**CHAPTER 1**

**INTRODUCTION**

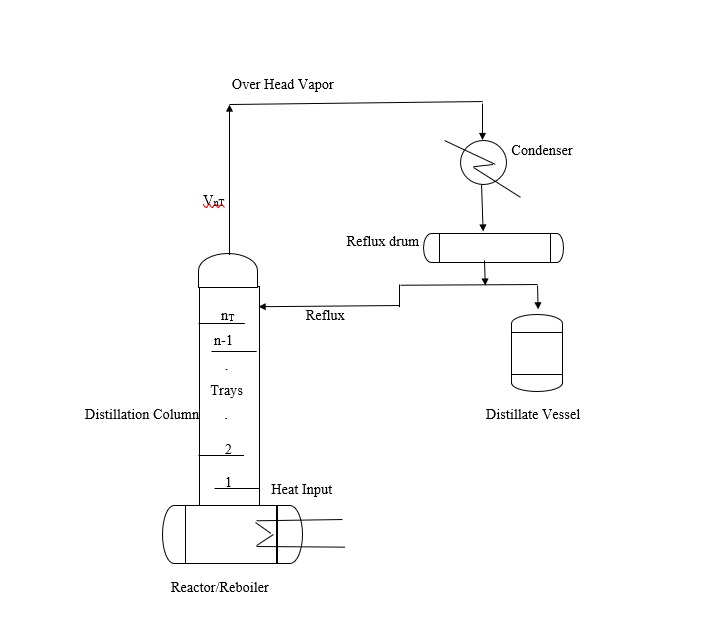
Fossil fuels meet almost 80% of the world's energy requirements [1]. However, as reported by the US Energy Information Administration, fossil fuels also account for almost 80% of greenhouse gases in the US [2]. This release of greenhouse gases is responsible for global warming, which has resulted in longer growing seasons and milder winters in certain regions of the world. Additionally, global warming has led to the recession of glaciers, loss of biodiversity, rise in sea levels, and other environmental changes that pose significant dangers. Given these apparent environmental changes, reducing the use of fossil fuels requires a strong global and regional commitment, technological innovation, and human willpower. To achieve environmental targets and reduce reliance on fossil fuels, various technological developments are advancing, including the exploration of renewable energy sources, invention of novel energy-efficient technologies, and improvement of energy efficiency in existing processing units.

In reactive batch distillation, a batch reactor can be combined with a distillation column. This combined unit operation is especially useful for those chemical reactions for which chemical equilibrium limits the conversion. By continuous separation of products from reactants while the reaction is in progress, the reaction can proceed to a much higher level of conversion that can be obtained without separation. Reboiler in conventional batch distillation column acts as reactor and reboiler at the same time (Figure 1). In this column the feed is charged into a large reboiler or reactor at the bottom of the rectifying column. If the reaction is exothermic, the heat of reaction can be used to provide the heat of vaporization and reduce the reboiler duty.

Thermal integration in distillation columns involves optimizing the heat transfer between the different sections of the column. The objective of thermal integration is to minimize the energy consumption of the distillation process while maintaining its efficiency. This is achieved by using the heat generated by the reboiler to provide energy to the condenser, instead of using external sources of energy. This process is called heat integration, and it results in significant energy savings. Generally, electrically driven vapor recompression columns, closed-cycle heat pump and bottom flashing are some of the widely used heat pumps for reactive batch distillation columns.

In conventional batch distillation operation, the column top temperature is smaller than the column bottom temperature. The heat is therefore added at the highest temperature and it is rejected at the lowest temperature, which leads to a large degradation of temperature levels. This is the principal reason for the low thermodynamic efficiency of the conventional distillation column.

To enhance the efficiency vapor recompression and bottom flashing are most feasible. Here we coupled the condenser and the reboiler in which the rejected heat from the condenser was supplied to the reboiler thus use of outside energy was reduced. Most work on vapor recompression and bottom flashing was focused on the separation of closed boiling mixtures in continuous distillation in the literature. But few papers have dealt with vapor recompression and bottom flashing in reactive batch distillation for esterification reaction. This research work proposes a systematic design, operational and simulation approaches for vapor recompression and bottom flashing techniques in reactive batch distillation columns for production of Butyl Acetate.



