Use of multiple inter-reboilers to achieve energy savings and improve thermodynamic efficiency of the distillation of N,N-dimethylformamide wastewater

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Abstract−The purification of N,N-dimethylformamide wastewater involves an energy intensive distillation process. We propose a novel energy-saving process scheme involving multiple inter-reboilers sed. In this scheme, ideal thermodynamic model non-random two liquid (NRTL) model was used to calculate the phase equilibrium using Aspen Plus platform. While the relationship between important process parameters and energy consumption by the distillation process was studied, several parameters such as the most suitable positions for the inter-reboilers and the most reasonable steam extraction rates were obtained. The feasibility was detected under the same separation duties and main technological structure. For 10 wt% DMF wastewater, the inter-reboilers were installed on the 37th, 38th and 39th plates, while the corresponding heat transferred values were 3,038 kW, 91 kW and 179 kW, respectively. In comparison to the conventional distillation process, an energy consumption of 77.43% and thermodynamic efficiency of 65.69% were obtained. For 20 wt% DMF wastewater, the inter-reboilers were installed on the 21st and 25th plate, while the corresponding values for the heat transferred were 1,632 kW, and 1,450 kW, respectively. In comparison to the conventional distillation process, the energy consumption can be reduced by 71.31%, while the thermodynamic efficiency can be improved by 47.10%.

Keywords: Inter-reboiler, Energy-saving, Distillation, Thermodynamic Efficiency, N,N-dimethylformamide Wastewater

INTRODUCTION

N,N-dimethylformamide (DMF) is a toxic, colorless and transparent liquid used as a chemical raw material having excellent performance in petrochemical industry, organic synthesis, pharmaceutical, pesticide, synthetic fiber, acrylic spinning, leather and other fields [1,2]. Due to strong polarity, DMF is soluble in water, ether, alcohol, unsaturated hydrocarbon and aromatic hydrocarbon, and therefore, is called the "omnipotent solvent" [3-6]. Its widespread application in chemical production has resulted in a large quantity of wastewater containing DMF. The exposure or inhalation of DMF can cause not only hematopoietic and liver related problems, but also bring about serious contamination to the atmosphere and water [7-9]. For relatively lower DMF concentrations in wastewater, simple recovery devices cannot directly gain economic benefits due to high energy consumption and costly recycling process [10].

Currently, there exists a wide-range of treatment methods for DMF wastewaters, including physical and chemical methods [11-14] (azeotropic distillation, adsorption, and extraction), biochemical method [15,16], supercritical water oxidation method [17] and chemical method [18-21] (alkalization). Azeotropic separation has the disadvantages of high energy consumption, low thermodynamic efficiency, low separation purity and high separation cost, and com-

plex operation process [22-24]. Biochemical method has the characteristic of longer processing time and incomplete degradation, whereas the supercritical water oxidation method requires relatively rigorous conditions, which makes it difficult to apply on a large scale. Additionally, adsorption involves adsorbent regeneration and separation of reverse extraction process, which is not only relatively complex but also incurs high operating cost. Currently, distillation is used more commonly for DMF recovery despite its high-energy consumption in a conventional process.

In distillation, top heat is removed by cool recycle water, whereas high level steam heats the bottom, which undoubtedly may lead to a large amount of energy being wasted along with lower thermodynamic efficiency [25,26]. However, it is even more significant that a number of energy-saving industrial application technologies are not mature enough to be carried out at a commercial scale.

In the petrochemical industry, energy consumption by the distillation process is one of the core parts in separation processes. Statistically, distillation accounts for 3% of the whole industrial energy consumption. However, due to phenomena like energy shortage, increasing prices and environment pollution which is threatening human health, energy saving and emission reduction in distillation have become priority topics in social practice and scientific research. It is of great urgency and practical-oriented significance to achieve these targets [27].

To achieve the goal of energy-saving and improved thermodynamic efficiency, lower level heat transfer must be employed to heat the bottom of the column. However, high temperature of the col-

umn's bottom cannot be heated by lower level steam having a temperature difference of 10 °C for heat transfer [28]. To achieve energy saving, Alcántara-Avila et al. proposed a strategy to add an intermediate condenser [29], and Chao et al. proposed a strategy to add an intermediate reboiler [30] to save energy. In this study, the strategy of adding multiple reboilers is proposed, which is more energy and thermodynamically efficient.

In this paper, the inter-reboiler technology constitutes the stripping section for the mixture, which is partially extracted and heated by the inter-reboiler with lower level steam, and after sufficient heat transfer, was sent back to the column for separation. The principle is to reduce the consumption of high level steam in the column. In this way, the inter-reboiler technology can improve the thermodynamic efficiency and significantly reduce the energy consumption.

THEORY AND FEASIBILITY MONITORING

The T-x-y diagram of the DMF-water system is at atmospheric pressure shown in Fig. 1. Note that DMF-water belongs to a typical large temperature difference system, which results in a large temperature difference between the top and bottom of the column.

