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Comprehensive Evaluation of Distillation Methods for Water-Ethanol Separation: Energy, Exergy, and CO_2 Emission Analysis with Process Enhancement

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ABSTRACT

Azeotropic distillation (AD) and extractive distillation (ED) are two conventional methods for separating azeotropic mixtures. However, these processes consume a high amount of energy like other distillation columns. As an entrainer may significantly affect energy consumption, selecting a proper entrainer is considered a key parameter in designing AD and ED processes. In this study, three different entrainers were used to separate the azeotropic water-ethanol mixture. Two entrainers, benzene and cyclohexane, were used in the AD process, and ethylene glycol was used as an entrainer in the ED process. According to the results, the ED process with ethylene glycol entrainer outperformed the AD process with benzene and cyclohexane entrainer in terms of separation and energy consumption. Therefore, in order to reduce the energy consumption of the ED process with ethylene glycol entrainer, feed splitting (FS) method was used. In addition, to further investigation on these two processes, exergy loss, CO_2 emission and TAC were calculated. Results showed that the ED process with heat integration reduced hot and cold utilities, respectively, by 94.33% and 60.56% relative to the AD process. Also, exergy loss, CO_2 emission and TAC in the proposed ED process decreased compared to the basic ED process to the extent of 32%, 7% and 80.1%.

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1. Introduction

Separation plays a vital role in all chemical processes [1]. One of the most convenient methods of separation in the industry is distillation [2]. However, distillation cannot be used to separate all mixtures. For example, when an azeotrope is formed from the mixture components, complete separation cannot be achieved by distillation [3]. Under such conditions, azeotropic distillation (AD), extractive distillation (ED), and pressure swing distillation (PSD) should be used, or even liquid-liquid extraction (LLE) is another possible method and membrane methods should be employed [4-8]. A third substance called entrainer is used in the AD and ED processes to break the azeotrope [9]. The entrainer should be recovered in another column. The problem is the high amount of energy consumed in distillation columns. At least two columns are required in the AD and ED processes (the main column and an entrainer recovery column), and azeotropic mixtures are difficultly separated[10]. Therefore, energy optimization of these two processes seems necessary. The water-ethanol mixture is one of the well-known azeotropic mixtures. Ethanol has numerous applications. Biofuels have been recently proposed to reduce the use of fossil fuels [11]. In 2009, the EU set the use of biofuels from renewable resources as a major goal up to 2020 [12].

Ethanol is the most well-known biofuel, which is extensively used as a liquid fuel in transportation[13]. However, high-purity ethanol should be used as fuel, and it is necessary to eliminate the water-ethanol azeotrope point [14]. Numerous studies have been conducted on water-ethanol separation. The optimization of the water-ethanol azeotropic mixture separation by Franke using the process of AD. The objective function was total annual cost (TAC), and continuous and discrete variables were considered optimization variables. According to the optimization results, discrete variables had a lower impact on TAC. Their results also revealed that when all variables are optimal, the heat duty may be reduced by 20% [15]. For separating the azeotropic mixture of water-ethanol Shang et al. used the ED process. A deep eutectic solvent was used as entrainer. According to their results, the solvent outperformed ionic liquids in terms of the emissions of CO_2, energy consumption, and TAC [16]. Pan et al., in the ED process, used an ionic solvent as an entrainer for the separation of the ternary water–ethanol-ethyl acetate mixture. The thermally coupling method was used for heat integration. The propositioned heat integration method indicated a reduction of heat duty by 12% following the findings [17]. Shi et al. used a heat pump for heat integration of direct and indirect ED columns to purify ethanol. TAC reduction by 43%was also noted down as an optimization of the direct process relative to the indirect process. The direct method also reduced energy cost and CO_2 emissions by 41% and 46% relative to the indirect method[18]. Li et al. proposed the extractive-azeotropic dividing wall column distillation for ethanol and water separation and produce isobutyl alcohol and ethanol. According to their results, this process along with heat integration, may reduce TAC by 37% [19]. Popescu et al. used the AD process for separating water and ethanol. One column was used instead of two columns in the proposed process, so that the column was simultaneously used for heterogenous azeotropic distillation. The proposed process reduced energy consumption by 50% [20].

Sun et al. presented research aimed at enhancing the separation of azeotropic mixtures via side-stream extractive distillation while concurrently addressing economic, environmental, and safety criteria. The primary objective is to design an intensified process architecture that integrates thermodynamic analysis (using quaternary phase diagrams) with multi-objective optimization (employing a multi-objective particle swarm optimization algorithm) and multi-criteria decision making (using the TOPSIS method with entropy weighting). The methodological framework encompasses a conceptual design, process parameter optimization, and performance evaluation in terms of reductions in annual costs, CO_2 emissions, and the process safety index. Key findings indicate that one of the proposed

configurations (SSED-1) achieves substantial reductions in both investment and operational costs—up to 19.21%—and decreases CO_2 emissions by up to 16.80% relative to conventional methods. These outcomes underscore the potential for advancing sustainable industrial processes and reducing energy consumption. While the study's comprehensive multidimensional approach and application of modern optimization algorithms are significant strengths, its computational complexity and the need for industrial-scale validation represent potential limitations [21].

