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REVIEW ARTICLE

Comparative analysis of liquid mixture separation flowsheets

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Abstract

Objectives. When developing separation flowsheets for liquid mixtures, preference is often given to a specific process or flowsheet. Although alternative separation variants are sometimes considered, these tend to be based on a single-phase process, usually distillation. And while review papers on the specifics of implementing a particular separation technique exist, these mainly focus on the specific process of extractive distillation, combination of distillation and splitting processes, and extraction. Moreover, studies comparing the separation flowsheets of mixtures of different physicochemical nature based on different processes and special methods are fragmentary. This study presents a comparative analysis of the processes and methods of liquid mixtures separation based on a critical review of the literature and the authors' own research results.

Methods. The study is based on the critical analysis of literature and mathematical modeling of phase equilibria using local composition equations via freely distributed software packages.

Results. Specific liquid mixture separation methods, including combining various processes in one flowsheet (including hybrid technologies), are compared in terms of their advantages and disadvantages.

Conclusions. Promising areas of further research in the field of synthesis of organic mixtures separation flowsheets through the use of various separation processes and methods are identified. The effectiveness of the various processes (extraction, splitting, special distillation techniques) is estimated at different stages of different number of components mixtures separation. A comparative analysis of extractive and heteroazeotropic distillation processes when separating mixtures of different initial composition highlights the areas of energy advantage of each process. The effectiveness of the flowsheets is estimated by combining extraction with other processes, depending on the stage of extractant regeneration.

Keywords
separation flowsheet, phase equilibria, distillation, splitting, pressure swing-distillation, extraction, separating agents, energy consumption

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ОБЗОРНАЯ СТАТЬЯ

Сравнительный анализ схем разделения жидких смесей, основанных на различных процессах и методах

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Аннотация

Цели. При разработке схем разделения жидких смесей часто отдается предпочтение конкретному процессу или схеме. В редких случаях рассматривается не один, а несколько альтернативных вариантов разделения, чаще основанных на одном фазовом процессе, как правило, ректификации. Имеющиеся на сегодня обзорные работы по особенностям реализации того или иного приема разделения сконцентрированы на конкретном процессе: экстрактивная ректификация, сочетание ректификации и расслаивания, экстракция. Комплексные исследования по сравнению схем разделения смесей разной физико-химической природы, базирующихся на разных процессах и специальных методах, не проводились. Научные публикации в данной области представлены фрагментарно и относятся к определенным объектам исследования. Целью настоящей работы является сравнительный анализ процессов и методов разделения жидких смесей на основе критического обзора литературы и результатов собственных исследований.

Методы. Работа базируется на критическом анализе литературы и математическом моделировании фазовых равновесий на основе уравнений локальных составов с использованием свободно распространяемых программных комплексов.

Результаты. Проведен обзор научной литературы, на основе которого показаны достоинства и недостатки конкретных приемов разделения жидких смесей, преимущества и ограничения на сочетание в одной схеме различных процессов (в том числе в гибридных технологиях).

Выводы. Перспективными направлениями дальнейших исследований в области синтеза и сравнения схем разделения смесей органических продуктов за счет использования различных методов и процессов разделения являются: оценка эффективности применения различных процессов (экстракции, расслаивания, специальных приемов ректификации) на разных этапах разделения смеси различной компонентности; сравнительный анализ процессов экстрактивной и гетероазеотропной ректификации при разделении смесей разного исходного состава (выделение областей энергетического преимущества каждого процесса); оценка эффективности реализации схем, основанных на сочетании экстракции с другими процессами, в зависимости от стадии регенерации экстрагента.

Ключевые слова

схема разделения, фазовые равновесия, ректификация, расслаивание, варьирование давления, экстракция, экстрактивные агенты, энергозатраты

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INTRODUCTION

A basis for the development of technological separation flowsheets is provided by information about the phase behavior of the system, including under changing external conditions (pressure, temperature), as well as in the presence of additional substances. The diversity of manifestations of phase behavior forms a basis for separation process flowsheets having different structures based on specific phase processes or their combinations. As such, each process has its own limitations related to the peculiarities of the phase equilibrium of the systems: for distillation process, such limitations include azeotropes, separatrix varieties and the proximity of component volatilities; for phase-separation processes, the presence of isopycnics (when implemented in a liquid separator); for extraction, the presence of solutropes; for crystallization, eutectic points, etc.

The separation techniques can be conditionally divided into two groups: the first being based on the use of the internal reserves of the system (the addition of a new component into the system is excluded), while the second always involves the addition of new substances. Here, by internal reserves, we refer to the specifics of phase behavior manifestation, which are influenced by the shift of the azeotrope under pressure (separation in a complex of columns operating at different pressures), the belonging of the azeotrope to the splitting region (a combination of distillation and splitting process), the effect of one of the components present in the mixture on the relative volatility of other components pairs (autoextractive distillation (AED)), different compositions of azeotropic and eutectic points (a combination of distillation and crystallization), and the curvature of separatrix manifolds. Such reserves represent possibilities for overcoming the limitations imposed by the nature of the mixture itself. The second group involves the addition of one or more new substances whose task is to transform the phase space and dynamic system of the separation process. These include processes such as extractive distillation (ED), re-extractive distillation (RED), (hetero)azeotropic distillation (HAD), as well as extraction and separation in the presence of an inert gas.

The peculiarities of the phase behavior of systems with different component compositions (primarily binary and ternary) have been fairly well studied along with the general patterns of their separation processes [1–3]. A large number of publications are dedicated to the study of various separation processes of specific mixtures, the vast majority of them pertaining to the ED process. This process has clear advantages over alternative separation methods: it is universal in terms of the properties of the initial mixture, has virtually no limitations, and is considered an energy-efficient separation technique. The main disadvantages of the process include the necessity for adding a new substance, which can negatively affect the quality of the obtained products, and the requirement of an additional column for agent regeneration. Summaries of the experience of using the ED process for the separation of mixtures of different physicochemical natures are presented in the previous studies [4–8].

Some energy-efficient separation methods combine distillation and splitting processes. Due to the common presence in reaction mixtures for the production of organic substances of components having limited mutual solubility (organic and water-organic mixtures), such separation methods are used quite frequently. A summary of the features of implementing this method for combining two processes and the effectiveness of its application for separating liquid mixtures is provided in studies [9, 10]. The authors discuss possible ways to reduce energy consumption in such flowsheets by using additional liquid separators (changing the qualitative composition of the reflux flow) and external decanters, as well as by reorganizing the flow structure of the flowsheet.

There are a few review articles dedicated to the efficiency of using extraction and various extractants (classical solvents, ionic liquids, deep eutectic solvents) for the separation of liquid mixtures [11–13]. Other studies discuss the effectiveness of applying the crystallization process [14–16]. However, while there are plenty of articles focused on developing separation technologies for specific mixtures, the scientific literature lacks studies on the broader use or combination of these methods.

