

Simulation Study for Separation of Acetone – Methanol Mixture by Extractive Distillation with Water as Entrainer via Heat Integration

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ABSTRACT

Separation of azeotropic mixtures often observed in pharmaceutical and specialty chemical industries, are of great practical and industrial interest which are impossible to separate by ordinary distillation, but may be effectively separated by adding a third component. The third component is usually called as entrainer /solvent. Present work is a simulation study on employing Extractive Distillation (ED) in separation of binary azeotropic mixture Acetone – Methanol with water as solvent. PRO/II simulation software is utilized for the said studies. Suitable thermodynamic model is chosen viz., UNIFAC, UNIQUAC and NRTL from VLE (Vapor Liquid Equilibrium) studies. Parametric studies viz., reflux ratio, solvent flow rate, feed tray location; number of trays is made in the present simulation to obtain the desired separation. Further, heat integration for the system is also carried out and results are compared with literature studies. Substantial reduction in energy consumption with and without heat integration is evaluated and reported.

KEY WORDS: Extractive Distillation, Solvent, Simulation, Parametric Study, PRO/II.

1. INTRODUCTION

Extractive distillation is a partial vaporization process, in the presence of a non-volatile and high boiling point separating mass agent that it is usually called solvent or separating agent, which is added to the azeotropic mixture to alter the relative volatility of the component without forming the azeotrope (Treybal, 1981, Black and Distler, 1972).

Generally, solvent is fed in the upper part of the column, above the feed stream and it remains in a considerable concentration in the liquid phase along the column. It is withdrawn as a bottom product with one of the components that are being separated and it is sent to a second regeneration column operated under vacuum. The most common solvents used in extractive distillation are glycols (Meirelles, 1992), glycerol (Lee and Pahl, 1985), gasoline (Chianese and Zinnamosca, 1990) and for the case of saline extractive distillation, acetate and inorganic salts such as: CaCl₂, AlCl₃, KNO₃, (CuNO₃)₂·3H₂O, Al(NO₃)₃·9H₂O, K₂CO₃ (Barba, 1985; Furter, 1992). Acetone is widely used in chemical industry as powerful solvent and as raw material in chemical synthesis of nail-polish removers, cement, lacquers and finishes, cleaners, paints, films and adhesives, cellulosic acetate films, plastics and coatings, de-waxing etc. Acetone forms azeotrope with methanol. At 1 atmosphere pressure, the azeotropic composition and temperature are 79.63 mol% and 328.3 K.

Several researchers explored the extractive distillation for the separation of acetone from methanol. Brondani (2015), studied industrial separation of benzene from a non-aromatic stream through extractive distillation with NFM as the solvent. Process was simulated using ASPEN Plus® with NRTL as thermodynamic model. Different sets of parameters for the NRTL model for many pairs of substances is identified. Pokhrel (2016), designed an Extractive Distillation (ED) column for separating maximum boiling azeotropic mixture of acetone-chloroform using DMSO as third component. Two different thermodynamic models, NRTL and UNIQUAC were explored for the study of overall process. PRO/II 8.2 is used to carry out simulation of different Case Studies considered. EDA Hosgor (2014), explored design and control of alternative distillation processes for separation of methanol–chloroform azeotropic mixture. The steady-state and dynamic simulations are carried out with Aspen Plus and Aspen Dynamics. Comparison studies are made with respect to homogeneous extractive distillation and pressure-swing distillation processes. Gil (2009), studied ED for acetone-methanol mixture using water as entrainer. Simulations are carried out in Aspen Plus software with UNIQUAC as thermodynamic model to calculate VLE for ternary system. Studies show the effect of feed molar ratio, reflux ratio, feed stage, feed solvent stage and solvent feed temperature on ED column design. Gil (2008), separated azeotropic mixture of ethanol-water by ED technique using ethylene glycol-calcium chloride mixture as third component. Aspen Plus® 11.1 is used for simulation with NRTL as the thermodynamic model. Parameter studies are made viz., feed molar ratio, molar reflux ratio, number of theoretical stages, feed stage, feed solvent stage and feed solvent temperature. Observations show that there is reduction in energy consumption when ethylene glycol-calcium chloride mixture is used as third component compared to conventional processes.

Yuan (2015), proposed Continuous Extractive Distillation (CED) to separate azeotropic mixtures of acetone–tetrahydrofuran, n-hexane–tetrahydrofuran, n-hexane–ethyl acetate and ethyl acetate–ethanol. Aspen Plus is used to carry out simulations. Results show that these azeotropic mixtures are separated successfully through solvent selection based on polarity principle and pseudo-binary VLE. Onkar (2015), explored the separation of THF-Water mixture with ED technique. Figueiredo (2011; 2015), explored the dehydration of ethanol-water mixture using ethylene glycol as solvent. Their main focus is to obtain the optimum condition for ED columns. Aspen plus is employed in simulation studies.

