

## WAYS OF RISING OF EFFICIENCY OF N-PARAFFIN DEHYDROGENATION REACTOR BLOCK FUNCTIONING WITH USE OF TECHNOLOGICAL MODELLING SYSTEM

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### Abstract

*In this article the applied using of technological modelling system of dehydrogenation process of n-paraffins has been examined. The products of this process are monoolefines, which are used to obtain synthetic detergents. In the basis of this system a formalized mechanism of hydrocarbons transformation on the Pt-catalyst surface lies. Using of the developed system allows to prolongate the catalyst durability due to optimisation of its usage mode and also to model different variants of plant reconstruction. After preliminary research of temperature influence, raw materials consumption and volume on the process efficiency, it was ascertained that one of the possible variants is switch into two-reactor scheme (it means using the second reserve reactor of paraffins dehydrogenation). Also the conditions of continuous functioning of two parallelly working reactors have been considered. Conformity of proposed variants of plant reconstruction has been economically proved. This helps to raise efficiency of synthetic detergents obtaining.*

**Key words:** mathematical modelling, dehydrogenation, reactor, catalyst, n-paraffin, n-olefines, efficiency, model, account

### 1. INTRODUCTION

Last years development of the consumer market actively stimulates development of the industry, including petrochemical one. In particular, manufactures of synthetic detergents and their raw-material base - linear alkyl benzenes (LAB), linear alkyl benzene sulphanates (LABS) rapidly develop. Thus the volume of import of synthetic washing-up liquids remains high enough: More than 60 % of the Russian release of synthetic washing-up liquids are supervised by transnational corporations.

By expert estimations, the American company Procter&Gamble, German Henkel, etc. supervise about 80 % of the market of detergents [1].

The factor of updating of a fixed capital is for Russian manufacturers of household chemical goods in 4 times less than necessary. The reached level of capital investments mismatches real needs of a chemical complex. Especially depresses that fact, that raw branches till now do not perceive chemical manufacture as the integral part - the share of ready chemical production in total amounts of sales of the Russian oil and gas companies makes only 3 % whereas at foreign partners it comes up to 30 %.

Narrow segmentation and a stable demand earlier did not allow the domestic companies to be fixed for a long time in this sphere, and they gave a up the place in the market to foreign competitors. But the situation changes also those foreign technologies which contained growth of the domestic industry, now work for its blessing. A striking example of it is the enterprise of Open Company «Kirishinefteorgsintez » (Open Company " Kinef ") and a factory the "LAB-LABS" included in its structure. The complex LAB and LABS manufacturing includes a number of technological blocks:

- 1) Plant of separation of n-paraffins by adsorption on zeolites;
- 2) Plant of of n-monoolefines obtaining by means of n-paraffins dehydrogenation on Pt-catalyst ;
- 3) Block of benzene alkylation with obtained monoolefines with the use of HF-catalyst;

4) Plant of LAB sulphidation and neutralization.

Such complex allows to receive about 180 tons of LAB with high ability to biological decomposition per day.

It is necessary to note, that for a petroleum-refining industry of all countries of the world last decade of XX century has passed under growing pressure of the ecological factor with accompanying toughening requirements of new technics to consumer properties of mineral oil [2] .

Ability for biological decomposition of synthetic detergents is the base criterion of quality of production which level, according to modern Russian ecological norms, should be not less than 88 %. The LAB obtained by Open Company " Kinef ", meets these requirements with a greater stock. The technology realized at this factory, allows to obtain the LAB with ability for biological decomposition up to 95 %.

Despite high quality, the domestic market of LAB is characterized by high deficiency of this production. Potential need of the domestic market in LAB is about 100 thousand tons per year.

Therefore increase of a production efficiency of alkyl benzenes is an important and actual problem.

The continuous monitoring of plants work with application of modern mathematical modelling methods is necessary for maintenance of optimum conditions of carrying out the processes.

The methodology of mathematical modelling developed by us for multicomponent catalytic processes of hydrocarbonic raw material processing is based on following concepts of the system analysis of chemical-technological processes:

- 1) Availability and reliability of the experimental data put in a basis of researches (on raw material, products and mode conditions of process);
- 2) Formalization of the mechanism of hydrocarbons transformation on the catalyst surface, based on association of components according to reactionary ability; drawing up of the transformation scheme of the incorporated components, finding out how reactionary ability of the hydrocarbons belonging to one gomological group depend on their physical and chemical properties;
- 3) An establishment of kinetic laws of processes of catalyst desactivation, based on the account of simultaneously proceeding reactions of coke formation, ageing and a poisoning with poisons of an active contact surface, and also influence of technological conditions of process and hydrocarbonic structure of processed raw material on speed of catalyst desactivation;
- 4) Construction of technological modelling system of process for monitoring and forecasting of work of plants, computer support and revealing of "narrow" places of manufacturing.

These concepts are most fully fulfilled on an example of catalytic reforming of gasolines [3-7].

At the same time, the developed methodology is applicable for a petrochemical complex of processes of LAB manufacturing processes, (fig 1) which capacities make 50 thousand tons of alkyl benzene.