Salman et al. investigated the azeotropic separation of ethyl propionate and ethanol using two distillation techniques: extractive distillation (EDM) with heavy entrainers (glycerol and ethylene glycol) and pressure swing distillation (PSDM). The research highlights the significance of effective separation for applications in biofuel production and pharmaceuticals. Utilizing Aspen Plus, the authors conducted detailed property and phase behavior analyses, coupled with temperature- and pressure-dependent modeling, to optimize process parameters and compare product purities, energy consumption, and total annual costs (TAC). Key findings reveal that EDM using glycerol achieves 99.90 mol% purity for ethyl propionate with considerable TAC reductions compared to EDM with ethylene glycol and PSDM, which, although offering lower reboiler duty, yields slightly inferior product purity. Notable strengths of the work include its comprehensive simulation approach and rigorous process optimization, while its complexity and the need for precise operating conditions are acknowledged as potential limitations [22].

Xing et al. investigated and optimizes a liquid-only transfer extractive dividing wall column (LEDWC) integrated with vapor recompression (VRC) and an intermediate reboiler (IR) for separating minimum-boiling point azeotropes. The primary objective is to reduce energy consumption, lost work, total annual cost (TAC), and gas emissions by introducing four novel configurations and optimizing TAC. The methodology involves steady-state simulation in Aspen Plus, thermodynamic modeling using the NRTL model, and comprehensive energy, exergy, economic, and environmental (4E) analyses. The results show that the RE-VRC-IR-LEDWC-CR configuration achieves savings of 39% in energy consumption, 31% in lost work, 25% in TAC, and 39% in gas emissions compared to conventional methods. Notable strengths of this research include its integrated 4E analysis and multi-stage optimization approach, although the complexity of implementation and industrial-scale transfer remain as potential limitations [23].

Guo et al. aimed at enhancing the recovery of ethanol and isopropyl alcohol from industrial and petrochemical wastewater, an extractive distillation process employing a double-entrainer system—[EMIM][BF4] and ethylene glycol—is proposed. The research investigates the system's ability to break the ethanol/water and isopropyl alcohol/water azeotropes, as well as to separate the near-boiling ethanol/isopropyl alcohol mixture. The main objectives include evaluating the energy-economic performance and reducing operating costs compared to single-entrainer methods, with research questions focusing on the effectiveness of heat pump-assisted pre-concentration (HPP) schemes in lowering water content and enhancing process efficiency. The methodologies employed encompass simulation using Aspen Plus V11, optimization of process parameters in the distillation columns (AWDC, EIDC, and SRC), and a comparative analysis between single- and double-entrainer processes. Results indicate a significant reduction in energy consumption and annual costs (up to over 50%) when applying the double-entrainer strategy combined with HPP. While the study's innovative and comprehensive approach offers substantial improvements in energy efficiency and cost reduction, the increased equipment expenses associated with the HPP configuration represent a relative drawback [24].

Liu et al. investigates the design and multi-objective optimization of a reactive-extractive dividing wall column (REDWC) integrated with Organic Rankine Cycle (ORC) systems while considering safety aspects. The research, focused on the separation of a ternary azeotropic mixture of ethyl acetate, ethanol, and water commonly found in industrial wastewater, addresses significant challenges in conventional separation methods. The primary objectives are to reduce the total annual cost (TAC), decrease CO_2 emissions, and improve the process route index (PRI) using a multi-objective genetic algorithm (MOGA); additionally, the ORC system is designed for waste heat recovery to enhance energy efficiency. Results show that the REDWC process outperforms the traditional DCRED method by lowering TAC and CO_2 emissions by up to 8.15% and 4.49%, respectively, and that integrating ORC with REDWC achieves further savings of over 27%, although incorporating feed preheating (REDWC-FP) yields no significant improvement over REDWC [25].

The research gaps in heat integration for the water-ethanol separation process include challenges in design, control, and integration with alternative energy sources. The complexity of equipment and high capital costs hinder the widespread industrial implementation of these technologies. Operational stability and sensitivity to feed variations require the development of advanced control strategies. Integrating these processes with renewable energy sources and waste heat still demands optimized and compatible designs. The combination of different separation methods to enhance energy efficiency has been less explored. A comprehensive economic and environmental assessment of these technologies at an industrial scale remains an outstanding research need.

This study aims to select the best (optimal) process and entrainer reducing energy consumption for the separation of the mixture of water and ethanol. The AD and ED processes were evaluated for this purpose. Benzene and cyclohexane were used as entrainer in the AD process and ethylene glycol in the ED process. Hence, three different processes were compared. Two distillation columns were used in both AD and ED processes. Energy optimization seems necessary because of the high amounts of energy consumed in distillation columns. Numerous methods have been recently proposed to reduce the energy consumption in the columns of distillation, among which pinch, heat pump, and VRC can be noted [26-28]. To reduce energy consumption, the method of feed splitting (FS) is employed. In it, the feed is split into two streams, and the preheating of the lower stream by a process stream significantly decreases the reboiler heat duty. As the upper feed stream comes through the column without any temperature change, The heat duty of the condenser changes negligibly. Consequently, energy consumption can be reduced significantly. In addition, CO_2 emission, exergy loss and TAC were calculated for the ED process and the ED process with heat integration to further investigate the effects of FS method on energy consumption, CO_2 emission, TAC and exergy loss.