The present article is dedicated to the analysis of literary sources that compare various separation methods: the combination of processes (extraction / distillation / phase separation), the use of special distillation methods (pressure-swing distillation (PSD), HAD, ED, AED, RED, (auto)extractive-heteroazeotropic distillation (AEHAD)). To identify the characteristics and conditions for the implementation of a particular process, a computational experiment is conducted using the Non-Random Two-Liquid model, whose parameters are taken from the database of the software package or the corresponding scientific article. All sets of parameters reproduce the features of the phase behavior of the systems with an error not exceeding 3%.

The presentation of the material requires the systematization and structuring of such basic concepts as process, method, technique, and separation mode. The processes discussed in the article include distillation, extraction and splitting (based on first-order phase transitions), as well as pervaporation (separation through a membrane). Here method refers to the way to achieve the required qualitative separation of a mixture (for example, ED, PSD, etc.), while techniques involve the combination of one (homogeneous flowsheets) or several (heterogeneous or hybrid flowsheets) methods within a single technological flowsheet (for example, the combination of distillation and extraction), fractionation, or pre-concentration of a multi-component mixture (through the use of a specific method or process). The mode of operation determines the nature of the equipment's functioning (periodic, continuous, stationary, dynamic), as well as the set of technological parameters that ensure a certain separation result (for example, pressure in the columns, reflux ratio, solvent flow rate, number of separation stages, etc.).

The structure of the separation technological flowsheet is determined by the specifics of the organization and direction of recycle flows. In the absence of the latter, the flowsheet has a linear appearance; if all the apparatuses of the flowsheet are covered by feedback (recycle flow), it represents a separation complex. The flowsheet can consist of a linear part and one or several complexes; the most complicated structure of the flowsheet is a "complex within a complex."

THEORETICAL ANALYSIS

The physicochemical principles of separation processes are well known and thoroughly covered in the studies [1–3, 11–13, 17]. Each of the processes, methods and separation techniques is characterized by its own limitations and possesses its own advantages and disadvantages, as mentioned above.

The first stage of the synthesis of separation flowsheet involves the study of phase behavior (thermodynamic-topological analysis of the phase equilibrium diagram structure) [2, 17]. This analysis comprises the following stages:

1) determine the structure of the phase portrait (the diversity of types of singular points formed not only by pure components but also by their mixtures is manifested on phase diagrams of liquid–vapor and liquid–solid equilibrium; splitting phase diagrams (containing different numbers of liquid phases) are characterized by the presence of various varieties of different dimensionalities (binoidal/critical), splitting phase simplices);
2) establish the presence of phase equilibrium constraints on specific processes;
3) determine the limiting capabilities (maximum achievable compositions) of processes (for example, by determining the boundaries of areas and sub-areas of distillation or crystallization, the boundaries of the splitting area);
4) determine the characteristics of the change in the structure of the diagram of a specific type of equilibrium (including the possibility of lifting constraints) when conditions (pressure, temperature) change;
5) assess the feasibility and prospects of combining different processes through comparative analysis of the structures of diagrams related to different types of equilibrium (for example, by superimposing one diagram onto another) or to one type under various conditions;
6) determine the qualitative compositions of impurity components (for example, when isolating pure substances in the distillate or the bottom of a distillation column based on the analysis of the arrangement features of distillation line bundles or when separating a mixture in a liquid separator, when one or both layers may potentially represent the product flow);
7) identify the mutual influence of the components of the base mixture on each other (changes in relative volatility, formation of (hetero)azeotropes with minimum boiling point, etc.) to assess the feasibility of using so-called "auto" processes (AED, autoheteroazeotropic distillation);
8) determine the behavior of the system in the presence of specially selected substances for processes such as ED and RED (change in relative volatility), HAD (formation of a new azeotrope with the lowest boiling point), distillation with a medium-boiling agent (separation of the mixture through conventional distillation by correctly selecting the desired distillation type), and extraction (formation of a two-phase splitting region).

Even at the stage of studying the phase behavior of the system, it is possible to discriminate some processes, methods of mixture separation, as well as the possibilities of their combination (approaches). Exclusion of variants is related not only to the principled impossibility of implementing a particular process but also to the feasibility of its implementation (for example, separation of a mixture in a column complex under different pressures with a small azeotrope shift, in an ED complex with a low-selectivity agent, in a hybrid flowsheet combining distillation and crystallization, with similar azeotrope and eutectic points compositions, etc.).

The variants for a multitude of synthesized separation flowsheets can further be discriminated based on certain qualitative or quantitative criteria: by the number of apparatuses, the amount of material flows (when the number of degrees of freedom in the separation flowsheets is greater than zero and part of the parameters should be arbitrarily set for material balance calculations), etc. The selected flowsheets after discrimination can be modified (structural optimization) by coupling material and thermal flows, using additional equipment (for example, separators, heat pumps), using complex dividing wall columns (with side section flows, with internal walls), etc.

The set of separation flowsheets formed at the final stage of structural optimization is subject to parametric optimization. The criteria for selecting optimal parameters (while achieving the required quality of product fractions) can include the total duty of column reboilers (for distillation separation flowsheets), total operating costs, total annual costs (TAC), environmental indicators (carbon dioxide emissions, water consumption, wastewater generation), as well as the criterion of best available technologies as a whole. It is important to note that, regardless of the choice of criterion, the comparison of variants should be considered correct, for which not only the required product quality is achieved, but also the adequacy (optimality) of the selected conditions and parameters, as well as selectivity of the chosen solvents are taken into account. For example, it would be incorrect to compare processes based on the addition of a new substance, one of which is low-selective, since in this case, one of the processes or flowsheets is placed in a priori unfavorable conditions.

(Auto)ED and (auto)HAD

Both methods involve the addition of a new substance, except for auto processes. ED and HAD are considered as energy-efficient separation techniques. For example, in the ED process, separation efficiency is enhanced by the multi-level feeding of the agent and the initial mixture. It creates an additional counterflow in the column and forms an area of maximum agent concentration in the middle part of the column since the work expended on separating the mixture is partially compensated by the work of mixing the two feed flows. The addition of a new substance in the HAD process leads to the formation of an unstable node type azeotrope, which favorably affects the reduction of temperature along the height of the column.

A brief description of the processes and separation flowsheets for specific mixtures is provided in Table 1.

