

Article

Comparison of continuous homogenous azeotropic and pressure-swing distillation for a minimum azeotropic system ethyl acetate/*n*-hexane separation☆

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ABSTRACT

Continuous homogenous azeotropic distillation (CHAD) and pressure-swing distillation (PSD) are explored to separate a minimum-boiling azeotropic system of ethyl acetate and *n*-hexane. The CHAD process with acetone as the entrainer and the PSD process with the pressures of 0.1 MPa and 0.6 MPa in two columns are designed and simulated by Aspen Plus. The operating conditions of the two processes are optimized via a sequential modular approach to obtain the minimum total annual cost (TAC). The computational results show that the partially heat integrated pressure-swing distillation (HIPSD) has reduced in the energy cost and TAC by 40.79% and 35.94%, respectively, than the conventional PSD, and has more greatly reduced the energy cost and TAC by 62.61% and 49.26% respectively compared with the CHAD process. The comparison of CHAD process and partially HIPSD process illustrates that the partially HIPSD has more advantages in averting the product pollution, energy saving, and economy.

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1. Introduction

Ethyl acetate and *n*-hexane, as industrially important organic solvents, are widely used in the textile and chemical industry due to their high solubility [1,2]. A large amount of wastewater will be produced in the production of chemical products [2]. The wastewater will increase the total annual cost and lead to environmental pollution if being discharged directly. Therefore, it is necessary and attractive to separate the ethyl acetate/*n*-hexane mixture and realize recycling resources. Since ethyl acetate (boiling point 77 °C) and *n*-hexane (boiling point 69 °C) form a minimum-boiling homogenous azeotrope (azeotropic point 65 °C) at atmospheric with azeotropic composition of 61 wt% of *n*-hexane [1–4]. However, ordinary distillation methods failed to effectively separate the azeotropic system [3,4]. As a result, some special types of distillation methods should be considered to separate the azeotropic system.

Many special distillation methods have been used for separating azeotrope in industry, such as azeotropic distillation [4–7], pressure-swing distillation [8–14], extractive distillation [7,8,15–19], and other

new separation techniques. Azeotropic distillation is widely used to separate azeotropes, which based on the change in vapor–liquid equilibrium by using a suitable entrainer [5–7,20–23]. Azeotropic distillation can be operated either in batch mode or in continuous mode. Compared with batch azeotropic distillation, the continuous azeotropic distillation process has advantages in stable of process and the well products quality. Besides, the entrainer can be recycled resulting reduction in the consumption of entrainer [1,5,7]. Zeng *et al.* [4] studied the separation of ethyl acetate and *n*-hexane by batch azeotropic distillation, and then developed a new calculation model for this process. Finally the experiment was used to evaluate the feasibility of separation for this azeotropic system. The results indicated that yields of ethyl acetate and *n*-hexane were 73.89% and 75.15% when the ratio of acetone and *n*-hexane was 1.15. Pressure swing distillation (PSD) is another popularly method in the separation of azeotropes when the azeotropic composition is very sensitive to the pressure. This method can avert the product pollution by the entrainer which must be added to the azeotropic system. Many scholars pointed out that PSD technology had some superiorities compared with other methods in some azeotrope separation [11,12,24,25].

Although many reports studied different distillation methods for the separation of azeotropes, there were few efforts that focus on the comparison of these separation methods for separating the ethyl acetate and *n*-hexane azeotropic system. Therefore, the feasibility of separating the azeotropic system using PSD and continuous homogenous azeotropic

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Table 1

Azeotropic point and composition of the ternary system of ethyl acetate/*n*-hexane/entrainer at 0.1 MPa

Entrainer	Boiling point/°C	<i>n</i> -Hexane/wt%	Ethyl acetate/wt%	Entrainer/wt%	Azeotropic point/°C
Methanol	64.9	72.0	28.0	50.6	62.3
Ethanol	78.0	79.0	21.0	58.6	71.8
Isopropanol	82.3	77.0	23.0	62.7	75.3
Acetone	56.2	46.5	53.5	49.7	68.3
Isobutanol	108.4	97.5	2.5	68.3	
Carbon tetrachloride	76.8	43.0	57.0	74.8	

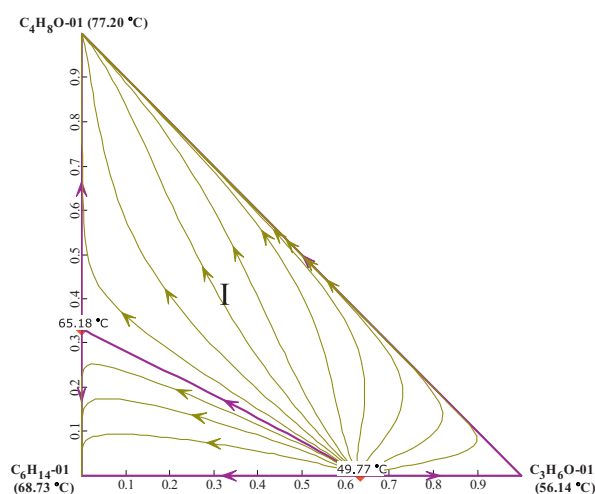
distillation (CHAD) and the comparison of the two methods from the viewpoint of economics are proposed in this work. First, the entrainer is selected by analyzing the residue curve maps in the CHAD process and the pressure sensitivity of this system is evaluated by the binary phase diagram in the PSD process. Then the operating conditions of

two distillation methods are optimized via a sequential modular approach to achieve the minimum total annual cost (TAC). Furthermore, the partially heat integrated pressure-swing distillation (HIPSD) is used to study the potential of large energy savings by analyzing the temperature difference of the condenser in high pressure column (HPC) and the reboiler in low pressure column (LPC). Finally, the optimized results of steady state are used to compare the PSD with CHAD processes and find process which is better suitable to separate the ethyl acetate and *n*-hexane azeotropic system.