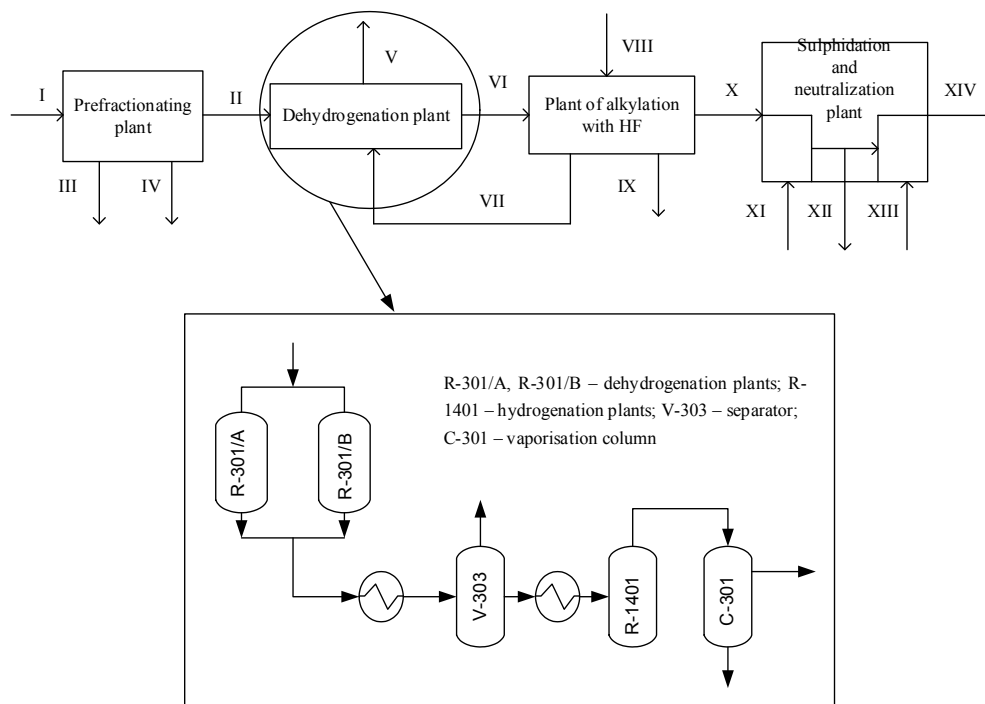


**Fig. 1.** Flowchart of lines of LAB, LABS production complex: I) n-paraffins from adsorption plant ; II)  $C_{10}$ – $C_{13}$ n-paraffins; III) fraction of  $C_{14}$ – $C_{17}$ n-paraffins; IV) fraction of  $C_{18}$  and higher n-paraffins; V) hydrogen containing gas; VI) mixture of n-paraffins and monoolefines; VII) recirculating n-paraffins; VIII) benzene from catalytic reforming plant; IX) heavy alkylate; X) LAB; XI) sulfur from an elemental sulfur set; XII) LABS; XIII) alkali; XIV) sodium salt of alkylbenzenesulphuric acid

One of possible ways of plant productivity increase is switching of a reserve reactor dehydrogenation reactor in parallel work with operating one.

## 2. SEARCHING OF HOW DOES CONCENTRATION OF OLEFINS DEPENDS ON TEMPERATURE, VOLUME AND EXPENDITURE OF RAW MATERIAL PASSED

The basic purpose of works at the present stage of researches is development of recommendations on technological parameters of carrying out of industrial dehydrogenation process with simultaneous work of two dehydrogenation reactors at the various expenditure of processed raw material and volumetric speed of raw materials supply (VSRMS).

Switching to parallel work of dehydrogenation reactors means decrease in raw material loading on one reactor from 75 m<sup>3</sup> per hour to 37,5 m<sup>3</sup> per hour, that will be reflected both on kinetic, and in a hydrodynamical operating mode of a reactor.

We have lead calculations of hydraulic resistance of the dehydrogenation catalyst layer at various loadings on raw material, tab. 1.

Table 1 - Influence of the raw material supply on hydraulic resistance of the catalyst layer

| Parameter              | $G_c=75 \text{ m}^3/\text{h}$<br>$l=7/1$ | $G_c=37,5 \text{ m}^3/\text{h}$<br>$l=7/1$ | $G_c=37,5 \text{ m}^3/\text{h}$<br>$l=8/1$ |
|------------------------|------------------------------------------|--------------------------------------------|--------------------------------------------|
| VSRMS, $\text{h}^{-1}$ | 22                                       | 11,3                                       | 11,3                                       |
| $\Delta P_{c,l}$ , kPa | 18                                       | 4,9                                        | 5,4                                        |

\* $l$  - molar ratio "hydrogen/raw material",  $G_c$  - raw material supply,  $\Delta P_{c,l}$  - hydraulic resistance of the catalyst layer, VSRMS - volumetric speed of raw materials supply.

Decrease in hydraulic resistance of the catalyst layer will allow to lower diffusion complications during the target dehydrogenation reaction passing and to maintain the Pt-catalyst in softer hydraulic mode.

Change of volumetric speed of raw materials supply will affect a degree of grain use of the catalyst which can be evaluated through the factor of efficiency (fig. 2):

$$\varphi = l \cdot \sqrt{\frac{k}{D_{\text{eff}}}} \quad (1);$$

$$\eta = \frac{3}{\varphi} \left( \frac{1}{\tanh \varphi} - \frac{1}{\varphi} \right) \quad (2),$$

where  $l$  – a radius of the catalyst particle, m;  $k$  - a constant of speed of chemical reaction,  $\text{s}^{-1}$  ( $\text{l mol}^{-1} \text{s}^{-1}$ ),  $D_{\text{eff}}$ -effective factor of diffusion,  $\text{m}^2/\text{s}$ ;  $\varphi$ -module of Tile;  $\eta$  factor of efficiency of grain use of the catalyst.

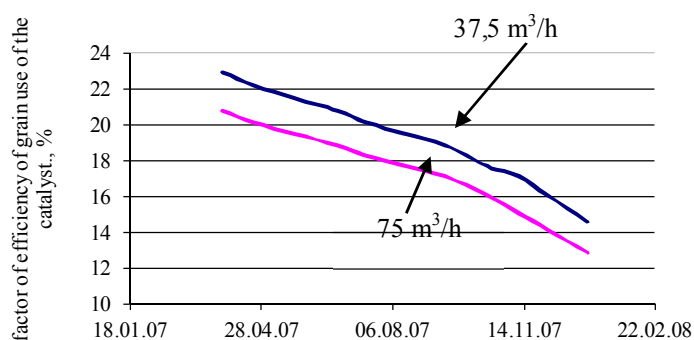


Figure 2. A degree of the catalyst use in dependence on loading on raw material and time of the catalyst operation

Change of temperature drop in a reactor depends on kinetics of the process, so, on type of the loaded catalyst and structure of raw material. With application of non-stationary mathematical model of process and calculation of its thermal balance it became possible to lead an estimation of change of temperature drop.

The mathematical model of industrial dehydrogenation process, developed by the department of chemical technology of fuel of Tomsk polytechnical university is sensitive to a chemical compound of raw material processed on the plant [9]. The kinetic description put in a basis of technological modelling system, is developed taking into account various reactionary ability of paraffin hydrocarbons C9-C14 in dehydrogenation reactions. The hydro dynamical component of model allows to consider influence of the raw material expenditure passed through a reactor on parameters of process. The generalized mathematical description of process represents system of the following differential equations of material and thermal balances, eq. (3) - (4).

$$G \frac{\partial C_i}{\partial z} + G \frac{\partial C_i}{\partial V} = (1 - \varepsilon) \sum_j w_j, \quad (3)$$

s.c.  $z=0$ :  $C_i=0$ , rp.y.  $V=0$ :  $C_i=C_{ex}$

$$G \frac{\partial T}{\partial z} + G \frac{\partial T}{\partial V} = -(1 - \varepsilon) \frac{\sum_j \Delta H_j r_i}{c_p \rho}, \quad (4)$$

s.c.  $z=0$ :  $T=T_0$ , rp.y.  $V=0$ :  $T=T_{ex}$ ,

where  $C_i$ - concentration of  $i$ -th hydrocarbon, mol/m<sup>3</sup>;  $U$ - linear speed of a stream, m/h;  $\varepsilon$ -pore volume of catalyst layer;  $r_i$ - speed of reaction, mol/m<sup>3</sup>·h;  $r_i = \sum_j W_j = \frac{dC_i}{dt}$ ;  $w_j$ -speed of  $j$ -th component in  $i$ -th reaction, mol/m<sup>3</sup>·h;  $T$  – temperature of the process, K;  $\Delta H_j$ - thermal effect of reaction, Joule/mol;  $c_p$  – thermal capacity of a mixture, Joule/mol·K;  $\rho$  -density of a mixture, kg/m<sup>3</sup>;  $T_0$  - start temperature (temperature of an environment), K;  $T_{ex}$ - temperature of an input in a reactor, K;  $C_{ex}$ -entrance concentration of hydrocarbon, mol/m<sup>3</sup>.

