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## EXTRACTION PROCESS OF WASTEWATER TREATMENT AND DEVELOPMENT OF PRINCIPLE FLOWCHART

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### Abstract

The possibilities of the extraction process for wastewater treatment, oil and oil products pollution have been studied, and principle flowchart has been developed. Physical and chemical problems accompanying the extraction process are analyzed. Some features of the extraction process are considered and a principle flowchart of wastewater treatment is developed. Liquid-phase extraction processes are carried out in various apparatuses, mixing devices, and in some cases using membrane technologies. It is shown that the processes of liquid-phase extraction occurring in mixing devices are complicated by many physical phenomena, for example, coalescence and crushing of extractant drops, separation and stratification of extract and refined sugar, accompanying various effects (Marangoni effects, Fasi effect) and interfacial mass transfer. To improve the parameters of interfacial transfer, in addition to various ones, the most acceptable is to increase the interfacial surface. In liquid phase extraction processes, the main factor is mass transfer between two immiscible liquids. Liquid extraction of organic compounds and oil from wastewater is a mass transfer process that occurs in an interfacial film and is carried out by dispersing the extract - it will dissolve in an aqueous medium and extraction in an interfacial film, characterized by diffusion transfers of the corresponding components to the interfacial surface.

*Keywords:* extraction; oil waste water; suspensions; interfacial surface; interfacial film; isotropic turbulence.

## ЕКСТРАКЦІЙНИЙ ПРОЦЕС ОЧИЩЕННЯ СТИЧНИХ ВОД ТА РОЗРОБКА ЙОГО ПРИНЦИПОВОЇ ТЕХНОЛОГІЧНОЇ СХЕМИ

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### Анотація

Вивчені можливості процесу екстракції для очищення стічних вод, забруднених нафтою і нафтопродуктами, та розроблено його принципову технологічну схему. Проаналізовані фізико-хімічні проблеми, що супроводжують процес екстракції. Розглянуті деякі особливості процесу екстракції та розроблено принципову технологічну схему очищення стічних вод. Процеси рідиннофазної екстракції здійснюються в різних апаратах, змішувальних пристроях, а в деяких випадках – з використанням мембранних технологій. Показано, що процеси рідиннофазної екстракції, які протікають в апаратах змішування, ускладнюються багатьма фізичними явищами, наприклад, коалесценцією і диспергацією крапель екстрагенту, розділенням і розшаруванням екстракту і рафінованого цукру, супутніми різноманітними ефектами (ефекти Марангоні, Фасі) і міжфазним масообміном. Для покращення параметрів міжфазного переносу, найбільш прийнятним поміж різних способів є метод збільшення міжфазної поверхні. У процесах рідинної фазової екстракції основним фактором є масообмін між двома незмішуваними рідинами. Рідинна екстракція органічних сполук і нафти зі стічних вод – це масообмінний процес, який відбувається в міжфазній плівці і здійснюється шляхом диспергування екстракту: його розчинення у водному середовищі та екстракції в міжфазній плівці, що характеризується дифузійним перенесенням відповідних компонентів до міжфазної поверхні.

*Ключові слова:* екстракція, нафтові стічні води; суспензії; міжфазна поверхня; міжфазна плівка; ізотропна турбулентність.

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## Introduction

Wastewater treatment from various organic, asphalt-resinous substances and solid particles is an important, complex and multifactorial environmental, economic and industrial problem for many chemical and oil refining industries [1]. Industrial wastewater treatment can be carried out by adsorption, plasma-membrane methods, as well as their combination, in particular, by the extraction-membrane method and various processes based on new instrumentation [2–4]. Industrial wastewater contains various impurities – oil products, suspensions, particles of the dispersed phase, which are insoluble in water and form heterogeneous systems with varying degrees of dispersion [5]. However, given the high productivity of wastewater, the most effective in this aspect are technologies based on adsorption and extraction processes, the complexity of which is determined by the appropriate choice of adsorbent and extractant. The liquid-phase extraction processes are located in different apparatuses: countercurrent flow spray columns, mixing devices, in some cases using membrane technology. The processes of liquid-phase extraction occurring in mixers are complicated by many physical phenomena: coalescence and crushing of extractant drops, separation and stratification of extraction and raffinate, accompanied by various effects (Marangoni effect, Faci effect, thermal diffusion) and mass transfer. The interfacial mass transfer in liquid-phase extraction processes is determined by the amount of the transferred substance from one phase to another and is determined by the general formula

$$M = \beta_1 F(C - C^*), \quad (1)$$

where,  $\beta$  – is the mass transfer coefficient,  $F$  – is the interfacial mass transfer surface,  $C$  and  $C^*$  – are the current and equilibrium concentration.