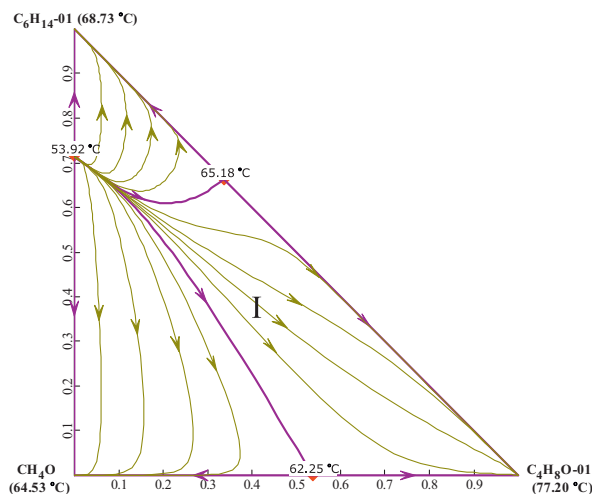
2. Continuous Homogenous Azeotropic Distillation

2.1. Design basis

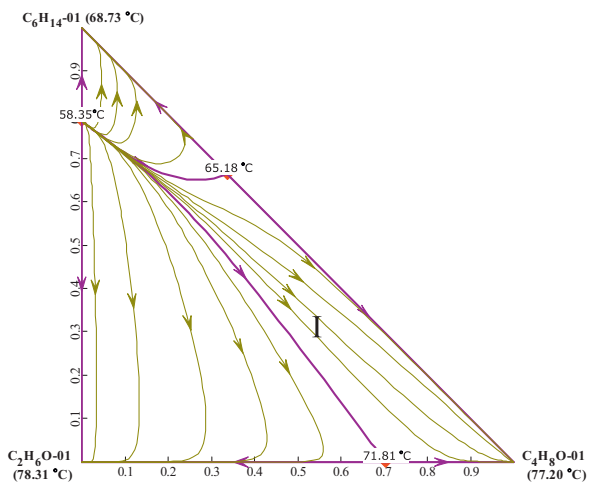
In the CHAD process, the Aspen Plus V7.2 software is used to simulate and optimize the process for separating the ethyl acetate and *n*-hexane azeotropic system. The possible entrainers are chosen by consulting the solvent handbook [4]. Residue curve maps for ethyl acetate/*n*-hexane/entrainer are obtained by implementing the *Property Analysis* function in the Aspen Plus. The feed rate of the wastewater is set to



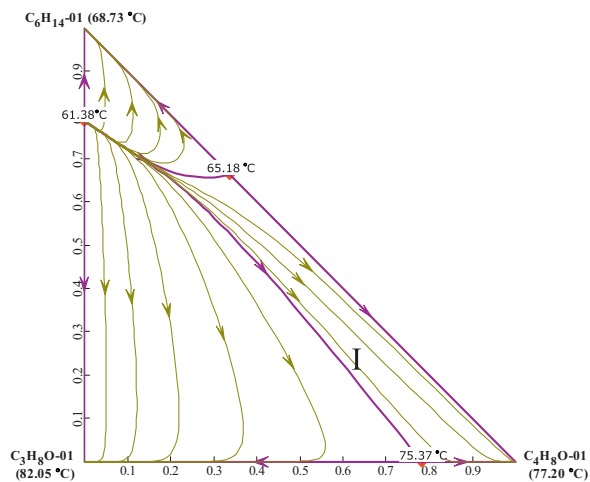
(a) ethyl acetate/*n*-hexane/acetone



(b) ethyl acetate/*n*-hexane/methanol



(c) ethyl acetate/*n*-hexane/ethanol



(d) ethyl acetate/*n*-hexane/isopropanol

Fig. 1. The residue curve maps of ternary system ethyl acetate/*n*-hexane/entrainer at 0.1 MPa. (a) ethyl acetate/*n*-hexane/acetone. (b) ethyl acetate/*n*-hexane/methanol. (c) ethyl acetate/*n*-hexane/ethanol. (d) ethyl acetate/*n*-hexane/isopropanol.

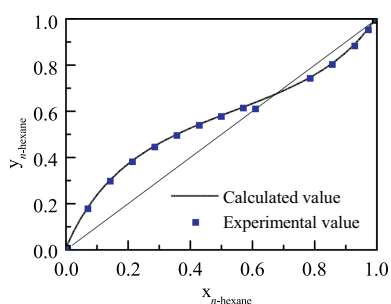


Fig. 2. The x - y diagrams of experiment and calculation for ethyl acetate/ n -hexane azeotropic system at 0.1 MPa.

Table 2
Binary interaction parameters of NRTL model

i	j	A_{ij}	A_{ji}	B_{ij}	B_{ji}	C_{ij}
Ethyl acetate	n -hexane	0	0	120.159	228.099	0.3

$1000 \text{ kg} \cdot \text{h}^{-1}$. According to the wastewater sampling analysis of some pharmaceutical factories, it consists of around 39.00 wt% ethyl acetate and 61.00 wt% n -hexane [1–4]. The specification of ethyl acetate and n -hexane product purities are set as 99.10 wt%.

2.2. Entrainer selection

In the CHAD process, it depends on the selection of entrainer that the separation process will be successfully completed. A criterion for entrainer selection is that the entrainer should generate a low-boiling azeotrope with one or two components in the mixture [4]. The larger the temperature difference of azeotropic points between the original and the new-forming azeotrope, the easier the separation. Generally, the temperature difference is not less than 10°C . Also, the easy recovery of the entrainer is another key factor in the selection entrainer. In general, the heterogeneous entrainer is the superior entrainer, because the minimum-boiling azeotrope produces two immiscible liquid phases can be easily separated for recycling *via* a decanter. For some azeotropic system, the homogenous entrainer will be used when the suitable heterogeneous entrainer cannot be chosen. In industry, the extraction

method is used to recovery the homogenous entrainer and the water is usually selected as the extractant [1,4]. Besides, the entrainer should have the advantages of cheapness, innocuity, lower dosage, and less heat consumption.

The following six possible entrainers are studied for separating the azeotropic system: methanol, ethanol, acetone, isopropanol, isobutanol, and carbon tetrachloride [1,4,26,27]. Table 1 demonstrates azeotropic point and composition of the ternary system of ethyl acetate/ n -hexane/entrainer at 0.1 MPa. From the Table 1, when acetone is selected as the entrainer, the temperature difference of the azeotropic points is 15.3°C between the new azeotrope (49.7°C) and original one (65°C). And acetone can be easily recycled by water [4]. Fig. 1 represents that the residue curve maps and distillation boundaries of ethyl acetate/ n -hexane/entrainer. From this figure, it can be seen that the acetone as entrainer has a larger distillation region (*i.e.*, I) compared with other entrainer. It indicates that the operation flexibility will be significantly increased and reduced in the energy cost based on the material balance. From Fig. 1(a), the high purity product of ethyl acetate can be obtained in the bottom stream of azeotropic distillation column while mixtures of n -hexane and acetone are obtained from the top of the column in the I. Following that, acetone is selected as suitable entrainer in the simulation of CHAD process.

2.3. Selection of physical property model

The precision and reliability of simulation results depend on choosing an appropriate physical property model. Fig. 2 shows the x - y diagram of experiment and calculation for ethyl acetate/ n -hexane azeotropic system at 0.1 MPa [11–14,25]. It is observed that the predicted vapor–liquid equilibrium data at 0.1 MPa with the NRTL model has good agreement with experimental data. Therefore, the NRTL physical property can be used to predict the vapor liquid equilibrium data at higher pressure. The binary interaction parameters are shown in Table 2.