Table 1. Brief characterization of mixtures and their separation processes using (auto)extractive distillation ((A)ED) and (auto) heteroazeotropic distillation ((A)HAD)

System	Specifics of vapor–liquid equilibrium (VLE)*, process separating agent (SA)	Comparison indicators	(A)HAD	(A)ED
Ethanol–water [18]	Positive azeotrope; for HAD—pentane for ED—ethylene glycol	Equipment ΣN ΣQ	2C+S 54 6942.8	2C 33 2007.8
		Equipment ΣN ΣQ	DWC 1C+S 60 5538.1	DWC 1C 36 1819.5
Propionic acid–water [19]	Positive azeotrope; for HAD—cyclohexanol for ED—*N*-MP	Equipment ΣN ΣQ	1C+S 18 376.8	2C 40 1816.4
Acetic acid–water [19]	– for HAD—vinyl acetate for ED—*N*-methylacetamide	Equipment ΣN ΣQ	1C+S 18 1070.0	2C 88 1832.3
1-Methoxy-2-propanol–water [19]	Positive azeotrope; for HAD—isopropyl acetate for ED—sulfolane	Equipment ΣN ΣQ	2C+S 36 1795.9	2C 52 2271.3
1-Methoxy-2-propanol–water [20]	Positive azeotrope; for HAD—isopropyl acetate; for ED—sulfolane / *N*-MP	Equipment ΣN ΣQ TAC × 10^6	2C+S 37 11870 34.5	2C / 2C 35 / 72 6600 / 9040 23.2 / 33.0
Ethanol–toluene–water [21]	3.3.1-2; for HAD—toluene for ED—glycerol	Equipment ΣN ΣQ TAC × 10^6	3C+S 92 1017.6 3.66	2C+S 42 455.2 1.96
		PCHMF flowsheet ΣN ΣQ TAC × 10^6	92 735.8 3.23	–

Table 1. Continued

System	Specifics of vapor–liquid equilibrium (VLE)*, process separating agent (SA)	Comparison indicators	(A)HAD	(A)ED
Methanol–water–methyl methacrylate [22]	3.2.0-2b; for AED—water for ED—phenol / DMSO and phenol	Equipment ΣN ΣQ	3C+S 58 2369.7	3C / 4C 89 / 107 4395.6 / 4477.4

Note: * azeotrope type or VLE diagram class according to L.A. Serafimov's classification [23]; *N*-MP is *N*-methyl-2-pyrrolidone; DMSO is dimethyl sulfoxide; *N* is a number of theoretical stages; *Q* is reboiler duty, kW; C — column, S — liquid separator; DWC is a divided-wall column—indicated for flowsheets that consider the same technology, but with the replacement of a conventional column with a complex one; PCHMF are partially coupled heat and material flows; HAD — heteroazeotropic distillation; ED — extractive distillation; TAC — total annual costs, USD/year.

A comparison of the separation flowsheets of the ethanol–water mixture showed that ED with ethylene glycol is a less energy-intensive separation variant (approximately 70% savings compared to HAD) [18]. Such a significant difference in energy consumption is due to the selection of a low-selectivity solvent (pentane) for the HAD process, which affected its consumption (for HAD 751.6 kmol/h, for ED 190 kmol/h per 100 kmol/h of the initial mixture) (Fig. 1, the ratio of the amounts of the pentane and aqueous phases is above 12). Moreover, pentane needs to be evaporated in the HAD process. Increasing the efficiency of this process is possible by using another agent (for example, benzene, cyclohexane [1]). Conversely, ethylene glycol is claimed to be a selective solvent for the ED process [24]. However, the comparison provided by the authors appears to be biased since the HAD process is placed in an a priori disadvantageous position.

The use of complex dividing wall columns allowed for a reduction in energy consumption for both HAD (by 20%) and ED (by 9.5%), which did not affect the cost ratio between the flowsheets. The difference in the placement of the dividing wall (at the top for HAD and at the bottom of the column for ED) is related to the implementation specifics of the process.

The study [19] shows that when separating mixtures of high-boiling components, HAD can be more energy-efficient than ED due to the reduction in process temperature through the formation of azeotropes with a minimum boiling point and the possibility of eliminating one of the columns (one of the layers leaving the separator represents the product flow). As the boiling point of the components in the base mixture increases, the efficiency of the ED process decreases. When selecting an extractive agent (EA) for the HAD process, it is necessary to comprehensively evaluate the characteristics of the liquid–liquid equilibrium (under separator operating conditions) and the liquid–liquid–vapor equilibrium due to the latter affecting the amount of the recycle flow [19]. It should be noted that for both processes, the present authors selected the most selective agents on the basis of literature analysis and their own research results to inform the comparison correctly.

When separating the 1-methoxy-2-propanol-water mixture of composition F (Fig. 2) in a two-column flowsheet, the HAD process is carried out in the second

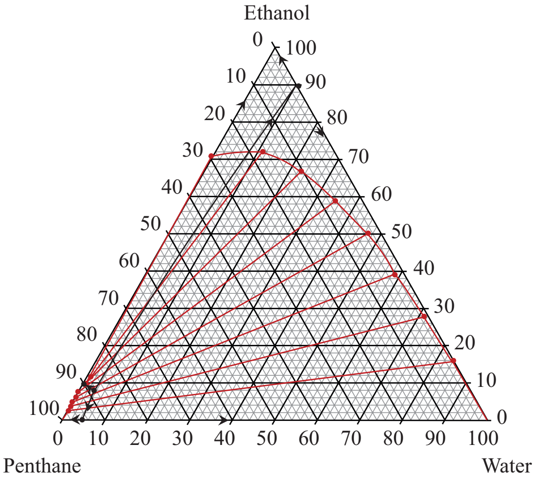
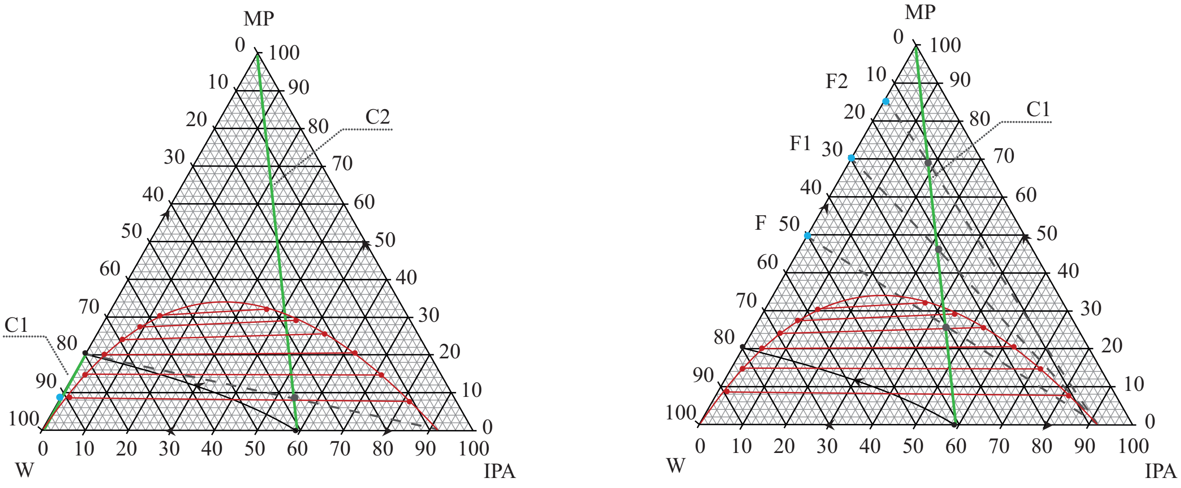


Fig. 1. Structure of the ethanol–water–pentane system phase diagram



(a) (b)