Accordingly, the high temperature system is in the bottom, where the inter-reboilers can be added and the mixture is extracted from the stripping section. The dropping liquid is heated to partial vaporization using low quality heat, while the mixture is returned to the column for further separation. For low temperature system in the bottom, the reboiler acts as a cold source recovery device for its own sake, where the inter-reboiler can also be added such that the mixture is extracted from the stripping section, which can be heated by low quality heat. Due to such an arrangement, the thermodynamic efficiency would show significant improvement.

In distillation, the mass transfer driving force is expressed by the difference between the operating and equilibrium lines. For smaller differences between the lines, the process becomes reversible, lead-
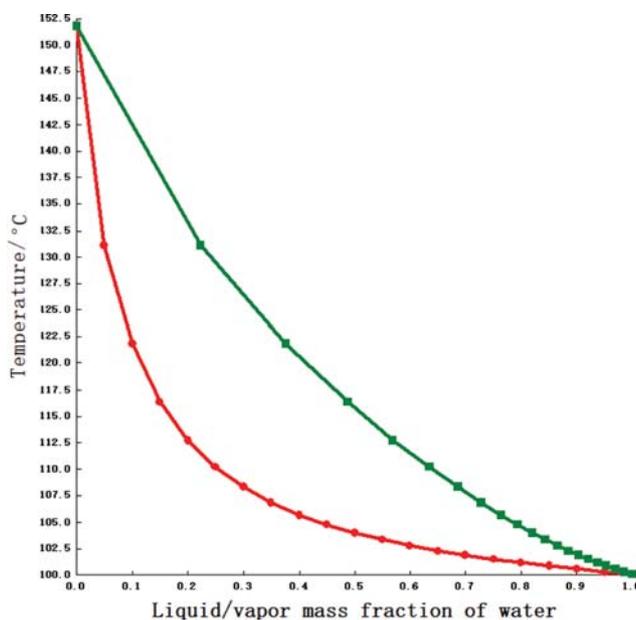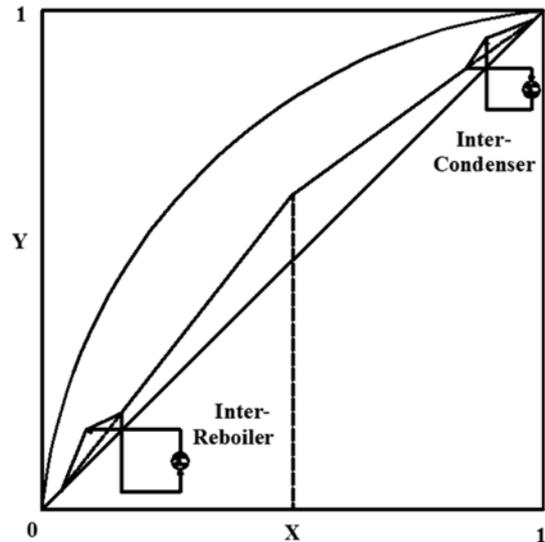

Fig. 2. McCabe-Thiele diagram with inter-reboiler and inter-condenser.

ing to higher thermodynamic efficiency. Ideally, the two lines should overlap perfectly, so that the energy loss would be 0 kW. Regrettably, it is impossible to attain the ideal state in an actual process.

The principle could be well explained by McCabe-Thiele considering a binary distillation process [31], which is shown in Fig. 2.

The operating line presents an obvious change, and is closer to the equilibrium line due to the installation of inter-reboiler. Meanwhile, high separation efficiency can be achieved by increasing the number of plates. Thus, adding an inter-reboiler increases the number of plates. Fortunately, the increase in number of plates is small, thus accounting for only a proportion of the capital cost. If this is the case, the economic benefit obtained from the operating costs is significantly greater than the increase in the equipment cost. Therefore, the addition of an inter-reboiler can always achieve good economic benefits under any circumstances.

SYSTEM DESCRIPTION AND EVALUATION OF THE MODEL

The wastewater was fed at a flowrate of 5,000 kg/h with the DMF mass fractions of 10 wt% and 20 wt% with the balance being the water. The product purity of the DMF was not less than 99.8 wt%. The top product was condensed by water. The bottom of the column was heated by 0.62 MPa saturated steam. For the inter-reboiler, the heating temperature was determined such that the minimum temperature difference with the extracted mixture was 10 °C.

Obviously, the choice of physical property method has a great impact on the accuracy of the simulation results. In this paper, NRTL model in Aspen Plus [32] was used to calculate and analyze the vapor-liquid equilibrium data, which shows remarkable agreement with the data reported in the literature [33]. Therefore, NRTL was chosen as the model for the simulation process.

1. Evaluation Index

The total annual cost (TAC) was calculated as an index to evaluate the whole process, which included the operating cost (OC)

Fig. 1. T-x-y diagram of the DMF-water system.

