

Original Article: Minimization of Energy Consumption in the Condenser of Acetone Recovery Process from Air


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ABSTRACT

Today, the recovery of acetone as an expensive and widely used solvent in the production units of methyl methacrylate, bisphenol A, MTBA, plastics and synthetic fibers is very important. One of the most common recovery methods is the absorption process (solvent separation), which is sometimes used in combination with surface absorption or membrane separation due to its cheapness and availability of knowledge. In this research, it was simulated with Aspen Plus V 7.2 software and the experimental rules of absorption and distillation columns, and with the Levi-Benn model to recover 99.5% of acetone from its mixture with air (10.3-687Mol/hour), the optimal conditions for the absorption tower, 20 equilibrium stages were obtained at atmospheric pressure with effluent and return gas from the separation section. In this situation, the molar ratio of the reflux of the distillation tower is 20.9 with the minimum energy consumption in the partial condenser at the rate of 2.3 Kw and 20 equilibrium stages.

Introduction

In the highly consumed market of petroleum and polymer products, simultaneously with the increase in production, in order to excel in the competition and gain more profit, it has become necessary to optimize the process with the aim of reducing energy costs and

product production. Reducing fuel consumption [1-3], preventing the wastage of raw materials or expensive products, maintenance and inspection of valuable equipment are among these measures. Unfortunately [4-6], currently, despite the increase in demand, acetone in Iran is only imported, but due to the fact that a number of

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acetone production projects will be used in the coming years, the need for more study and research in this direction seems necessary [7].

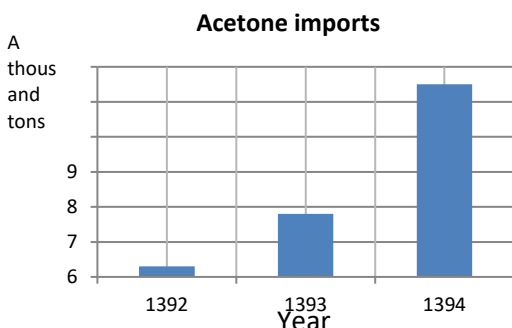


Figure 1. Statistics of Iran Customs Department

Recovering from a solution or converting non-extractable acetone to other valuable materials is the subject of various researches, including the work of Yu and his colleagues entitled "Optimization of acetone-methanol extractive distillation column with the aim of achieving minimum energy consumption for product flow" [8-10]. Purity percentage and acetone recovery rate are assumed to be certain values and the total annual cost is calculated and minimized by changing the influencing parameters based on the method mentioned in this reference [11-13]. In the production process, phenol acetone is produced as a side product, which can be avoided by optimizing the process. On the other hand, environmental requirements have imposed huge costs on industrial owners, which has caused other methods such as condensation, biological purification and reactive separation and, if necessary, incineration, to be used to prevent the release of more air pollutants such as solvents or volatile compounds [14].

Alarming statistics of the US Environmental Protection Agency show that the total release of volatile compounds including acetone was 31 million tons in 1975. Although acetone does not contribute significantly to air pollution, its negative impact on the environment cannot be ignored [15]. Condensation of acetone is carried out at a

constant atmospheric pressure with minimum diffusion in the environment.

The air stream containing acetone enters the condenser (its approach temperature is 10°C) and leaves it with a temperature of -100°C. Then I enter an adiabatic flash, and liquid acetone is obtained from the bottom of the drum. In this research with Aspen Plus, we show that in order to obtain liquid acetone with a purity of 99.3 and a recovery of more than 99.5% from the mixture of air-acetone 10.3-687Mol/hour, a condenser with a power of 1 KW is needed, but the cost of the cycle condensation is relatively high [1]. Also, the results of the simulation with Aspen Plus prove that by decreasing the temperature from -70 to -100°C, although the recovery increases, the purity of the acetone liquid coming out of the drum gradually decreases. Graphical sensitivity analysis shows that a further decrease in temperature increases the concentration of liquid air in the output product [2]. A quick estimation way to determine the condensation temperature is to use Antoine's relationship assuming the homogeneity of the mixture of air and acetone:

$$P_T \cdot y_i = \gamma_i \cdot P_i^\circ \cdot x_i \quad \gamma_i = 1, P_T = 1 \text{ Atm},$$

$$x_i = 1 \quad \ln P_i^\circ = A - B/(T + C)$$

The molar fraction of acetone in the outlet gas flow is calculated and the constants of Antoine's relation are obtained from Perry's handbook:

$$A = 14.55, B = 2940.5, C = 237.12,$$

$$y_i = 0.000075 \quad \text{We will have: } T = 88 \text{ }^\circ\text{C}$$

Therefore, it is practically impossible to separate acetone from air by increasing the pressure at ambient temperature [7].