Shen (2015), carried out feasibility study of azeotropic mixture separation based on a topological analysis with thermodynamic knowledge of residue curve maps, uni volatility and uni distribution curves, and extractive profiles. You (2015), investigated multi objective genetic algorithm optimization for extractive distillation process of acetone - methanol minimum azeotropic mixture with heavy entrainer as water. The process includes the extractive and entrainer regeneration columns, and studies yielded the optimum energy cost objective function and total annual cost and maximum efficiency indicators. Chen (2015), proposed an enhanced method of separating pyridine-water system by n-propyl format or di isopropyl ether (DIPE) as solvent. A hybrid extraction–distillation system design is proposed for this separation task. From the study DIPE is found to be a more practically favorable solvent for reaching equilibrium in each stage for liquid–liquid separation. Luo (2014), explored two methods fully heat-integrated pressure-swing distillation and extractive distillation in the separation of binary minimum-boiling homogeneous azeotropic system of Di isopropyl ether (DIPE) and isopropyl alcohol (IPA) with 2-methoxyethanol as an entrainer. Luyben (2008; 2013) discussed the use of pressure-swing distillation for separating the azeotropic mixtures of acetone-chloroform and acetone-methanol respectively in comparison with ED. Results revealed that ED process is far better in both capital investment and energy consumption. Ivonne (2003), evaluated the use of entrainers for the separation of heterogenous binary azeotropic mixtures by extractive distillation. Process feasibility is assessed from the iso volatility Curves and the volatility order diagram. The study of azeotrope formation establishes the distillation task sequence using a batch rectifying column. The evaluation is carried out for the case study in the separation of acetonitrile – water mixture with hexylamine or butyl acetate.

2. EXPERIMENTAL WORK

Scope of work: Present work aims to carry out simulation studies for separating the azeotropic mixture of Acetone – Methanol by extractive distillation technique in PRO/II simulation software. The solvent used is water. Parametric analysis for achieving an improved separation viz heat integration is made, subsequently parameters namely optimum reflux ratio, feed tray location, solvent-to-feed ratio, number of trays in ED column are reported which lead to lower heat duties.

Thermodynamic Model: In the present study, the azeotropic mixture is subjected to Extractive Distillation (ED) to separate the components of the mixture into near pure components. This is achieved by selecting a suitable thermodynamic model from the data bank available in PRO/II software. The very usefulness of these models lies in the estimation of liquid phase properties.

The system Acetone –Methanol (SIMSCI Names: ACETONE - METHALC) is a blend of ketone and alcohol. Separation of this system is studied by Luyben (2008) with the extracting solvent as WATER (SIMSCI Name – H₂O). Present work attempts to make a parametric study followed by heat integration for this system to obtain the desired purity with lower heat duty in re-boiler.

As a first step, judgment of suitable thermodynamic model is made from isobaric VLE analysis. VLE data for the system Acetone – Methanol is generated viz. three models namely NRTL (Renon and Prausnitz, 1968), UNIQUAC (Abrams and Prausnitz, 1975) and UNIFAC (Fredenslund, 1975). The azeotropic temperature and composition at 101.325 kPa pressure is compared with literature data (Horsley, 1973). From the close monitoring of the data UNIFAC model is considered for the simulation study. Table 1 depicts the azeotropic data generated.

Table.1. Azeotropic data for Acetone (329.44K) – Methanol (337.85K) System

Model	P ^{az} (atm)	T ^{az} (K)	x ^{az} (mole %)
Horsley (1973)	1	328.50	0.7907
NRTL	1	328.48	0.79879
UNIQUAC	1	328.487	0.79877
UNIFAC	1	328.364	0.796353

Isobaric VLE data (T-x-y and x-y plots) is depicted in Fig.1(a) and Fig.1(b), which stands in close comparison with the literature.