Time of stay of reagents in the reactionary zone, depending on the hour expenditure of raw material  $G$ , volume of the catalyst  $V$ , in conditions of a stable loading of plant on raw material is replaced with « resulted time » or total volume of the processed raw material,  $z = Gt$ , m<sup>3</sup>,  $t$  – time, hour.

Forecasting of a variant of double-reactor scheme of work of operating plant can be most effectively executed with application of the developed by authors kinetic model adequately describing real industrial paraffins dehydrogenation process [3].

Experimental researches allow to estimate only influence of change of volumetric speed of raw materials supply on process parameters.

At the same time it is necessary to consider a degree of dehydrogenation catalyst deactivation, that, in its turn, makes the contribution to change of a technological mode and, accordingly, qualitative and quantitative characteristics of a target product.

Experiments of such duration cannot be lead, since, the degree of the catalyst deactivation depends both on a temperature mode of catalyst operating, and on structure and volume of the raw material passed through the reactor.

So, the estimation of the catalyst activity change lead on mathematical model, has shown, that for 3-6-day period of plant work (about 7, 2 thousand m<sup>3</sup> of raw material passed) the catalyst activity concerning target paraffins dehydrogenation reaction will decrease on 0,04-0,11 %. Concerning collateral reactions of olefines dehydrogenation - on 0,01-0,02 %, paraffins hydrocracking - on 0,21-0,58 %, isoparaffins dehydrocyclization - on 0,07-0,21 %; the content of coke on the catalyst thus is estimated as 0,022 %, that practically does not influence a course of process.

On fig. 3-8 the results of calculations on model of concentration of olefines and diolefines in a product flow depending on temperature, the expenditure and volume of the raw material passed through a reactor are presented.

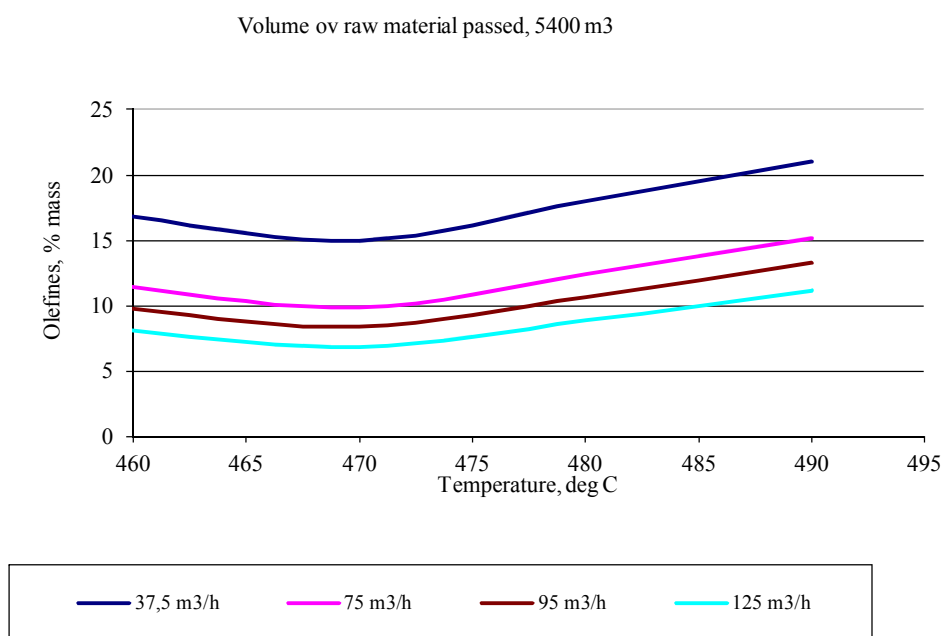


Figure 3. - Change of concentration of olefines depending on temperature and expenditure of raw material at 5 thousand m<sup>3</sup> of raw material passed (about 3 days of plant work)

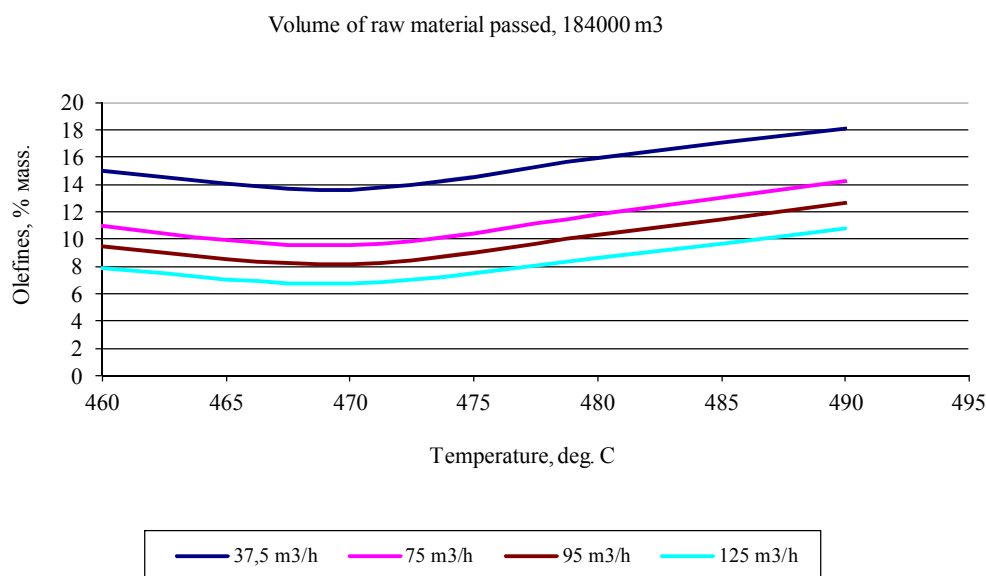


Figure 4. - Change of concentration of olefines depending on temperature and expenditure of raw material at 184 thousand m<sup>3</sup> of raw material passed (more than days of plant work)