Therefore, to improve the interfacial transfer, among other parameters, the most acceptable is to increase the interfacial surface. In liquid-phase extraction processes, in order to increase the interfacial surface between the extractant and the main phase, it is necessary to crush the extractant liquid into smaller drops that do not mix with the main phase. In this aspect, the most effective is to carry out the process of liquid-phase extraction in mixing devices, where it is possible to control the droplet size by choosing the number of revolutions of the stirrer, since, as noted in [6–8], the droplet size during crushing is inversely proportional to the number of revolutions:

$a \sim (nd)^{-1.75}$ , where  $n$  – is the stirrer speed. By changing the number of revolutions of the stirrer, it is possible to control the size of the droplets during their coalescence, which is necessary for the separation and stratification of the extract and raffinate.

The purpose of this study is to analyze the physical and chemical phenomena that accompany the processes of liquid-phase extraction in mixing devices, and their use in wastewater treatment from asphalt-resinous compounds and the solid phase.

*Liquid-phase extraction in mixing devices. Process theory.* In the processes of liquid-phase extraction in mixing devices, in order to increase the interfacial surface, the solvent is crushed to the minimum droplet size. At high speeds of the revolutions per minute in the mixing devices, the flow becomes turbulent, and the intensity of turbulence is such that one should speak of developed or isotropic turbulence. The size of the minimum drops under conditions of isotropic turbulence depends on many parameters, among which it is important to note the number of revolutions of the stirrer, turbulence parameters (specific energy dissipation, intensity and scale of turbulence) and the physicochemical properties of the medium and extractant droplets (density, viscosity). In [6–9], such a dependence, which ensures the maximum value of the interfacial surface, is expressed by an equation of the form:

$$\alpha_{\min} = k_R (nd)^{-1.75} \left( \frac{\sigma^3 v_c}{\rho_c \rho_d^2} \right)^{1/4}, \quad (2)$$

here:  $k_R$  – is the coefficient determined from the experimental data;  $v_c$  – is the viscosity of the medium;  $\rho_d$  – is the particle density;  $\sigma$  – is the coefficient of surface tension.

The numerical value of the coefficient  $k_R$  is determined based on experimental data. The change in the average droplet size over time is expressed by the equation

$$\frac{da}{dt} = -q_R (a_{\min} - a) \quad (3)$$

where  $a$  – is the current size of the drops, the coefficient  $q_R \sim \varepsilon_R^{1/6} \phi$  depends on the specific dissipation energy and the current volume fraction of the drops. Assuming that in an isotropic turbulent flow, the specific energy dissipation is expressed by the following dependence  $\varepsilon_R \sim n^3 d^3$ , from equation (3) the dependence of the change in the average droplet size over time depending on their volume fraction and other parameters is determined. If

the scale of turbulence is more than the Kolmogorov scale  $\lambda > \lambda_0$ , then the frequency of crushing of the extractant can be estimated by the formula [9; 10]:

$$\omega(a) \approx C_1 \varphi_0 \left( \frac{\varepsilon_R}{a^2} \right)^{1/3} \exp \left( -C_2 \frac{\sigma}{\rho_c \varepsilon_R^{2/3} a^{5/3}} \right), \quad (4)$$

$\varphi_0$  – volume fraction of particles,  $C_1, C_2$  – coefficients,  $\varepsilon_R$  – energy dissipation in turbulent flow.

This formula was obtained in a theoretical aspect using the equations of mass transfer in a turbulent flow, and experimentally confirmed in [11–13].

