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Research paper

Performance investigation of vapour recompressed batch distillation for separating ternary wide boiling constituents☆

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ABSTRACT

The vapour recompression scheme (VRC) has been very effective in continuous distillation for energy intensification. The applicability of this scheme for the separation of multicomponent wide boiling constituents in batch distillation is a major challenge, because of the unsteady nature of the batch. In this study, the vapour recompression scheme has been implemented for the separation of multicomponent wide boiling constituents in the batch distillation. For the optimal usage of energy from compressed vapours manipulation of top tray vapour or external energy is done. A comparative study of the vapour recompressed batch distillation having a variable speed single compressor (SVRBD) and double stage compressor (DVRBD) with conventional batch distillation in terms of energy savings and total annualized cost is done. The VRC schemes achieve an energy savings of 50% and 10.03% total annualized cost (TAC) for SVRBD and DVRBD achieve 52% energy savings and 12.21% TAC with a payback period of 10 years.

1. Introduction

The overexploitation of fossil fuels has enhanced the concentration of greenhouse gases in the atmosphere. It has increased the natural greenhouse effect, global warming. Its adverse effects are melting of glaciers, rise in sea levels, changes in patterns and amount of precipitations, droughts, floods, epidemics, malignancy, etc. This demands an urgent need for reduction in usage of fossil fuels.

Distillation is one of the major separation technology used in chemical and allied industries. It consumes an estimate of 60% of energy in chemical industries [1]. The thermodynamic efficiency of conventional distillation is very low around 5–20% [2]. Since distillation is an energy consumer of fossil fuels, the need for energy intensification of the process is severe. To improve the energy efficiency of distillation processes, several energy integration techniques have been proposed.

Coupling of condenser and reboiler using a heat pump is an effective method for energy intensification of distillation. The most frequently used heat pumps for distillation columns are electrically driven vapour recompression types. They include the direct vapour recompression, closed cycle heat pump and bottom flashing. In the direct vapour recompression column (VRC), the overhead vapour is compressed in a compressor and then it is used as an internal source of energy in liquid reboiling. In the closed cycle heat pump, an external refrigerant is used to facilitate the energy transfer between the top vapour stream and the bottom liquid stream. In bottom flashing, the bottoms product is expanded and used as a coolant in the condenser, where after it is compressed and returned to the column. Among these three heat pumping arrangements, VRC is the most popular and commonly used configuration and is suitable for close boiling constituents [3].

Since 1960 s, the application of vapour recompression heat pump system is observed in continuous columns [4–6] because of its ability to greatly improve the energy and economic performance of the system. The use of this heat pumping system in batch column, unlike the continuous distillation, is not so straightforward mainly because of unsteady state nature of the batch processing.

Batch processing has continued to be an important technology owing to the operational flexibility that it offers. This operational flexibility of batch distillation processes makes them particularly suitable for smaller, multiproduct or multipurpose operations. Manufacturing in the pharmaceutical and specialty fine chemical industries are examples of small, multiproduct operations, where products are typically required in small volumes, and subject to

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Nomenclature

L_1	Flow rate of liquid leaving 1st tray,
V_B	Vapour boil-up rate and
V_{n_T}	Overhead vapour rate
C	Total number of components,
D	Distillate rate,
H_n^L	Enthalpy of a liquid stream leaving nth tray,
H_n^V	Enthalpy of a vapour stream leaving nth tray,
$k_{n,j}$	Vapour-liquid equilibrium coefficient with respect to nth tray and jth

Component

L_n	Flow rate of a liquid stream leaving nth tray,
m_n	Liquid holdup on nth tray,
$P_{n,j}^0$	Vapour pressure of component j in nth tray,
P_T	Total pressure,
Q_R	Reboiler duty,
R	Reflux rate,
V_n	Flow rate of a vapour stream leaving nth tray,
$x_{n,j}$	Mole fraction of component j in a liquid stream leaving nth tray,
$y_{n,j}$	Mole fraction of component j in a vapour stream leaving nth tray
$\gamma_{n,j}$	Activity coefficient of component j in nth tray.

short product cycles and fluctuating demand. The thermodynamic disadvantage of batch distillation over continuous fractionation, which results in lower energy efficiency, has long been known.

The first configured thermally integrated batch distillation scheme consisted of a rectification tower surrounded by a jacketed reboiler (or still pot) [7]. The advantages of this internally heat integrated scheme was systematized and clarified through numerical simulations after a decade [8]. Even though the capital cost is higher compared to conventional batch distillation (CBD), it achieves a total annualized savings of 27.93% in 1.38 years payback period. A VRBD scheme which is a combination of internally heat integrated distillation with concentric reboiler (IHIBDCR) and VRC for the separation of wide boiling constituents was developed [9]. The scheme was illustrated using binary wide boiling constituents of methanol and water features the advantages of (IHIBDCR) and the hybrid scheme in reducing operating costs. The impacts of heat integration by performance indicators such as energy consumption, total annualized cost and CO_2 emissions was studied and simulated using a binary mixture of benzene and toluene [10]. The three parameters were compared with the CBD. The proposed VRBD secured significant reduction in the performance indicators. The investigation of vapour recompression scheme in batch distillation in separation of binary wide boiling constituents (acetone/water) found 68.89% reduction in energy use and around a 67.58% reduction in operating cost [11].