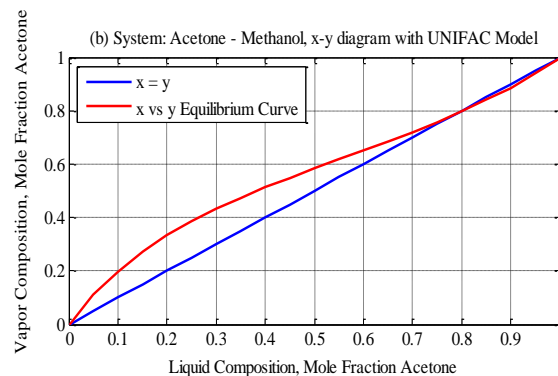
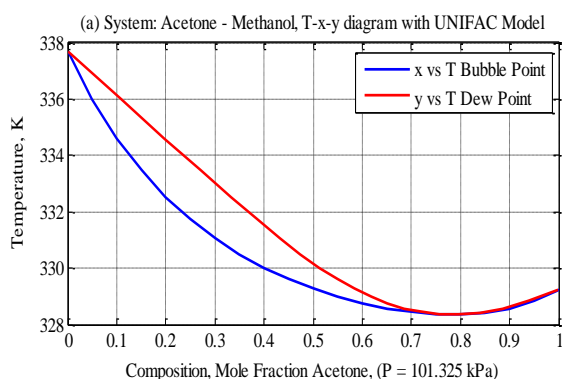


Figure 1(a). VLE for Acetone (329.44K) – Methanol (337.85K) by UNIFAC Model at P = 101.325 kPa

Figure 1(b). VLE for Acetone (329.44K) – Methanol (337.85K) by UNIFAC Model at P = 101.325 kPa

Simulation Study: Acetone – Methanol forms a minimum boiling azeotrope with azeotropic temperature of 328.3 K (55.2°C) and 79.63 % (mol) of Acetone at a pressure of 101.325 kPa. Separation of such constant-boiling mixture needs a third component either as an entrainer in azeotropic distillation or as a solvent in extractive distillation.

Focus of present simulation study is on Extractive Distillation technique to separate the mixture into pure components Acetone and Methanol using water as solvent. The simulation aims at reducing the heat duty of re-boiler in Extractive Distillation (ED) column. For this mixture, Residue Curve Map (RCM) analysis, simulation studies using PRO/II, parametric study with heat integration is made. Temperature and composition profiles are also observed.

Luyben (2008), revealed that water is a suitable solvent for separation of Acetone – Methanol mixture viz. extractive distillation. For the separation of 50% Acetone – 50% Methanol mixture by extractive distillation, knowledge of distillation sequencing is needed.

From Fig.2, RCM for the ternary system, Acetone – Methanol - Water, it is evident that Acetone can be obtained as pure component in the ED Column as top product and mixture of Methanol plus Water is obtained as bottom product. This bottom product can be separately sent to Recovery Column (RC) for recovery of Water.

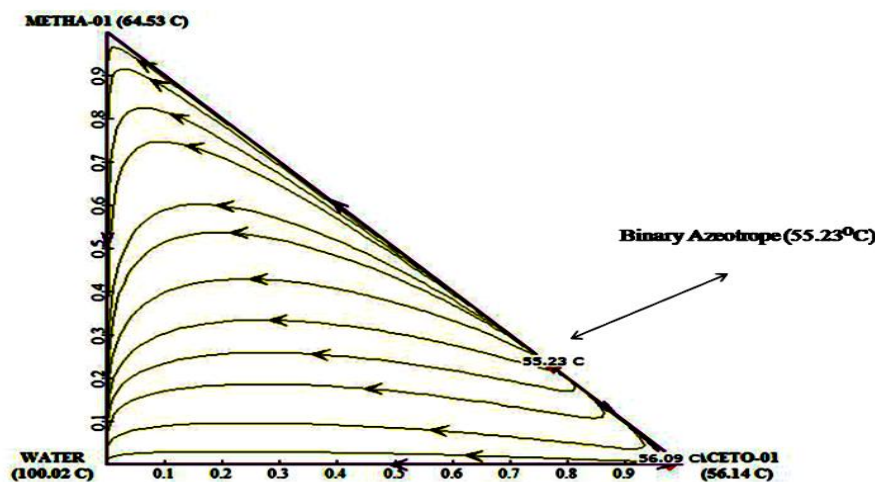


Figure 2. Residue Curves Map (RCM) for Acetone –Methanol – Water

Fig. 3, indicate information flow diagram for extractive distillation in separating Acetone – Methanol mixture for a feed flow of 540 k mol/hr entering at 303 K and a Reflux Ratio (RR) of 3.44 in ED column (Luyben, 2008). Acetone is obtained as a distillate and Water –Methanol mixture as bottom product. The bottoms are separated in a solvent Recovery Column (RC). Number of trays in ED Column are 57 and RC Column are 27. Feed to the ED column is entered at 40th tray and solvent at 25th tray.