2.4. Primary simulation

Fig. 3 illustrates the process flowsheet of the CHAD process with two columns and one decanter using the acetone as entrainer. It can be clearly seen that the mixture of ethyl acetate and n -hexane is fed at the middle of the azeotrope distillation column (denoted as ADC). The total entrainer feed with a recycled entrainer feed called as D2 from

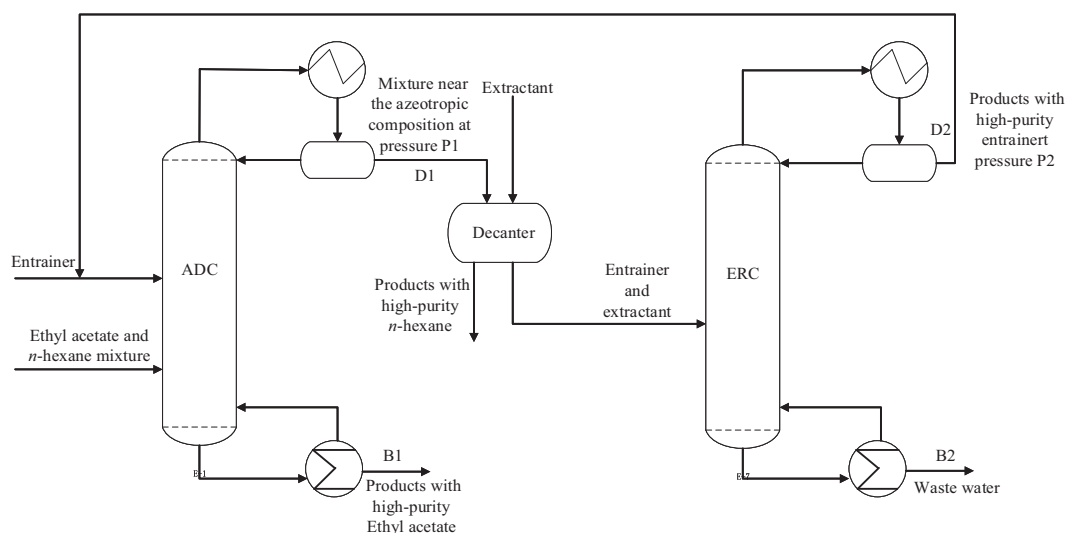


Fig. 3. The CHAD process flowsheet for ethyl acetate/ n -hexane separation.

the overhead of the entrainer recover column (ERC) is mixed with the make-up entrainer stream as an input for the ADC. Product with high-purity ethyl acetate can be obtained in the bottom stream of ADC while the azeotropic mixture of acetone and *n*-hexane is removed from the top of the ADC. Then, the stream D1 feeds into the decanter and removes the acetone from the mixture of *n*-hexane/acetone by water as heterogeneous solvent. Two liquid phases (*i.e.*, aqueous and organic phases) are separated *via* a decanter because of the immiscibility of *n*-hexane with water. The organic phase is the high-purity *n*-hexane product and the aqueous phase involves the water and acetone. The mixture of water and acetone is fed at the middle of the ERC. At the same time, the waste water can be recycled to the decanter.

The operating pressures of the ADC and ERC are set as 0.1115 and 0.1013 MPa, respectively. The total number of stage for the ADC is 38, while the ERC is 36. The fresh feed and the entrainer feed stage locations are 26th and 11th in the ADC, respectively. The aqueous phase stream from the decanter is fed at 10th of ERC. The flow rates of D1 and D2 are 1344.37 kg·h^{−1} and 414.58 kg·h^{−1}, respectively. The product purity is obtained by using the *Design Spec/Vary* function of Aspen Plus and adjusting the corresponding operated variables. The computational results show that the condenser duty is −340.90 kW and the reboiler duty is 372.34 kW in the ADC, while the ERC are −354.22 kW and 2133.47 kW. The reflux ratios which were obtained by implementing *Design Spec/Vary* function in Aspen plus for ADC and ERC are 1.14 and 4.98, respectively.

2.5. Process optimization

TAC is introduced to evaluate the economics of the separation processes. Therefore, the CHAD process was strictly optimized to achieve minimum TAC by sequential modular approach. The relevant appropriate method is provided by Luyben [24]. Conventionally, TAC is consisted of energy and equipment costs of heat exchangers and column vessel. The parameters of the column and exchangers are calculated by Aspen Plus. The number of stage is determined using Aspen Plus, with the last tray being the reboiler and the first tray being the reflux drum. The operating time of the distillation system is set at 8000 h per year and the payback period is assumed as 3 years [24]. The total heat transfer coefficients of the condenser and reboiler are 0.568 kW·K^{−1}·m^{−2} and 0.852 kW·K^{−1}·m^{−2}, respectively [24]. Additional costs including pipes, valves, reflux drums, and pumps can be neglected due to it is much tinier compared with the heat exchangers, column vessels and

Table 3	
Relevant calculation formula of equipment and energy consumption and utilities prices	
Parameters	Data
Condensers	$K_c = 0.852 \text{ kW} \cdot \text{K}^{-1} \cdot \text{m}^{-2}$ Capital cost = $7296 \times Ac^{0.65}$ Heat transfer area: $Ac = \frac{Q}{(K \times \Delta T)}$
Reboilers	$K_R = 0.568 \text{ kW} \cdot \text{K}^{-1} \cdot \text{m}^{-2}$ Equipment cost = $7296 \times Ac^{0.65}$ Heat transfer area: $Ac = \frac{Q}{(K \times \Delta T)}$
Column vessel cost	Length: $H = 1.2 \times 0.61 \times (N_T - 2)$ Capital cost = $17640 \times D^{1.066} \times L^{0.802}$
Energy cost	LP stream: 7.78 USD·GJ ^{−1} MP stream: 8.22 USD·GJ ^{−1} Cooling water: 0.354 USD·GJ ^{−1}
TAC	$\frac{\text{total equipment cost}}{\text{payback period}} + \text{annual energy cost}$
Payback period	3 years

energy costs [11,12,26]. Table 3 lists the relevant calculation formula of equipment and energy consumption and utilities prices [24,26,27]. In the CHAD process, the key design variables including the reflux ratio of two columns (*RR*₁, *RR*₂) and the ratio of distillate flow (*D*₁) to feed flow are adjusted to reach the separation requirement. This process