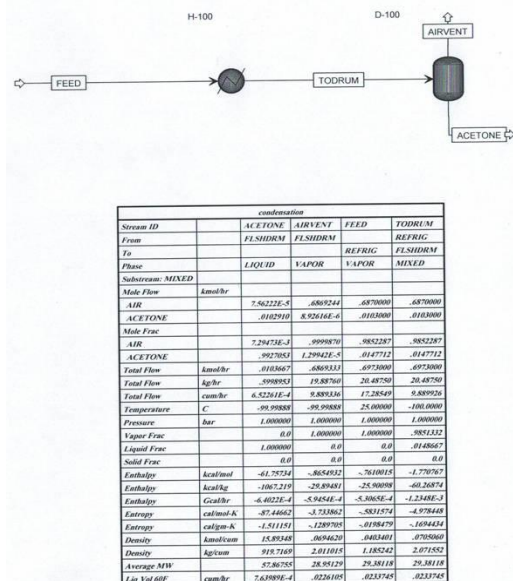


Figure 2. Acetone condensation process flow diagram

Another method is membrane penetration, which is very suitable for separating hydrocarbons in small amounts and high concentrations [30-32]. At first, polymer membranes such as silicone rubber halo fibers were used, which were not well received due to low solvent permeability. Ceramic membranes also have low thermal and chemical resistance [3].

Finally, alumina-gamma membranes modified with La_2O_3 were used in the processes of recovering volatile compounds such as acetone from nitrogen due to their high efficiency [6]. Hinchliffe and Porter have made a good comparison between distillation and membrane for the separation of water and acetone, which is described in table 1. They showed that the cost of the membrane with the total cost of separation equal to the

distillation of the aforementioned system is 1300 pounds per square meter and for low water concentrations (below 10%) the membrane technology is more economical. Today, new methods are used that are more compatible with the environment. Biological purification is a process that is recycled with the help of bacteria in a bio scrubber or bio filter [7].

For example, Zeminsky and his colleagues biochemically decomposed acetone in a laboratory bio filter fluidized bed consisting of two columns, one containing coconut fibers and the other plexi glass pieces. The concentration of acetone fluctuated from 0.3 to 2.5 g/m^3 in the air flow with a volume flow between 0.1 and 0.3 m^3/hour . In such conditions, the removal capacity of acetone was 95.8 m^3/hour with an air flow speed of 36 m/hour . Surface adsorption is an exothermic process in which the concentration of acetone increases due to the passage of air-acetone from a column containing solid materials with an active contact surface (such as activated carbon) on molecular active sites, then it is separated from it by the surface removal process [3].

Surface adsorption mechanisms are divided into two categories: Reversible physical and chemical, which form a chemical bond and can be irreversible. In this process, the phenomenon of permeation from the gas layer determines the kinetics of surface absorption. The absorption capacity of the adsorbent surface is a function of the concentration of the separating substance in the carrier gas at constant temperature and equilibrium conditions, which are formulated in the form of isotherm functions. Table 2 briefly describes some of them [13].

Table 1. Comparison of the cost of distillation and membrane

Feed rate = 7.2 tone/h		Product specification = 99% Product Purity and Recovery of Acetone	
		Molecular sieve membrane design	
		Mode = Pervaporation Inlet temperature = 708 C Feed side pressure = 1.013 bar a (atmospheric pressure) Permeate pressure = 8 m bar a Condensing temperature = 408 C Heating/condensing load = 871 kW Power for vacuum system = 1.1 kW Membrane area = 105 m ²	
	85 £k/year	Capital charge for vacuum system	13 £k/year
	296 £k/year	132.87 £k/year	
	£8.28/ton acetone		

The American Environmental Protection Agency states that this method is suitable when the concentration of acetone in the air is more than 2500 ppm. The following tables show the costs of surface adsorption in combination with other processes.

Table 2. Annual installation and operating costs of the surface adsorption process

ESTIMATED INSTALLED COSTS ^a OF ADSORPTION SYSTEMS			
Adsorber capacity, SCFM – Based on 25% lower explosive limit	1,000	10,000	20,000
With solvent recovery, \$	74,000	162,300	280,000
With thermal incineration/ no heat recovery, \$	89,500	202,000	344,000
With thermal incineration/ primary heat recovery, \$	101,500	255,000	431,000

^aCost data updated to first quarter of 1975.