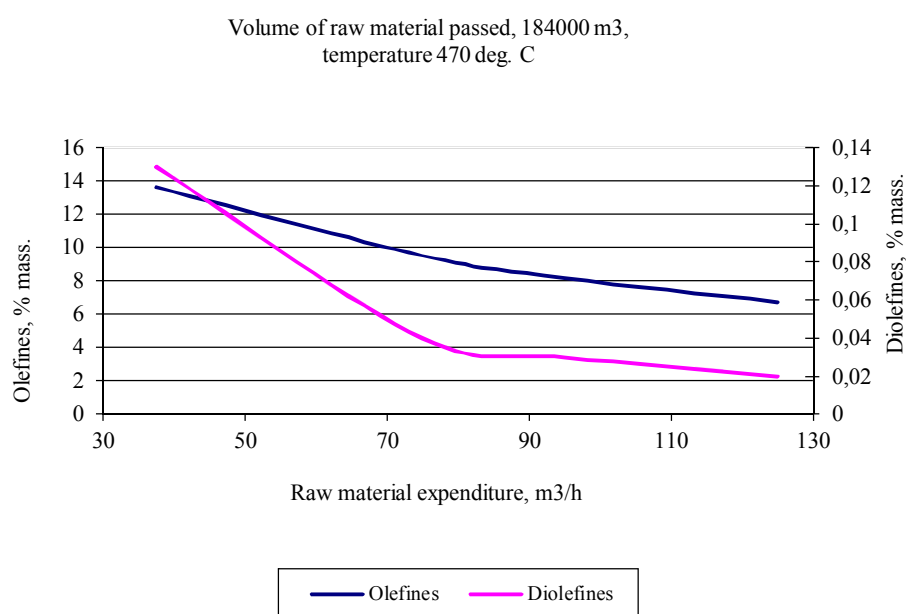


Figure 5 - Change of concentration of olefines and diolefines depending on temperature and the expenditure of raw material at 184 thousand m<sup>3</sup> of raw material passed (more than days of plant work)

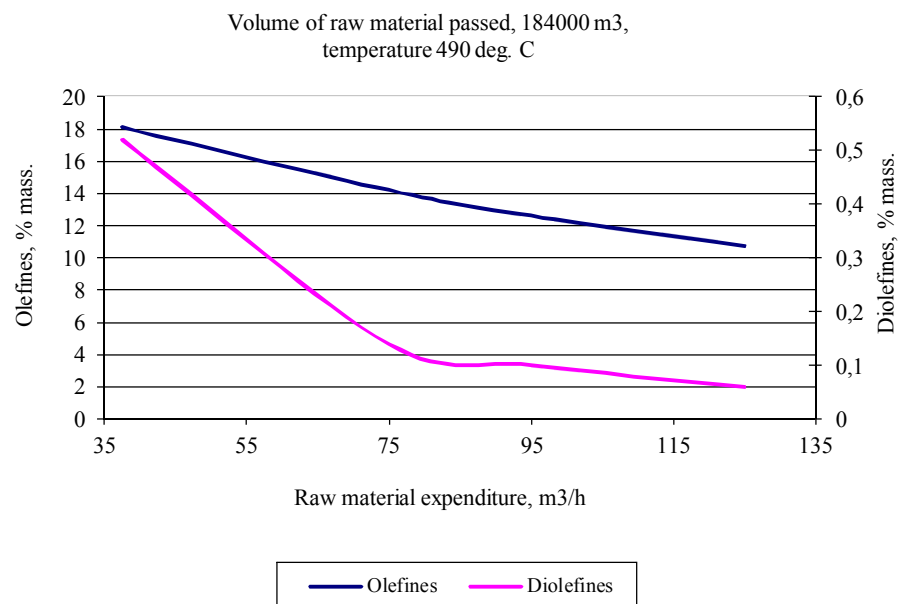


Figure 6- Change of concentration of olefines and diolefines depending on temperature and the expenditure of raw material at 184 thousand m<sup>3</sup> of raw material passed (more than days of plant work)

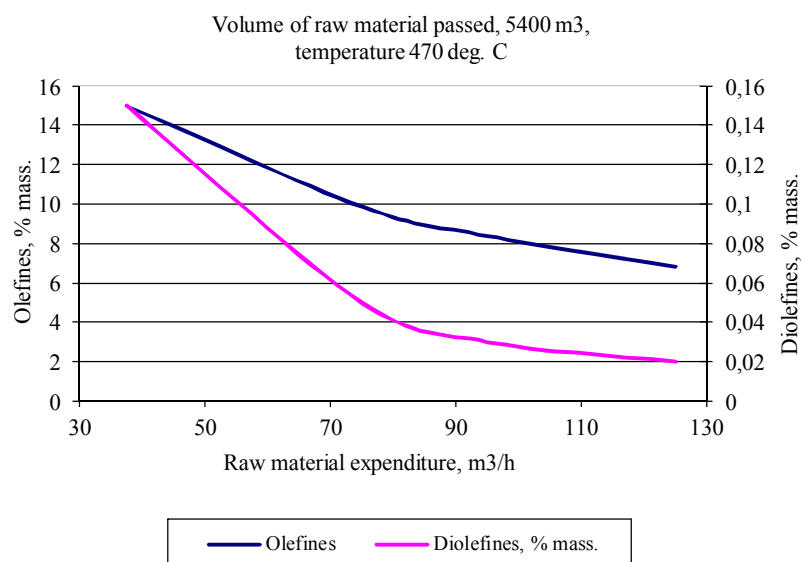


Figure 7- Change of concentration of olefines and diolefines depending on temperature and the expenditure of raw material at 5,4 thousand m<sup>3</sup> of raw material passed (about 3 days of plant work)

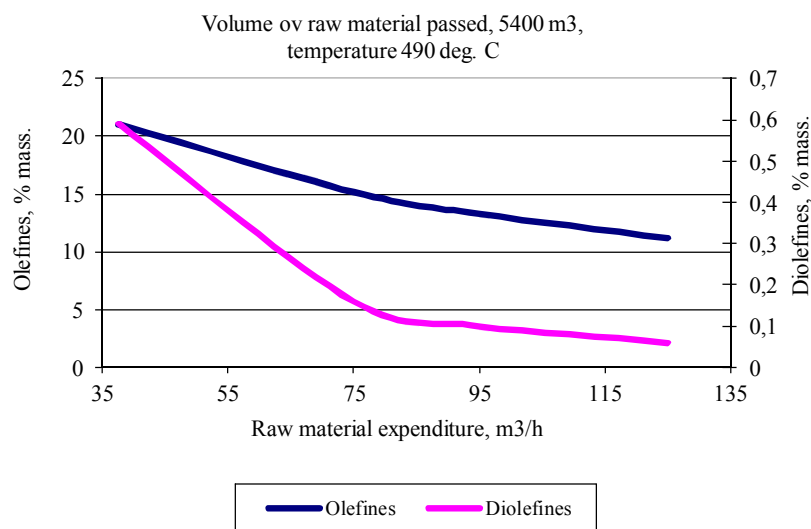


Figure 8- Change of concentration of olefines and diolefines depending on temperature and the expenditure of raw material at 5,4 thousand m<sup>3</sup> of raw material passed (about 3 days of plant work)

As it shown in fig. 3-8, with growth of temperature concentration target and by-products in mixture increases. At the same time it falls with increase of raw material expenditure and volume of raw material passed through dehydrogenation reactor (with strengthening of the catalyst), that does not contradict classical conceptions about heterogeneous catalysis. Inclusion of a reactor of diolefines by-products hydrogenation in the technological scheme allows to increase quantity target olefines, acting on a stage of benzene alkylation with monoolefines, tab. 2-3.