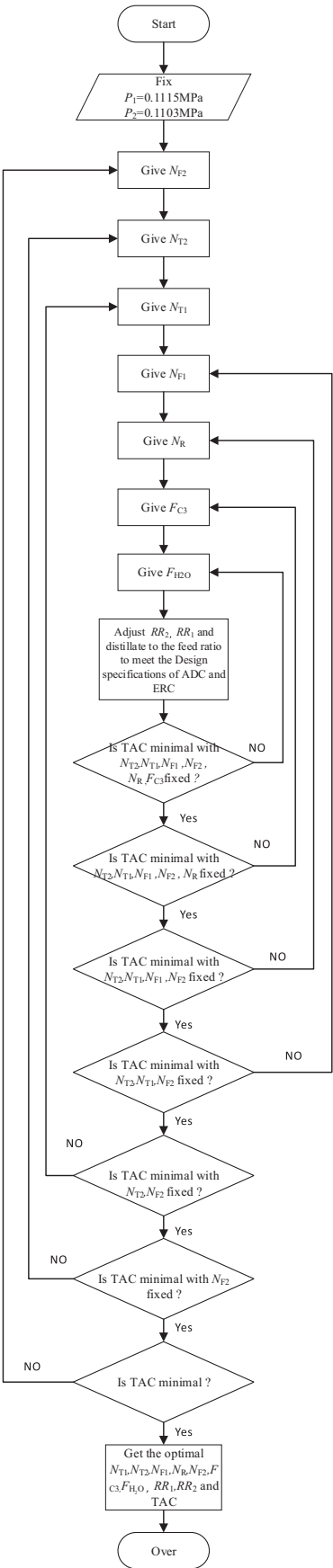


Fig. 4. The sequential iterative optimization procedure for CHAD process.

still has seven degrees of freedom which includes the total number stages of ADC (N_{T1}) and ERC (N_{T2}), the fresh feed stage location (N_{F1}), the feed stage location of entrainer (N_R), the aqueous phase feed stage location (N_{F2}), the flow rate of entrainer (F_{C3}) and the flow rate of water (F_{H_2O}), when the feed conditions and operating pressures have been fixed.

To obtain optimal operating parameters of this process with the minimum TAC, an optimization model of sequential iterative procedure is shown in Fig. 4. In the optimization process, the flow rate of water

(F_{H_2O}), the flow rate of entrainer (F_{C3}), the feed stage location of entrainer (N_R), and the fresh feed stage location (N_{F1}) are selected as the inner iterative loop, the total number of ADC (N_{T1}) and ERC (N_{T2}), the aqueous phase feed stage location (N_{F2}) are selected as the outer iterative loop.

The effects of operating parameters on TAC in the CHAD process are shown in Fig. 5. The relationship between extraction (H_2O) flow rate and TAC is displayed in Fig. 5(a). From this figure, the TAC decreases firstly and then increases when the F_{H_2O} increases. The optimal F_{H_2O} is

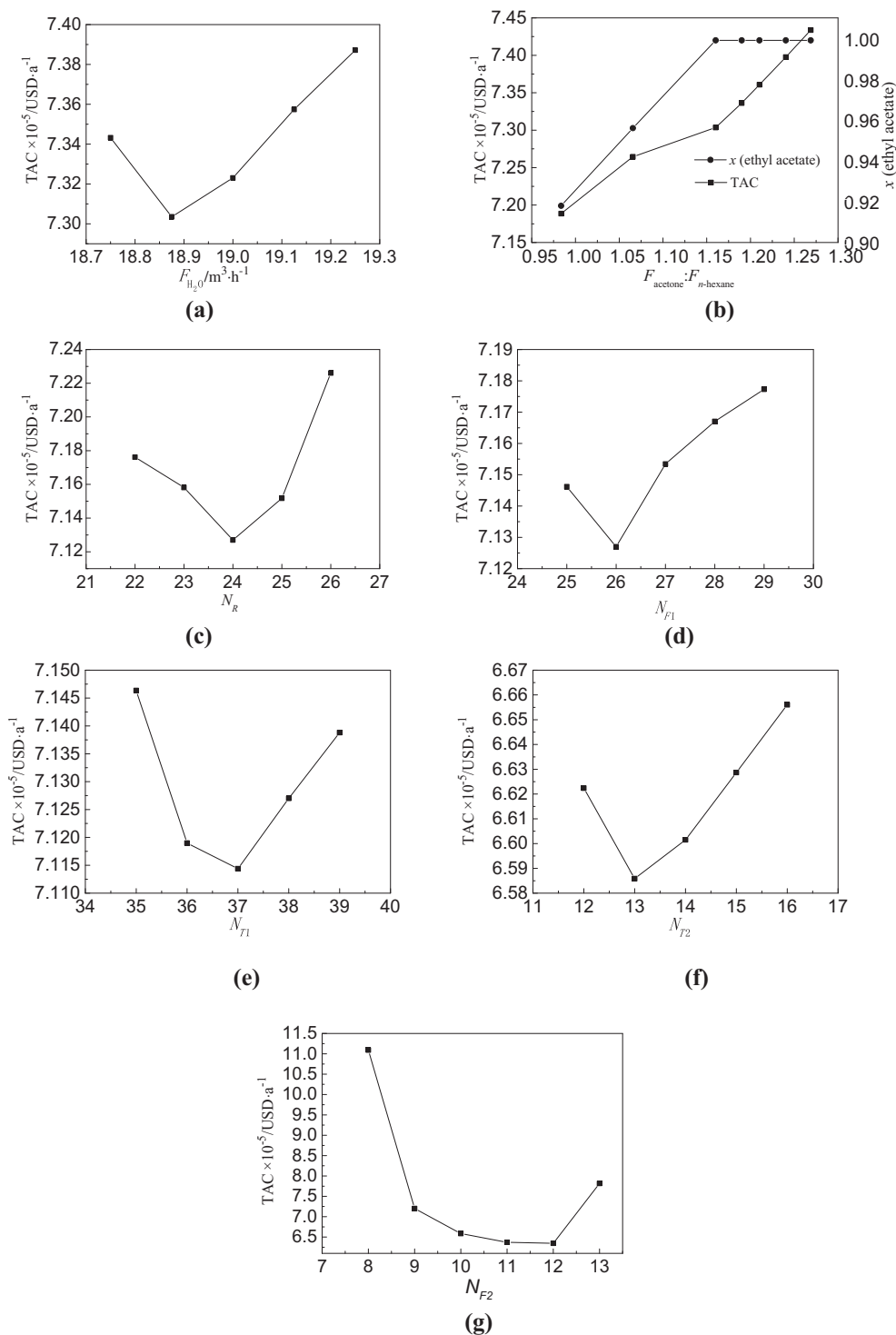


Fig. 5. The effects of operating parameters on TAC in the CHAD process.