**Figure 1. Reactive Batch Distillation Column**

# **CHAPTER 2**

**REVIEW OF LITERATURE**

# Distillation is the most widely used unit operation for the separation of liquid mixtures in the chemical and process industries. It accounts for 95% of the fluid separation in chemical sector [3]. Among the total energy used in the chemical industry nearly 60% of the energy is utilized for distillation alone. The chemical process sector requires a minimum TAC and CO2 emissions in energy –efficiency systems.

## 2.1. Continuous Flow Distillation

To increase the thermodynamic efficiency of distillation and reduce the energy consumption various thermally coupled continuous distillation arrangements have been explored. Electrically driven vapor recompression columns, closed-cycle heat pump and bottom flashing are some of the widely used heat pumps for continuous distillation column. In Vapor-Recompression the top tray vapor is compressed and the internal energy released is used to heat reboiler liquid. It is the most favoured scheme [14][16][17]. In Bottom flashing the bottom stream is expanded and then utilized as a condenser coolant. Later it is compressed and returned to the column. The internally heat integrated distillation column (HiDiC) is another arrangement which has acquired a lot of attention. The idea behind this is combination of heat pump assisted VRC scheme and heat integration between two diabatic sections [11].

## 2.2. Batch Distillation

Batch distillation is one of the oldest separation processes. It is used in the chemical industries and has several advantages. It is often used in industries where high purity products are required than continuous distillation [3].

The conventional batch distillation column consists of a bottom receiver/reboiler, rectifying column (either a tray or packed column) placed over the reboiler, connected to a total condenser or a partial condenser system and distillate receivers. In this column, the charge is loaded into the reboiler at the beginning of the process and heated to its boiling point. Vapour flows upwards in the column and condenses at the top. After some time, a part of the overhead condensate is withdrawn continuously as distillate, and the other part is returned to the column section as reflux. The liquid in the reboiler is increasingly depleted of the more volatile components. As the amount of liquid in the reboiler decreases, the concentration of high boiling constituents increases. Study have been made for improved energy efficiency and cost- effectiveness of batch distillation for separating wide boiling constituents with integration of vapor recompression column [16]. This research group found vapor recompressed system gives energy saving of 50% and total annual cost saving of 10.03%. Using bottom flashing in batch distillation works have been carried out for separation of closed boiling constituents [20][21].

## 2.3. Reactive Batch Distillation

It has been a common practice to carry out the reaction and separation in the industrial processes sequentially in separate unit operations. With a reversible reaction system

the traditional process consists of a reactor followed by a sequence of distillation columns (Figure 1). The mixture A and B is fed to the reactor, where the reaction takes place in the presence of a catalyst and reaches equilibrium. Conventional reactive batch distillation (Figure 1) operates in two modes, the startup phase and the production phase [16]. In the startup phase, the system attains a steady state where the lighter component reaches maximum purity. Once the system attains a steady state after that immediate production phase starts. In the production phase the product (lightest component) is withdrawn. A distillation column is then required to separate the products C and D while the unreacted components A and B are recycled back to the reactor. In recent years, the development and application of integrated processes combining the mechanisms of reaction and separation in one single unit has attracted growing interest in the chemical and process industry.

The benefits of reactive distillation are simplification of the separation system that leads to significant capital savings and heat integration benefits [4][19]. If the reaction is exothermic, the heat of reaction can be used to provide the heat of vaporization and reduce the reboiler duty. There are so many instances where reactive distillation is of great importance in the chemical industries. One of the frequently treated processes is the esterification of acetic acid with ethanol [4][22]. VRC implementation to reactive batch distillation for production of Ethyl Acetate was studied [4] and they found an energy savings of 37.08% and 22.88% reduction in operating cost with payback period of 2.78 years.

**CHAPTER 3**

**METHODOLOGY**

The distillation model has been created with the help of ODEs and algebraic equations [23] in the MATLAB environment using the first principal approach are given below. Accordingly, the MESH (Material balance, vapor-liquid Equilibrium, mole fraction Summation and Heat balance) equations are derived.

First the equations for the condenser and accumulator are presented. Then the equations for the plates in the column and the reboiler are presented: The plates are counted from the top to the bottom. ‘n’ refers to plates and ‘i’ refers to components.