In this case, the processes of stratification and separation of the emulsion are of particular interest [14]. The stratification of the emulsion into extract and raffinate, formed in the process of liquid-phase extraction, is associated with the size of the extract droplets and the rate of their settling ( $\rho_d > \rho_c$ ) or surfacing ( $\rho_d < \rho_c$ ) in the field of mass forces. As was noted in [6], the rate of settling or surfacing for small droplet sizes is determined by the Stokes law, however, with an increase in their size, a deviation from this law is observed depending on the Reynolds numbers of the particle  $Re_d = \frac{Ua}{\nu_c}$ , Morton  $Mo = \frac{g\eta_c^4 \Delta\rho}{\rho_c \sigma^3 \rho_c}$ , and

Weber  $We = \frac{\rho_c U^2 a}{\sigma}$ , which is explained by the

deformation of the droplet shape [9–13]. For low concentrations and small particle sizes with a degree of entrainment in a pulsating medium equal to  $\sim 1$ , a similar equation for the rate of gravitational settling of drops in an isotropic turbulent flow is expressed as

$$\frac{V_s}{U_\lambda} = Stk_\lambda \left( \frac{\nu_c}{\varepsilon_R^3} \right)^{1/4} g \left( 1 - \frac{\rho_c}{\rho_d} \right), \quad (5)$$

where  $Stk_\lambda = \tau_p / \tau_\lambda$  – is the Stokes number,  $U_s$  – is the Kolmogorov velocity scale,  $U_\lambda$  – is the velocity of turbulent fluctuations,  $U_\lambda = \left( \frac{\varepsilon_R^3}{\nu_c} \right)^{1/4} \cdot \tau_\lambda$  scale

$\lambda_0, \tau_\lambda = \lambda / U_\lambda = (\nu_c / \varepsilon_R)^{1/2}$  turbulence time scale or period of turbulent fluctuations of scale  $\lambda_0$ ,  $V_s$  – settling rate of Stokes droplets,  $\tau_p, \tau_\lambda$  – relaxation and pulsation time respectively. Comparison of expression (5) with experimental data [13; 14] in the range  $24.5 \leq Re_\lambda \leq 42.7$ ,  $0.92 \leq Stk_\lambda \leq 3.2$  and  $0.435 \leq \lambda \leq 0.606$  gives a fairly satisfactory relative error of  $\sim 6$ – $8$  %. Using the experimental data [15], the expressions for the dependences of the specific energy dissipation  $\varepsilon_R$  and the scale of turbulent fluctuations on the volume fraction of

drops ( $1.5 \cdot 10^{-6} < \varphi \leq 1.5 \cdot 10^{-4}$ ) can be expressed by the following empirical correlations:

$$\varepsilon_R = 340 - 2.96 \cdot 10^6 \varphi + 3.75 \cdot 10^{10} \varphi^2 - 10^{14} \varphi^3 \quad (6)$$

$$\lambda = 0.6 + 1714.5 \varphi - 4.94 \cdot 10^7 \varphi^2 + 2 \cdot 10^{11} \varphi^3.$$

The settling or floating of drops is complicated by such physical phenomena as: a) the Magnus effect, which acts on particles during their rotational motion; b) the Marangoni effect, which is characteristic of drops and bubbles and leads to the emergence of convective flows on the surface; c) the complexity of the internal circulation flow inside the drops, as well as the presence of various physical processes on the surface of the drops – evaporation (Fassiy effect, Stefan flow from the surface) and condensation growth of drops. The calculation of the rate of settling or surfacing of solid particles and deformable drops and bubbles must correspond to the hydrodynamic conditions under which the extraction process proceeds. The surfacing rate of extract droplets can be calculated using the formulas given in [9], and the separation thickness can be calculated using the formula

$$\frac{d\delta}{dt} = k(\delta_\infty - \delta), \quad t=0, \delta(0)=0, \quad (7)$$

where  $k = \frac{V_p}{\sqrt[3]{D_p \tau_p}}$ ,  $V_p$  – is the particle (coke)

settling rate,  $D$  – is the diffusion coefficient,  $\delta_\infty$  – is the maximum thickness of the supernatant layer. The relaxation time, or the time of complete stratification, for drops can be determined for liquids with different densities based on experimental studies, assuming that  $t_p \sim \Delta\rho L^2 / \eta_c$  ( $L$  – is the characteristic size). At constant physical and chemical properties, the volume of the surfaced layer can be determined by the equation