Table 2

Concentration of target and by-products (without taking into account a stage of hydrogenation) at various temperature and raw material expenditure (molar ratio "hydrogen/hydrocarbons" = 7/1, volume of the raw material passed-184 thousand m<sup>3</sup>)

| Products, % mass.                                | Temperature, °C |       |
|--------------------------------------------------|-----------------|-------|
|                                                  | 470             | 490   |
| Raw material expenditure, 37,5 m <sup>3</sup> /h |                 |       |
| Olefines                                         | 12,74           | 16,32 |
| Diolefines                                       | 1,27            | 2,63  |
| Aromatic hydrocarbons                            | 0,34            | 0,4   |
| Raw material expenditure, 75 m <sup>3</sup> /h   |                 |       |
| Olefines                                         | 9,19            | 13,28 |
| Diolefines                                       | 0,56            | 1,33  |
| Aromatic hydrocarbons                            | 0,29            | 0,38  |
| Raw material expenditure, 95 m <sup>3</sup> /h   |                 |       |
| Olefines                                         | 7,96            | 11,93 |
| Diolefines                                       | 0,4             | 1,01  |

|                                                 |      |       |
|-------------------------------------------------|------|-------|
| Aromatic hydrocarbons                           | 0,22 | 0,31  |
| Raw material expenditure, 125 m <sup>3</sup> /h |      |       |
| Olefines                                        | 6,62 | 10,31 |
| Diolefines                                      | 0,27 | 0,72  |
| Aromatic hydrocarbons                           | 0,23 | 0,33  |

Table 3

Concentration of target and by-products (with taking into account a stage of hydrogenation) at various temperature and raw material expenditure (molar ratio "hydrogen/hydrocarbons" = 7/1, volume of the raw material passed-184 thousand m<sup>3</sup>)

| Products, % mass.                                | Temperature, °C |       |
|--------------------------------------------------|-----------------|-------|
|                                                  | 470             | 490   |
| Raw material expenditure, 37,5 m <sup>3</sup> /h |                 |       |
| Olefines                                         | 13,62           | 18,1  |
| Diolefines                                       | 0,13            | 0,52  |
| Aromatic hydrocarbons                            | 0,29            | 0,35  |
| Raw material expenditure, 75 m <sup>3</sup> /h   |                 |       |
| Olefines                                         | 9,52            | 14,2  |
| Diolefines                                       | 0,04            | 0,14  |
| Aromatic hydrocarbons                            | 0,24            | 0,33  |
| Raw material expenditure, 95 m <sup>3</sup> /h   |                 |       |
| Olefines                                         | 8,17            | 12,61 |
| Diolefines                                       | 0,03            | 0,1   |
| Aromatic hydrocarbons                            | 0,22            | 0,31  |
| Raw material expenditure, 125 m <sup>3</sup> /h  |                 |       |
| Olefines                                         | 6,73            | 10,76 |
| Diolefines                                       | 0,02            | 0,06  |
| Aromatic hydrocarbons                            | 0,19            | 0,28  |

Simultaneous rise in temperature of process and increase in time of contact of reagents (increase of "rigidity" of process), undoubtedly, will lead to growth of by-products concentration and high coke accumulation. It will negatively influence quality of obtained production and will reduce durability of the catalyst.

Thus, executed with application of mathematical model researches about how does raw material expenditure influence parameters of n-paraffins dehydrogenation process have allowed to reveal following laws:

- a) With increase of raw material expenditure and, hence, volumetric speed of raw materials supply (VSRMS) it is necessary to rise temperature for manufacturing of demanded quantity of a target product - olefin, but without changing a process "rigidity";
- b) With rise in temperature and decrease of VSRMS concentration of olefines in a product mixture increases; that can be explained by thermodynamics of target dehydrogenation reaction. Calculations on model show increase in concentration of olefines, on average on 20 %, with decrease of VSRMS up to  $10 \text{ h}^{-1}$  and temperature  $490^\circ\text{C}$ ;
- c) With rise in temperature and decrease of VSRMS, together with increase of target product increases the level of by-products - diolefines and aromatic hydrocarbons also; that is connected with increase in time of reagents contact;
- d) Selectivity of process decreases with growth of conversion and increases with increase of raw material expenditure.

These laws correspond to theoretical conceptions and are experimentally proved.

Thus, for forecast estimations of work efficiency of parallel working reactors it is necessary to consider influence of several factors, in particular, operating parameters of the given process is not only time of reagents contact, but also temperature of industrial process carrying out.

With decrease in loading on raw material (from 75 till  $37,5 \text{ m}^3/\text{h}$ ) the decrease in rate of rise of input temperature in a dehydrogenation reactor on the average on  $20^\circ\text{C}$  is possible in view of equally set concentration of a target product (at carrying out forecast calculations on model 9,61 % mass.), fig. 9.

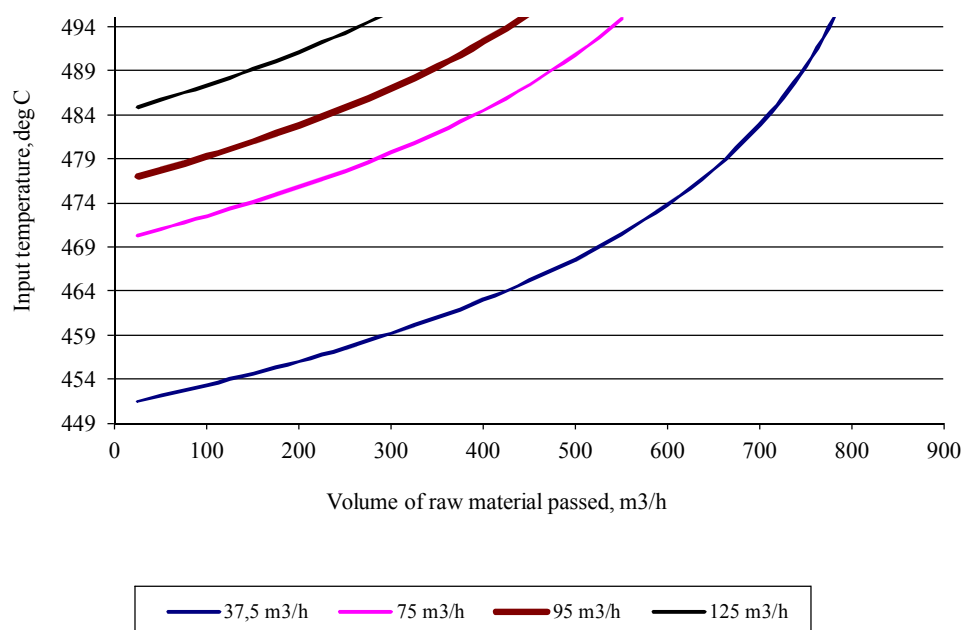


Figure 9. -Change of an input temperature in a dehydrogenation reactor at the various raw material expenditure

Possible variant of decrease in rate of coke accumulation on the catalyst, and also growth of by-products concentration is the increase of molar ratio "hydrogen/raw material". Results of modelling researches are presented in tab. 4-5.