ESTIMATED ANNUAL OPERATING COSTS ^a OF ADSORPTION SYSTEMS			
Adsorber capacity, SCFM – Based on 25% lower explosive limit	1,000	10,000	20,000
With solvent recovery, \$/yr	13,200	10,479 ^b	37,200 ^b
With thermal incineration/ no heat recovery, \$/yr	23,400	64,300	123,200
With thermal incineration/ primary heat recovery, \$/yr	25,600	82,000	141,600

^aCost data updated to first quarter of 1975.

^bIndicates a savings.

Table 3. Applications of acetone separation methods from air

Process	Basic principle	Remarks
Incineration	The solvents are completely destroyed by: thermal oxidation (>10 g/m ³) catalytic oxidation (3-10 g/m ³)	Only waste heat recovery possible
Condensation	High concentration of solvents (>50 g/m ³) are lowered by direct or indirect condensation with refrigerated condensers	- Effective in reducing heavy emissions, - good quality of the recovered solvents, - difficult to achieve low discharge levels
Absorption	Scrubbers using non-volatile organics as scrubbing medium for the solvent removal. Desorption of the spent scrubbing fluid	Advantageous if: - scrubbing fluid can be reused directly without desorption of the absorbed solvent - scrubbing with the subcooled solvent itself possible
Membrane permeation	Membrane permeation processes are principle suited for small volumes with high concentrations e.g. to reduce hydrocarbon concentration from several hundred g/m ³ down to 5-10 g/m ³	To meet the environmental protection requirements a combination with other cleaning processes (e.g. activated carbon adsorber) is necessary
Biological treatment	Biological decomposition by means of bacteria and using of bio-scrubber or bio-filter	Sensitive against temperature and solvent concentration
Adsorption	Adsorption of the solvents by passing the waste air through an adsorber filled with activated carbon. Solvent recovery by steam or hot gas desorption	By steam desorption the resulting desorbate has to be separated in a water and an organic phase using a gravity separator (or a rectification step)

Darabi and his colleagues optimized the energy production process of acetone with pinch technology. The result of their work shows that by using pinch, the consumption of steam and cooling water can be reduced and the size of heat exchangers can be reduced. In the burning method, the solvent is passed through thermal or catalytic oxidation by passing through a fluidized bed at a temperature between 1000 and 500°C to be burned and the products of water and carbon dioxide are obtained. One of the advantages of the catalytic oxidation of acetone is the low consumption of fuel compounds for burning acetone compared to thermal oxidation, which is carried out at a temperature between 1200

and 800°C, but there is a significant fixed cost and the possibility of the catalyst being poisoned by hydrocarbons. This method is recommended when there is a mixture of acetone with flammable hydrocarbon solvents. Ming Chi Ye et al.'s research on the thermal oxidation of the mixture of acetone, ethanol and toluene also confirms this [14].

Neto et al showed that inlet temperature, space velocity, type of material and concentration of volatile compounds have a significant effect on process efficiency and the ease of oxidation of acetone is less than ethanol and more than toluene, ethyl acetate and normal hexane [17]. Solvent burning method is passed through thermal or catalytic oxidation by passing through a flooded bed at a temperature between 1000 and 500 °C to be burned and the products of water and carbon dioxide are obtained [10].

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Neto et al showed that inlet temperature, space velocity, type of material and concentration of volatile compounds have a significant effect on process efficiency and the ease of oxidation of acetone is lower than ethanol and higher than toluene, ethyl acetate and normal hexane. In this research, due to the advanced process knowledge of the above methods, the high cost of refrigeration devices, and easy access to the information sources of absorption units (such as reference 3, which explains the methods of rapid estimation of the amount of water and the number of stages of absorption theory), acetone recovery through absorption and distillation is

simulated and optimized. Before describing this activity, it is appropriate to mention some of the conducted researches along with a brief description of the process [14].

As shown in figure 3, acetone and air are separated from each other by a cheap solvent such as water, then the solution rich in acetone is sent to the distillation tower to obtain acetone as a pure product, and if the effluent is expensive, it is absorbed into the tower. To separate acetone from air, Marki and his colleagues used an absorption column filled with water solvent and passing the acetone-rich solution through a permeable membrane. The results of their research show that the absorption process and the membrane are far more environmentally friendly and reduce energy costs in comparison [15]. It is more acceptable with the common method of absorption and distillation. They have used Sherwood's number to determine the amount of absorption as follows:

$$ShG = (KG \cdot deq) / DG = a \cdot (Re G)^b \cdot (ScG)^c$$

While Stanley Wallace uses the absorption factor to calculate the intensity of currents. Zhishan and his colleagues also simulated and optimized a process with 230 KMol/hour of water with Aspen Plus software. Jiang also uses Aspen Plus software to provide good results for the specifications of the input feed to the distillation tower with the help of sensitivity analysis [16]. He has shown the effect of additives such as surface activators, dispersants and salts on increasing absorption [17].