$18.875 \text{ m}^3 \cdot \text{h}^{-1}$. The mass fraction of ethyl acetate and TAC at different entrainer ratios is presented in Fig. 5(b). The TAC continues to increase as the increase of entrainer ratio, and the mass fraction of ethyl acetate increases firstly and then levels off. The most suitable entrainer ratio is 1.16. Fig. 5(c), (d) describes the influence of entrainer and fresh feed stage location in the ADC on the TAC. The N_R and N_{F1} are key parameters during the design process. The TAC decreases firstly and then increases with the increase of the N_R or N_{F1} . The optimized N_R and N_{F1} are 24th and 26th, respectively. Fig. 5(e), (f) shows the effects of N_{T1} and N_{T2} on TAC. The total number of stage is also an important factor for the TAC. The total number of stage can effectively reduce operation cost, but the equipment investment will increase correspondingly. The results indicate that the TAC is minimal when N_{T1} and N_{T2} are 37 and 13, respectively. Fig. 5(g) exhibits the effect of water-acetone solution feed stage location on TAC. As the feed location moves down, the TAC is drastically decreasing, then almost unchanged and finally increasing. As you can see from the figure, the TAC of the feed stage location between 11th stage and 12th stage is largely the same. Compared with the 12th stage, the 11th stage selected as the feed stage location will be more suitable from the point of the safety and stable operation. The calculated energy cost is $5.13475 \times 10^5 \text{ USD} \cdot \text{a}^{-1}$ and TAC for this process is $6.37024 \times 10^5 \text{ USD} \cdot \text{a}^{-1}$, respectively. Fig. 6 shows the optimal flowsheet with detailed information which includes stream data, equipment size, reflux ratio, operating pressure (P), inner diameter (ID) and heat duties, and so on. The optimal process of CHAD for separation of ethyl acetate/*n*-hexane is shown in Fig. 6.

3. Pressure Swing Distillation

3.1. Design basis

In the PSD process, the *RadFrac* models built in the Aspen Plus are used in the simulation and optimization of the process for separating ethyl acetate/*n*-hexane azeotropic. The vapor–liquid equilibrium of the ethyl acetate and *n*-hexane system under different pressures are obtained by using NRTL thermodynamic model in the *Analysis/Property* function. The flow rate, component, and purities of the wastewater of the ethyl acetate and *n*-hexane are set as the same in the CHAD process.

3.2. Determination of pressure

The feasibility of PSD process depends on the pressure sensitivity of the azeotropic system. A principle is that the variation of azeotropic composition should be no less than 5% (preferably more than 10%)

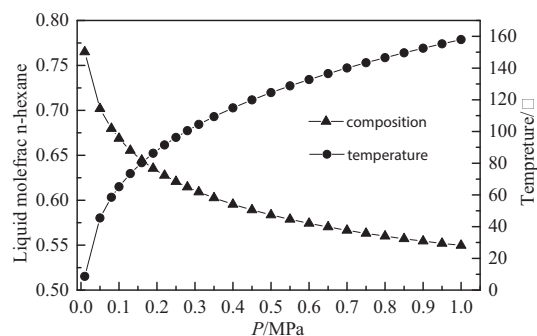


Fig. 7. Effect of pressure on azeotropic composition and temperature.

over a moderate pressure range of less than 1.0 MPa between the two pressures [14,28–31]. Fig. 7 shows the effect of pressure on the azeotropic composition and temperature of the ethyl acetate/*n*-hexane binary system. From Fig. 7, it can be clearly seen that the mole fraction of *n*-hexane significantly decreases in the scope of 0–1.0 MPa. To maintain the temperature differences between the stream in the heat exchangers and utilities are not less than 20 °C, the pressure of LPC and HPC are set as 0.1 and 0.6 MPa, respectively. In these circumstances, the mole fraction of *n*-hexane changes about 10%, the temperature of the stream in the condenser of LPC and HPC are 64.78 °C and 132.76 °C, while the temperature in the reboiler stream of LPC and HPC are respective 82.83 °C and 141.09 °C. It means that PSD process for the separation of ethyl acetate/*n*-hexane azeotropic system is feasible.

3.3. Selection of distillation sequence

The conventional PSD process includes two columns: low pressure column (LPC) and high pressure column (HPC). There are two separating sequences including high pressure–low pressure (HP–LP) and low pressure–high pressure (LP–HP). Fig. 8 shows the PSD flowsheet of different separating sequences. The fresh stream (F) and the recycled stream (R) from the second column are mixed, and then feed to the first column. One of the high purity products is obtained from the bottom and the mixture which near the azeotrope from the top of the first column is fed into the second column. Then, another high purity product is driven from the bottom of the second column and the stream R which close to the azeotrope is recycled back into the first column.

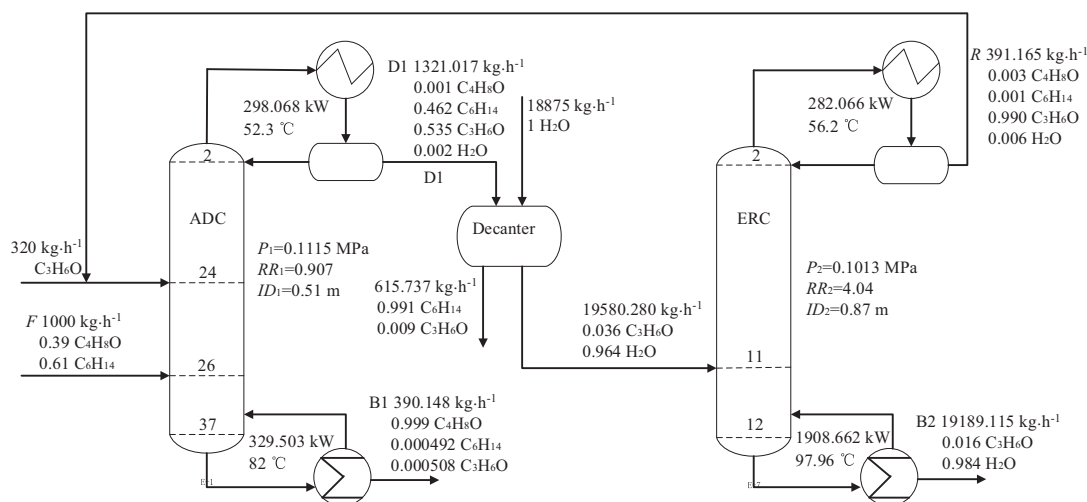


Fig. 6. The optimal flowsheet for CHAD process.