$$V = V_0 [1 - \exp(-kt)], \quad (8)$$

where  $V, V_0$  are the limiting and current volume of the surfaced phase. In [20], the calculated data according to equation (8) are compared with experimental data for oil–water ( $k = 0.2272$ ) and water–toluene ( $k = 0.45t - 1/2$ ) systems at an oil to water ratio of 30:1. At the same time, the amount of layered raffinate  $Q$  with organic residues can be determined by the expression  $Q = S_A \rho_{iz} \delta(t)$ , where  $S_A$  is the cross-sectional area of the apparatus,  $\rho_{iz}$  is the average density of the raffinate. In this case,  $\delta_\infty$  can be determined according to the expression  $\delta_\infty = \frac{Q_R}{S_A \rho_{iz}}$ , where  $Q_R$

is the amount of solvent and oil products contained in the emulsion, and the value of the coefficient  $k$  is estimated from experimental data. It should be noted that in practice, as a result of incomplete separation and separation of the extract from the raffinate, the condition  $Q < Q_R$  can be observed. Then, the separation efficiency can be estimated by the formula

$$\eta = \frac{Q}{Q_R} \cdot 100\% \quad (9)$$

If we take into account that the spreading of a liquid with a lower surface tension coefficient on the surface of a liquid with a higher surface tension, which is observed in the processes of liquid-phase extraction, leading to the appearance of a convective flow in the liquid, forms the basis of the Marangoni effect. The Marangoni effect is characteristic of liquid-liquid and liquid-gas systems and is the result of the appearance of a concentration gradient in mass transfer phenomena. The Marangoni effect plays a significant role in mass transfer processes (rectification, absorption, liquid extraction) [15; 16], and also affects the coalescence and crushing of droplets and their stratification in the flow. It displays itself in a change in the mass transfer coefficients due to the appearance of interfacial convection and in a change in the surface of the phase contact. In addition, the Marangoni effect affects the stability of interfacial films depending on the conditions and properties of the contacting liquids, and the stability of the films can increase or decrease depending on the value and sign of the concentration gradient of surface tension. The Marangoni effect is a partial correction to the surface tension coefficient, although it can have a significant effect on the nature of the flow and on the velocity distribution in the interfacial film [9]. The presence of two-dimensional pressures and the complexity of their distribution on the surface of a liquid thin film formed between two joined drops show that, when it becomes thinner, the presence of the Marangoni effect has a retarding effect on the rupture of the interfacial film between two drops and, to some extent, contributes to the stabilization and inhibition of coalescence processes in the system liquid-liquid. The Marangoni effect contributes to the temporary stabilization of the interfacial film, since at any point where the film becomes thinner due to the influence of external forces, a local increase in surface tension occurs, which counteracts

thinning and rupture. The Marangoni effect during the settling or floating of drops creates additional convective currents on the surface of the drops due to the difference in surface tension at different points on the surface of the drops. This significantly changes the circulation flow of the liquid inside the drop, as a result of the formation of many local convective flows on its surface, penetrating deep into the volume of the drop, which creates a certain chaos in the circulation flow and has a decelerating effect on the rate of settling and floating of drops [9]. In liquid phase extraction processes, the main factor is mass transfer between two immiscible liquids. Liquid extraction of organic compounds and oil from wastewater is a mass transfer process that occurs in the interfacial film and is carried out by dispersing the solvent extract in an aqueous medium and extraction in the interfacial film, characterized by diffusion transfers of the corresponding components to the interfacial surface.