Table 1. Utility and other equipment costs used to calculate TAC

Symbol	Utility	Price
C_B	Steam	32.89 \$/ton
C_C	Column	1195.82 \$/m^3
C_A	Heat exchange	127.06 \$/m^2
r_B	Latent heat of water at 0.3 MPa	2177.6 kJ/kg
Cp	Specific heat capacity of water	4.18 kJ/kg·°C
Cw	Cost of cooling water	2.7 \$/ton

and investment cost (IC). TAC [34] can be calculated using Eqs. (1)-(3).

$$TAC=OC+IC \tag{1}$$

$$OC=C_B\cdot(7200\times3600)\cdot(Q_B/r_B)/1000 \tag{2}$$
$$+C_W\cdot(7200\times3600)\cdot(Q_C/4.18\times6)/1000$$

$$IC=\{C_C\bullet[1.35\times(0.35\times N_T)\bullet\pi D^2/4]+C_A\bullet A_T\}/\beta \tag{3}$$

For a clear understanding of the proposed configuration, the detailed information is given in Table 1.

SIMULATIONS AND OPTIMIZATION

1. Conventional Distillation Process

The simulation for the conventional distillation process is shown in Fig. 3. In the simulation process, the DSTWU module is first used for the simple distillation calculation, the reflux ratio is set to 1.2 times of the minimum reflux ratio, the light component water of the top product is 0.99 (mass fraction), and the heavy component DMF content is 0.01 (mass fraction). The initial number of stages and the reflux ratio were obtained by DSTWU' module. Then we used the RadFrac module for strict distillation calculation; the simple calculation results were brought into the "RadFrac" module. The "Design Spec" was used to change the reflux ratio (0-2) and the top material flow rate (2,000-5,000 kg/h), the purity of the bottom material of the column bottom was controlled 0.998. Finally, we could obtain the required reflux ratio and get the minimum TAC corresponding to the different the number of stages. The wastewater with the DMF content of 10 wt% and 20 wt% was heated to bubble point by the preheater. Water in the top product
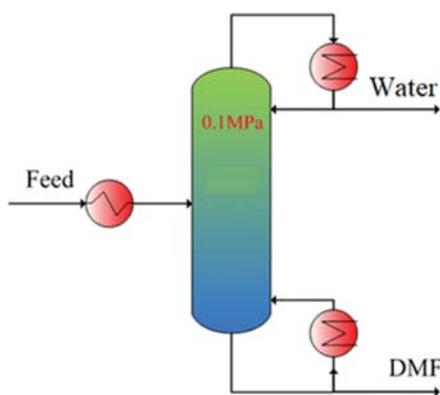
Fig. 3. Conventional single column distillation process.
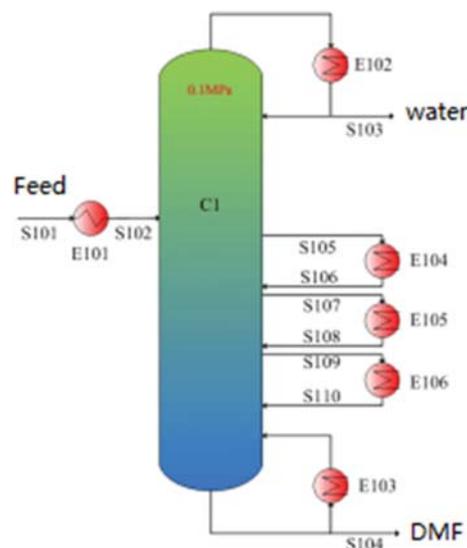
Fig. 4. 10 wt% DMF distillation with integrated inter-reboilers.

and DMF in the bottom product had purity of at least 99.8 wt%, respectively. The optimized parameters are shown in Table 2.

2. 10 wt% DMF Wastewater

2-1. Process Introduction

As all distillation columns with classic separation are operated at ambient pressure, the use of inter-reboilers was explored as an attractive process alternative for energy-saving. For such an innovation, the schematic of the process is shown in Fig. 4.

The fresh feed was fed at a mass flowrate of 5,000 kg/h having 10 wt% DMF and 90 wt% water. Under ambient temperature and pressure, the mixture flowed into the preheater, and once the bubble point was reached, it was returned to the column again. The top product had a purity of 99.8 wt% water, while the bottom product had a purity of 99.8 wt% DMF.

2-2. Position of the Inter-reboilers

To keep the same plate number and product purity, the inter-reboiler was fixed at the appropriate position between the feed and bottom in the stripping section. For the sake of optimal inter-reboiler position, the temperature distribution in the column must be acquired. From the perspective of energy-saving principle, the addition of inter-reboiler will undoubtedly affect the operating line, thus subsequently affecting the operational parameters. The most suitable location should not be far from the bottom; however, the temperature distribution indicates that when the position of the inter-reboiler is closer to the bottom, the heating medium required is higher and the reflux ratio needed is also larger. For these reasons, the inter-reboiler should not be too close to the bottom. For wastewater containing 10 wt% DMF, the process of the optimization of the position of inter-reboiler is described as the following curve (as shown in Fig. 5).

Fig. 5 shows that a steep temperature rise existed at the 37th plate, due to which, the first inter-reboiler was set as the 37th plate. Further considering the temperature distribution curve, other inter-reboilers were installed at the 38th, 39th and 40th plates.