Few papers have been published on application of vapour recompression in batch distillation (VRBD) with close boiling constituents [12–16]. The traditional vapour recompressed continuous distillation is not an economically attractive option for wide boiling constituents. In this paper vapour recompression technique in batch distillation is explored for the separating of ternary wide boiling constituents in terms of energy and cost savings.

In the present work, introduce the vapor recompression technique in a batch distillation column. The proposed VRBD arrangement consists an isentropic compressor that runs at a fixed as well as variable speed. It is noticing the variable speed VRBD additionally involves the manipulation of compression ratio (CR). The Objective is to ensure the optimal use of internal heat energy, an

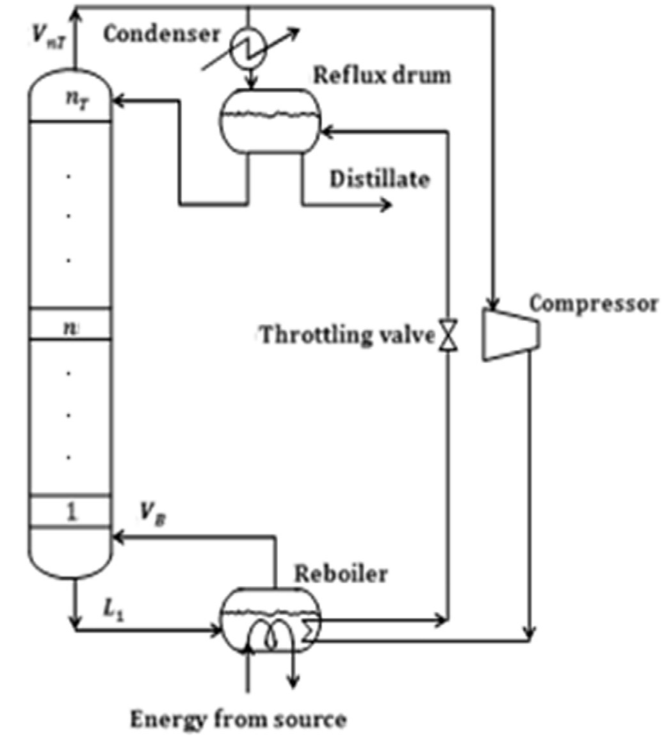

Fig 1. Schematic representation of Single Stage Vapour Recompressed Batch Distillation (SVRBD).

open-loop control policy is proposed for the VRBD that adjusts either the heat supply from external source to the reboier or top vapor splitting. Developing two alternative configurations of the VRBD column, i.e., Variable speed Single stage compressor (SVRBD) and Variable speed Double stage compressor (DVRBD) to identify the best heat integrated scheme in terms of energy and total annualized cost (TAC) savings.

2. Principle and configurations of vapour recompressed batch distillation

The conventional batch column that consists of a rectification tower consists of reboiler at the bottom and total condenser at the top. The trays are numbered from bottom-up, indicating the bottom tray as the 1st Stage and the topmost tray as the $n_T{}^{\text{th}}$ Stage. The batch operation in a distillation column operated in two phases, the startup phase and the production phase [17].

The SVRBD consists of electrically driven Single stage Variable speed compressor and a throttling valve [15]. Schematic representation of SVRBD is shown in Fig 1. In SVRBD, vapour leaving the top tray of the column is compressed to a desired pressure using a single stage isentropic compressor The work done by the compressor is an energy expense. In order to reduce it, instead of a single stage compressor, a double stage compressor is used to make a Double stage vapour recompressed batch distillation (DVRBD). Compression at high temperature with energy expense causes evaporation of lubricating oil and thereby increasing wear and tear in the compressor. In order to prevent that double stage compressor is used. DVRBD consists of two electrically driven Variable speed compressors, an intercooler and a throttling valve. Intercooler cools down vapour to low temperature. It cools down the compressed vapour from first compressor to the inlet temperature of the first compressor without any pressure drop and supplies vapour to the second compressor. The intercooler is assumed as ideal one. Intercooling helps to reduce the work done by the

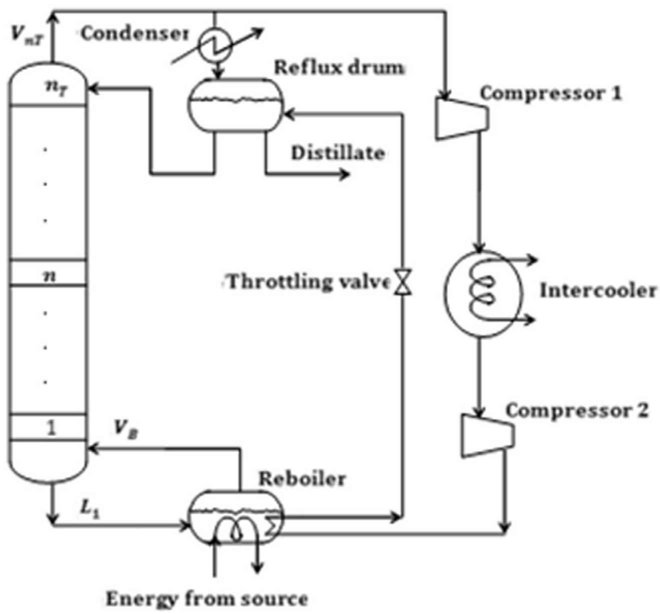

Fig 2. Schematic representation of Double Stage Vapour Recompressed Batch Distillation (DVRBD).

compressor to achieve required discharge qualities. Schematic representation of DVRBD is shown in Fig 2.