Materials and methods

First, a simple design of the absorption process to separate acetone from air is simulated with Aspen Plus, which includes two absorption and separation columns, as can be seen in figure 3. Then, based on the Levy-Ben model, the operational parameters are optimized and in this condition, assuming a certain value for the purity of the product, the return flows to the absorption tower are established to minimize the energy costs [8].

Design of absorption and removal units

Estimation of physical properties: In this research, to determine the specifications of the absorption tower by Aspen Plus, Van Laar's equation of state (or the two-parameter Margols) proposed by Lichten-Belt et al. was used and the number of stages of the absorption tower theory was set to 10 as an initial guess [9].

Bubble point calculations: Although the database of Aspen Plus V 7.2 software is used in this research, it should be noted that the advantage of reference 3 empirical rules compared to Aspen Plus software calculations with DSTWU or Distl block is that uses bubble point method with equal chemical potential of component i : $Y_i P_{oi} = X_i P_{oi}$

At low pressures, the fugacity coefficient is assumed to be equal to one and the ideal liquid phase is assumed to obtain the unknowns by Raphson's convergence method. This eliminates the error caused by the constant assumption of the relative volatility (α) that exists in the quick estimation methods of Aspen Plus software with DSTWU or Distl. The equation $Y_i = \alpha x_i / \sum \alpha x_i$ shows that if α is not constant, the calculation error will be significant [10].

Secondly, DSTWU and Distl blocks can only have one input, so calculating the Absorption Tower Shortcut is not possible in this way. In contrast, the bubble point method requires a long time to calculate the entire column. In such a situation, the use of Aspen Plus software at the same time as experimental rules and printed information in a specific field will be preferred. Kaghazchi and Rahbar have presented a formula for calculating the minimum value with mathematical modeling for the stone absorption tower, which is in agreement with the Douglas reference 3 method:

$$L_{min} = V_{N+1} [Y_{N+1} - Y_1] / [Y_{N+1}/K_N - x_0], X_0 = 0 \quad L_{min} = V_{N+1} \cdot K_N$$

The mole fraction of the dissolved component KN

Douglas proposed the simplified equations of Kremser to determine the stages of absorption theory and the relationship $L = 1.4 Mg$ to calculate the amount of water in isothermal absorption. While the following steps are used to model the absorption tower in Aspen Plus. First, the pressure of the absorption towers in both methods is one atmosphere and its changes along the length of the tower are ignored, and we ignore the return flows for simplicity. Then, by improving the process parameters, the impact of each is checked.

The non-ideality of VLE: The ideality of the liquid phase occurs only when the components are completely similar [11].

The minimum number of steps: For the absorption tower, as shown in figure 2, the minimum amount of solvent must be obtained first, then the minimum number of trays is obtained using the equilibrium curve and work line or reference relations 3.

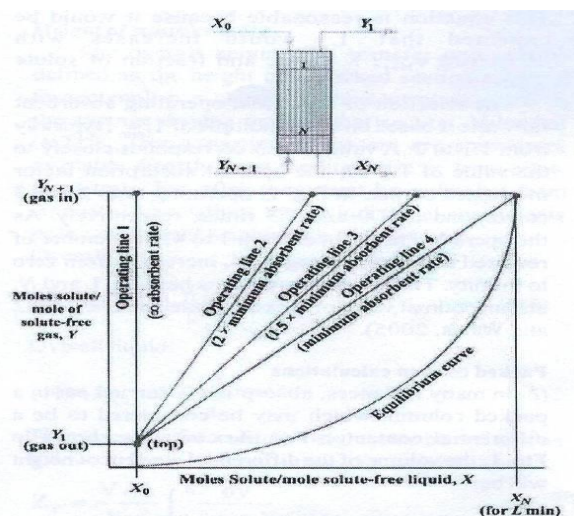


Figure 3. Equilibrium curve and working line of attraction tower

Also, from the Fenske equation, assuming that α is constant, the minimum number of steps is obtained as follows:

$$P_{n+1} = \frac{\log \left[\frac{(x_{d,lk}/x_{d,hk})(x_{b,hk}/x_{b,lk})}{\alpha_{lk,hk}} \right]}{\log \alpha_{lk,hk}}$$

lk : Light key hk : Heavy key

d : Top section b : Bottom section

The mole fraction values of acetone and water at the top and bottom of the absorption and distillation tower are obtained from the mass balance or reference 3. To calculate the relative volatility, use the geometric mean of the relative volatility of the first and last tray, which is in good agreement with the simulation results with Aspen Plus, or use a simulation with simple assumptions, for each tray we have: $\alpha = K_{lk}/K_{hk}$

In order to determine the minimum number of stages of the distillation tower, we simulate it with the DSTWU block in Aspen Plus software, which uses Winn's shortcut methods to calculate the minimum number of stages and Underwood's to calculate the minimum reflux.