The efficiency of mass transfer in extractors is determined by the size of the solvent droplets obtained as a result of their crushing under stirring conditions. The process of liquid-phase extraction of organic impurities in waste water with a solvent proceeds under conditions of intensive mixing and dispersion, due to which the conditions of isotropy of the turbulent flow in the extractor are achieved. If we exclude the mutual solubility of the components in organic impurities contained in wastewater, then for large values of the  $Pe > 10^3$  number, the mass transfer process can be considered steady with sufficient accuracy for practical purposes and can be considered in approximation to the diffusion boundary layer. In this case, the distribution of the substance concentration on the surface of a single extractant droplet is described by an equation of the form

$$\tilde{V}_r \frac{\partial \tilde{C}}{\partial r} + \frac{\tilde{V}_\theta}{r} \frac{\partial \tilde{C}}{\partial \theta} = \frac{\partial}{\partial r} \left[ D_T(r) \frac{\partial \tilde{C}}{\partial r} \right], \quad (10)$$

(the boundary conditions are  $r \rightarrow \infty, \tilde{C} = \tilde{C}_0, r=R, \tilde{C} = \tilde{C}_p$ ), where  $\tilde{C}_0$  and  $\tilde{C}_p$  – are the average concentrations of organic matter in the volume of the medium and on the surface of the solvent drop. Omitting the details of solving this type of equation, we finally determine the mass transfer coefficient between a turbulent flow and a drop in the form [17]:

$$\lambda > \lambda_0, \quad \beta_L = C_{11}(\varepsilon_R R)^{1/6} U^{1/2} \left( \frac{\alpha}{1+\gamma} \right)^{1/2},$$

$$\lambda < \lambda_0, \quad \beta_L = C_{12} \left( \frac{\varepsilon_R R}{v_c} \right)^{1/4} (UR)^{1/2} \left( \frac{\alpha}{1+\gamma} \right)^{1/2}, \quad (11)$$

Let's define the Sherwood number for ( $\lambda > \lambda_0$ ), using the expression

$$Sh = \frac{\beta_L R}{D_T} = C_{12} Sc^{1/2} \left( \frac{v_c^3}{\varepsilon_R R^4} \right)^{1/4} \left( \frac{Re_d}{1+\gamma} \right)^{1/2}. \quad (12)$$

Similarly, for ( $\lambda < \lambda_0$ ) we define the number Sh as

$$Sh = C_{11} Sc^{1/2} \left( \frac{v_c^3}{\varepsilon_R R^4} \right)^{1/4} \left( \frac{Re_d}{1+\gamma} \right)^{1/2}. \quad (13)$$

Omitting simple transformations taking into account gravity (floating or settling of drops), equation (14) can be transformed to the form

$$\frac{Sh}{Sc^{1/3}} = C_{13} \left[ \left( \frac{\varepsilon_R R^4}{v_c^3} \right)^{1/3} \right]^k, \quad (14)$$

where  $C_{13} = \frac{Sh}{Sc^{1/3}} \frac{Re^{11/2}}{(1+\gamma)^{1/2}}$ ,  $C_{11}$ ,  $C_{12}$ ,  $C_{13}$  are empirical coefficients,  $k$  – is the exponent, which is estimated on the basis of experimental data.

In mixing devices under conditions of isotropic turbulence for a liquid-liquid system, the turbulent diffusion coefficient, according to [18], is determined by the expression  $D_T \approx \varepsilon_R / v_c^{1/2} \lambda^2$ . Under conditions of intensive mixing, it can be assumed that the length of the path of movement of liquid particles is equal to the radius of the mixer, i.e.  $\lambda \approx d/2$ . Then, assuming that  $\varepsilon_R = f(Re_d) \frac{n^3 d^5}{V_0}$ , the turbulent

$$D_T \approx \frac{1}{4} \sqrt{f(Re_d)} \left( \frac{n^3 d^7}{v_c V_0} \right)^{1/2}, \quad (15)$$

where  $f(Re_d)$  – is the resistance coefficient of the medium, depending on the  $Re_d$  number,  $V_0$  – is the volume of liquid in the mixer.

As follows from equation (15), the turbulent diffusion coefficient is directly proportional to the rotation frequency  $n^{3/2}$  and the stirrer diameter, and inversely proportional to the medium viscosity  $v_c^{1/2}$ , i.e. with an increase in the viscosity of the medium (at low temperatures), the coefficient of turbulent diffusion decreases. In principle, the effective diffusion coefficient is defined as  $D_E = D + D_T$ , and under conditions of intensive mixing we have  $D_T \gg D$ , i.e. the turbulent diffusion coefficient is much (several orders of magnitude) larger than the molecular

diffusion coefficient. All necessary calculations were performed using the software package "OptimMe" [17].