Fig. 2. Balance lines of the heteroazeotropic distillation (HAD) flowsheet of the mixture 1-methoxy-2-propanol (MP)–water (W) in the presence of isopropylacetate (IPA): (a) HAD is realized in the second column C2, (b) HAD is realized in the first column C1

column [20]. The first column (C1) is intended for the partial separation of water to subsequently reduce the load on the azeotropic distillation column. In the second column (C2), the flows of distillate from the first column and the organic layer leaving the liquid separator are mixed, followed by separation into pure 1-methoxy-2-propanol and a mixture of the azeotropic composition of water–isopropyl acetate. The balance lines are shown in Fig. 2a. Due to the low solubility of isopropyl acetate in water, residual water can be separated in the form of a product flow (if necessary, this flow can be further purified in an additional distillation column). Comparing the structure of flowsheets that differ in the organization of the HAD in column C2 or C1, the possibility of reducing the ratio of distillate and bottom flows in the HAD column (the balance green line of column C2 in Fig. 2a) compared to the same ratio in the HAD column in column C1 (the balance green line of column C1 in Fig. 2b) should be noted. However, if the water concentration in the aqueous layer of the separator meets the specified requirements, the second column in the classical HAD flowsheet will be absent.

For the ED process of the considered binary mixture, two selective solvents were chosen: sulfolane and *N*-methyl-2-pyrrolidone (*N*-MP), with the selectivity of the former being 1.5 times higher, which resulted in lower agent consumption and lower energy costs for the complex compared to *N*-MP. The energy costs of the ED flowsheets with *N*-MP and HAD are comparable (a difference of 4% in favor of ED), while for ED with sulfolane they are significantly lower (by more than 30%). This example clearly illustrates how significantly the separation results in terms of energy costs can differ when comparing ED and HAD flowsheets with effective solvents.

Comparing the results of studies [19] and [20], in which the same mixture is subject to separation but with different compositions, it can be noted that with an increase in the concentration of 1-methoxy-2-propanol in the initial mixture (compositions F1 and F2 in Fig. 2b), the efficiency of applying the HAD process with isopropyl acetate increases, while the ED process with sulfolane decreases. This is explained by the decrease in isopropyl acetate consumption and consequent distillate- bottom ratio in the HAD column (Fig. 2b) (the product flows are represented in the column's bottom). Thus, when comparing the ED and HAD processes, it is important to consider not only the boiling points of the components of the base mixture (light or heavy boiling) [19], but also its composition, which will directly affect the solvent consumption in HAD.

For the separation of the ethanol–toluene–water mixture, two methods have been proposed: HAD, including with partially coupled heat and material flows (PCHMF), and ED [21]. The first method is called HAD by the authors, as the material flows of the flowsheet are organized in such a way that a ternary heteroazeotrope is separated in the distillate of the first column (ethanol in the bottom) (however, no new agent is added into the system). ED turned out to be the more advantageous variant (lower energy and TAC). This is due to the necessity of using only two distillation columns; a liquid separator was used to separate the water–toluene pair (a combination of ED and liquid phase splitting). However, the cost calculations for this process did not take into account the necessity of using expensive refrigerants for condensing the vapors of the distillate from the regeneration column (the top temperature of the column is 8°C). It should be noted that the initial structure of the phase equilibrium diagram of the ternary system is unfavorable for the implementation of the method based on the combination of distillation and phase splitting: near the ternary azeotrope in the region with a stable node—ethanol, there is a touching of the separatrix (Fig. 3), which ultimately leads to the difficulty of isolating this alcohol and achieving the azeotropic composition (the column efficiency was 75 theoretical stages). The ethanol separation column in the HAD flowsheet is also characterized by the highest energy costs (more than twice compared to the other columns in both flowsheets), i.e., this method is placed in an a priori disadvantageous position. The application of PCHMF allowed for a reduction in energy costs for the HAD flowsheet, but it did not make it competitive with the ED process.

When separating the methanol–methyl methacrylate–water mixture, separation flowsheets based on the use of additional substances were compared: ED, AED, extraction [22]. The use of water in the first stage of the extraction process is ineffective due to the extract flow consisting of a ternary mixture with a predominant content of water and methanol (with an impurity of about 1% methyl methacrylate), while the raffinate flow is binary (methyl methacrylate with an impurity of about 5% water). To isolate all components in pure form, additional methods will need to be employed, which will negatively impact energy costs. The lowest energy costs are observed for the AED flowsheet with water. Water not only increases the volatility of methyl methacrylate, but also lowers the process temperature due to the formation of a positive azeotrope, which positively affects energy consumption. The latter for the AED column is four times lower compared to the ED column with phenol or dimethyl sulfoxide (DMSO). Concerning the temperature profile of the AED column between the levels of the feed mixture and EA, a decrease in process temperature is observed, averaging around 56°C, with; here, the distillate temperature is approximately 80°C, while the bottom temperature is about 85°C. With the temperature profile of the ED column with DMSO, on the contrary, an increase in temperature averaging up to 100°C is observed.

Despite AED being more advantageous overall, it is necessary to pay attention to the flowsheet with phenol, as it acts as a stabilizer for methyl methacrylate to prevent its polymerization. The higher quality of the target product in these flowsheets should additionally be noted.

The combination of the two methods discussed, provided that EA is a component of the base mixture (AEHAD in one apparatus), is generally the most efficient option in terms of energy consumption for separation [22, 25, 26].

ED and PSD

The characteristics of the processes and separation flowsheets of specific mixtures are presented in Table 2.

For almost all the mixtures in Table 2, whose separation is planned in the ED or PSD complex, an economic analysis is conducted with the determination of annual costs. The correlation between the ratio of energy costs for the separation of these processes and the TAC criterion is quite justified considering the fact that operating costs generally contribute more to the annual totals. An exception is the mixture of dipropyl ether–*n*-propanol [28] for which the ED process shows lower energy costs for column reboilers, but higher TAC. Such a difference is due to the contribution of capital costs to the TAC indicator (to achieve the required product quality, the ED complex requires a greater total number of theoretical separation stages, 67 *vs* 24). The authors [28] note that the difference in annual costs will decrease with an increase in production volume due to the increased load on the column reboilers.