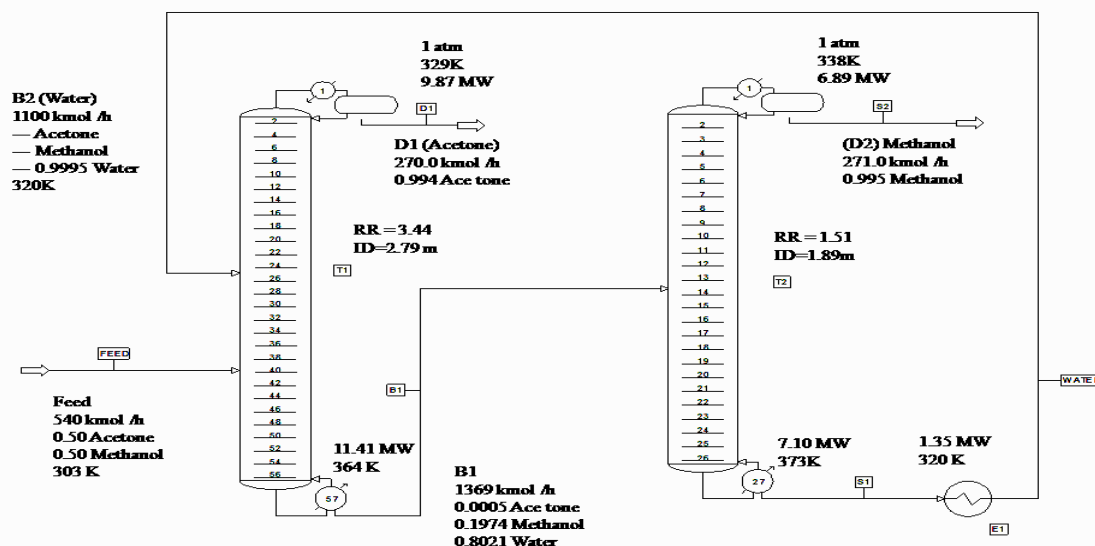


Figure 3. Information Flow Diagram in Simulation Study for Acetone –Methanol - Water (Luyben, 2008)

Fig. 4, indicate information flow diagram for the proposed simulation study. 540 k mol/hr feed of Acetone -Methanol mixture is pre-heated with the recovered solvent (Water) to its bubble point (329 K) and entered into the ED column at 48th tray. The solvent subsequently enters ED column at 35th tray. Reduction in Reflux Ratio (RR) from 3.44 (Luyben, 2008) to 2.78 (mole basis) further lowered heat duty of ED Column re-boiler keeping the same top product purity.

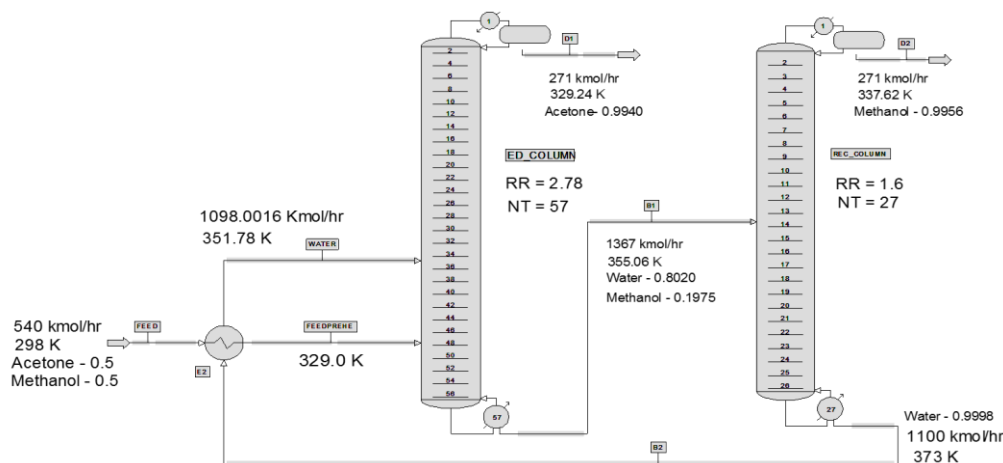


Figure 4. Information Flow Diagram in Proposed Simulation Study for Acetone –Methanol – Water with Heat Integration

Distillation Column data for simulation: Input data, equipment parameters and performance specifications for both ED and RC Columns are explained as under. Output from simulation is also mentioned.

ED Column @ 101.325 kPa: Equimolar mixture of acetone and methanol enters at a rate of 540.0 Kg-mol/hr after preheating to 329.00K with the recovered solvent from RC Column. Recovered Solvent enters at 348.63K temperature at the rate of 1098.00 kg-mol/hr. ED column consists of 57 trays with feed charging at 48th tray and solvent at 35th tray along with total condenser and kettle type re-boiler with negligible pressure drop. Reflux Ratio of 2.78 (on mole basis) and distillate flow rate of 271 kg-mol/hr are given as specifications. Top Product containing 99.40% Acetone at 329.24 K is obtained flowing at 271 kg-mol/hr and a ternary mixture at the rate of 1367.00 Kg-mol/hr with composition of Acetone 4.6098e-04, Methanol 0.1975 and Water 0.8020 at 355.06K is obtained as bottom product. This bottom product is then sent to rectification column for methanol - water separation.