2. Case study

AD and ED processes were used in this study to separate water and ethanol. Table 1 reports feed specifications[29].

Table 1. Feed conditions

Conditions		**Feed**
Temperature [K]		298
Pressure [kPa]		101.3
Mass flow rate [kg/h]		1000
Mass fractions [%]	Ethanol	0.5
	Water	0.5

It should be noted that feed conditions are the same in all three processes. At 101.3 kPa, water and ethanol combine to produce a minimum-boiling homogenous azeotropic combination. Fig. 1 shows the diagram of equilibrium of the mixture at 101.3 kPa, plotted using Aspen Plus.

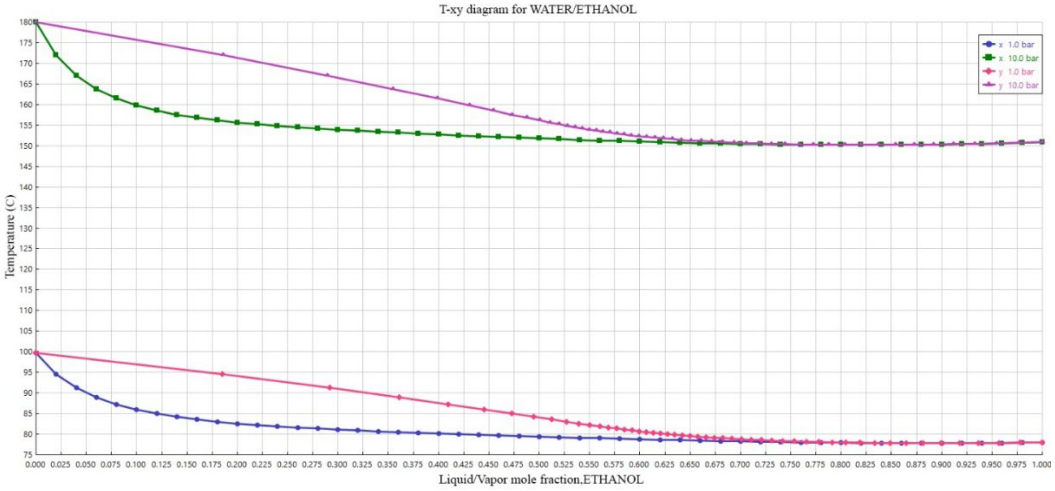

Fig. 1. The equilibrium diagram of the water-ethanol mixture at different pressures

3. Methods

Processes were simulated in Aspen HYSYS V.11. The exact modified HYSIM Inside-Out method was used to increase the accuracy of calculation and run columns. The NRTL equation was used for determining the liquid-phase activity. The SRK equation of state (EOS) was used for the vapor phase fugacity coefficient. Table 2 list the binary parameters of the NRTL model[30].

Table 2. Binary parameters of the NRTL model

Material	Material				
	Ethanol	**Water**	**Benzene**	**Ethylene glycol**	**Cyclohexane**
Ethanol	-	1332.3	991.50	-203.769	1253.4
Water	-109.63	-	4843.3	771.662	4042.5
Benzene	334.15	3719.4	-	1300.41	286.97
Ethylen glycol	1644.0	1715.8	108821	-	9407.1
Cyclohexane	545.70	4720.4	-33.395	1976.75	-

3.1. Heat integration of processes

The feed splitting (FS) method was used for heat integration of processes. The feed is first split into two streams. The lower feed stream is preheated by the bottom product of the column or a hot process stream, reducing the reboiler heat duty. The temperature of the top input stream does not vary as it enters the column, and the condenser's heat duty remains almost unchanged [31].

3.2. Environmental evaluation

CO_2 emission is a key parameter in the environmental assessment of chemical processes. The increasing emission of CO_2 and the subsequent warming of the earth due to the greenhouse effect have raised global concerns. Therefore, reducing CO_2 emission is considered an effective factor in the design of chemical processes. CO_2 emission in distillation processes are measured using equation 1 [32]:

$$CO_{2,emission} = \frac{Q_{fuel}}{NHV} \times \frac{C\%}{100} \times \alpha \quad (1)$$

Where α is the molar mass ratio of CO_2 and equals to 3.67, NHV is the net heating value for heavy fuel oil and equals to 39771 kJ/kg, C% is the carbon content, which is 86.5%, and Q_{fuel} is the duty of fuel and calculated from equation 2 [32]:

$$Q_{fuel} = \frac{Q_{Proc}}{\lambda_{proc}}(h_{proc} - 419) \times \frac{(T_{FTB} - T_0)}{(T_{FTB} - T_{stack})} \quad (2)$$

Where Q_{Proc} (kJ/h) is the heat duty of the reboiler, λ_{Proc} (kJ/kg) is the latent heat of low pressure steam (LPS) or medium pressure steam (MPS), h_{Proc} (kJ/kg) is the enthalpy of LPS or MPS, T_0 is the reference temperature (T_0=25°C), T_{FTB} is the flame temperature (T_{FTB}=1800°C), and T_{Stack} is the stack temperature (T_{Stack}=160°C). CO_2 emission for electrical power is equal to 51.1 kgCO_2/GJ.