2-3. Optimization of the Extraction Quantity

The extraction quantity cannot be arbitrarily set, and therefore a
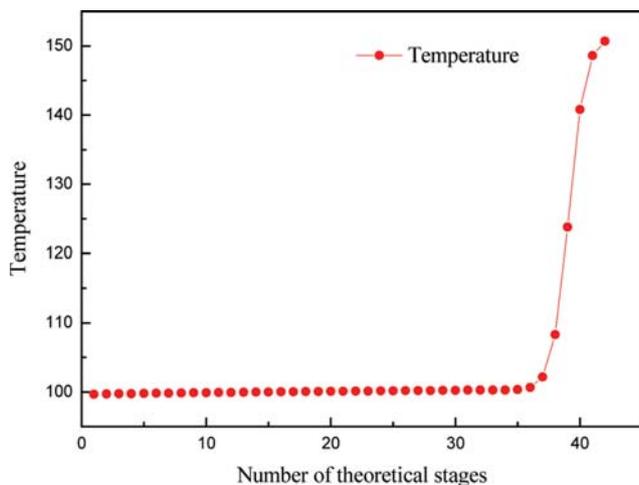
Fig. 5. Distribution of temperature in the distillation column.

comprehensive consideration must be given to both vapor-liquid flow rate and total annual cost. The optimization process for the inter-reboiler installed at 37th plate is described in Fig. 6(a). Clearly, the optimal extraction rate is 50% of the total plate flow. Similarly, the extraction rate was optimized for other inter-reboilers. The connection between the extraction quantity at 37th, 38th, 39th and 40th plates and total annual cost has been described in Fig. 6(b)-6(d).

After the addition of two inter-reboilers, the most reasonable extraction rates can be significantly different from each other. Fig. 6(b) indicates that 30% of the total flow on 37th and 90% of the total flow on 38th plates was the most appropriate. Supposing three inter-reboilers were added, Fig. 6(c) illustrates that 90% of the flow on the 37th, 10% of the flow on the 38th and 10% of the flow of on the 40th plates were reasonable. For the relatively more complicated case of four inter-reboilers, Fig. 6(d) demonstrates that 90% of the total flow on the 37th, 10% of the total flow on the 38th, 10% of the total flow on the 40th and 10% of the total flow on the 41th plates were the most reasonable.

Through the analysis of TAC under optimal conditions, values in Table 2 show that the strategy of using three inter-reboilers was economical and feasible, and the detailed parameters are shown in supporting information (Table S1). In this scenario, low quality steam was used to heat the extraction mixture and to further decrease the consumption of the steam. Additionally, with this three inter-reboiler strategy, the energy consumption can be reduced by 79.1% in comparison to the conventional distillation process.

Fig. 4 shows that 90 wt% of the total amount of the mixture (7,705.1 kg/h) was taken from the 37th plate through pipe (S105) and was heated by 0.17 MPa saturated steam using the heat exchanger (E104). The consumption of 3038.07 kW energy ensures that the mixture will be heated to 108.6 °C. About 543.1 kg/h mixture was extracted from the 38th plate and was heated by 0.17 MPa saturated
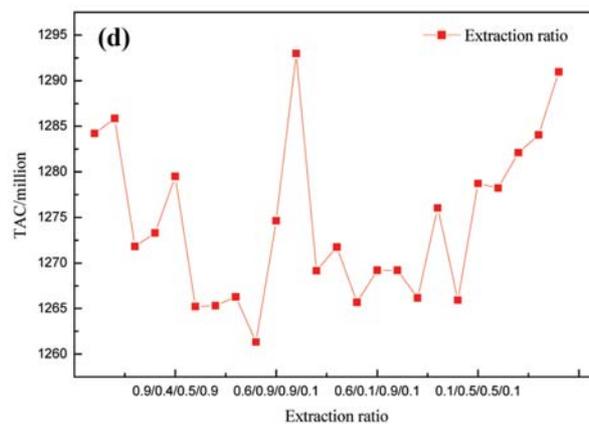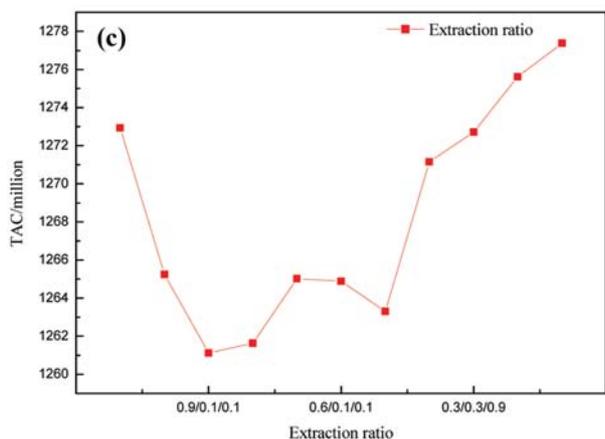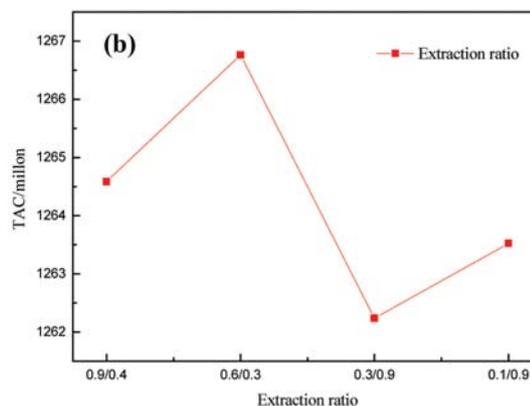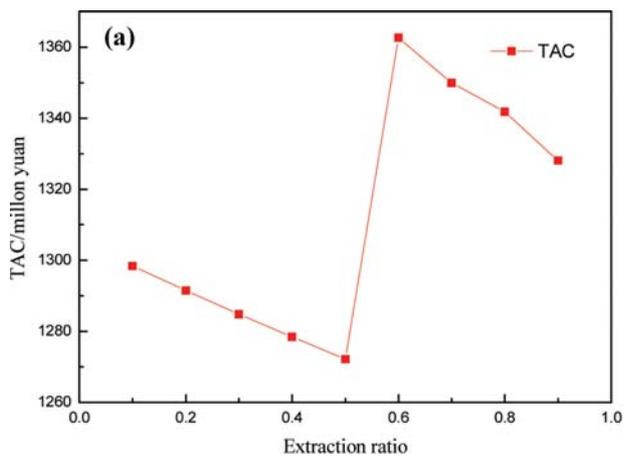