RC Column @ 101.325 kPa: Bottom product from ED column is directly fed for rectification into Recovery Column (RC Column). This column operates with negligible pressure drop having 57 trays and feed entering into 27th tray. Total condenser and kettle type reboiler are used with UNIFAC as thermodynamic model. Reflux Ratio of 1.61(on mole basis) and distillate flow rate of all components 271 kg-mol/hr on a wet basis are given as specifications. Top product with 99.57% mole of methanol at 337.62K at the rate of 271.00 kg-mol/hr is obtained. 1096.00 Kg-mol/hr, liquid containing 99.99% water at temperature 373.11K is obtained as bottom product.

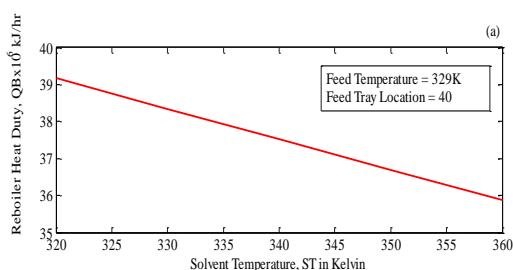
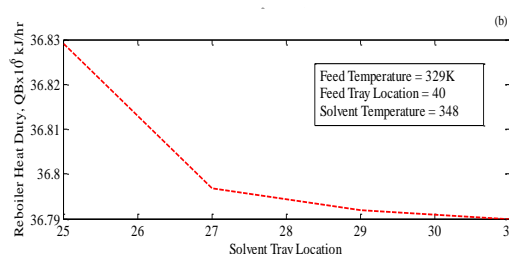
Parametric Study: Feed, Acetone – Methanol mixture is preheated to 329 K in the present study with the recovered water. A parametric study is made for this scenario to lower the re-boiler heat duty of the ED Column. All other parameters are kept constant as per literature (Luyben 2008). Table. 2, indicate the change in re-boiler heat duty and ED Column top product purity with variation in solvent temperature for feed temperature of 329 K and feed tray location at 40. From the trend observed (refer Fig 5(a)) solvent temperature is taken as 348 K for further simulation. With the feed temperature at 329 K, feed tray location at 40 and solvent temperature of 348 K, trend in re-boiler heat duty is verified for variation in solvent tray location (Table. 3 and Fig. 5(b)).

Table. 2. Variation in Solvent (water) Temperature

Run No.	1	2	3	4	5
Feed Temperature (K)	329	329	329	329	329
Feed Tray Location	40	40	40	40	40
Solvent Temperature (K)	320	330	340	350	360
Reboiler Heat Duty, Q_B (M*kJ/hr)	39.1727	38.3461	37.5189	36.6905	35.8666
x_D	0.99452	0.99445	0.99442	0.99439	0.9943

Table. 3. Variation in Solvent (water) Tray Location

Run No.	1	2	3	4
Feed Temperature (K)	329	329	329	329
Feed Tray Location	40	40	40	40
Solvent Tray Location	25	27	29	31
Solvent Temperature (K)	348	348	348	348
Reboiler Heat Duty, Q_B (M*kJ/hr)	36.829	36.7969	36.7919	36.7900
x_D	0.9947	0.9949	0.9952	0.9949

**Figure. 5(a). Re-boiler heat duty for Varied Solvent Temperature****Figure. 5(b). Re-boiler heat duty for Varied Solvent Tray Location**

With the above studies, feed tray location is varied to check whether the re-boiler heat duty is further lowered, but a slight change is seen. However, feed tray of 48 and solvent tray location of 31 is fixed. Table. 4 and Table. 5, represent the effect of feed tray location and solvent tray location on heat duty. Finally, tray locations for feed and solvent are arrived at 48 and 35 respectively. Fig. 5(c) and Fig. 5(d), show the trend in establishing the simulation.