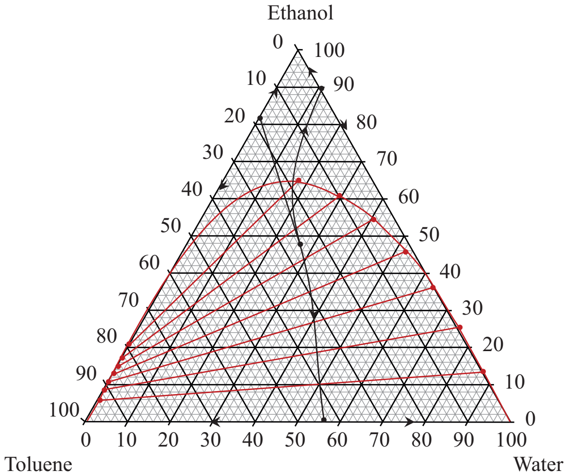


Fig. 3. Structure of the ethanol–toluene–water system phase equilibrium diagram

Table 2. Brief characterization of mixtures and their separation processes using ED and PSD

System	Specifics of VLE*	Comparison indicators	PSD	ED
Isopropanol–diisopropyl ether [27]	Positive azeotrope; for ED—2-methoxyethanol	Equipment	2C	2C
		ΣN	56	106
		ΣQ	4293.6	3027.1
		TAC × 10^6	–	1.964
		PCHMF flowsheet		
		ΣN	56	120
		ΣQ	4085.2	2677.1
		TAC × 10^6	1.851	2.149
Dipropyl ether–*n*-propanol [28]	Positive azeotrope; for ED—2-methoxyethanol	Equipment	2C	2C
		ΣN	24	67
		ΣQ	1058.3	816.7
		TAC × 10^6	0.632	0.896
		Capital expenses × 10^6	1.479	2.364
Acetonitrile–*n*-propanol [29]	Positive azeotrope; for ED—*N*-MP	Equipment	2C (PCHMF)	1C (DWC)
		ΣN	41	38
		ΣQ	2133.3	2716.7
		TAC × 10^6	0.920	1.18
Tetrahydrofuran–water [30]	Positive azeotrope; for ED—DMSO	Equipment	2C	2C
		ΣN	29	35
		ΣQ	1985.9	1391.7
		Capital expenses × 10^6	0.430	0.473
		TAC × 10^6	0.600	0.469
		PCHMF flowsheet		
		ΣN	29	35
		ΣQ	1450.4	1324.9
		Capital expenses × 10^6	0.426	0.470
		TAC × 10^6	0.479	0.454
		FCHMF flowsheet		–
		ΣN	29	
		ΣQ	1414.16	
		Capital expenses × 10^6	0.458	
		TAC × 10^6	0.487	
Isobutanol–isobutyl acetate [31]	Positive azeotrope; for ED—*n*-butylpropionate	Equipment	2C	2C
		ΣN	46	80
		ΣQ	1275.0	930.6
		Capital expenses × 10^6	2.79	4.62
		TAC × 10^6	1.26	1.69
Methanol–chloroform [32]	Positive azeotrope; for ED—*n*-propanol	Equipment	2C	2C
		ΣN	49	156
		ΣQ	3220.0	10270.0
		Capital expenses × 10^6	1.61	7.34
		Energy expenses × 10^6	0.737	2.29
		TAC × 10^6	1.27	4.73
		Including VDC		
		Capital expenses × 10^6	1.48	7.27
		Energy expenses × 10^6	0.737	2.29
		TAC × 10^6	1.23	4.71
		Savings when using the VDC column	10.36%	1.01%

Table 2. Continued

System	Specifics of VLE*	Comparison indicators	PSD	ED
Acetone–methanol [32]	Positive azeotrope; for ED—water	Savings when using the VDC column	4.18%	1.85%
Acetone–chloroform [32]	Negative azeotrope	Savings when using the VDC column	2.57%	1.07%
Benzene–cyclohexane [32]	Positive azeotrope	Savings when using the VDC column	0.57%	0.36%
Isopropanol–diisopropyl ether [32]	Positive azeotrope	Savings when using the VDC column	7.96%	0.00%
Methanol–isopentane–*n*-pentane–1-pentene–2-pentene [33]	Positive azeotropes in binary subsystems; for ED—water	Equipment ΣN ΣQ Capital expenses $\times 10^6$ Energy expenses $\times 10^6$ TAC $\times 10^6$	2C 20 46700.0 4.25 6.92 8.34	2C 44 14730.0 2.44 2.22 3.04
Acetone–methanol [34]	Positive azeotrope; for ED—water	Equipment ΣN ΣQ Capital expenses $\times 10^6$ Energy expenses $\times 10^6$ TAC $\times 10^6$	3C 79 2733.3 0.327 1.77 −8.52**	2C 62 1197.2 0.175 0.776 −9.34**
		Including PCHMF Capital expenses $\times 10^6$ Energy expenses $\times 10^6$ TAC $\times 10^6$	0.327 1.37 −8.92**	–
Acetone–methanol [35]	Positive azeotrope; for ED—water	Equipment ΣN ΣQ Capital expenses $\times 10^6$ Energy expenses $\times 10^6$ TAC $\times 10^6$	2C 114 23220.0 3.51 3.35 4.52	2C 83 18510.0 3.03 2.72 3.75
		Including PCHMF Capital expenses $\times 10^6$ Energy expenses $\times 10^6$ TAC $\times 10^6$	2.65 2.45 3.33	3.00 1.98 2.98
Acetone–chloroform [36]	Positive azeotrope; for ED—DMSO	Equipment ΣN ΣQ Capital expenses $\times 10^6$ Energy expenses $\times 10^6$ TAC $\times 10^6$	2C 78 21508.0 3.65 3.11 4.33	2C 34 2628.0 0.65 0.74 0.95
Methanol–chloroform [37]	Positive azeotrope; for ED—1-propanol	Equipment ΣN ΣQ Capital expenses $\times 10^6$ Energy expenses $\times 10^6$ TAC $\times 10^6$	2C 49 3350.0 0.89 0.43 0.73	2C 156 9070.0 2.62 1.34 2.22
		Including PCHMF Capital expenses $\times 10^6$ Energy expenses $\times 10^6$ TAC $\times 10^6$	0.76 0.31 0.56	–

Table 2. Continued

System	Specifics of VLE*	Comparison indicators	PSD	ED
Methyl acetate–methanol–acetic acid–acetic anhydride [38]	Positive azeotrope: methylacetate–methanol for ED—ethylene glycol	Equipment ΣQ	4C 67000.0	4C 24500.0
Isopropanol–water [39]	Positive azeotrope; for PSD—tetrahydrofuran for ED—DMSO	Equipment ΣN ΣQ Capital expenses × 10^6 Energy expenses × 10^6 TAC × 10^6	2C 75 7916.2 1.94 1.77 2.42	2C 75 2097.5 0.78 0.54 0.80
		Including PCHMF Capital expenses × 10^6 Energy expenses × 10^6 TAC × 10^6	1.65 1.10 1.65	–

Note: * azeotrope type is indicated; N is a number of theoretical stages; Q is reboiler duty, kW; C — column, S — liquid separator, DWC is a divided-wall column, VDC is a variable diameter column; P(F)CHMF are partially (fully) coupled heat and material flows; economic indicators are presented in USD/year; ** authors indicate profit.

Comparison of separation flowsheets (based on ED or PSD) of binary mixtures with different physicochemical properties (with positive or negative deviations from ideality) (Table 2) showed a nearly uniform distribution between variants where PSD [29, 31, 32, 39] or ED [33–36, 38] is more advantageous. The costs of separation flowsheets for the tetrahydrofuran-water mixture (ED with DMSO and PSD) turned out to be comparable [30].