## 3.1 Modelling Equations for Reactive Batch Distillation

For Condenser:

Mass Balance:

Component Balance:

Energy Balance:

+

Equilibrium:

For Reboiler:

Mass Balance:

Component Balance:

Energy Balance:

+

Equilibrium:

For Trays:

Mass Balance:

Component Balance:

Energy Balance:

+

Equilibrium:

Tray n

(holdup=)

**Notations**

* *xn* is the mole fraction (or composition) of the more volatile component (here Butyl Acetate) in a liquid stream leaving nth tray
* *yn* is the mole fraction of Butyl Acetate in a vapor stream leaving nth tray
* *xD,i* is the mole fraction of ith component in the liquid distillate
* *xB,i* is the mole fraction of ith component in the liquid of the still-pot
* *Ln*is the liquid flow rate leaving nth tray (kg-mol/min)
* *Vn* is the vapor flow rate leaving nth tray (kg-mol/min)
* *R* is the reflux flow rate (kg-mol/min)
* *D* is the distillate flow rate (kg-mol/min)
* *VB* is the vapor boil-up rate (kg-mol/min)
* is the liquid holdup on nth tray (kg-mol/min)
* is the liquid holdup in the reflux drum (kg-mol/min)
* is the liquid holdup in the still-pot (kg-mol/min)
* is the enthalpy of distillate (J/g-mol)
* is the enthalpy of a liquid stream leaving nth tray (J/g-mol)
* is the enthalpy of a vapor stream leaving nth tray (J/g-mol)
* *QR* is the heat input to the still-pot (kJ/min)
* The dot symbol (.) represents the time derivative. The time derivative of the multiplication of two variable, like *m* and *x*, is written as .

## 3.1.1 Approach 1: A Systematic Design, Operation and Simulation Approach for Conventional Reactive Batch Distillation

*Step 1*: Input data on the column size (number of total trays and column diameter), weir dimensions (weir height and weir length), feed (components, flow rate, composition and temperature), feed tray location, pressure profile, tray efficiency, reaction information, vapor pressure data, etc.

*Step 2:* Input data for variables at time t = 0. The variables include the liquid-phase compositions (xi) and liquid holdups (m) for all trays.

*Step 3:* Either input the values of reflux rate, distillate rate, bottoms rate and reboiler duty (or vapor boil-up rate) or manipulate these variables employing the suitable controllers. The start-up phase runs under total reflux condition and at production phase, the manipulated inputs are adjusted using the controllers.

*Step 4:* Compute the equilibrium vapor-phase composition and temperature (T) for each tray based on bubble point algorithm. The Newton–Raphson convergence method can be used in the bubble point calculations. Subsequently, calculate the actual vapor-phase composition (yi) employing Murphree relationship shown below.

*Step 5:* Calculate the liquid-phase and vapor-phase enthalpies for each tray using the algebraic form of equations.

*Step 6:* Compute the reaction rate.

*Step 7:* Calculate the internal liquid flow rates using the Francis weir formula and the vapor flow rates solving the energy balance equations. Note that the three different approaches for calculating the vapor flow rate are discussed later. In the open-loop simulation, we use the steady state values of reflux rate and reboiler duty, and we calculate the distillate rate and bottoms rate assuming constant holdups in reflux drum and column base, respectively.

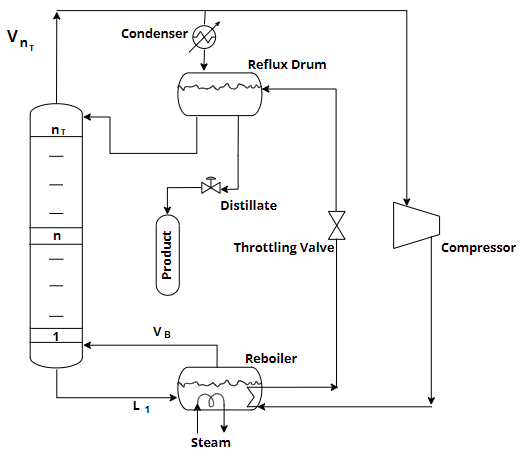
*Step 8:* Calculate the liquid holdup on each tray for the future time step (t + t) by solving the total mole balance equation.