Table. 4. Variation in Feed Tray Location

Run No.	1	2	3	4	5
Feed Temperature (K)	329	329	329	329	329
Feed Tray Location	42	44	46	48	50
Solvent Tray Location	31	31	31	31	31
Solvent Temperature (K)	348	348	348	348	348
Reboiler Heat Duty, Q_B (M*kJ/hr)	36.7870	36.7861	36.7858	36.7858	36.7904
x_D	0.9956	0.9958	0.9959	0.9959	0.9948

Table. 5. Variation in Solvent (water) Tray Location fixing Feed tray @ 48

Run No.	1	2	3	4	5
Feed Temperature (K)	329	329	329	329	329
Feed Tray Location	48	48	48	48	48
Solvent Tray Location	31	33	35	37	39
Reboiler Heat Duty, Q_B (M*kJ/hr)	36.7858	36.7829	36.7808	36.7797	36.7809
x_D	0.9959	0.9959	0.9959	0.9957	0.9950

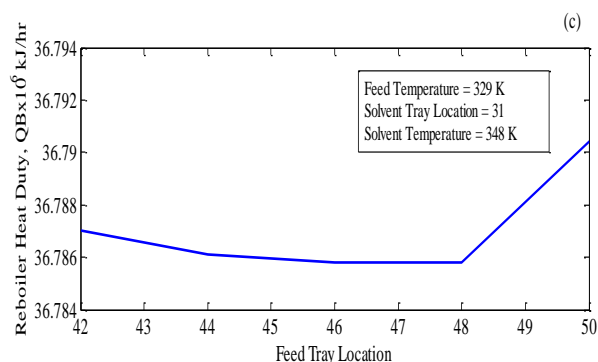


Figure. 5(c). Re-boiler heat duty for Varied Feed Tray Location

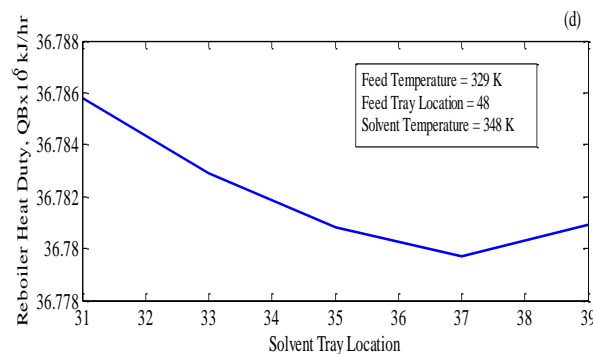


Figure. 5(d). Re-boiler heat duty for Varied Solvent Tray Location

Table. 6, shows an attempt made by simulation for the decrease in re-boiler heat duty with respect to change in solvent flow rate. This instead may lead to high consumption of solvent flow rate. Again a decrease in reflux ratio in ED Column from 3.44 to 2.78 displayed a drastic decrease in the re-boiler heat duty with the desired Acetone composition of 99.4 mol % Acetone (Table. 7). The parametric study with respect to reflux ratio is represented pictorially Fig. 6(a) and Fig. 6(b).

Table.6. Variation in Solvent Flow Rate fixing Feed tray @ 48, Feed Temp @ 329K and Solvent Tray @35, Solvent (water) Temperature

Run No.	Feed Tray and Temperature	Solvent Tray and Temperature	Solvent Flow Rate (kmol/hr)	Reboiler Heat Duty, Q_B (M*kJ/hr)	x_D (%)
1	48 329 K	35 348 K	1100	36.8324	0.9959
2			1200	36.7422	0.996
3			1300	36.7076	0.996
4			1400	36.6765	0.996
5			1500	36.6484	0.9961
6			1600	36.6218	0.9961
7			1700	36.5935	0.995
8			2000	34.1729	0.99

Table. 7. Variation in Reflux Ratio

Run No.	Feed Tray and Temp (K)	Solvent Tray and Temp (K)	Solvent Flow Rate (kmol/hr)	Reflux Ratio	Reboiler Heat Duty, Q_B (M*kJ/hr)	Condenser Heat Duty, Q_C (M*kJ/hr)	x_D
1	48 329 K	35 348 K	1100	3.44	35.365	-35.113	0.99582
2				3.39	34.971	-34.719	0.99577
3				3.34	34.577	-34.325	0.9957
4				3.29	34.182	-33.931	0.99563
5				3.24	33.788	-33.788	0.99553
6				3.19	33.394	-33.144	0.99542
7				3.14	33.000	-32.751	0.9953
8				3.09	32.606	-32.357	0.99516
9				2.99	31.819	-31.571	0.99501
10				2.94	31.425	-31.178	0.99466
11				2.89	31.031	-30.785	0.99447
12				2.84	30.637	-30.392	0.99427
13				2.79	30.243	-29.999	0.99404
14				2.74	29.85	-29.607	0.99375
15				2.69	29.456	-29.214	0.99339
16				2.64	29.062	-28.821	0.99297
17				2.59	28.669	-28.429	0.99247