Table 4

Concentration of target and by-products (with taking into account a stage of hydrogenation) at various temperature and raw material expenditure (molar ratio "hydrogen/hydrocarbons" = 8/1, volume of the raw material passed-184 thousand m<sup>3</sup>)

| Products, % mac.                                 | Temperature, °C |       |
|--------------------------------------------------|-----------------|-------|
|                                                  | 470             | 490   |
| Raw material expenditure, 37,5 m <sup>3</sup> /h |                 |       |
| Olefines                                         | 13,37           | 16,31 |
| Diolefines                                       | 1,14            | 2,44  |
| Aromatic hydrocarbons                            | 0,34            | 0,41  |
| Raw material expenditure, 75 m <sup>3</sup> /h   |                 |       |
| Olefines                                         | 8,62            | 12,76 |
| Diolefines                                       | 0,48            | 1,18  |
| Aromatic hydrocarbons                            | 0,28            | 0,38  |
| Raw material expenditure, 95 m <sup>3</sup> /h   |                 |       |
| Olefines                                         | 7,39            | 11,33 |
| Diolefines                                       | 0,34            | 0,89  |
| Aromatic hydrocarbons                            | 0,25            | 0,35  |
| Raw material expenditure, 125 m <sup>3</sup> /h  |                 |       |
| Olefines                                         | 6,09            | 9,68  |
| Diolefines                                       | 0,22            | 0,62  |
| Aromatic hydrocarbons                            | 0,22            | 0,31  |

Calculations on model show, that the thermodynamics of target reaction will not allow to raise essentially an output of a target product in a unit of time with decrease in loading on raw material. Undoubtedly, more economically justified and effective is the variant of simultaneous increase of raw material expenditure.

For achievement of the maximal economic benefit maintenance of raw material expenditure at level of 50 m<sup>3</sup>/h (100 m<sup>3</sup>/hour for both reactors) is expediently.

Thus, realization of following recommendations is necessary:

- 1) temperature of an input in a reactor 460-487 °C;
- 2) molar ratio «H<sub>2</sub>/hydrocarbons» 8/1.

Finally, the choice of any of variants of work is caused by indemnification of the costs connected with maintenance of the chosen technological mode, increase in an output of the LAB in a unit of time, tab. 6.

Table 6

Variants of a technological mode at operation of the plant including two dehydrogenation reactors

| Raw material expenditure,<br>m <sup>3</sup> /h | Input<br>temperature, °C | Duration of the<br>catalyst run,<br>days | Average<br>manufacturing<br>of LAB, t/day. | By-products<br>(diolefines) by the end<br>of run, % mass. |
|------------------------------------------------|--------------------------|------------------------------------------|--------------------------------------------|-----------------------------------------------------------|
| Molar ratio 7/1                                |                          |                                          |                                            |                                                           |
| 75 (for 1 reactor)                             | 470                      | 263                                      | 190                                        | 0,84 on 263-th day                                        |
| 50 (100 for 2 reactors)                        | 462                      | 479                                      | 256                                        | 1 on 479-th day and<br>0,85 on 263-th day                 |
| 37,5 (75 for 2 reactors)                       | 456                      | 611                                      | 201                                        | 1,21 on 611-th day and<br>1,02 on 263-th day              |
| Molar ratio 8/1                                |                          |                                          |                                            |                                                           |
| 75 (for 1 reactor)                             | 470                      | 319                                      | 178                                        | 0,74 on 319-th day and<br>0,69 on 263-th day              |
| 50 (100 for 2 reactors)                        | 462                      | 604                                      | 245                                        | 0,92 on 604-th day 0,71<br>on 263-th day                  |
| 37,5 (75 for 2 reactors)                       | 456                      | 833                                      | 187                                        | 1,03 on 833-th day on<br>0,74 on 263-th day               |

It is necessary to note, that the quantity of the LAB is estimated without taking into account loss of a target product as a result of ablation with polyalkylbenzenes (PAB). On average 8-10 % the LAB are carried away with PUB (by results of laboratory analyses).

Thus, economic benefit of realization of switching to parallel work of dehydrogenation reactors can be connected:

- 1) With increase in duration of plant work at one loading of the Pt-catalyst from 263 till 833 days (the greatest possible increase);
- 2) With increase in target product output in unit of time up to 201 th day of LAB. Simultaneous achievement of both parameters is obviously possible with increase in loading of raw material on 2 reactors till 100 m3/hour and increase of molar ratio «H<sub>2</sub>/hydrocarbons» up to 8/1 (increase in run till 604 day and daily average manufacturing of LAB up to 245 tones).

At the same time, the increase in loading on a hydrogenation reactor and, accordingly, on a reactor of hydrogenation, can cause decrease of conversion at diolefines hydrogenation. The rules of the enterprise provides a working temperature interval for this reactor 180-220 °C. Calculations on model have shown, that the increase in loading on raw material on a reactor from 75 till 100 m3/h (approximately from 56000 up to 75000 kg/hours) will not essentially affect conversion of diolefines at an intermediate stage of LAB manufacturing; target concentration of diolefines will not exceed 0,12 % mass during all run.

### 3. ORGANISATION OF CONTINUOUS PLANT WORK

One more effective variant of plant operation can become the organization of continuous work of reactors as at increase in loading till 100 m3/hour (50 m3/hour on one reactor), and at its preservation at a level of 75 m3/hour (37, 5 m3/hour on one reactor).

At carrying out forecast calculations the following prospective scheme of plant start-up has been chosen:

- 1) simultaneous start-up of dehydrogenation reactors at loading 50 m<sup>3</sup>/hour on one reactor (319 days);
- 2) a stop of one of reactors, for example, R-301/A, at achievement of input temperature in a reactor 487 °C; thus the second reactor R-301/B continues working at smooth rise of input temperature from 487 up to 491 °C and increase in loading of raw material at this reactor from 50 till 75 m<sup>3</sup>/h (a maximum of 30 day, a minimum of 2 weeks), depending on when will R-301/A be overloaded with fresh catalyst );
- 3) Switching off a reactor R-301/B; simultaneous inclusion in work of reactor R-301/A at loading of raw material 75 m<sup>3</sup>/hour with the fresh loaded catalyst with starting temperature of an input 470 °C (15 days);
- 4) inclusion of reactor R-301/B in work, increase in loading on raw material till 100 m<sup>3</sup>/h (50 m<sup>3</sup>/h on one reactor) (about 1 year), etc.

Results of calculation of a similar technological mode on model are shown on figures.12-13 (at loading of 50 m<sup>3</sup>/hour on one reactor).

Let's accept following designations of sites on a curve: 1, 3 - parallel inclusion of reactors, fig.10; 2 - monoreactor operating mode, fig. 11.