*Step 9:* Compute the liquid-phase compositions on all trays for the future time step (t + t) by solving the component mole balance equations.

*Step 10:* To continue the process simulation for the next time step, go back to step 3.

# **3.2 Vapor Recompression in Reactive Batch Distillation (VRCRBD)**

Figure 2 shows the principal operation of a VRC with a single compressor in reactive batch distillation conceptually. The primary goal of this VRC is to recover heat from top vapor, lowering hot and cold utility consumption. For that we had to use the heat released in the condenser effectively, here we used an external compressor to the existing reactive batch distillation column for compressing the vapor from top tray and send to the reboiler where the latent heat of compressed top vapor is used for liquid re-boiling which will reduce the external utility requirement. The main criteria here is to evaluate the additional energy component to the compressor involved, because thermal energy is significantly cheap as consumption of electricity is much more expensive that thermal utility



**Figure 2. Vapour Recompression in Reactive Batch Distillation**

## 3.2.1 Approach 2: A Systematic Design, Operation and Simulation Approach for VRCRBD

*Step 1*: Carry out *Steps 1* through *10* of Approach 1**.**

*Step 2*: The vapor flow through the plate holes and reaches to the top tray.

*Step 3*: The compressor receives the vapor from the upper tray. The upper tray vapor () is compressed to the necessary pressure or temperature (temperature lift). The latent heat is released by compressed vapor, which is then used as heat input to the reboiler and condensed against the reboiler liquid.

* The vapor is compressed so that the compressed overhead vapor and bottom liquid () have a reasonable temperature difference () for total condensation in the reboiler-condenser. In order to maintain a of for total condensation of overhead vapor in the reboiler-condenser and to avoid compressor operation at maximum CR throughout the batch operation, a variable speed compressor (manipulation of CR) has been used.

Manipulation of CR: We know the , and (top tray temperature) values, then calculate the compressor outlet temperature ( from Equations below, and then find out the corresponding CR to maintain constant value which means CR is variable.

*Step 4*: Use the following Equation to calculate the energy released by compressor vapor.

An iterative scenario including vapor flow rate modification has been used to achieve the same dynamical performance between the standard CRBD and VRCRBD while retaining the same reboiler heat duty of 52307.8 kJ/min.

*Manipulation of vapor flow rate*: Split the top vapor into two halves if the latent heat is released by the overhead vapor () exceeds the required reboiler heat obligation. One part goes to the compressor to raise the pressure, then to the bottom receiver for condensation and the rest goes to the condenser (). Compressor work is calculated by Equations as shown below. If the compressed vapor latent heat () is less than the required reboiler heat duty, external heat must be supplied to the reboiler.

If

else

end

Where, polytropic coefficient, – inlet and outlet pressures of compressor, respectively

*Step 5*: The complete condensed liquid at high pressure is then throttled to top tray pressure and reaches the reflux drum. A throttling valve depressurizes the high pressure condensed liquid before it is returned to the reflux drum.

*Step 6*: *Steps 7* through *14* of Approach 1 should be repeated.

Finally, Calculate the Energy savings and TAC savings from:

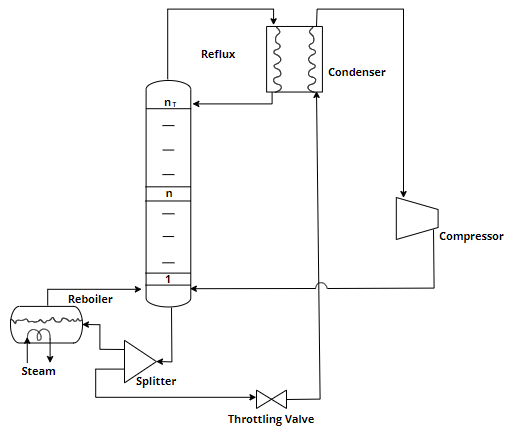
# **3.3 Bottom Flashing in Reactive Batch Distillation (BFRBD)**

The Figure 3shows the principal operation of BFRBD with a single compressor in a binary reactive batch distillation conceptually. The primary goal of BFRBD is to integrate the heat released in condenser to evaporate the liquid from reboiler so that we can save energy supplied to reboiler and also the cost of reboiler.