It should be noted that such a comparison is not always correct. For example, when separating a methanol–chloroform mixture in an ED complex, a low-efficiency (low-selectivity) *n*-propanol [32, 37] EA was chosen, which affected the process indicators and equipment parameters: the total number of theoretical trays (N) in the ED flowsheet was 156 stages, while the reflux ratios in the columns were 7.64 and 7.9, respectively; here, the agent consumption was 600 kmol/h per 100 kmol/h of the initial mixture. The conducted own calculation of the vapor-liquid equilibrium in the derivative ternary system methanol–chloroform–*n*-propanol showed that the composition simplex of the base binary system is divided into composition areas for which the relative volatility of the mixture components changes differently with increasing EA consumption. In the range of 0 to 0.5 mole fractions of chloroform, the relative volatility of the chloroform–methanol pair decreases with increasing propanol concentration: in the range of 0.50–0.56 mole fractions, it passes through an extremum, while in the range of 0.56–0.66 mole fractions (azeotropic composition), it increases (Fig. 4). The patterns of change in the relative volatility of binary mixtures in the presence of EAs are presented in study [40]. The authors of study [32] consider an equimolar composition as the initial one, for which the change in relative volatility is insignificant, which explains the result of the ED process calculation.

It is possible that the choice of a more selective solvent may change the cost ratio picture of the considered flowsheets.

This study discusses the efficiency of using varied-diameter columns (VDC) when the flow of liquid and vapor in different sections of the column significantly differs. This is illustrated by the example of five binary systems (considering different diameters of the rectifying and stripping sections in several cases allows for a reduction in capital costs and achieving an overall economic effect when comparing TAC in the range of 0–10%). Similar research was conducted in study [37].

In the study [36], the PSD separation flowsheet of the acetone-chloroform mixture turned out to be significantly more energy-intensive due to the low sensitivity of the azeotrope composition to pressure changes (about 0.02 mole fractions with a pressure difference in the columns exceeding 9 atm).

While the transformation of ED and PSD flowsheets to structures with PCHMF generally leads to reduced separation costs, this does not affect the ratio of operational to annual costs between separation variants. However, if the difference in separation

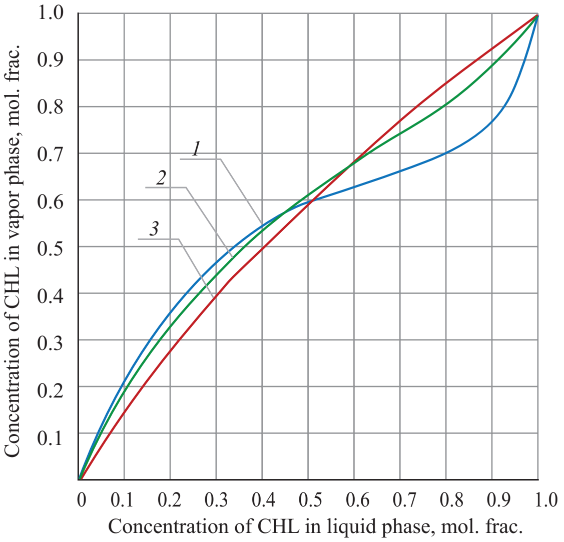


Fig. 4. Phase equilibrium curve of binary and pseudo-binary chloroform–methanol system at different amounts of *n*-propanol, mol %: (*1*) 0, (*2*) 30, (*3*) 80

costs in the initial flowsheets is significant, then using PCHMF reduces it; moreover, the effect of such structural optimization is generally more pronounced for a more energy-intensive flowsheet [35]. In the study [30], partial flow coupling for both flowsheets allowed for a reduction in operational (annual) costs by 4.8% (3.4%) and 26% (20%) for ED and PSD, respectively. Full heat integration for the flowsheet based on pressure variation led to a reduction in operational (annual) costs by 27% (18.7%). The lesser effect of full heat integration for the PSD flowsheet is associated with the need to use additional equipment and the increase in capital costs by 6.4% compared to the classical flowsheet.

In the study [29], structural optimization of flowsheets was carried out by using a dividing wall column (DWC) for the ED flowsheet of the acetonitrile–*n*-propanol mixture and partial flow coupling for the PSD complex. The application of a complex column compared to the classical ED complex allowed for a 7.94% reduction in TAC and a 12.8% reduction in PCHMF for the PSD complex. The optimization of the flowsheet structure did not change the cost ratio of the flowsheets, which is higher for the ED process. Dynamic process modeling (in nonstationary and transitional modes) showed greater resistance to disturbances (changes in flow rate and feed composition) of the PSD flowsheet with PCHMF compared to ED (DWC).

When purifying methanol from C5 fraction components (during the production of *tert*-amyl ether), the process of extractive heteroazeotropic distillation with water [33] serves as an alternative to the PSD process. Water is not only an extractive heteroazeotropic agent, which serves to lower the overall temperature of the process in the column by forming azeotropes with a boiling point minimum, but also possesses the properties of an extractant (elements of extraction are present), which ensures greater energy efficiency of this process.

The phase diagram of the methyl acetate–methanol–acetic acid–acetic anhydride mixture is characterized by a simple structure of the phase equilibrium diagram: there is a single azeotrope having a minimum boiling point, whose composition changes with varying pressure. Considering two main PSD and ED separation methods, the authors [38] synthesized 12 flowsheets of different structures. The latter differ in the use of ED, varying pressure, and the column of conventional distillation at different stages of separation. Such a comprehensive study not only allowed for the comparison of different separation methods, but also determined the effectiveness of their application in separating mixtures of varying component compositions: quaternary, ternary, binary. As a rule, the latter decreases with an increase in the number of components in the mixture being separated. When using ED (PSD), the total columns reboiler duty was: binary mixture, 24.5 MW (67.0 MW); ternary, 37.4 MW (70.3 MW); quaternary, 37.6 MW (92.5 MW) [38].

Special attention should be given to the results of the authors [39] on the separation of the isopropanol–water mixture. Since the composition of the azeotrope is practically independent of pressure, the authors add a new substance (tetrahydrofuran) into the system in order to use PSD. This forms new azeotropes (tetrahydrofuran–water at 0.1 and 0.6 MPa and tetrahydrofuran–isopropanol at 0.6 MPa), whose composition is sensitive to changes in external conditions. In place of pure ether, a mixture of ether with water in a composition close to azeotropic is added in the initial mixture. The separation of the mixture occurs in a two-column complex. A comparison with the classical ED flowsheet using DMSO demonstrated the inefficiency of this method even with partial flow coupling. The amount of agent in the PSD process (95 kmol/h per 100 kmol/h of the initial mixture) is twice that of the solvent in the ED process (45 kmol/h), while tetrahydrofuran is always present in the distillate flow, which negatively affects energy costs. Furthermore, water must also be present in the recycle flow (163 kmol/h).

Comparison of flowsheets based on the combination of various processes and separation methods

The article [41] compares two separation variants for the diisopropyl ether–isopropanol–water mixture: PSD and a combination of distillation and splitting, including the use of partial heat and material flow integration. The comparison of classical flowsheets without heat integration demonstrated the advantages of the flowsheet based on pressure variation over the second variant (reduction of energy, capital, and TAC by 5.3%, 6.3%, and 5.5%, respectively) in terms of a smaller recycle flow, lower reflux ratios, and a total number of theoretical stages. The application of partial heat integration for the PSD flowsheet further reduced energy consumption by 15% and TAC by 11%.