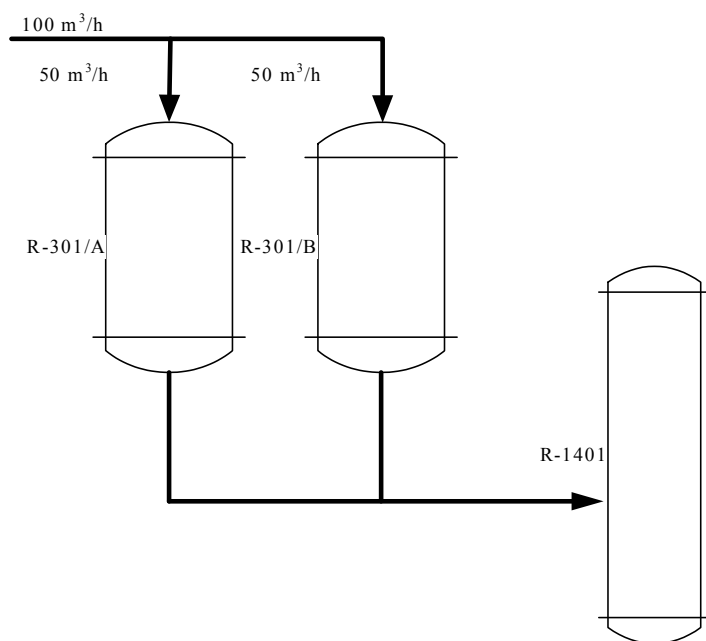


Figure 10. - parallel inclusion of dehydrogenation reactors

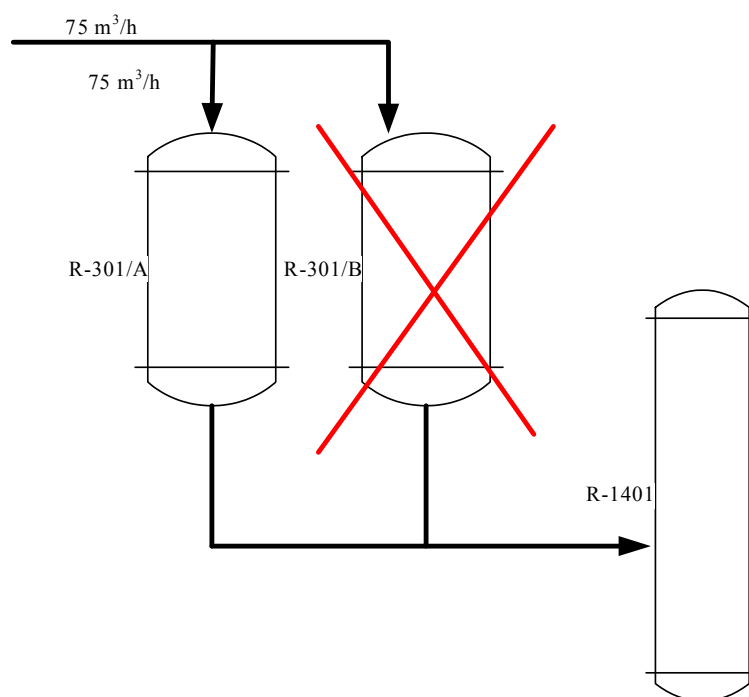


Figure 11- monoreactor operating mode

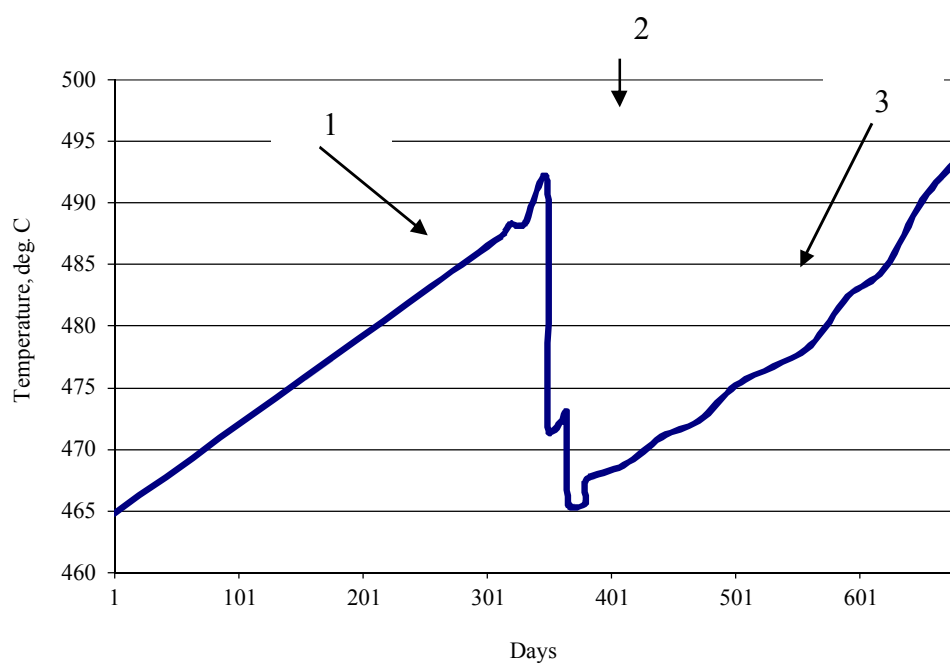


Figure 12. - Temperature of an input of raw material at the organization of continuous work of hydrogenation reactors

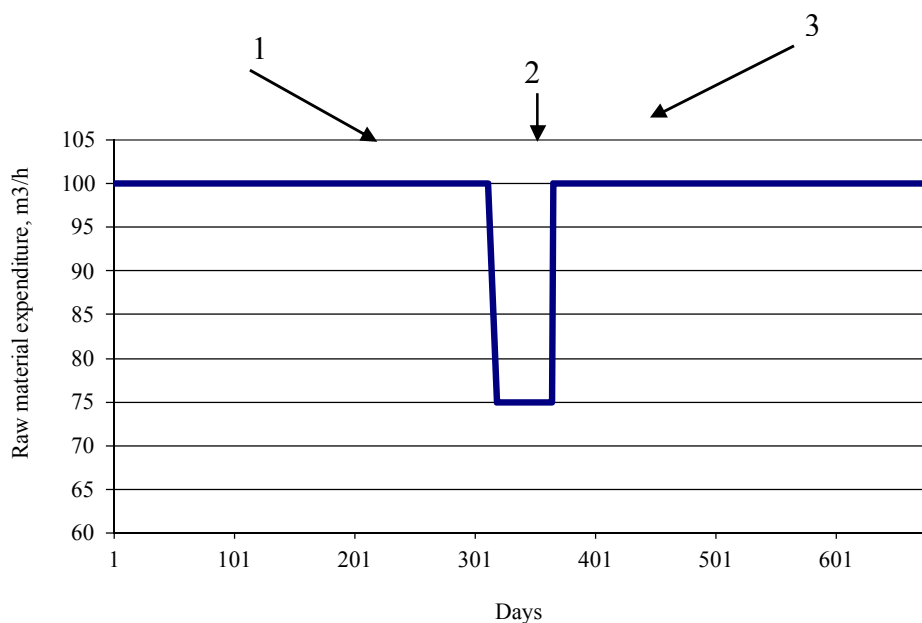


Figure 13. - Raw material expenditure at the organization of continuous work of hydrogenation reactors

If it is not possible to increase loading on raw material on both reactors till 100 m³/h, operating modes of plant and the basic parameters of efficiency at work on former loading (75 m³/hour) will look as follows, fig. 14 (at loading 37,5 m³/hour on one reactor).