3.3. Exergy

Unlike energy analysis, exergy accounts for both quantity and quality of various energy forms, so exergy has been recommended as a measure of system performance[33]. Exergy is the maximum available work obtained by bringing a system to equilibrium with its environment [28]. Exergy of a material stream which flows through a system can be divided into chemical exergy and physical exergy. Total exergy (Ex_{total}) of a material stream, physical exergy (Ex_{ph}), and chemical exergy (Ex_{ch} for ideal gas and ideal mixture) are measured using equations 3 to 5[33]:

$$Ex_{total} = Ex_{ph} - Ex_{ch} \quad (3)$$

$$Ex_{ph} = (H - H_0) - T_0(S - S_0) \quad (4)$$

$$Ex_{ch} = RT_0 \sum_{i=1}^{n} n_i \ln x_i \quad (5)$$

Except the exergy that enters or leaves the system through a material stream, the heat that enters or leaves the system can also have exergy and calculated from equation 6 [33]:

$$E_Q = Q \times (1 - \frac{T_0}{T}) \quad (6)$$

Exergy loss can be calculated from equation 7:

$$Ex_{loss} = \sum_{i=1}^{n} (Ex_{in})_i - \sum_{i=1}^{n} (Ex_{out})_i - \Delta Ex_{system} \quad (7)$$

In steady state condition, ΔEx_{system} is equal to zero.

3.4. TAC

In this research, in order to compare the economic cost of the two simulated processes, the parameter of Total Annual Cost (TAC) has been used. The Total Annual Cost has been calculated through equation 8 [32]:

$$TAC = Operating + (\frac{Capital\,\mathrm{Cos}\,t}{PaybackPeriod}) \quad (8)$$

This paper considers a payback period of 5 years and a man-hour value of 2480 hours. The utility costs are equal to: Cooling water = 0.354 ($/GJ), Low pressure steam (LP) (5 barg, 160°C) = 13.28 ($/GJ) and Medium pressure steam (MP) (10 barg, 184°C) = 14.19 ($/GJ). Additionally, the equipment costs are based on dollars[32]. The equations for calculating equipment costs are provided in the table below.

Table 3. Cost equation of equipment [32]

Equipment	Cost equation
Column	$\frac{M\&S}{280}\times 101.9\times D_c^{1.066}\times L_c^{0.802}\times(2.18+3.67)$
Tray	$\frac{M\&S}{280}\times 4.7\times D_c^{1.55}\times L_c\times(1+1.8+1.7)$
Heat exchanger	$\frac{M\&S}{280}\times 101.3\times A^{0.65}\times(2.29+F_c)$
Pump	$26700\times(\frac{24\times F_F\times 3600}{50000})^{0.53}$

4. Process simulation

PSD, AD, and ED processes were used for the separation of the mixture of water and ethanol. Benzene and cyclohexane were used as entrainers in the AD process and ethylene glycol in the ED process.

4.1. Pressure swing distillation

Two columns with distinct pressures are expended in the PSD process. Fig. 1 displays the equilibrium diagram of the water-ethanol mixture at two different pressures. As shown in Fig. 1, the azeotrope point is not considerably shifted by changing pressure. Therefore, PSD is not considered a suitable method for separating the water-ethanol mixture.

4.2. Azeotropic distillation with benzene entrainer

Fig. 2 shows the AD process of water-ethanol with benzene as an entrainer.

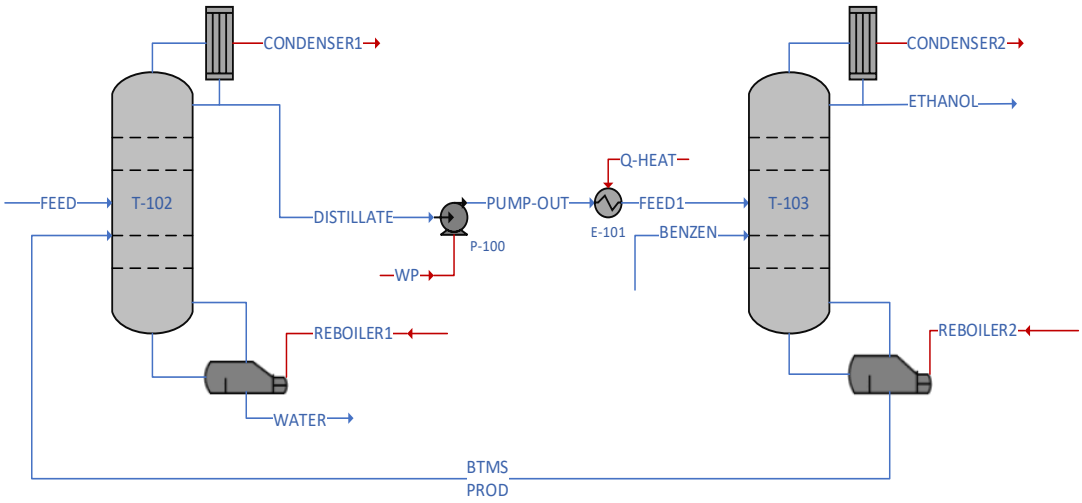

Fig .2. Azeotropic distillation of water-ethanol mixture with benzene as entrainer