Considering that the use of this method to achieve 99% purity of acetone must be done with an error guess for the amounts of water and air in the products at the top of the tower and it requires spending time. Therefore, we use the RadFrac block with the design and vary commands for the purity of the acetone product and the boiling heat load. In such conditions, the temperature profile and the amount of acetone in the effluent is a good guide to reach the appropriate process parameters.

Process and economic optimization

So far, dozens of models and algorithms have been proposed with the aim of reducing the annual costs of an industrial separation unit, which often focus on reducing energy costs and reducing the size of main equipment such as distillation column or heat exchanger. Therefore, it is very effective to improve the parameters that cause less thermal load in the boiling and cooling or to increase the separation coefficients, and the following

things were investigated and tested in this research.

Vapor pressure: The ease or difficulty of separating the components of a solution depends on their vapor pressure.

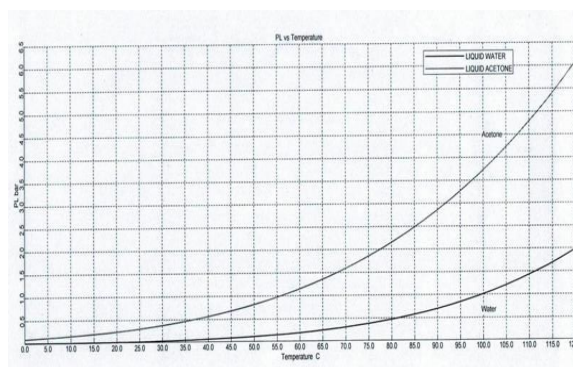


Figure 4. Acetone and water vapor pressure curve (Aspen Plus V7.2 software database)

Therefore, as an initial guess, we increase the pressure of the distillation tower to 1.5 Bar A. Therefore, as an initial guess, we increase the pressure of the distillation tower to 1.5 Bar A. Of course, using the logarithmic curve of steam pressure in terms of temperature (Kelvin) in the figure (5) below is an easier way to check the pressure.

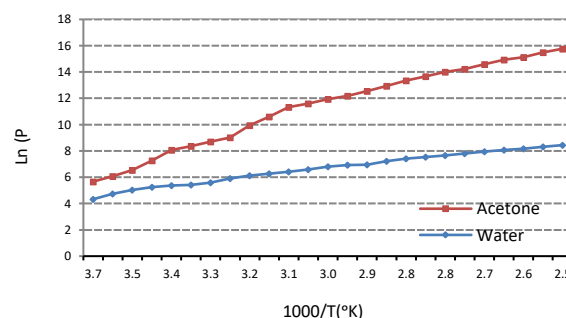


Figure 5. The curve (almost line) of the natural logarithm of pressure in terms of temperature, (Aspen Plus V7.2 software database)

Two-component liquid-vapor equilibrium phase diagrams: Since the pressure in a distillation column is almost constant, one Txy diagram and one xy diagram are sufficient for the analysis of two-component systems.

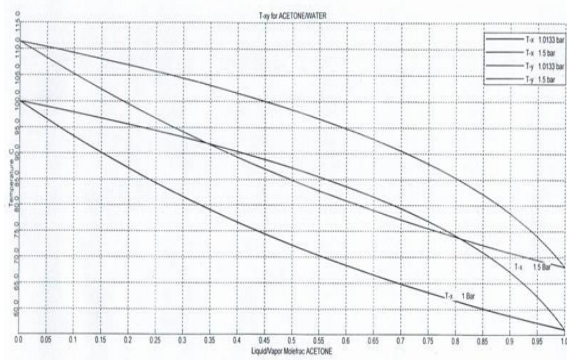


Figure 6. Acetone-water equilibrium diagram

Bar A 1.5 and 1 water (database of Aspen Plus V7.2 software)

Diagram 3 shows that at a pressure of 1.5 Bar A, mole fraction of acetone will be equal to one at a temperature of about 68°C. Therefore, as a preliminary guess, the feed temperature of the distillation tower will be equal to 68 °C. The noteworthy point is that due to the high molar fraction of water in the column, introducing the solution at this temperature leads to an increase in the amount of reflux. Therefore, finding an optimal temperature will be on the agenda.

Estimated rules: There are two estimated rules to optimize the design of the distillation process, which cannot be used simultaneously because the stabilization of one of them completely determines the design of the tower.