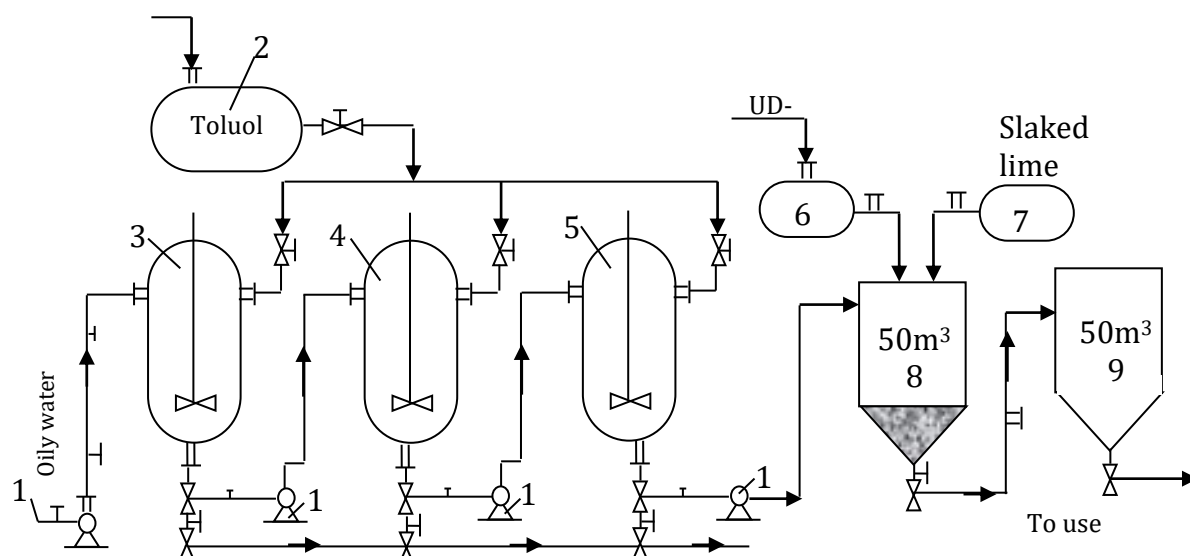
## Experimental

*Development of technology for wastewater treatment from asphaltene-resinous substances.* Asphaltenes in crude oil are solid particles with a size of 1–1.5 nm, which, at a certain concentration, form nanoaggregates, clusters of nanoaggregates, and then a viscoelastic framework that gives oil certain rheological properties of non-Newtonian liquids as a result of coagulation and agglomeration. The formation of aggregates from asphaltene particles also occurs in pipelines with an intense turbulent flow of oil. As noted in [19–21], as a result of experimental studies and analysis by ultrasonic methods and further by diffusion measurements by NMR methods, it was found that at an asphaltene concentration in oil of more than 4–10 (mass fraction), they coagulate with each other, forming nanoaggregates. If the concentration of asphaltenes in oil reaches 5 mg/l (mass fraction –  $5 \cdot 10^{-3}$ ), then nanoaggregates combine into clusters, which, at a concentration of more than  $10^{-2}$  (mass fraction), flocculate, forming a viscoelastic framework [22–25]. It was found that asphaltenes in toluene do not form aggregates and clusters, since the solid particles of asphaltenes in toluene first soften, then form a viscous mass and completely dissolve in it. It should be noted that particles of asphalt-resinous substances dissolve very well in aromatic hydrocarbons, which is an important factor for creating a technology for wastewater treatment from oil products.

Based on the above reasoning and experimental studies, a technological scheme for the wastewater treatment from asphalt-resinous substances and solid impurities using liquid-phase extraction was developed and proposed [26; 27]. At the same time, the extraction unit is proposed in the form of three extractors operating in the mode of sequential supply of waste water and cross supply of the extractant solvent. The figure shows such a flow chart, according to which waste water is pumped by pump 1 through three extractors 3–5 connected in series. A solvent, toluene, is partially supplied

from tank 2 to each extractor, which, as a result of intensive mixing, is dispersed in the form of droplets 10–100  $\mu\text{m}$  in size to increase the interfacial surface. Dispersion is carried out by

changing the stirrer speed in the range of 1500–2500 rpm, since the droplet size is proportional to the stirrer speed.