As shown in Fig. 2, the feed enters tray 18 of the first column. This column contains 20 trays, operating at 101.3 kPa. The top product contains the water-ethanol azeotropic mixture leaving the first column entering tray 23 of the second column after being preheated while the bottom exit stream is pure water. This column contains 25 trays operating at 101.3 kPa like the first column. For the breakage of the azeotrope point, Benzene is used in this column as an entrainer. Benzene enters the first tray of the second column. The top exit stream is almost pure ethanol, and the bottom product containing water and ethanol is returned to the process inlet

4.3. Azeotropic distillation with cyclohexane entrainer

The schematic of this process is similar to the previous one, with cyclohexane used instead of benzene. The feed enters tray 18 of the first column. Like the AD process with benzene as entrainer, the first and second columns respectively contain 20 and 25 trays, both operating at 101.3 kPa. The bottoms liquid outlet of the first column is pure water, and its overhead liquid outlet contains the water-ethanol azeotropic mixture entering the second column after being preheated. Cyclohexane is used in this column so that cyclohexane enters the first tray. The overhead liquid outlet of the second column is almost pure ethanol, and the bottoms liquid outlet containing water and ethanol is returned to the process inlet.

4.4. Extractive distillation with ethylene glycol as entrainer

The extractive distillation of the water–ethanol combination using ethylene glycol as an entrainer is shown in Fig. 3. As shown in Fig. 3, the feed enters tray 18 of the first column. Like the other two processes, the first and second columns contain 20 and 25 trays, operating at 101.3kPa. Ethylene glycol enters the first tray of the first column as an entrainer. The overhead liquid outlet of the first column is pure ethanol, and its bottoms liquid outlet contains the water-ethanol azeotropic mixture entering the next column. Pure water exits the top of the second column, and pure ethylene glycol exits the column and is returned to the process inlet after being precooled.

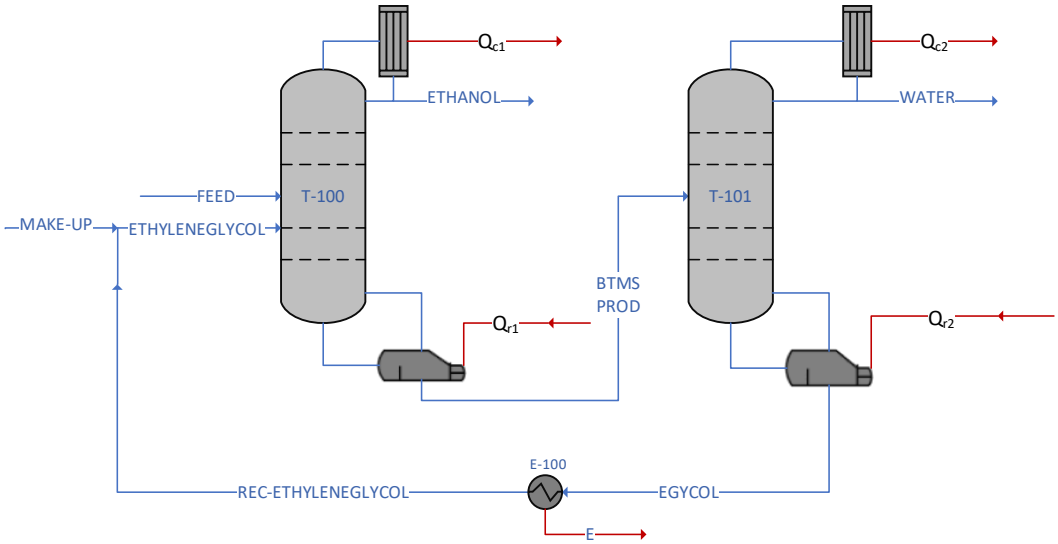

Fig. 3. Extractive distillation of the water-ethanol mixture with ethylene glycol as entrainer

5. Results and discussion

Water and ethanol mixture separation by the AD process with benzene as entrainer (P1), AD process with cyclohexane (P2), and the ED process with ethylene glycol as entrainer (P3) was exactly simulated. Fig. 4 compares ethanol purity at three distinct procedures.

As indicated in Fig. 4, the ED process with its entrainer being ethylene glycol outperformed the AD processes with benzene and cyclohexane as entrainers. In terms of energy usage, three processes are compared. Fig. 5 compares hot and cold utility consumption of the three processes. As shown in Fig. 5, the ED process with ethylene glycol as entrainer (P3) significantly reduces hot and cold utility consumption relative to both AD processes with benzene and cyclohexane as entrainers. P3 is thus chosen as the most energy-efficient procedure.