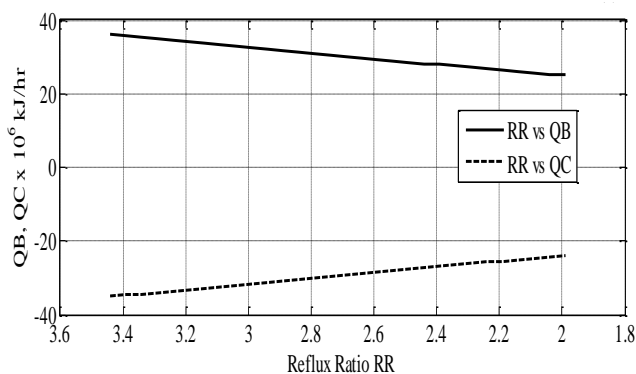


Figure 6(a). Reflux Ratio vs QB, QC

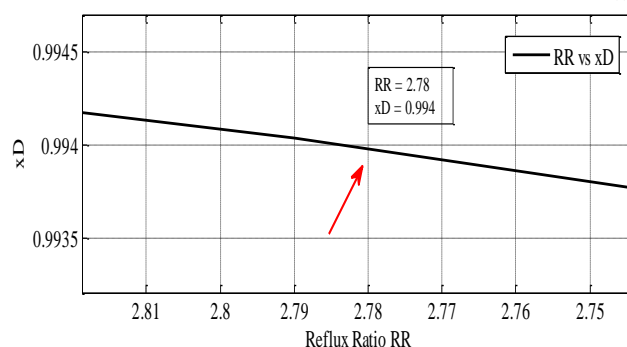


Figure 6(b). Reflux Ratio vs xD

3. RESULTS AND DISCUSSION

The results comparing with that of the literature are presented in Table 8.

Table 8. Comparison of Proposed Simulation to literature for Acetone – Methanol System

Parameter	Luyben (2008)		Proposed Simulation with heat integration	
	ED Column	Recovery Column	ED Column	Recovery Column
F (Feed) kmole/hr	540	1369	540	1369
x_{Acetone}	270	0.6845	270	0.6845
x_{Methanol}	270	270.24	270	270.24
x_{Water}	-	1098.07	-	1098.07
Feed tray Location	40	14	48	14
Solvent tray Location	25	-	35	-
Feed (Solvent), kg mole/hr	1100	1098.07	1100	1098.07
$x_{\text{Solvent (DMSO)}}$	1	0.8021	1	0.8021
Distillate, kg mole/hr	270.0	271	270.0	271
Reflux Ratio	3.44	1.51	2.78	1.51
N_T	57	25	57	25
x_D	99.40	99.50	99.40	99.57
Q_B, M kJ/hr	41.076	25.56	30.1646	25.9653
Q_C, M KJ/hr	-35.532	-24.804	-29.9207	-24.9041

DISCUSSION

From Table. 7, it can be observed that employing heat integration in extractive distillation technique to separate Acetone – Methanol System using water as solvent resulted in lowering the ED Column re-boiler heat duty (Q_B) from 41.076 M kJ/hr to 30.166 MkJ/hr. During the simulation study to achieve the above result parameters viz., feed and solvent tray locations, reflux ratio variation have also been made.

4. CONCLUSION

To separate Acetone – Methanol System, (i) water is used as solvent (ii) UNIQUAC thermodynamic model is employed (iii) after rigorous simulation, with the feed to ED Column pre-heated to its bubble point by the recovered solvent, re-boiler heat duty is observed to be reduced from 41.06 M*kJ/hr to 30.66 M*kJ/hr with the parameters fixed as $N_T = 57$, Feed Tray Location = 48, Solvent Tray Location = 35, Reflux Ratio in ED Column = 2.78 (as against the literature value of 3.44).

REFERENCES

- Abrams DS, Prausnitz, JM, Statistical Thermodynamics of Mixtures: A New Expression for the Excess Gibbs Free Energy of Partly or Completely Miscible Systems, AIChE J, 21, 1975, 116-128.
- Barba D, Brandani V, Giacomo G, Hyperazeotropic, Ethanol salted - out by extractive distillation, Theoretical evaluation and experimental check, Chemical Engineering Science, 40, 1985, 2287 – 2292.
- Black C, Distler DE, Dehydration of aqueous ethanol mixtures by extractive distillation, In: R.F. Gould, Extractive and azeotropic distillation, Advances in Chemistry Series, 115, 1972, 1-15.

Brondani LB, Flores GB, Soares RP, Modeling and simulation of a benzene recovery process by extractive distillation, *Brazilian Journal of Chemical Engineering*, 32 (1), 2015, 283-291.

Chen Yi-Chun, Li Kung-Ling, Chen Cheng-Liang, Chien I-Lung, Design and Control of a Hybrid Extraction–Distillation System for the Separation of Pyridine and Water, *Ind. Eng. Chem.*, 54, 2015, 7715–7727.

Chianese A, Zinamosca F, Ethanol dehydration by azeotropic distillation with mixed solvent entrainer, *The Chemical Engineering Journal*, 43, 1990, 59– 65.