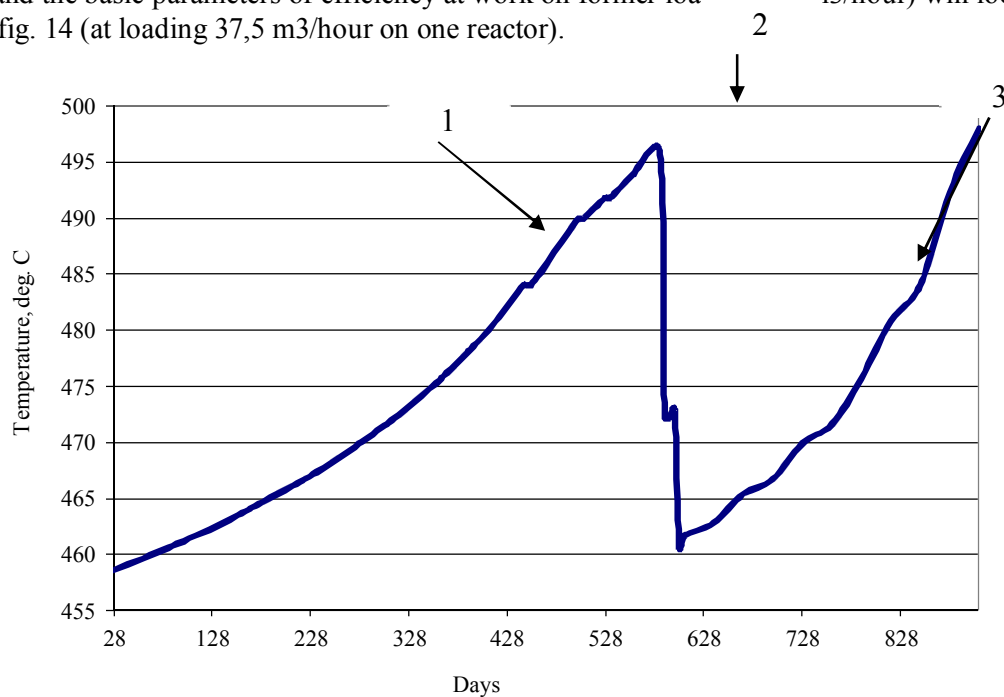


Figure 14. - Temperature of an input of raw material at the organization of continuous work of dehydrogenation reactors

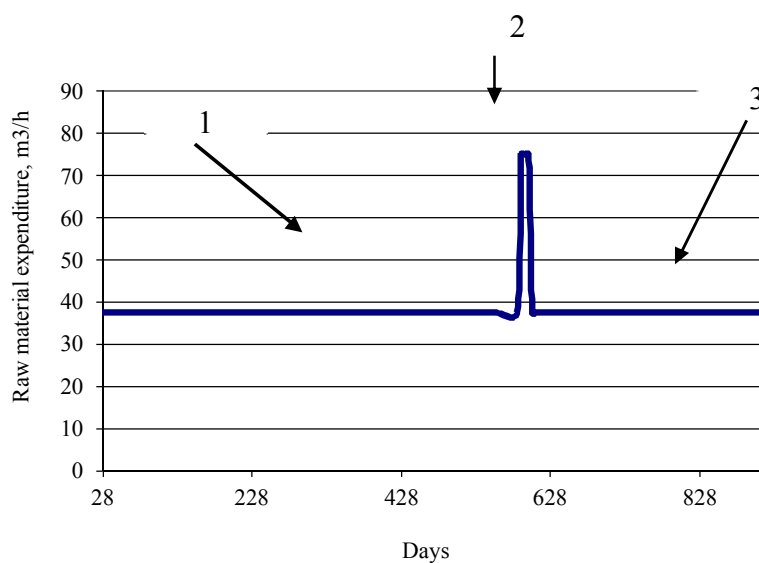


Figure 15. - Raw material expenditure at the organization of continuous work of dehydrogenation reactors

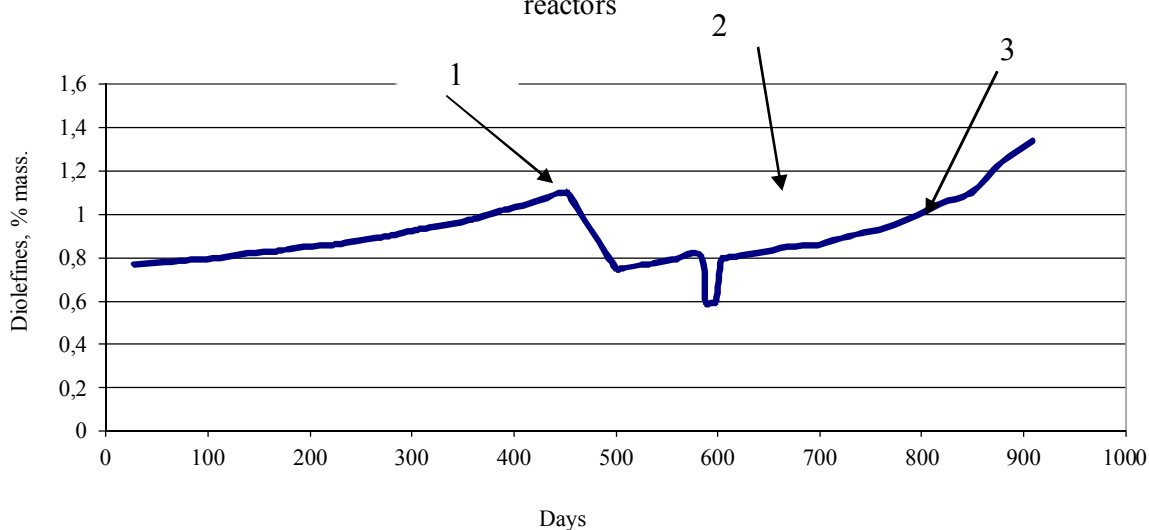


Figure 16. -Diolefines at the organization of continuous work of dehydrogenation reactors

Thus, simultaneous inclusion of two dehydrogenation reactors into work can be economically justified as from the point of view of increase in target product LAB manufacturing in a unit of time, especially with increase in "rigidity" of process, and with the organization of continuous plant work.

## CONCLUSIONS

The researches carried out on mathematical model, have shown, that increase of economic efficiency linear alkyl benzenes obtaining plant work at its switching to parallel work of dehydrogenation reactors can be reached in some cases:

□□ At simultaneous increase in loading of raw material □ up to 100 m<sup>3</sup>/h and hydrogen-containing gas expenditure the maximal effect connected both with increase of duration of catalyst run, and with increase in daily average LAB manufacturing will be reached;

□□ The organization of continuous work of reactors.

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