For the separation of the ternary mixture water–benzene–isopropanol, two flowsheets have been considered: the first is based on varying the pressure in the columns (PSD), while the second is a combination of (preliminary) phase separation, conventional distillation, and pressure variation [42]. For both flowsheets, the possibility and efficiency of using partial heat integration are considered. The amount of the recycle flow formed from the distillate flows of the distillation columns for the flowsheet with phase splitting is significantly higher (179.42 *vs* 95.45 kmol/h per 100 kmol/h of the feed mixture), which ultimately had a negative impact on the energy costs of the flowsheet (the latter being 20% higher when compared to flowsheets without heat integration). An analysis of the phase equilibrium diagram structure and the position of the balance lines shows the inefficiency of the method based on preliminary phase separation. An increase in the values of reflux ratios and the number of theoretical stages in distillation columns (101 compared to 74 theoretical plates in the PSD flowsheet) is achieved using large distillate and bottom flow ratios in columns. This is due to the need to reach the boundaries of separation areas to ensure the possibility of transitioning to the required distillation area. At the same time, capital costs are more than 40% higher. Already at the stage of studying the phase diagram, it was possible to abandon the method based on the combination of different processes. The use of preliminary phase separation is generally not an effective separation technique if all components are present in the equilibrium liquid layers in the separator [9]. This example also shows that the use of partial heat integration does not yield a significant effect in any cases: for PSD flowsheets, this effect is around 2.8%, while for the flowsheet based on preliminary phase separation the corresponding figure is about 6%.

In the study [43], a comparison of two methods for dehydrating acetonitrile obtained through the ammoxidation of ethanol is presented: varying pressure in distillation columns (1–7 bar and 1–10 bar) and a combination of extraction (with dichloromethane) and distillation. As the calculation results showed, the lowest capital costs at comparable operating costs are observed for the PSD flowsheet (1–10 bar), while the highest are for the extraction flowsheet; however, in the hybrid flowsheet, the required quality of acetonitrile (99.99 wt %) is not achieved. It should be noted that the extractant is not used in its pure form, but with a certain amount of acetonitrile due to its low selectivity, which reduces the efficiency of the extraction process and increases the recycle flow (the extractant in the ternary mixture is separated in the bottom of the distillation column).

Different separation flowsheet structures for the toluene–methanol–water mixture based on the use of the ED single separation method are associated with varying effects of the selected solvents on the relative volatility of the components: in the presence of *N*-methyl-2-pyrrolidone, the volatility of methanol increases relative to the other components, while in the presence of ethylene glycol, it is toluene that demonstrates the highest relative increase in volatility [44]. As a result of the separation of methanol after the agent regeneration column due to their low mutual solubility, the separation of the toluene–water mixture occurs in the liquid separator (the flowsheet includes two columns and a separator). If toluene is separated in the distillate of the column, an additional distillation column will be required for the subsequent separation of the water–methanol mixture, which will negatively impact both the energy costs of the flowsheet and the total number of theoretical stages N (the flowsheet contains three distillation columns). This is confirmed by a computational experiment in the study: for the flowsheets with *N*-MP and ethylene glycol, the $\Sigma N/\Sigma Q$ was 91/4.15 and 137/4.47, respectively. The TAC were also higher for the flowsheet with ethylene glycol.

The problem of purifying components that are present in trace amounts, especially when dealing with a multicomponent mixture, can be addressed by using ED (RED) or RED processes. In the study [45], two separation flowsheets based on the combination of ED or RED and single extraction are proposed for the separation of a six-component mixture of water–hydroxyacetone–isopropylbenzene–alpha-methylstyrene–phenol–2-methylbenzofuran. The structures of the flowsheets are almost identical except for the method of introducing the agent into the first column (diethylene glycol as a high-boiling or acetone as a low-boiling solvent) and its regeneration (diethylene glycol in the bottom, acetone in the column distillate). The flowsheet with acetone was found to be significantly inferior to the flowsheet with diethylene glycol with double the energy costs. However, the authors state that the competitiveness of the flowsheet may be due to the repeated use of the excess acetone generated during the cumene method of phenol production in the technological chain.

A large amount of experimental data on liquid–liquid equilibrium in ternary and quaternary systems is presented in both the Russian and foreign scientific literature (as a rule, a basic binary mixture is considered in the presence of one or two potential extractants for the extraction process). Conclusions about the feasibility of using a particular solvent—and consequently the implementation of the extraction process—are based solely on known separation criteria (distribution and separation coefficients). The articles do not evaluate the feasibility of implementing such a separation option considering the solvent regeneration unit. Considering that the energy-intensive distillation process is typically used for solvent regeneration, the efficiency of the combination of these two processes can be significantly reduced due to the presence of thermodynamic equilibrium constraints of the liquid–vapor (liquid–liquid–vapor) mixture representing the extract.

The separation of the acetonitrile–water mixture in a flowsheet based on a combination of extraction (chloroform as the extractant) and distillation turned out to be a less energy-efficient process compared to the ED flowsheet with DMSO (by more than 30%) and comparable to the PSD flowsheet (a difference of 1.7%) [46]. The high-energy costs are due to the need to separate (evaporate) chloroform in the distillate of the extractor regeneration column.

Another variant for separating this mixture [47] involves a pre-concentration column for the mixture (partial separation of water) followed by its separation in an ED complex with ethylene glycol (the total duty on the column reboilers was 4556 kW, TAC was $1.56 \cdot 10^6$ USD/year). Due to the restructuring of material flows, the flowsheet was transformed into a two-column complex by combining the concentration column and the EA regeneration column, which allowed for a 16% reduction in heat duty and a 13% reduction in TAC compared to the initial version. The extraction flowsheet for this mixture with *n*-propyl chloride (with some amount of acetonitrile) includes 4 units: an extractor, a liquid separator, and two distillation columns. The regeneration of the extractant is carried out by a combination of distillation and liquid phase splitting. This flowsheet turned out to be more advantageous both in terms of energy consumption (2972 kW) and annual costs ($1.02 \cdot 10^6$ USD/year).

Comparing the results of the studies [46, 47], it can be assumed that the ratio between the considered separation options may be changed by abandoning the mixture concentration column and selecting a more selective agent for the ED process.

When investigating the possibilities of the extraction process of ethanol mixtures with hydrocarbons (hexane and heptane) using ionic liquids, the authors of study [11] indicate a positive economic effect when organizing a combination of extraction and distillation compared to classical separation ED or HAD methods (more detailed information is not provided in the article).