**Fig. 6. Optimization of the extraction quantity.**
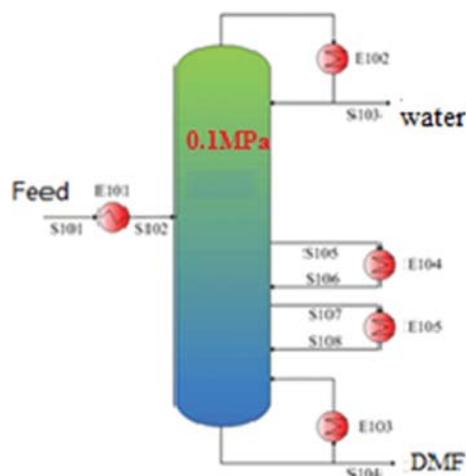
Fig. 7. 20 wt% DMF distillation based upon two inter-reboilers.

steam using E105 to the temperature of 115.4 °C. After heating, the mixture was returned to the column. Later, 0.2 MPa saturated steam was used to heat 391.5 kg/h of the mixture that was extracted from the 39th plate and passed through E106. Finally, 527 kg/h of the mixture was extracted from the 40th plate and was heated using 0.36 MPa saturated steam to the temperature of 140 °C.

3. 20 wt% DMF Wastewater

Compared to the 10 wt% wastewater, the distillation process for 20 wt% DMF wastewater was similar. Fig. 7 shows the flowsheet representing the distillation process used for 20 wt% DMF wastewater, and having two inter-reboilers.

Under ambient conditions, the mixture containing 20 wt% DMF (balance being the water) flowed into the preheater E101 with the flow rate of 5,000 kg/h. Upon reaching the bubble point, the mixture was returned to the column for further separation.

The positions of the inter-reboilers were optimized according to the same process as used for the 10 wt% DMF distillation. The results showed that the most appropriate locations for the inter-reboilers were 21st, 25th and 26th plates. Fig. 8 shows the results for the quantity of mixture extracted from the inter-reboilers.

For only one inter-reboiler, the most reasonable extraction quantity was 50% of the total flow on the 21st plate (as shown in Fig. 8(a)). Furthermore, for two inter-reboilers, 30% of the total flow on the 21st plate was the most reasonable, whereas 70% of the total flow on the 25th plate was found to be the most optimal (see Fig. 8(b)). Similarly, for three inter-reboilers, 30% of the total flow on the 21st plate, 50% of the total flow on the 25th plate and 10% of the total flow on the 26th plate were the most suitable (see Fig. 8(c)).

