 13,1 | 9,5 |
| 0,05 | 0,45 | 83,5 | 56,7 | 49,6 | 15,5 | 5,9 |
| 0,10 | 0,9 | 80,5 | 58,9 | 49,1 | 14,3 | 4,8 |
| 0,15 | 1,35 | 81,8 | 55,1 | 46,6 | 15,8 | 4,4 |
| 0,25 | 2,0 | 53,4 | 61,2 | 33,4 | 14,1 | 2,4 |

**TABLE 7**. The influence of zirconium content in the highly active (MoO3)x∙(ZnO)y/HSZ catalyst on the composition of gases during the aromatization of liquefied hydrocarbon gases at 600°C

|  |  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- | --- |
| **Zr/Zn ratio, mole** | **Zr content, % mass** | **Composition of gas products** | | | | | |
| **H2** | **CH4** | **C2H6** | **C2H4** | **C3H8** | **C4** |
| 0 | 0 | 53,6 | 23,8 | 15,6 | 2,4 | 5,8 | 1,4 |
| 0,05 | 0,45 | 50,3 | 23,9 | 16,8 | 2,6 | 6,8 | 2,6 |
| 0,10 | 0,9 | 49,8 | 23,9 | 14,9 | 3,3 | 6,8 | 3,9 |
| 0,15 | 1,35 | 51,9 | 20,6 | 12,2 | 3,7 | 9,2 | 3,3 |
| 0,25 | 2,0 | 39,5 | 21,1 | 12,6 | 4,2 | 20,4 | 3,1 |

The catalytic conversion of propane over the highly active (MoO3)x∙(ZnO)y∙(ZrO2)z catalyst begins at 450 °C and increases with rising temperature, reaching 100% at 600 °C. Aromatic hydrocarbons are formed in significant quantities at 500 °C and reach a maximum yield of 52.5% at 600 °C.

##### CONCLUSION

Based on the research results, a catalyst with the following composition was selected as the optimal highly active catalyst: 5.0% Mo, 2.0% Zn, and 2.0% Zr. Thus, it has been confirmed that, based on the aforementioned findings, it is not advisable for the zinc content in the catalyst composition to exceed 4%.

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