Eda Hosgor, Tugba Kucuk, Ilayda N. Oksal, Devrim B. Kaymak, Design and Control of Distillation Processes for Methanol-Chloroform Separation. *Computers and Chemical Engineering*, 67, 2014, 166-177.

Figueiredo MF, Brito KD, Ramos WB, Vasconcelos LGS, Brito RP, Optimization of the Design and Operation of Extractive Distillation Processes, *Separation Science and Technology*, 50, 2015, 2238–2247.

Figueiredo MF, Guedes BP, Optimal design of extractive distillation columns-A systematic procedure using a process simulator, *Chemical Engineering Research and Design*, 89, 2011, 341–346.

Fredenslund A, Jones RL, Prausnitz JM, Group-Contribution estimation of activity coefficients in non-ideal liquid mixtures, *AIChE J*, 21 (61), 1975, 1086–1099.

Furter WF, Extractive Distillation by Salt Effect, *Chemical Engineering Communications*, 116 (1), 1992, 35-40.

Gil ID, Diana C Botla, Pablo Ortiz, Oscar F Sanchez, Extractive Distillation of Acetone/Methanol Mixture Using Water as Entrainer, *Ind. Eng. Chem. Res.*, 48, 2009, 4858–4865.

Gil ID, Uyazan A, Aguilar J, Rodríguez G, Caicedo L, Separation of ethanol-water mixtures by extractive distillation with salt and solvent as entrainer: Process simulation, *Brazilian Journal of Chemical Engineering*, 25 (1), 2008, 207-215.

Horsley LH, Azeotropic Data III, *Advances in Chemistry Series No. 116*, 1973, American Chemical Society, Washington, DC.

Ivonne Rodriguez Donis, Jhoany Acosta Esquijarosa, Vincent Gerbaud, Xavier Joulia, Heterogeneous Batch-Extractive Distillation of Minimum Boiling Azeotropic Mixtures, *AIChE Journal*, 49, 2003, 3074-3083.

Lee Fu Ming, Pahl Robert H, Solvent screening study and conceptual extractive distillation process to produce anhydrous ethanol from fermentation broth, *Ind. Eng. Chem. Process Des. Dev.*, 24 (1), 1985, 168–172.

Luo H, Liang K, Li W, Li Y, Xia M, Xu C, Comparison of Pressure-Swing Distillation and Extractive Distillation Methods for Isopropyl Alcohol/Diisopropyl Ether Separation, *Ind. Eng. Chem. Res.*, 53, 2014, 15167–15182.

Luyben WL, Comparison of Extractive Distillation and Pressure-Swing Distillation for Acetone-Methanol Separation, *Ind. Eng. Chem. Res.*, 47 2008, 2696-2707.

Luyben WL, Comparison of extractive distillation and pressure-swing distillation for acetone/chloroform separation, *Computers and Chemical Engineering*, 50, 2013, 1-7.

Meirelles A, Weiss S, Herfurth H, Ethanol dehydration by extractive distillation, *Journal of Chemical Technology & Biotechnology*, 53, 1992, 181–188.

Onkar AD, Sathe VS, Mahajan Yogesh S, THF-Water mixture separation using Extractive and Azeotropic distillation: Review, Modeling and simulation in Aspen plus, *International Journal of Science and Engineering*, 3, 2015, 10-16.

Pokhrel Manish, Asante Daniel Owusu, Jung Ho Cho, Separation of Acetone-chloroform maximum boiling azeotrope using Dimethyl sulfoxide, *International Journal of Applied Engineering Research*, 11, 2016, 7772-7775.

Renon H, Prausnitz JM, Local Composition in Thermodynamic Excess Functions for Liquid Mixtures, *AIChE J.*, 14, 1968, 135-144.

Shen WF, Benyounes H, Song J, Thermodynamic Topological Analysis of Extractive Distillation of Maximum Boiling Azeotropes, *Brazilian Journal of Chemical Engineering*, 32, 2015, 957 – 966.

Treybal E Robert, *Mass – Transfer Operations*, Third Ed., McGraw – Hill Book Company, Singapore, 1981, 457.

You Xinqiang, Ivonne Rodriguez-Donis, Vincent Gerbaud, Investigation of Separation Efficiency Indicator for the Optimization of the Acetone–Methanol Extractive Distillation with Water, *Ind. Eng. Chem. Res.*, 54, 2015, 10863–10875.

Yuan Shenfeng, Zou Cancan, Chen Zherong, Yang Wendong, Study On the Separation of Binary Azeotropic Mixtures by Continuous Extractive Distillation, *Chemical Engineering Research and Design*, 9 (3), 2015, 113–119.