The first rule: The total number of trays is twice the minimum number. To start the work, it is appropriate to use the aforementioned rule, but by increasing the number of trays in the absorption tower to more than two times, at least if the purity of the final acetone product is kept constant, the recovery percentage of the acetone output from the top of the distillation tower will increase. The reason is the decrease in the amount of dissolved air in the liquid exiting the absorption tower.

The second rule: The return current is 1.2 times the minimum return current. In the

simulation with Aspen Plus, the required number of theoretical steps can be obtained by increasing the amount of reflux to achieve the same purity.

Optimum feed tray: The criterion is to minimize the cost of boiling energy. It is clear that adding feed to the upper trays of the tower increases the amount of acetone recovery, but we may have to increase the reflux to keep the purity of the acetone product constant.

Return flows: In using return flows, attention should be paid to the energy required for the rotation of materials, the negative effect of temperature and the economic savings resulting from reducing the solvent loss of valuable materials.

Energy integration: Although the conditions have improved, energy integration should be done in a more comprehensive way due to the use of steam and cooling water and ignoring the thermal energy of the effluent of the distillation tower. Increasing the pressure from Bar A 1.5 to 2 in the distillation tower, although it causes more acetone extraction, but it increases the energy loss and the level of the converters. The results of this work assuming atmospheric absorption column pressure, 0.99 product purity, 2.5 KW boiling heat load and 10.3Mol/hour input acetone.

Economic potential and return on investment

For economic optimization, it is necessary to ensure the cost-effectiveness of the plan first. For this purpose, the economic potential is used as follows: Economic potential = Value of products - Value of raw materials - Capital and operational cost of equipment such as reactor and compressor

In this research, a one-year period with 8100 hours is considered, capital costs with coefficients of India (hand), operation from reference 3 and value of weight unit or mole of fuel, products and raw materials based on market rate and production utility.

Petrochemical industries of Iran. This estimate is a study and has an error of about ± 25 . Then the profit is calculated after tax deduction. The common measure of ROI, return on investment, is used for optimization. These calculations are repeated until the maximum profit is obtained.

Discussion

Design results: The preliminary design shows that the amounts of water in table 4 are significantly different, which can be due to the

assumptions of the experimental method proposed by Douglas ($L=1.4Mg$), such as the assumption that the tower is isothermal. A very important point is that with the increase in the amount of water, the amount of dissolved air in the flow of the solution exiting from the bottom of the absorption tower increases, which has a negative effect on the separation and recovery of acetone from the distillation tower.

Table 4. Comparison of water solvent amount for acetone recovery from air

Description	Solvent amount of fresh water (Mole/hour)	Equation of state	The number of theory trays Attraction tower	Recovery rate of absorption tower %	Absorption and distillation process	
	193	Van Laar and Redlich Kwang	10	99	Simulation and design with Aspen Plus	The first method
	202		16	99.9		
L = 1.4 Mg $\mu_m = -\gamma p^0/PT$ and Kremser's equation	1943	-----	10	99	Douglas' empirical rules	The second method
	1943	-----	16	99.9		

Air to acetone mixture: 687 to 10.3mol/hour

The number of trays was also done for an absorption tower with 10 theoretical stages (initial guess) and the arithmetic mean α equal to 11.45 was obtained. Sensitivity analysis with Aspen Plus shows the effect of increasing the number of steps on the amount of water. By reducing the number of theoretical steps to 2, the amount of solvent increases greatly.

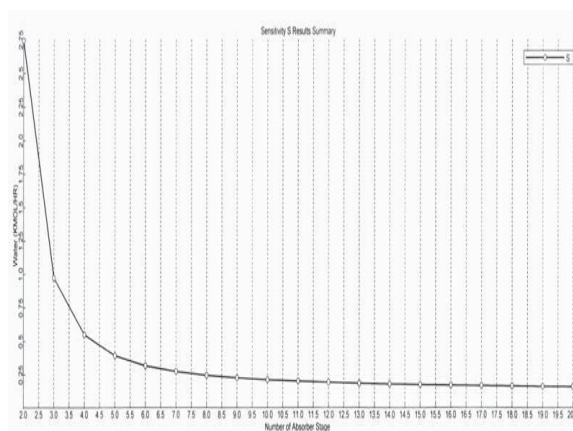


Figure 7. The diagram of the relationship between the minimum number of stages and the amount of water solvent in the absorption tower with the recovery percentage of 0.995

acetone from air. Atmospheric pressure - molar ratio of air to acetone 687 to 10.3. The number of K Mol/hour reaches 2.75.