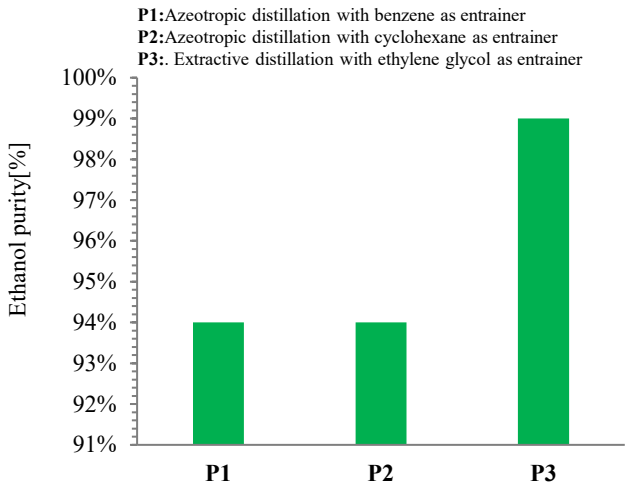

Fig. 4. Comparison of ethanol purity in different processes

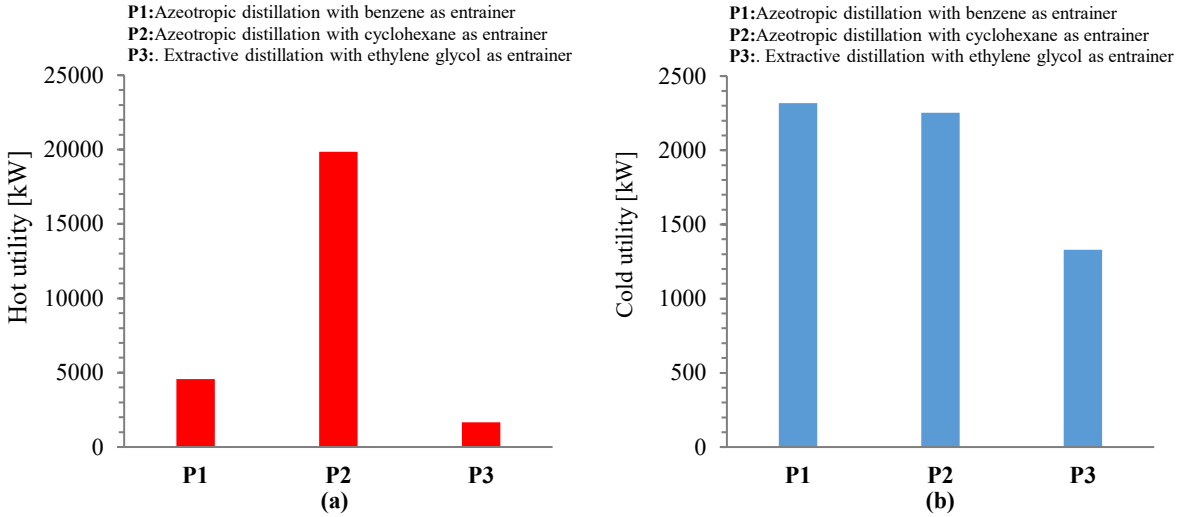

Fig. 5. Energy consumption in different processes: (a) hot utility, and (b) cold utility

According to the results, the process of ED with ethylene glycol as entrainer (P3) outperforms both AD processes with benzene and cyclohexane as entrainers in terms of separation and energy consumption. The FS method is used for heat integration of the ED process with ethylene glycol as entrainer (P4), shown in Fig. 6. As shown in Fig. 6, the feed is separated into two before entering the distillation column. A feed stream at 24.85°C enters tray 16, and the other stream is preheated by the bottom product of the first column and enters tray 19 of the first distillation column at 70°C. Ethylene glycol enters the first tray of the first column as an entrainer. The top product of the first column is pure ethanol, and the outlet of the heat exchanger is also split into two streams as the feed to the second column. One feed stream at the same outlet temperature of the heat exchanger (78.08°C) enters tray 22, and the other feed stream is preheated by the bottom liquid outlet of the second column and enters this column at 106.6°C on tray 24.

Table 3 reports the heat duties of condenser and reboiler in processes P3 and P4. where Q_{C1}, Q_{C2}, and E respectively represent the heat duties of condensers in the first and second columns and the cooler, and Q_{r1} and Q_{r2} respectively denote heat duties of reboilers in the first and second columns. Fig. 8 compares energy consumption in the ED process with ethylene glycol as an entrainer without (P3) and with (P4) heat integration.