Another example of the use of hybrid technologies consists in the combination of distillation and membrane separation (in particular, pervaporation). The principle of membrane separation differs from the previously discussed phase processes; it is based on the selective extraction of substances as they pass through the membrane. The processes occur in different apparatuses arranged sequentially (in any order) or using an external pervaporation unit [48]. The authors of study [48] discuss the efficiency of hybridization flowsheets through the combination or integration of membrane and distillation separation processes, which can provide a significant positive economic effect compared to previously considered methods. These conclusions are also supported by the results of studies [49–58] on the example of systems: methanol–methyl *tert*-butyl ether [49], isopropanol–water and propylene–propane [50], isopropanol–water (24% savings compared to azeotropic distillation) [52], ethanol–water [53], ethanol–water

(66% savings compared to heteroazeotropic distillation) [54], dimethylformamide–water [55], isobutanol–water [56], and methyl acetate–methanol (24% savings compared to ED and PSD, 31.7% compared to PSD with CSM) [57].

The diversity of phase behavior in liquid mixtures [1–3, 9, 14, 59], including in the presence of specially selected substances, determines the polyvariance of the choice of processes and methods, as well as the structures of separation technological flowsheets. The development of qualitative and quantitative criteria for the discrimination of processes, methods, and flowsheets at various stages of structural and parametric optimization remains an urgent objective. These include: analysis of physicochemical information about the properties of mixture components; the structure of the phase diagram (including the construction and analysis of the balance lines); assessment of the selectivity and feasibility of introducing EA; selection of separation processes and methods; determination of material flow rates when solving the balance problem (correct assignment of free variables); synthesis of separation flowsheets based on a specific process or the application of hybrid technologies, analysis of the operational equipment (meeting specific criteria: energy, economic, environmental) and separation results parameters; evaluation of the effectiveness of process or flowsheet management.

To minimize heat losses and reduce energy costs, many efforts are directed towards various types of heat integration, such as partial [21, 30, 35, 37, 39–42, 60–63], full [30, 64], internal [65], or external [66, 67]. An effective energy-saving measure consists in the use of complex columns (with side flows, DWC) [18, 29, 63] and heat pumps [68, 69]. The cost ratio (operational and annual) between separation options is not affected by restructuring the flows of the flowsheet or using complex columns; if the difference in separation costs in the initial flowsheets is significant, it can be reduced by one of the structural optimization options (the more optimal the structure and operating conditions of the initial flowsheet, the less effective the methods related to structural restructuring).

It is necessary to separately highlight the study of dynamic modes (control and stability to perturbations) of various processes [28, 70–75], which also influence the final choice of separation technology for a specific mixture.

CONCLUSIONS

The presented analysis of the scientific literature confirms that when developing separation flowsheets for specific mixtures, authors generally prefer a specific method, with ED being more frequently considered as the universal and energy-efficient separation method. The number of studies comparing alternative methods is significantly smaller; moreover, these are typically techniques based on a specific process (usually distillation). Flowsheets based on various mass transfer processes (hybrid technologies) are rarely considered. The criterion for comparing alternative separation variants is the total heat duty on the column reboilers and the TAC. As a rule, there is a complete correlation between these criteria. In rare cases, the contribution of capital costs to annual totals can be decisive, leading to different ratios of energy costs and TAC. Such a situation is observed if the flowsheets differ significantly in the total number of stages.

As confirmed by one of the processes (flowsheets) being placed in an a priori disadvantageous position (nonselective agents are chosen; the geometry of the phase diagram elements does not favor the use of the chosen method; the azeotrope shifts slightly when the pressure changes; etc.), the comparison of flowsheets is not always objective. Such variants can be excluded from consideration at the stage of studying the structure of the phase equilibrium diagram, synthesizing separation flowsheets, and calculating the material balance (up to the stage of parametric optimization).

In certain cases, the authors may reasonably prefer flowsheets characterized by higher energy consumption, which may be due to the physicochemical properties of the components of the base mixture or the newly added substances, as well as the specifics of the separation technology implementation.

The presented comparison of effective separation methods (cases where nonselective agents are chosen or the phase diagram structure is unfavorable for the chosen method are excluded) confirms that ED remains the most energy-efficient separation method. However, in some cases, its competitiveness significantly decreases compared to alternative methods, particularly HAD. The latter should be preferred for the separation of mixtures containing high-boiling components. The ratio of energy costs for the HAD and ED flowsheets will be significantly influenced by the composition of the initial mixture, which directly affects the solvent consumption, especially in the HAD process (agent is present in the azeotropic mixture in the column distillate). The latter process will be more advantageous if the content of the component in the initial mixture with which the solvent forms a heteroazeotrope exceeds the concentration of the second component. It is likely that the composition simplex of a binary system can be conditionally divided into composition regions for which the cost ratio of the processes of HAD and ED (with the same agents) will be different. The significant reduction in energy consumption of the flowsheets by organizing ED and HAD in auto mode—and even more so by combining them—is due to the absence of the need to use an additional column for agent regeneration, as well as the lowering of the separation process temperature inside the column due to the formation of positive azeotropes.

Another effective separation technique is the combination of distillation and splitting process. The feasibility and practicality of implementing this method are significantly influenced by the specific arrangement of azeotropes, separatrix manifolds, and splitting simplices. The unfavorable arrangement of the latter when solving the balance problem leads to inflated values of columns distillate flows and difficulties in achieving certain compositions (for example, when touching the separatrix). If this technique can be effectively implemented independently (without the involvement of other specialized methods) then comparisons with other methods are not conducted since any other method is likely to be less effective.

A reduction in energy costs can be achieved by using a liquid separator (splitting) in complex flowsheets that combine different methods and processes. This is explained by the fact that phase separation is a spontaneous process; moreover, in some cases, one or both flows leaving the separator can represent product flows due to the almost complete insolubility of the components in each other.

When separating multicomponent mixtures, a question arises about the feasibility of using a particular method or process at different stages of separation (when separating mixtures with different numbers of components). The analyzed sources do not fully answer this question: a detailed study has been conducted only for one system and two methods (ED and PSD). The efficiency of these processes in separating a mixture of specific composition decreases with the increase in the number of components in the mixture being separated.

The comparison of extraction-based flowsheets with other separation variants is scarcely mentioned in other studies. When assessing the feasibility of combining extraction and distillation processes, it is important to consider the possibility of regenerating the extractant in the bottom of the column, especially in cases where the ratio of the amounts of the initial mixture to the extractant is ≤ 1.

Based on the conducted review of scientific literature, the following promising directions for further research in the synthesis of separation flowsheets for organic product mixtures through the use of various separation methods and processes can be formulated:

(1) evaluation of the effectiveness of applying different processes (extraction, liquid phase splitting, distillation, special distillation methods) at different stages of separation;
(2) comparative analysis of ED and HAD methods in the separation of mixtures with different initial compositions (identification of areas of energy advantage for each method);
(3) evaluation of the effectiveness of implementing flowsheets based on the combination of extraction with other processes, depending on the stage of extractant regeneration.

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Authors' contributions

A.V. Frolkova—management, formulation of the scientific concept of the study, critical literature review, and writing the text of the article.

A.N. Novruzova—collecting and processing the scientific literature, mathematical modeling the phase equilibria, and preparing illustrations and tables for the article.

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