The proposed reactive batch distillation with bottom flashing basically couples the top vapor with bottom liquid from the reboiler in the overhead condenser. As the liquid in the bottom is hotter than the vapor at the top to help the condensation happen from the condenser adiabatic flashing of the bottom liquid is done with the help of throttling valve. The pressure of the bottom liquid is lowered to a level below the saturation temperature of the top vapor which helps to cool down the vapor. This operation mainly performed two operations:

1. It recovered the latent heat of top vapor in the overhead condenser by vaporizing the flashed liquid.

2.Cold vapor from the condenser is compressed back with the help of compressor which reduced the external energy use.



**Figure 3. Bottom Flashing in Reactive Batch Distillation**

**3.3.1 Approach 3: A Systematic Design, Operation and Simulation Approach for BFRBD**

*Step 1*: Carry out *Steps 1* through *10* of Approach 1**.**

*Step 2*: The liquid flow through the reboiler reaches to the throttle valve.

*Step 3*: The throttle valve receives the liquid from the reboiler. The liquid from reboiler

() is flashed through throttle valve to the necessary pressure or temperature (temperature drop). The flashed liquid now contains liquid and vapor. The liquid after the flash will exchange the heat with top tray vapor and evaporates

* The liquid is flashed so that the flashed liquid and bottom liquid () have a reasonable temperature difference () for total condensation of the top tray vapor in condenser. In order to maintain a of for total condensation of the top tray vapor in condenser and to avoid compressor operation at maximum CR throughout the batch operation, a variable speed compressor (manipulation of CR) has been used.

We know the , (top tray temperature), then calculate the flash outlet temperature (as shown below

*Step 4:* After flashing the liquid we get VTV and LTVwhich will exchange heat with top tray vapor by vaporization. The amount of flash required to condense the top tray vapor is calculated as follows.

Here we use nested loop as we need to converge PTV (pressure after flash) and vapor fraction after flash to attain the temperature of TTV.

*Outer Loop*

For i=1:500

Assume PTV<PB

(Here PTV, j0 can be calculated from Antoine equation which is a function of temperature)

*Inner Loop*

For j=1:500

If Qavailable>QR then

End

Now, if yes then Break

Inner loop end

Now, if yes then Break

end of outer loop

In above algorithm latent heat of vaporization is calculated using Clausius Clapeyron equation as follows

And latent heat of vaporization for the mixture is calculated using mixing rule

*Step 5:* After vaporizing the liquid from the reboiler completely using throttle valve and in condenser, we need to compress the vapor at TTV to TB (reboiler temperature) to maintain same dynamics as conventional reactive batch distillation. As mentioned in *step 3* here we have to use variable speed compressor which means CR is variable.

To compress the vapor to TB the work done by the compressor is calculated as follows

Where, polytropic coefficient, – inlet and outlet pressures of compressor, respectively.

*Step 6*: *Steps 7* through *14* of Approach 1 should be repeated.

Finally, Calculate the Energy savings and TAC savings from:

**3.4 Performance Evaluation Indices**

# **3.4.1 Energy Savings**

In reactive batch distillation, energy consumption is a key performance measure for evaluating the energy efficiency of heat pump assisted in VRCRBD and BFRBD. The sum of the total reboiler heat duty plus three times the compressor duty yields total energy consumption of vapor recompressed batch distillation.

(The factor 3 in the above equation is to account for converting the compressor duty into thermal energy which produces an equivalent amount of electrical power. This is determined empirically). The conventional multicomponent reactive batch distillation energy consumption is calculated from:

When comparing vapor recompressed multi component reactive batch distillation (VRMRBD) to CMRBD, the corresponding energy savings are

## 3.4.2 Total Annual Cost

TAC has been chosen as one of the most important indicators of economic performance. The TAC is determined by the following Equation

TOC stands for total annual operating cost, while TCI stands for total capital investment. The capital investment includes the installation cost of apparatus (reboiler, condenser, column shell, column trays, intercoolers and compressors). The operating costs includes the electricity, cooling water and steam (Babu et al., 2012). In this study, the modernized price of the Marshall and Swift (M&S) index of 1704.9 was used. While heat pumps offer a high energy efficiency rating for separating high temperature lifts, it is difficult to achieve economic benefits over a shorter payback period. Additional costs such as reflux drums, pumps, valves, and piping can be eliminated because they are far less expensive than column vessels and heat exchangers. The TCI and TOC were calculated using correlation methods from (Douglas, 1988) are given below:

Capital fixed cost for Reactive Batch Distillation includes the following

* Column Shell Cost
* Column Tray Cost
* Condenser Cost
* Reboiler Cost
* Compressor Cost

Operating Cost for Reactive Batch Distillation includes the following

* Cooling water cost to cool the vapor entering condenser
* Stream cost to heat up the feed in reboiler

Electricity cost to the compressor

**3.4.2.1 Capital Costs (MOC: SS)**

Where, DC is the column diameter (ft), LC The column height (ft), =1704.9, and the coefficients Cin =2.18, Cm =3.67, Cp=1.0, **A** is the heat transfer area (ft2), and the coefficients Cin =2.18 Cm =3.67 and Cp=1 and Ct=1.8

Note: Here, Heat exchanger cost is applicable for both reboiler and condenser.

**3.4.2.2 Operating Cost**

* =
* =
* =

**CHAPTER 4**

**RESEARCH OBJECTIVES**

1. Modelling and Simulation of Conventional Reactive Batch Distillation for production of Butyl Acetate.
2. To Implement Bottom Flashing and Vapor Recompression Techniques in Conventional Reactive Batch Distillation for production of Butyl Acetate to improve Energy Savings and reduce the Total Annual Cost.

**CHAPTER 5**

**RESULTS AND DISCUSSION**

# **5.1 Conventional Reactive Batch Distillation**

Production of Butyl Acetate system was investigated in this work. This system is wide boiling/zeotropic because the temperature difference between the lightest and heaviest components is significant. The column specifications utilized in computer simulations are listed in Table 1. The computer program was developed and simulated in MATLAB with a time step of 0.01min. In the simulations, the initial compositions in all stages and reflux drum are equal to that of the feed mixture.

To analyze the performance of Vapor Recompression System and Bottom Flashing, A conventional reactive batch distillation model (without vapor compressor) has been simulated using a mixture of Acetic Acid, Butanol, Butyl Acetate, Water.

**Table 1. Design and Operating Parameters** [A K Jana & D maiti, 2013]

|  |  |
| --- | --- |
| **System** | **Acetic Acid/Butanol/Butyl Acetate/Water** |
| Total feed charge, Kmol | 200.7 |
| Stages | 16 |
| Feed Composition | 0.45/0.45/0.05/0.05 |
| Tray holdup in each tray (Start-up), Kmol | 0.9 |
| Reflux drum holdup, Kmol | 6 |
| Tray efficiency, % | 75 |
| Heat input to the still pot, kJ/min | 52307.8 |
| Reboiler composition (Startup phase), mol fraction of Butyl Acetate | 0.459 |
| Distillate rate(fixed), kmol/min | 0.19066 |

**5.1.1 An Equimolar Feed Composition**

Equimolar feed composition of 0.45-0.45 Acetic Acid and Butanol is considered and detailed results are explained and presented below.



**Figure 4. Distillate composition throughout the batch operation**



**Figure 5**. **Reboiler composition throughout the batch operation**

Initially, the feed was added into the reboiler and a part of it, 3% was distributed to the reflux drum and 0.9 kmol of feed was distributed onto each tray. Then, process was simulated in MATLAB where, the startup phase lasted till the composition of Butyl Acetate in the reboiler reached 0.459, after which the production phase started which was stopped when the average reboiler composition of Butyl Acetate reached 0.98.

When Feed composition was 0.45-0.45 Acetic Acid-Butanol, the total Batch time was 1386.9 min. The startup phase lasted till 657.3 minutes and then the production phase started.

From the Figures 4, 5 we can interpret the same results that startup phase lasted for 653.7 minutes and then total batch time was 1386.9 minutes. We can also interpret that when heat is given initially in production phase the high volatile component in the distillate increases and in the reboiler it decreases, this is because separation takes place as we give heat and in production phase since we are withdrawing the product, we can observe that there is a slight decrease in the concentration, we will stop this process when the average composition of Butyl Acetate collected reaches 0.98.