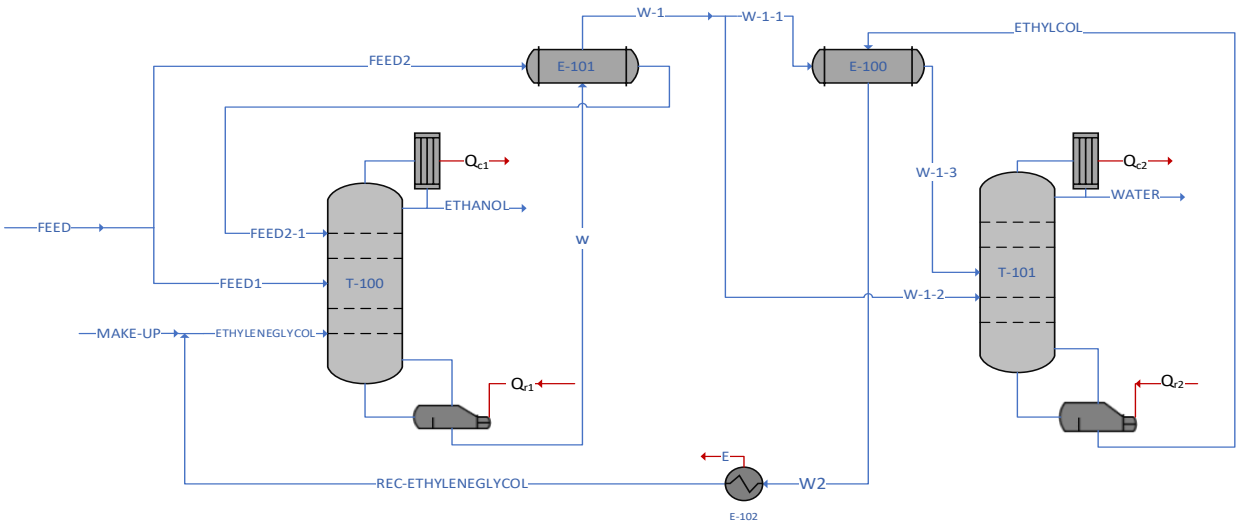

Fig. 6. The heat integrated ED process with ethylene glycol as entrainer

Table 4. Heat duties of condenser and reboiler in P3 and P4

Parameter	Value	
	Without heat integration [kW]	With heat integration [kW]
Q_{C1}	757.7	627.6
Q_{r1}	894.3	712.8
Q_{C2}	393.5	121.9
Q_{r2}	772.7	412
E	177.7	139

As shown in Fig. 7, the heat integrated ED process significantly reduced energy consumption and therefore is chosen as the best (optimal) process in this study in terms of energy consumption. The FS method considerably lowered the energy consumption of the ED process. The results indicated that the FS-ED process reduced hot and cold utility consumption by 49% and 50%, respectively, compared to the conventional ED process. Based on previous studies, the reduction in cold and hot utility consumption using the feed splitting method ranges between 25% and 60% [31, 34].

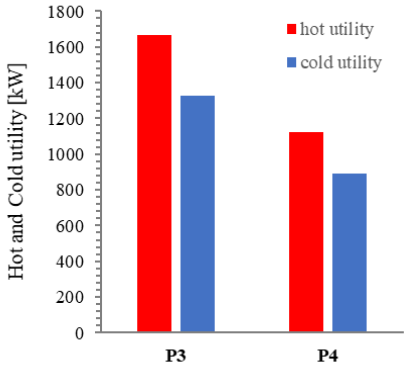

Fig. 7. Comparison of energy consumption in P3 and P4

To further investigation on ED process and heat integrated ED process, exergy loss and CO_2 emission were calculated. As shown in Fig. 8, as a result of the reduction in energy consumption, CO_2 emission is also affected and CO_2 emission decreased from 577.5 kg/h for the ED process without heat integration to 388.31 kg/h for the ED process with heat

integration. The results obtained for exergy loss, similar to the results of CO_2 emission, show the reduction of exergy loss of the ED process with heat integration compared to the ED process without heat integration. As shown in Fig. 8, exergy loss decreased from 1559 kW for the ED process without heat integration to 1456 kW for the ED process with heat integration. This reduction in exergy loss is mostly due to reduction of hot and cold utilities of the ED process with heat integration which leads to a reduction in exergy of heat. Table 5 shows exergy of heat of both process. Moreover, as a result of energy consumption reduction, operating cost of heat integrated process decreased from 208151$ to 139505$. Also, Due to the reduction in the reboiler heat duty, the amount of steam returning to the column decreases. Since the column diameter is calculated based on the gas phase, the column diameter is also reduced, capital cost of proposed process decreased by almost 84%.