Optimization results

60	40	20	15	12	10	9	8	The number of theoretical stages
21.8	21.8	21.9	22	22.7	27.7	34.9	Divergent	Reflux molar ratio

Table 5. The effect of the number of stages of absorption tower theory on the parameters of the separation process

N	ΔT of the Absorber	Water (Solvent)	Acetone From the Absorber	Reflux Ratio of the Distillation Tower	Purity of Acetone Product	Acetone From the Distillation Tower
	(°C)	Mol/hour	Mol/hour		%	Mol/hour
5	6	393	.01025	18.7	99	0.085
10	6	213	.01025	20.9	99	0.098
15	7	178	.01025	21.6	99	0.098
20	7	163	.01025	21.9	99	0.099

The steep slope expresses the infinite reflux rate changes. Figure 8 shows that for the number of stages less than 9, the distillation tower does not converge, and by increasing the number of stages to 20, the molar ratio of reflux decreases from 28 to 22, and by adding the number of theoretical stages to 60, the amount of reflux will not change further, as stated by the estimated rule. It has relative compatibility. The important point is that by fixing the number of trays, increasing the amount of reflux decreases the recovery percentage, which is not desirable.

Fortunately, the simulation results in figure 5 show that the number of stages in the distillation tower is very close to the aforementioned estimation method. But increasing the trays of the distillation tower can have a negative effect on the amount of reflux. The simulation results are prepared in table 5 with the assumption of Bar A pressure of 1.5 in the distillation tower and the number of fixed stages is 20.

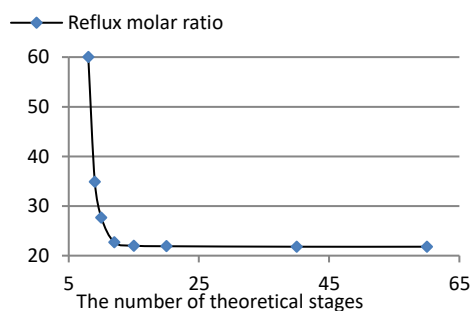


Figure 8. Relation of reflux with the number of theoretical stages of distillation tower

The purity percentage of acetone product is 99%, the recovery percentage of acetone in the absorption tower is 99.5% and the constant thermal load of the boiler is 2.5 KW.

Therefore, the simultaneous use of the two aforementioned laws is not logical as

mentioned earlier. Table 6 also shows that the increase in product purity is not related to the amount of reflux, which is in agreement with reference 12.

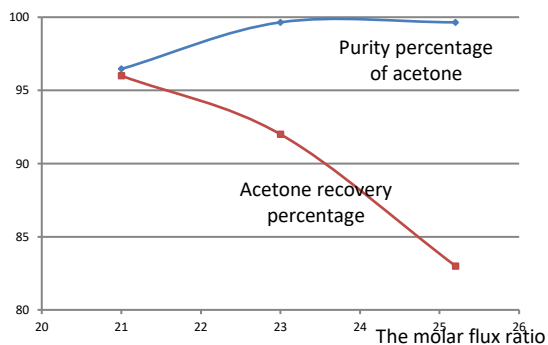


Figure 9. Variations of purity percentage and acetone recovery in distillation tower with reflux molar ratio.

The number of stages is 20, the feed tray is 9 and the thermal load of the boiler is 2.5 KW

The result of changing the feed tray in Aspen Plus V 7.2 software from 6 to 15 has no effect on the heat load of the boiler and condenser. The reason for this phenomenon is the high amount of water in the tower and as a result a relatively constant temperature profile in a part of the tower. The effect of returning the effluent of the distillation tower to the beginning of the process at atmospheric pressure by Aspen Plus V7.2 software (Figure 9) shows that the amount of water required is reduced to less than 10% of the state without return effluent, but the flow of water entering the absorption and distillation towers increases by about 2%, which does not have much effect on the increase in diameter and the capital cost of the process. In both cases, by keeping the number of theoretical steps equal to 20 distillation towers, the molar ratio of reflux increases from 22.8 in the case without return effluent to 28.8, which can be seen in the increase of re boiler load in table 6.