**Figure 6. % conversion profile throughout the batch operation**

**Table 2: Compositions of Reboiler and Distillate: Start Up and Production** **Phase**

|  |  |  |  |  |
| --- | --- | --- | --- | --- |
| Component | Steady State Composition of Reboiler | End Composition of Reboiler | Steady State Composition of Distillate | End Composition of Distillate |
| Acetic Acid | 0.1153 | 0.0113 | 4.84×10-4 | 0.0320 |
| Butanol | 0.1156 | 0.0085 | 2.3×10-4 | 0.0282 |
| Butyl Acetate | 0.459 | 0.98 | 4.93×10-5 | 0.9161 |
| Water | 0.3101 | 2.3335×10-4 | 0.9992 | 0.0237 |

**5.2 Bottom Flashing in Reactive Batch Distillation**

Equimolar composition of Acetic Acid and Butanol are taken and performance of Reactive Batch Distillation Column was investigated and results are shown and explained below. For a fair comparison, the simulations were run using the same design and operational variables as the Conventional Reactive Batch Distillation but with Bottom Flashing.



**Figure 7. Compression ratio profile throughout the batch operation**



**Fig 8. Energy profile throughout the batch operation**

In Bottom Flashing, since it’s an external heat integration system the composition profiles in reboiler and condenser remains the same as shown in the figures 4 and 5. As we start the process in startup phase there will be rapid increase in the temperature difference between top tray and the reboiler and hence the compression ratio also increases rapidly showing the same trend. In production phase there will be a increase in the temperature difference first and decreases gradually which results in decrease of Compression Ratio as shown in fig 7.

Since we have to supply 52307.8KJ/min of heat to the reboiler, in case when the heat released by the condensation of the top tray vapor is less than 52307.8KJ/min then we will supply the deficit heat through steam as external energy as shown in the equation above. If there is excess of heat available then there is no need of any external energy therefore QE =0. In the figure 8 shows the available energy (QCV), required energy (QR) and supplied external energy (QE).

**5.3 Vapor Recompression in Reactive Batch Distillation**

The performance of a single-stage vapor recompressed reactive batch distillation column was investigated and results are shown and explained below. For a fair comparison, the simulations were run using the same design and operational variables as the conventional reactive batch but with vapor recompression.



**Figure 9. Compression ratio profile throughout the batch operation**



**Figure 10. Energy profile throughout the batch operation**

In vapor recompression, since it’s an external heat integration system the composition profiles in reboiler and condenser remains the same as shown in the figures 4, 5. As we start the process in startup phase there will be decrease in the temperature difference between top tray and the reboiler initially and hence the compression ratio also decreases showing the same trend. In production phase there will be a steady (linear almost) increase in the temperature difference resulting in steady increase in the compression ratio as shown in fig 9.

Since we have to supply 52307.8KJ/min of heat to the reboiler, in case when the heat released by the condensation of the top tray vapor is less that 52307.8KJ/min, we will supply the deficit heat through steam as external energy as show in the above equations. If there is excess of heat available, there is no need of any external energy, therefore, QE =0. In the figure 10 shows the available energy (QCV), required energy (QR) and supplied external energy (QE)

# **5.4 Economic Performance Comparison**

The total annual calculation is calculated for all 3 systems of CRBD, VRCRBD, BFRBD are tabulated below for the case of equimolar feed. The below are the costs of the utility that is used to calculate TAC\*\* **(Xinqiang You et al.).**

**\*\*Cooling water cost**: 0.032$, **Steam Cost**: 17$ and **Electricity Cost**: 0.0775$.

**Table 3. TAC Savings between CRBD AND BFRBD**

|  |  |  |
| --- | --- | --- |
| **Component** | **CRBD** | **BFRBD** |
| **Capital Cost ($)** | | |
| **Column Shell** | 117350 | 117350 |
| **Column Tray** | 11345 | 11345 |
| **Compressor** | 0 | 492460 |
| **Condenser** | 76215 | 74910 |
| **Reboiler** | 348680 | 1090.9 |
| **Total Capital Cost** | 553590 | 697155.9 |
| **Operational Cost ($)** | | |
| **Cooling water** | 18540 | 0 |
| **Steam** | 193380 | 27.1095 |
| **Electricity** | 0 | 58905 |
| **Total Operational Cost** | 211920 | 58932.1 |
| **TAC** | 396440 | 291320 |
| **TAC savings%** |  | **26.515** |