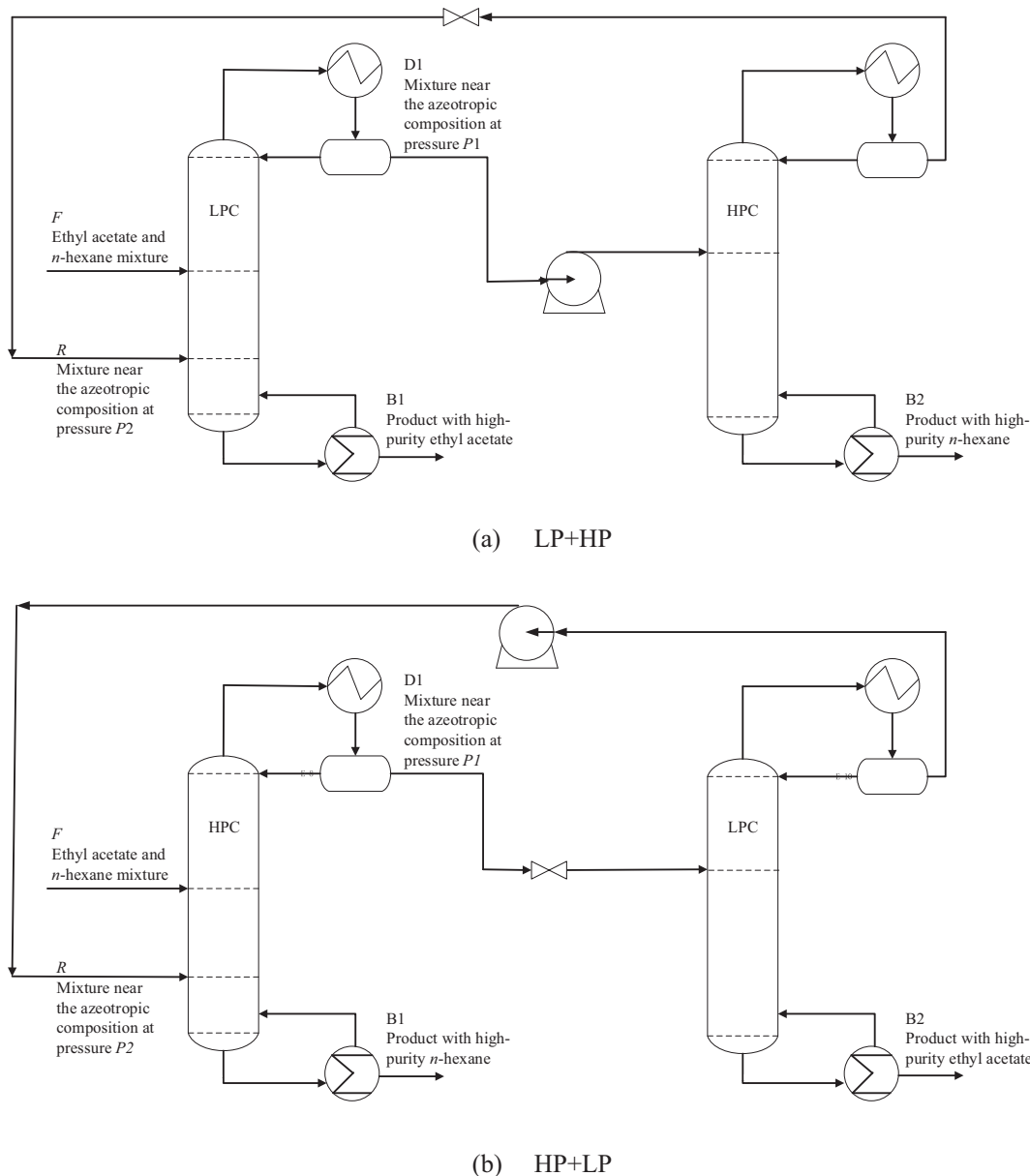


Fig. 8. The PSD flowsheet of different separating sequences. (a) LP + HP. (b) HP + LP.

The distillation sequence of the PSD process has important influence on the economy of the separation process. The suitable separating sequence is selected by comparison the LP-HP and HP-LP sequences

Table 4
The result of the TAC for PSD process with different distillation sequences

Variables	Pressure-swung distillation	
	LP-HP	HP-LP
P_1/P_2	0.1 MPa/0.6 MPa	0.6 MPa/0.1 MPa
N_{T1}/N_{T2}	20/26	26/20
$N_{F1}/N_R/N_{F2}$	11/13/10	8/12/12
$RR1/RR2$	1.95/1.08	1.95/1.184
$ID_1/ID_2/m$	0.82/0.88	1.086/1.184
Q_C/kW	1019.881/733.230	1266.233/1484.28
Q_R/kW	896.125/920.058	1600.684/1212.192
$TAC \times 10^{-5} / USD \cdot a^{-1}$	6.78	10.04

based on the minimum TAC. The total number of stage for the LPC and HPC are respective set as 20 and 26; the stage pressure drop is set as 0.72 kPa. The operating pressure of the LPC is 0.1 MPa, while the specification pressure of HPC is selected as 0.6 MPa. The heat duties of the reboiler and condenser in LPC and HPC are obtained after running the simulation. Table 4 lists the TAC of PSD process with different distillation sequences. As you can see from Table 4, the LP-HP sequence is more economical compared with HP-LP sequence.

3.4. Process optimization

3.4.1. PSD without heat integration

The PSD process is optimized by sequential modular approach, the distillate-to-feed ratio of LPC and the reflux ratio of the HPC are adjusting to maintain the purities of ethyl acetate and *n*-hexane which are both set at 99.9 wt%. Based on the sequential iterative optimization algorithm, the reflux ratio of LPC ($RR1$), the feed stage location of HPC (N_{F2}), the fresh material feed stage location (N_{F1}) and the feed stage

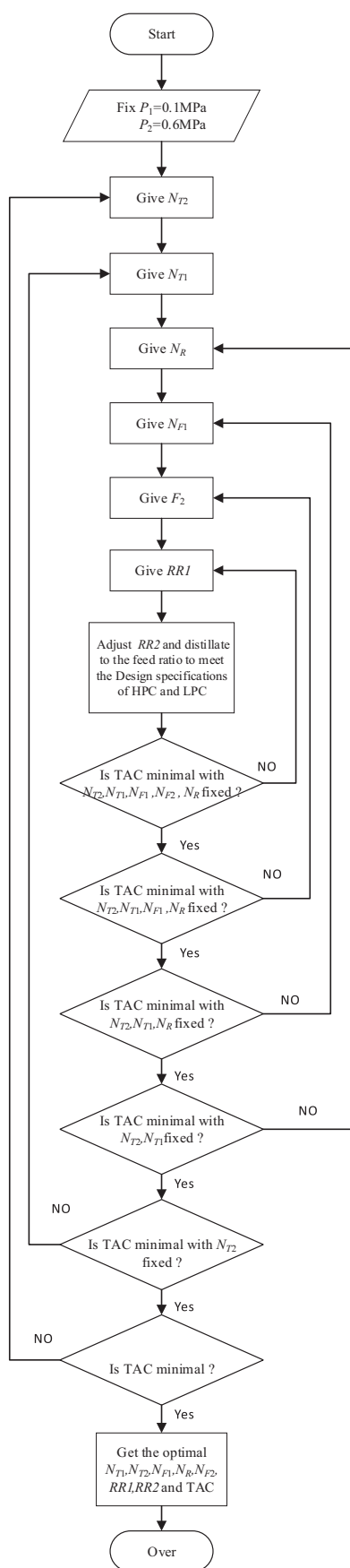


Fig. 9. The iterative optimization diagram of PSD without/with partial heat integration.

location of recycle-stream (N_R) are select as the inner iterative loop, the total tray numbers (N_{T1}) and the number of total trays (N_{T2}) are selected as outer iterative loop. The optimized operating parameters and the minimum TAC can be obtained by multiple iterative vary of the parameters from the inner to the outer loop. Fig. 9 shows the iterative optimization diagram of PSD without/with partial heat integration.