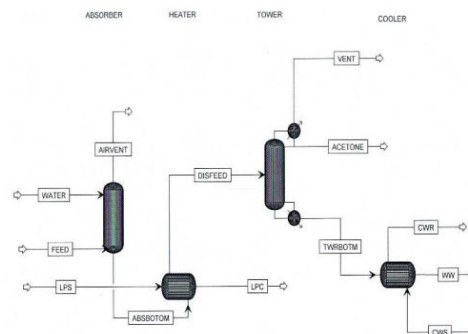


Figure 10. Flow diagram of the adsorption and distillation process of acetone recovery without/with return effluent, (A) No effluent return,

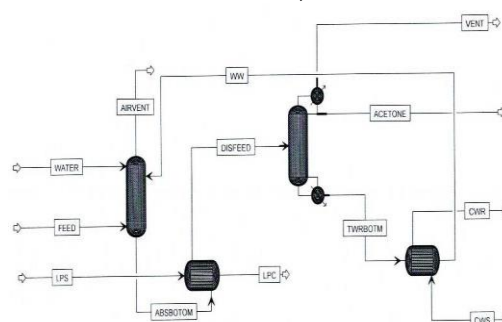


Figure 11. Flow diagram of the adsorption and distillation process of acetone recovery without/with return effluent, (B) With return effluent and compensating water 17Mol/hour

According to the last line of table 5 and the cheapness of water compared to fuel, it can be predicted that if the number of theoretical stages of the absorption tower is 10 and the pressure tower is atmospheric, the economic potential of the process with return effluent will not be positive. Unless the sewage treatment costs prevent it. In order to improve these conditions, we first increase the pressure to Bar A 5.1 to increase the amount of recovery at the same boiling heat load and purity. The simulation results show that the thermal load of the condenser has also decreased. Then, the number of steps in the absorption tower was changed to 20 to increase the molar fraction of acetone in the aqueous solution sent to the distillation tower. This helps to improve product recovery.

Table 6. Comparison of process heat exchangers with/without effluent returning to the beginning of the unit at atmospheric pressure

Heat exchanged (KW)		Heat transfer surface (m ²)		
Return with sewage	No return effluent	Return with sewage	No return effluent	
.200	.201	.0024	.0025	Distillation tower feed preheater
.261	.257	.015	.0136	Effluent cooling unit
2.8	2.32	---	---	Condenser
3	2.5	---	---	Boiling
6.261	5.278			total

Again, the amount of fresh water consumed is reduced by establishing the flow of effluent back to the beginning of the process and keeping the purity of the product and the thermal load of the boiler constant [18]. The reason for this is the more positive effect of pressure and the number of absorption steps. Tables 7A, B and C show the results of improved simulation. Restoring the flow of

exhaust gases from the distillation tower (Purge Gas) is the last step of optimization. This flow is sent to the bottom tray of the absorption tower with a temperature of 35°C to separate the rest of the acetone from the air flow. The result is shown in the last line of table 7 [19].

Table 7. The effect of improving process parameters on energy consumption and the final percentage of acetone product recovery

[A]

Case Ref.	Absorber		Cooler		Heater	
	N	Fresh Water Inlet	Duty	A	Duty	A
		Mol/hour	KW	m ²	KW	m ²
Appendix A of file 2	10	214	0.257	0.0136	0.202	0.0024
Appendix A of file 7	10	214	0.278	0.0101	0.1996	0.0025
Appendix A file 8	20	163	0.208	0.0073	0.157	0.0019
Appendix A file 9	20	18	0.203	0.00067	0.158	0.0019
Appendix A file 10	20	18	0.247	0.0136	0.162	0.0020
Appendix A of file 11	20	16	0.128	0.0099	0.147	0.0026
Appendix A file 12	20	16	0.103	0.010	0.145	0.003

[B]

Case Ref.	Pump		Dist Tower		
	Brake hp	N	P	Duty of Condenser	Reflux Molar Ratio
			Bar A	KW	
Appendix A of file 2	0	20	1	2.32	22.8
Appendix A of file 7	7.55512e-5	20	1.5	2.28	20.9
Appendix A file 8	5.9680e-5	20	1.5	2.34	21.9
Appendix A file 9	6.0117e-5	20	1.5	2.34	22
Appendix A file 10	1e-4	20	2	2.3	21
Appendix A of file 11	1.2e-4	20	2	2.29	20.9
Appendix A file 12	1.2e-4	20	1.6	2.3	20.9

[C]

Case Ref.	Dist Tower		
	Recycle (Mol/hour)		Recovery
	WW	Purge Gas	Mole/hour
Appendix A of file 2	---	---	7.980
Appendix A of file 7	---	---	9.760
Appendix A file 8	---	---	9.883
Appendix A file 9	150	---	9.85
Appendix A file 10	155	---	10
Appendix A of file 11	154.8	---	10
Appendix A file 12	151.9	11.4	10.25

Conclusion

The integration results in table 7 show that the feed flow of the distillation tower should be preheated to a temperature of 59°C by the hot effluent flow from it and cooled to a temperature of 36°C in front of the return effluent to enter the absorption tower. For the optimal state process shown in the last row of table 7 was examined for the amount of profit and the ROI index.

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