**Figure. Principle flowchart of wastewater treatment, oil and oil products pollution by liquid phase extraction method: 1 – pumps; 2 – tank for solvent; 3,4,5 – extractors; 6 – capacity for demulsified; 7 – capacity for slaked lime; 8 – mixing tank; 9 – rinsing tank**

The amount of solvent (toluene) in relation to water should be taken as 1 : (30–35), and its distribution between the extractors is as follows: I – 50 %; II – 30 %; III – 20 %. Thus, the hardware layout of the unit of cross-extraction of petroleum wastewater is presented in the form of three continuous extractors-mixers with the supply of toluene to each extractor. Solvent crushing is an important factor in increasing the efficiency of the liquid phase extraction process. As the number of revolutions of the stirrer increases, the secondary and tertiary crushing of the solvent drops down to their minimum size occurs. Based on engineering logic, the solvent should be fed to the area of intensive mixing of each extractor. It should be noted that with an increase in the frequency of grinding, on the one hand, the rate of the mass transfer extraction process increases due to the increase in the interfacial surface, on the other hand, the increase in the number of particles in the volume of the flow increases the probability and number of collisions between the solvent droplets and the absorbed substance from the wastewater. This factor leads to an increase in the rate of coalescence of drops and the probability of separation of the extract and raffinate phases. After the system of extractors, the mixture is fed into sump 5, where, due to the expansion of the flow, there is a stratification of the extract from the main water, in which still suspended particles of the solid phase remain.

Processes of stratification of products after extraction in settling devices due to gravity or density difference between the extractant and water ( $\Delta\rho \approx 200 \text{ kg/m}^3$ ) play an important role for the entire flowchart, since the qualitative separation of these immiscible phases can become limiting for the entire process of extraction treatment. If the liquid – the extractant – is in a dispersed form, i.e. in the form of droplets, so in order to effectively separate these phases, it is important to create conditions for collision, coalescence and coarsening of droplets. It should be noted that the rate of coalescence, the rate of their floating (droplet size, viscosity and density of the medium) and the coefficient of effective diffusion of toluene droplets in water have a significant influence on the efficiency of the separation of droplets. Further, after extraction, the wastewater contained in the bottom of the extractor is fed into mixing tank 8. Waste water contains only finely dispersed solid phase, consisting mainly of particles of sand, clay, mineral salts, etc. The size of the particles of the solid phase in water is so small (almost colloidal size) that they do not settle under the action of body forces. Therefore, a coagulant enters from tank 6, and slaked lime from tank 7. Wastewater with solid phase from the bottom of tank 5 is fed into mixing tank 8, where the coagulant is supplied from tank 6. The universal demulsifier UD–1 is used as a coagulant, which is a dark blue



liquid with a specific gravity of  $d_4^{20} = 1.2737$  g/cm<sup>3</sup> and a boiling point of 216–218 °C. UD-1 was obtained by the condensation of monomethylamine with phenylformaldehyde in the presence of iron trichloride followed by its treatment with a phenylcarbonyl compound of ferrocene. UD-1 is highly soluble in both organic and inorganic solvents. When dissolved in water, UD-1 does not hydrolyze or decompose. With intensive stirring in apparatus 8, a “cotton-like” substance is formed in the volume of waste water. As a result, a solid precipitate is formed, which is freely deposited on the bottom of the apparatus 8 and is removed from there. Purified clear water from apparatus 8 is fed into clean

water tank 9 and can be used for 9 practical purposes. Purified industrial water, can be used to feed into the oil reservoir in order to maintain reservoir pressure in the well or for other purposes, in particular, for irrigation.

## Conclusions

Thus, the proposed science-based technology for waste oil water treatment from asphaltene-resinous substances using solvent recycling is ecologically and economically beneficial, since oil waste contained in water and polluting the environment is neutralized. The most important thing is that the proposed process is waste-free and energy-efficient.

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