The influences of $RR1$, N_{F2} , N_{F1} , N_R , N_{T1} and N_{T2} on the TAC are displayed in Fig. 10. Fig. 10(a) depicts the effect of $RR1$ in LPC on TAC. The results reveal that the optimal reflux ratio $RR1$ is 1.1. Fig. 10(b), (c), (d) display the influences of N_{F2} , N_{F1} and N_R on the TAC. In the optimization process, the TAC increases in the order of N_{F2} , N_{F1} , N_R . The results show that the TAC is minimal with N_{F2} , N_{F1} , N_R are 7th, 8th, 16th respectively. Figs. 10(e) and 9(f) present the influence of N_{T1} and N_{T2} on TAC. For the LPC, the TAC decreases first and then increases, when the total number of stage changes from 24 to 28. The optimal N_T is 26. The impact of N_{T2} on TAC trend has a similar with N_{T1} . The result shows that optimal N_{T2} is 25. The energy cost of the PSD is $3.2425 \times 10^5 \text{ USD} \cdot \text{a}^{-1}$ and the TAC is $5.04518 \times 10^5 \text{ USD} \cdot \text{a}^{-1}$.

Fig. 11 shows the detailed information of material flow and equipment for the optimized PSD process. It can be obviously observed from Fig. 11 that the condenser duty is 726.479 kW and the reboiler duty is 603.11 kW in the LPC. Meanwhile, for the HPC, the condenser duty is 539.362 kW and the reboiler duty is 726.133 kW. The temperature difference between the condenser of the HPC (132.15 °C) and the reboiler of the LPC (81.83 °C) is large. It indicates that the operating costs and the capital investment can be effectively reduced through partly-heat integration between HPC and LPC.

3.4.2. PSD with partial heat integration

In the optimal operating conditions, the temperature difference is about 50 °C between the streams in the reboiler of LPC and the condenser of HPC. Consequently, the PSD process with partially heat integration is proposed to increase the utilization efficiency of energy.

Fig. 12 shows the flowsheet of partially HIPSD process. The heat duties are not balanced for this process because the condenser duty is 539.362 kW in the HPC and the reboiler duty in the LPC is 603.11 kW. Hence, an auxiliary reboiler would be required for the LPC with the duty of 63.748 kW to provide the remaining heat. The TAC of the HIPSD process is $3.232 \times 10^5 \text{ USD}$ per year, the energy cost is $1.9199 \times 10^5 \text{ USD}$ per year, and the total equipment cost of the partially HIPSD process is $1.312 \times 10^5 \text{ USD}$ per year.

4. Comparison of PSD and CHAD Methods

In this paper, two kinds of special distillation methods are used to separate the ethyl acetate/*n*-hexane azeotropic system. Table 5 provides a head-to-head comparison of economics among the optimized PSD, partially HIPSD, and CHAD processes. From the results, the energy cost of the partially HIPSD process has reduced by 40.79%, and meanwhile, the TAC of this process has reduced by 35.94% compared with conventional PSD. The energy cost and the TAC of the partially HIPSD process has further reductions of 62.61% and 49.26% compared with CHAD process. The main reason is that the reboiler of the LPC in the partially HIPSD consumes energy by the overhead steam of the HPC instead of low pressure steam and the partially HIPSD process saves the condenser of the HPC.

From the point of economic benefit, the partially HIPSD is much more attractive for ethyl acetate/*n*-hexane azeotropic system separation than the CHAD process and conventional PSD process. And it is obvious that the partially HIPSD has some advantages in averting the product pollution compared with the CHAD process.

5. Conclusions

Both CHAD and PSD methods are proposed to effectively separate the ethyl acetate/*n*-hexane system. Then, acetone is selected as the

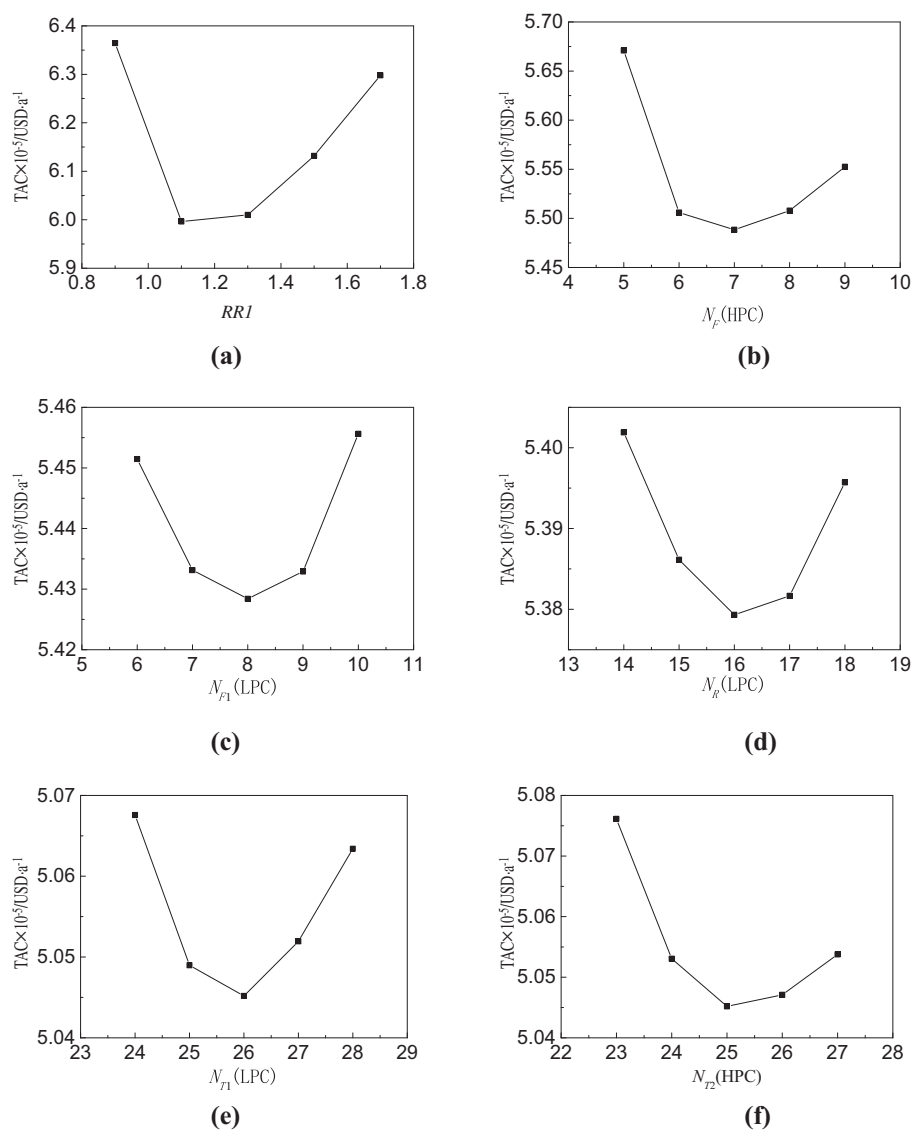


Fig. 10. The effects of operating parameters on TAC in the PSD process.