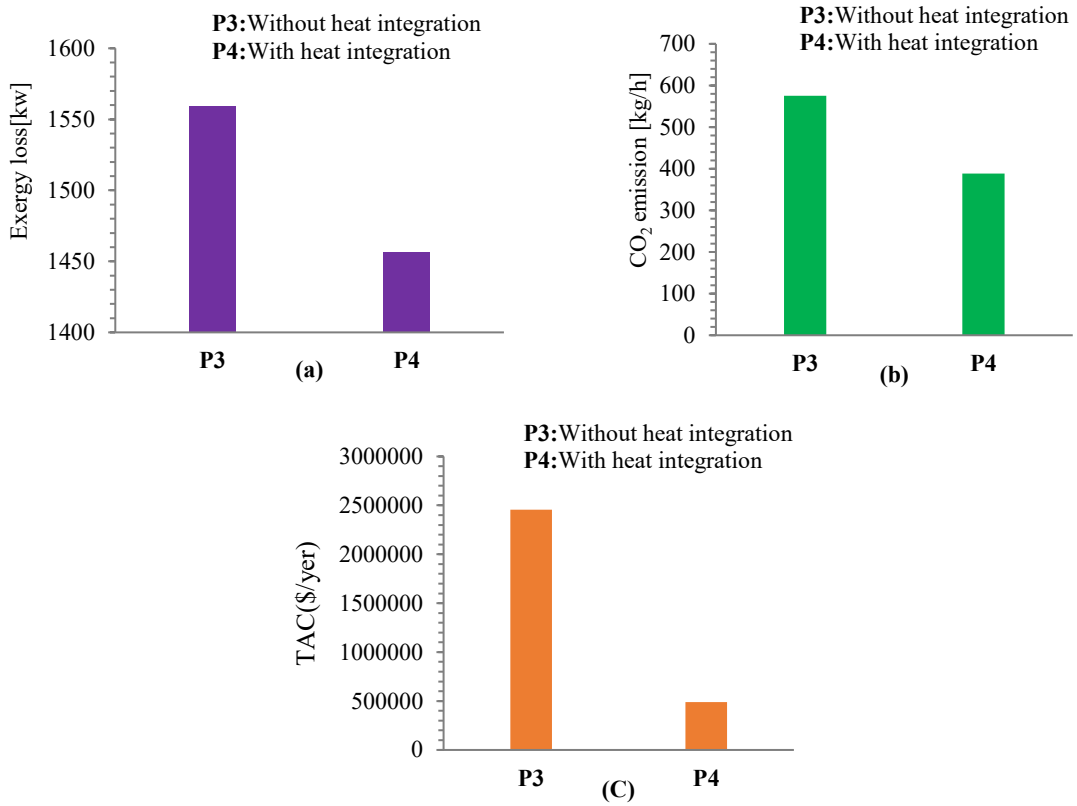

Fig. 8. Comparison of exergy loss and CO_2 emission in P3 and P4, (a) exergy loss, (b) CO_2 emission, and (c) TAC

Table 5. Exergy of heat of the P3 and P4

Energy stream	Exergy of heat [kW]	
	P3: Without heat integration	P4: With heat integration
Q_{C1}	90.75	74.05
Q_{r1}	192.24	153.22
Q_{C2}	76.10	23.81
Q_{r2}	271.90	144.97
E	27.68	21.65

6. Conclusion

This study investigated the minimum-boiling homogenous azeotropic separation of the water and ethanol mixture by four different processes: PSD, AD with benzene as entrainer, AD with cyclohexane as entrainer, and ED with ethylene glycol as entrainer. The azeotrope point of the water-ethanol mixture didn't significantly change with pressure; thus, the PSD process is not appropriate to separate this mixture. Hence, AD and ED processes were considered. The

procedures were precisely simulated before being compared in terms of separation and energy usage. Finally, the best (optimal) process was selected and heat integrated by the FS method. The most important outcomes are listed below.

1)The ED process with ethylene glycol as entrainer outperformed the AD processes with benzene and cyclohexane as entrainers in terms of separation and consumption of energy.

2) The ED process with ethylene glycol as entrainer increased ethanol purity by 6% relative to the two AD processes with benzene and cyclohexane as entrainers.

3) The ED process with ethylene glycol as entrainer reduced hot and cold duty consumption respectively by 92% and 43% relative to the AD process.

4) Hot and cold utilities contribute to a major part of process costs. According to the outcomes, the ED process with ethylene glycol as an entrainer significantly reduced process costs of water-ethanol separation relative to the AD process. In addition to economic issues, it is of environmental importance, as less hot and cold utility consumption is more compatible with the environment.

5) The FS method significantly reduced the energy consumption of the ED process. According to the results, the FS-ED process reduced hot and cold utility consumption, respectively, by 49 and 50% than the conventional ED process.

6) Due to the reduction of hot and cold utilities in heat integrated ED process, especially hot utility which is low pressure steam and medium pressure steam, CO_2 emission decreased approximately 32%.

7) Reduction in hot and cold utility of ED process with heat integration resulted in exergy loss reduction and it is due to the decreased exergy of heat in heat integrated ED process.

8) Compared with ED process without heat integration, ED process with heat integration's operating cost, capital cost, and TAC reduced by 32.97%, 84.46%, and 80.1%, respectively.

Nomenclature

Symbols

E	Cooler duty [kW]
Q_{c1}	Heat duty of first condenser [kW]
Q_{c2}	Heat duty of second condenser [kW]
Q_{r1}	Heat duty of first reboiler [kW]
Q_{r2}	Heat duty of second reboiler [kW]

Abbreviation

AD	Azeotropic distillation
ED	Extractive distillation
EOS	Equation of state
FS	Feed splitting
LLE	Liquid-liquid extraction
PSD	Pressure swing distillation

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