By means of analysis and calculations, the strategy of two inter-reboilers resulted in significant reduction in the use of high qual-
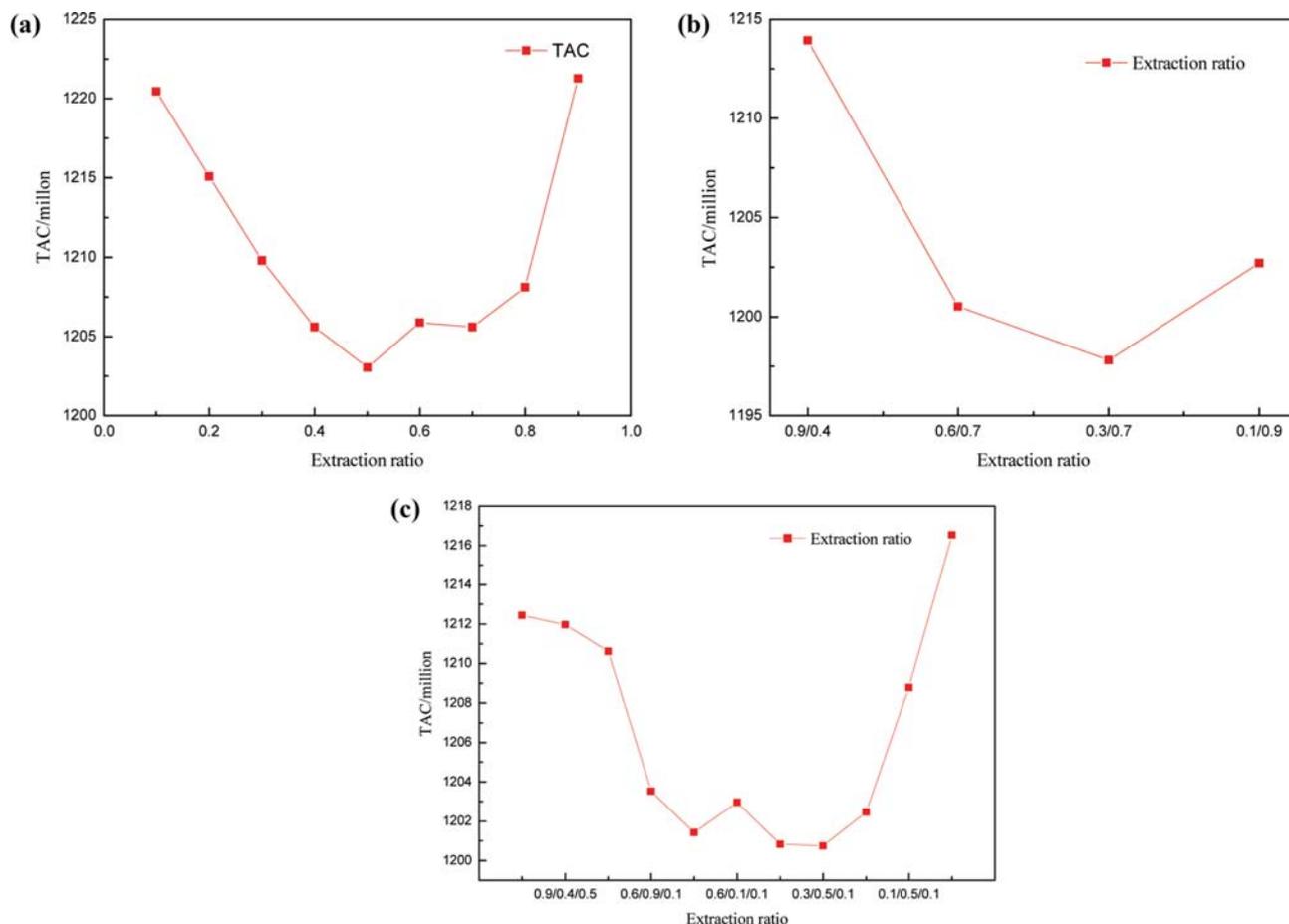

Fig. 8. Optimization of the extraction quantity.

Table 2. Summary results of the simulation

Parameter	Conventional distillation process		10 wt% DMF wastewater				20 wt% DMF wastewater		
	10 wt% DMF	20 wt% DMF	1	2	3	4	1	2	3
N_T	42	25	45	45	45	45	28	28	28
N_F	34	19	34	34	34	34	19	19	19
R	0.46	0.54	0.47	0.47	0.47	0.47	0.54	0.54	0.54
D (m)	1.75	1.80	1.25	1.12	1.15	1.16	1.15	1.08	1.07
Q_C (kW)	4146	3874	4062	4068	4076	4076	3859	3826	3826
Q_R (kW)	4272	4322.14	2309	1088	964	905	1586	1240	1154
TAC ($\times 10^4$ \$)	1343.93	1344.94	1272	1262	1261	1261	1205	1197	1200

Table 3. Summary for thermodynamic efficiency

Parameter	Conventional distillation process		10 wt% DMF wastewater				20 wt% DMF wastewater		
	10%DMF	20%DMF	1	2	3	4	1	2	3
Minimum work (kJ/h)	9.5×10^4	1.5×10^5	1.2×10^5	1.0×10^5	1.0×10^5	2.5×10^4	7.1×10^5	1.2×10^6	1.2×10^6
Exergy loss (kJ/h)	1.5×10^6	1.5×10^6	9.7×10^6	6.2×10^5	4.7×10^5	1.1×10^5	3.4×10^6	4.8×10^6	4.6×10^6
η/%	6.3	10	12	16	21	22	20	25	26

ity steam. Under the same requirements, the amount of low quality steam used to heat the extracted mixture was optimized so that the energy consumption can be reduced by 71.3% in comparison to the conventional distillation process.

Fig. 7 shows that 30% of the total flow (3,692.6 kg/h) was extracted from the 21st plate, and was heated by 0.17 MPa saturated steam. Consumption of 1,638.2 kW energy heated the mixture to 105.2 °C. Similarly, 0.23 MPa saturated steam was used to heat 4,835.8 kg/h of the flow, which was extracted from the 25th plate, and consumed 1,081.8 kW of energy. Additionally, 764 kg/h of the mixture was extracted from the 26th plate and was heated by 0.36 MPa saturated steam using E102. To achieve same separation results, all parameters were optimized, and are presented in Table 2. The detailed parameters, including the extraction quantity, extraction stage and other key points, can be seen in supporting information (Table S1).