As shown in Table 3 when the payback time is 3 years for bottom flashing, annual savings in cost is around 26.515% and energy saving is around 75.93%. The high savings in BFRBD is because of the savings in reboiler, as reboiler holdup and duty decreases. We can also observe that the cooling water utility can be eliminated as flashed liquid will act as hot utility. Heating utility i.e., stream usage decreases as we are meeting a part of energy requirement through bottom flashing.

**Table 4. TAC Savings between CRBD and VRCRBD**

|  |  |  |
| --- | --- | --- |
| **Component** | **CRBD** | **VRCRBD** |
| **Capital Cost ($)** | | |
| **Column Shell** | 117350 | 117350 |
| **Column Tray** | 11345 | 11345 |
| **Compressor** | 0 | 252500 |
| **Condenser** | 76215 | 4860.1 |
| **Reboiler** | 348680 | 348680 |
| **Total Capital Cost** | 553590 | 734735.1 |
| **Operational Cost ($)** | | |
| **Cooling water** | 18540 | 268.5493 |
| **Steam** | 193380 | 2477.8 |
| **Electricity** | 0 | 26083 |
| **Total Operational Cost** | 211920 | 28829.34 |
| **TAC** | 395000 | 286550 |
| **TAC savings%** |  | **30.9** |

As shown in Table 4 when the payback time is 3 years for Vapor Recompression Technique, the annual savings in cost is around 30.9% and the energy saving is around 88%. Even though compressor is much costly we are saving at condenser as usage of condenser is almost eliminated and therefore cooling water utility drastically decrease. Heating utility i.e., steam usage decreases as we are meeting a part of energy requirement through Vapor Recompression.

|  |  |  |  |
| --- | --- | --- | --- |
| **Component** | **CRBD** | **VRCRBD** | **BFRBD** |
| **Capital Cost ($)** | | |  |
| **Column Shell** | 117350 | 117350 | 117350 |
| **Column Tray** | 11345 | 11345 | 11345 |
| **Compressor** | 0 | 252500 | 492460 |
| **Condenser** | 76215 | 4860.1 | 74910 |
| **Reboiler** | 348680 | 348680 | 1090.9 |
| **Total Capital Cost** | 553590 | 734735.1 | 697155.9 |
| **Operational Cost ($)** | | |  |
| **Cooling water** | 18540 | 268.5493 | 0 |
| **Steam** | 193380 | 2590.6 | 77.2542 |
| **Electricity** | 0 | 26083 | 58905 |
| **Total Operational Cost** | 211920 | 28829.34 | 291320 |
| **TAC** | 396440 | 273740 | 291320 |
| **TAC savings%** |  | **30.95** | **26.51** |
| **Energy Savings(%)** |  | **88** | **75.93** |

**Table 5. Economic Performance Comparison**

Table 5 shows the economic comparison between CRBD, VRCRBD and BFRBD. For fair comparison, we have considered 3 years as payback period for all the cost calculation. It is evident from the results that VRCRBD outperforms than the CBD and BFRBD, as it gives more cost and energy savings.

**CHAPTER 6**

**SUMMARY AND CONCLUSION**

After carrying out systematic design, operation and simulation approach for Conventional, Vapor Recompression, Bottom Flashing on Reactive Batch Distillation the following can be concluded.

Start Up phase time for the production Butyl Acetate is 653.7 min and total batch time is 1386.9min.

Implementing Vapor Recompression to Conventional Reactive Batch Distillation gives us energy saving around 88%.

Vapor Recompression has a total annual saving of 30.95% for payback period of 3 years.

Implementing Bottom flashing to conventional batch distillation gives us energy saving around 75.93%.

Bottom flashing has a total annual saving of 26.51% for payback period of 3 years.

In comparison between VRBD and BFBD we can conclude that Vapour Recompression gives us better energy saving, about 12% more than BF and also total annual savings in VRC is 4.44% higher than Bottom Flashing.

**CHAPTER 7**

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