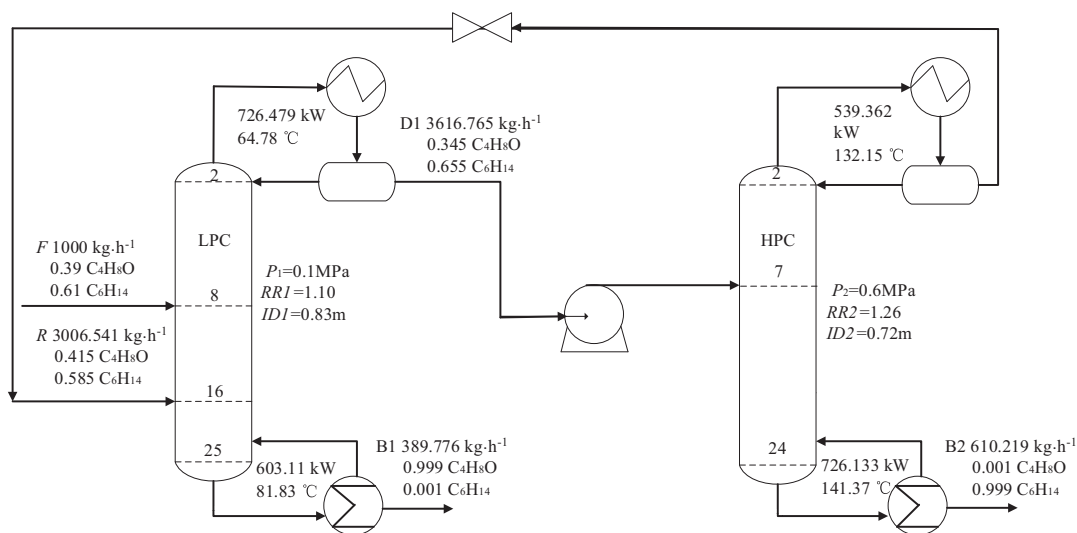


Fig. 11. The optimal flow sheet of the for the conventional PSD process.

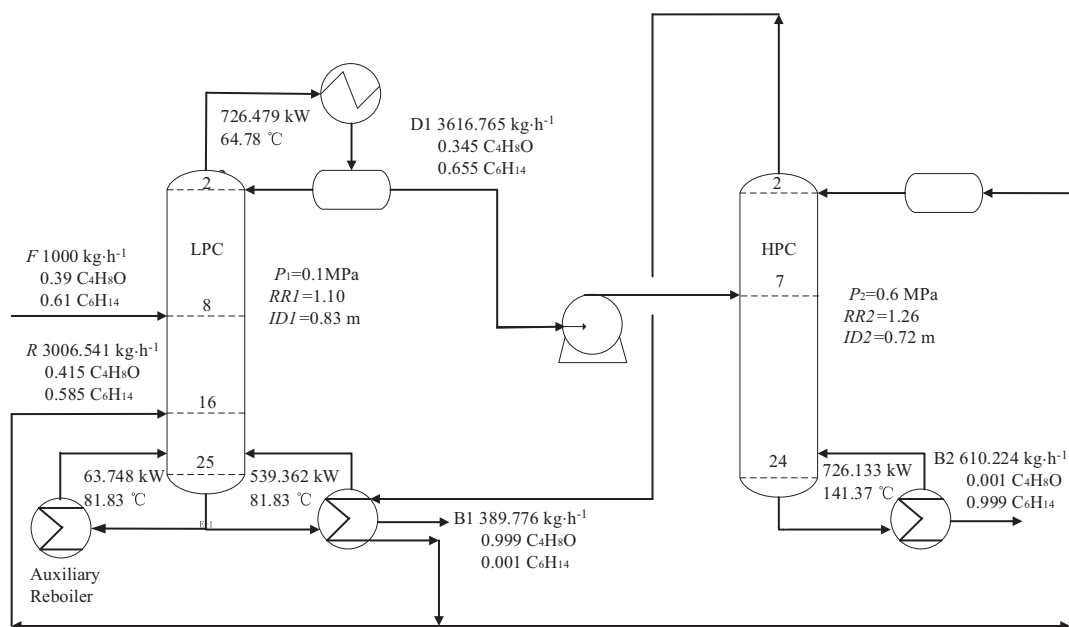


Fig. 12. The optimal flowsheet of partially HIPSD process.

Table 5

Optimal results for partially HIPSD, conventional PSD and CHAD processes

Parameter	Partially HIPSD		Conventional PSD		CHAD	
	LPC	HPC	LPC	HPC	ADC	ERC
N_T	26	25	26	25	37	13
N_F	8	7	8	7	26	11
N_R	16		16		24	
RR	1.10	1.26	1.10	1.20	0.907	4.04
Reboiler heat/kW	63.748	726.133	603.111	726.133	329.503	1908.662
Total energy cost $\times 10^{-5}/\text{USD} \cdot \text{a}^{-1}$	1.920		3.243		5.135	
Total equipment cost $\times 10^{-5}/\text{USD} \cdot \text{a}^{-1}$	1.312		1.802		1.235	
TAC $\times 10^{-5}/\text{USD} \cdot \text{a}^{-1}$	3.232		5.045		6.370	

appropriate entrainer through analyzing residue curve maps in the CHAD process. And the operating pressure of PSD is determined via the x - y diagram of acetate/ n -hexane system. Finally, two distillation processes are optimized by a sequential modular approach with TAC as the objective function.

In the CHAD process, the optimal number of stage is 38 for the ADC and 13 for the ERC. The feed stage location is 26th for N_{F1} , 24th for N_R , 11th for N_{F2} . The reflux ratio of ADC and ERC are 0.907 and 4.04, respectively. In the partially HIPSD process, the optimal number of stage is 26 for LPC and 25 for HPC. The feed stage locations of N_{F1} , N_R and N_{F2} are 8th, 16th and 7th. The reflux ratio of LPC is 1.10 and the HPC is 1.26. The pressures of LPC and HPC are 0.1 and 0.6 MPa, respectively.

The results show that the partially HIPSD process has 35.94% and 40.79% reduction of TAC and energy consumption compared with the conventional PSD process. Furthermore, the partially HIPSD process has more greatly reduced the energy cost and TAC by 62.61% and 49.26% respectively compared with the CHAD process. Besides, the partially HIPSD process effectively avoids product pollution caused by introducing a third component which must be added into ADC in the CHAD process. It is obvious that the partially HIPSD process has more advantages in averting the product pollution, energy saving, and economy among the three processes.

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