4. Calculation of the Thermodynamic Efficiency

The core of the thermodynamic efficiency is mainly about the irreversible process. Several reasons are responsible for the exergy loss, which leads to a thermodynamic efficiency less than 1. On one hand, the excess reflux, which is the difference between the actual and minimum reflux ratio, contributed to it, while the temperature difference existing in different unit operations, including preheater, reboiler and condenser contributed to it. Finally, some exergy is lost to overcome the flow resistance of the mixture of different components. The inter-reboiler did well in improving the thermodynamic efficiency [35-37]. According to the second law of thermodynamics, minimum separation work, net power consumption and thermodynamic efficiency are calculated using Eqs. (4)-(6).

$$W_{min, T0} = \Delta B_{separation} = \Sigma_{out} n_k (H_k - T_0 S_k) - \Sigma_{input} n_j (H_j - T_0 S_j) \quad (4)$$

$$W_{net} = Q_R \left(1 - \frac{T_0}{T_R}\right) - Q_C \left(1 - \frac{T_0}{T_C}\right) \quad (5)$$

$$\eta = \frac{\Delta B_{separation}}{W_{net}} \quad (6)$$

where n_k is the total extracted mole flow, n_j is the total incoming mole flow, T_0 is the surrounding temperature, S is the entropy and H is the enthalpy. In a conventional distillation column, the heat added to the reboiler (Q_R) at the temperature of T_R represents required work, whereas the heat removed from the condenser (Q_C) at the temperature of T_C is the lost work. Table 3 presents a summary for thermodynamic efficiencies of various inter-reboiler strategies (process schemes discussed in the current work), and the energy loss of specific positions is shown in supporting information (Table S2).

Surprisingly, the thermodynamic efficiency can be improved by 39.3% for 10 wt% DMF distillation process with three inter-reboilers. Furthermore, the thermodynamic efficiency can be improved by 47.1% for 20 wt% DMF distillation process with two inter-reboilers.

CONCLUSION

Under the same conditions, the total heat in conventional distillation is equal to the sum of inter-reboiler heat and reboiler heat, whereas the only thing that changes is the additional use of low quality steam. In terms of 10 wt% DMF, the optimal energy-saving effect can be achieved by adding three inter-reboilers, which reduces the energy consumption by 77.43% and improves the thermodynamic efficiency by 65.69%. For 20 wt% DMF, the optimal energy-saving effect can be achieved by adding two inter-reboilers, which reduces the energy consumption by 71.31% and improves the thermodynamic efficiency by 47.10%. For the system with large temperature difference, the optimum use of the inter-reboiler can decrease the energy consumption and improve thermodynamic efficiency, which is of great practical significance.

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NOMENCLATURE

β : equipment depreciation/5
N_T : number of stage
N_F : feed stage
R   : reflux ration
QC   : condenser duty
QB   : reboiler duty
AT   : total heat exchange area
D   : diameter of column
Cw   : cost of electricity
ES1 : first extraction stage
ES2 : second extraction stage
ES3 : third extraction stage
ES4 : fourth extraction stage
ET1 : first extraction temperature
ET2 : second extraction temperature
ET3 : third extraction temperature
ET4 : fourth extraction temperature
IP1 : first inter-reboiler duty
IP2 : second inter-reboiler duty
IP3 : third inter-reboiler duty
IP4 : fourth inter-reboiler duty
EQ1 : first extraction quantity
EQ2 : second extraction quantity
EQ3 : third extraction quantity
EQ4 : fourth extraction quantity
TAC : total annual cost
OC : operation cost
IC : investment cost

SUPPORTING INFORMATION

Additional information as noted in the text. This information is available via the Internet at http://www.springer.com/chemistry/journal/11814.

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The different quality steams have different prices, In this paper, the price is 32.89 \$/ton for higher quality steam, and 21.43 \$/ton for low quality steam, respectively. Here, taking 0.1-0.9DMF conventional distillation process and 10 wt% DMF distillation process with three inter-reboilers as an example, the detailed calculation data is as follows.

The 0.1-0.9 DMF conventional distillation process

First, the operating cost (OC) is as follows:

$$OC = C_B \cdot (7200 \times 3600) \cdot (Q_B/r_B)/1000$$
$$+ C_W \cdot (7200 \times 3600) \cdot (Q_C/4.18 \times 6)/1000 = 32.89$$
$$\times (7200 \times 3600) \times (4272/2177.6)/1000 + 2.7 \times (7200 \times 3600)$$
$$\times (4146/4.18 \times 6)/1000 = 1327.84 \times 10^4 \, (\$)$$

Second, the investment cost (IC) is as follows:

$$IC = \{C_C \cdot [1.35 \times (0.35 \times N_T) \cdot \pi D^2/4] + C_A \cdot A_T\}/\beta$$
$$= \{1195.82 \cdot [1.35 \times (0.35 \times 42) \cdot \pi 1.75^2/4] + 127.06 \cdot 5878.91\}/5$$
$$= 16.09 \times 10^4 \, (\$)$$

Finally, The total annual cost (TAC) is as follows:

$$TAC = OC + IC = 1327.84 \times 10^4 + 16.09 \times 10^4 = 1343.93 \times 10^4 \, (\$)$$

The 0.1-0.9 DMF adding three inter-reboilers

First, the operating cost (OC) is as follows:

Table S1. Summary results of the simulation

Parameter	Conventional distillation process		10 wt% DMF wastewater				20 wt% DMF wastewater		
	10 wt% DMF	20 wt% DMF	1	2	3	4	1	2	3
N_T	42	25	45	45	45	45	28	28	28
N_F	34	19	34	34	34	34	19	19	19
ES1	-	-	37	37	37	37	21	21	21
ES2	-	-	-	38	38	38	-	25	25
ES3	-	-	-	-	39	39	-	-	26
ES4	-	-	-	-	-	40	-	-	-
ET1 (°C)	-	-	101	103	102	103	103	102	102
ET2 (°C)	-	-	-	104	112	105	-	115	118
ET3 (°C)	-	-	-	-	115	113	-	-	132
ET4 (°C)	-	-	-	-	-	130	-	-	-
IP1 (kW)	-	-	1963	983	3038	3018	2736	1632	1641
IP2 (kW)	-	-	-	2201	91	174	-	1450	732
IP3 (kW)	-	-	-	-	179	88	-	-	795
IP4 (kW)	-	-	-	-	-	87	-	-	-
R	0.46	0.54	0.47	0.47	0.47	0.47	0.54	0.54	0.54
D (m)	1.75	1.80	1.25	1.12	1.15	1.16	1.15	1.08	1.07
EQ1 (kg/h)	-	-	4280	2568	7705	7705	7385	3692	3692
EQ2 (kg/h)	-	-	-	6348	543	543	-	6770	4835
EQ3 (kg/h)	-	-	-	-	3915	391	-	-	764
EQ4 (kg/h)	-	-	-	-	-	527	-	-	-
Q_C (kW)	4146	3874	4062	4068	4076	4076	3859	3826	3826
Q_R (kW)	4272	4322.14	2309	1088	964	905	1586	1240	1154
TAC ($\times 10^4$ \$)	1343.93	1344.94	1272	1262	1261	1261	1205	1197	1200

Table S2. A summary for thermodynamic efficiency

Parameter	Conventional distillation process		10 wt% DMF wastewater				20 wt% DMF wastewater		
	10%DMF	20%DMF	1	2	3	4	1	2	3
Minimum work (kJ/h)	9.5×10^4	1.5×10^5	1.2×10^5	1.0×10^5	1.0×10^5	2.5×10^4	7.1×10^5	1.2×10^6	1.2×10^6
Preheater loss (kJ/h)	9.2×10^4	9.5×10^4	7.3×10^4	5.8×10^4	7.7×10^4	1.3×10^4	2.6×10^5	3.9×10^5	4.4×10^5
Condenser loss (kJ/h)	7.5×10^5	6.4×10^5	4.3×10^5	2.3×10^5	1.8×10^5	3.9×10^4	1.4×10^6	2.1×10^6	1.7×10^6
Reboiler loss (kJ/h)	4.2×10^5	4.4×10^5	2.3×10^5	1.4×10^5	7.3×10^4	1.2×10^4	9.6×10^5	1.0×10^6	9.4×10^5
Inter-reboiler loss (kJ/h)	-	-	8.3×10^4	2.3×10^4	7.2×10^2	5.9×10^3	1.7×10^5	1.2×10^5	3.2×10^5
Inter-reboiler (kJ/h) loss (kJ/h)	-	-	-	6.4×10^4	1.7×10^4	9.5×10^3	-	2.7×10^5	2.2×10^5
Inter-reboiler (kJ/h) loss (kJ/h)	-	-	-	-	2.5×10^4	6.6×10^3	-	-	6.3×10^5
Inter-reboiler loss (kJ/h)	-	-	-	-	-	8.9×10^3	-	-	-
Column (kJ/h) segment (kJ/h)	2.1×10^5	1.5×10^5	1.6×10^5	1.1×10^5	8.8×10^4	1.8×10^4	5.3×10^5	8.7×10^5	8.4×10^5
Exergy loss (kJ/h)	1.5×10^6	1.5×10^6	9.7×10^6	6.2×10^5	4.7×10^5	1.1×10^5	3.4×10^6	4.8×10^6	4.6×10^6
η/%	6.3	10	12	16	21	22	20	25	26

$$OC=C_B\cdot(7200\times3600)\cdot(Q_B/r_B)/1000$$
$$+C_W\cdot(7200\times3600)\cdot(Q_C/4.18\times6)/1000=32.89$$
$$\times(7200\times3600)\times(964/2177.6)/1000+21.43$$
$$\times(7200\times3600)\times(3308/2584.6)/1000+2.7$$
$$\times(7200\times3600)\times(4076/4.18\times6)/1000=1241.89\times10^4\ (\$)$$

$$IC=\{C_C\bullet[1.35\times(0.35\times N_T)\bullet\pi D^2/4]+C_A\bullet A_T\}/\beta$$
$$=\{1195.82\bullet[1.35\times(0.35\times45)\bullet\pi 1.15^2/4]$$
$$+127.06\bullet7363.48\}/5=19.24\times10^4\ (\$)$$

Finally, the total annual cost (TAC) is as follows:

$$TAC=OC+IC=1241.89\times10^4+19.24\times10^4=1261.13\times10^4\ (\$)$$