3. Mathematical modelling of vapour recompressed batch distillation

The equilibrium stage model of a batch distillation column consists of a set of ordinary differential equations (ODEs) coupled with algebraic equations/correlations. The assumptions for conventional batch distillation (CBD) are perfect mixing and equilibrium on all trays, constant Murphree vapour phase tray efficiency, fast energy dynamics, negligible tray vapour holdups, total condensation with no subcooling in the condenser, Non-linear Francis-weir formula for tray hydraulics, variable liquid holdup in each tray and Raoult's law for vapor-liquid equilibrium (VLE). Modelling equations [17] in CBD includes the MESH equations (Material balance, vapor-liquid Equilibrium, mole fraction Summation and Heat balance) of reboiler (subscript '*B*'), intermediate trays (subscript '*n*'), condenser and reflux drum (subscript '*D*') is shown in Table 1.

The time derivative of the multiplication of two variables, say *m* and *x*, is represented by $\dot{m}\dot{x}=\frac{d(mx)}{dt}$. For simulating this differential algebraic equation (DAE) system, the computational steps [17] have been sequentially followed in our work.

VRBD consists of all the modelling equations of a conventional distillation column with that consists of equations for compressor duty (Q_{Comp}) [18].

$$Q_{Comp} = 3.03 \times 10^{-5}\,\frac{\mu}{\mu-1}\,V_{n_T}\,P_i\left[\left(\frac{P_c}{P_i}\right)^{\frac{\mu-1}{\mu}} - 1\right] \tag{15}$$

in this equation, the pressure (inlet pressure P_i and outlet pressure, P_c) is in $\mathrm{lb_f/ft^2}$, and the vapour in flow rate to the compressor (V_{nT}) is in $\mathrm{ft^3/min}$. The polytropic coefficient of species *j* (μ_j) is temperature dependent and the value of μ is calculated from:

$$\frac{1}{\mu-1} = \sum_{j=1}^{C} \frac{y_j}{\mu_j - 1} \tag{16}$$

For the minimum work required by the compressor in DVRBD the discharge pressure of the first compressor is optimized that pressure is the inlet pressure to second compressor as intercooler has no pressure drop. It is the interstage pressure whose optimum value (P_{opt}) for minimum work is given by

$$\frac{P_{opt}}{P_i} = \frac{P_c}{P_{opt}} = (P_c \times P_i)^{\frac{1}{2}} \tag{17}$$

P_c is the discharge pressure of second compressor and P_i is the inlet pressure of the first compressor. Under the minimum work condition, compression ratio of both compressors are same. Then the compressor duty for DVRBD becomes:

$$Q_{Comp} = 2 \times 3.03 \times 10^{-5}\,\frac{\mu}{\mu-1}\,V_{n_T}\,P_i\left[\left(\frac{P_c}{P_i}\right)^{\frac{\mu-1}{2\mu}} - 1\right] \tag{18}$$

in order to ensure the optimal use of internal heat source under the VRC framework, in this section, an open-loop control algorithm is devised. This control strategy is synthesized with classifying the VRBD into two schemes, namely the fixed speed VRBD and the variable speed VRBD, and both of these schemes are presented below.

To perform a meaningful comparison between the VRBD and its conventional, the input conditions (e.g., feed charge and composition, and reboiler duty) and output specifications (e.g., product purity) to keep the same. Among them, the reboiler heat load, which is maintained constant throughout the entire batch operation. For complete condensation of overhead vapour (at T_{n_T}) in the reboiler (at T_B), it is reasonable to maintain a thermal driving force, ΔT_T $(= T_{n_{TC}} - T_B)$ of 20 °C [18], where $T_{n_{TC}}$ denotes the temperature of compressed overhead vapour. Accordingly, we select the two operating criteria for the VRBD as: (i) $\Delta T_T \geq 20$°C, and (ii) constant Q_R. In order to meet these objectives, now aim to formulate an open-loop control policy with a suitable variable manipulation mechanism.

(i) $\Delta T_T \geq 20$°C: Based on our assumption stated earlier, we can operate the heat integrated scheme with a bounded $\Delta T_T(\geq 20$°C) to ensure the complete condensation of vapour after compression. To attain the criteria, the column can also be operated with an exact ΔT_T of 20 °C. Accordingly, the VRBD operation into two modes as: fixed speed VRBD ($\Delta T_T \geq 20$°C) and variable speed VRBD ($\Delta T_T \geq 20$°C).

In the case of VRBD that should run at a fixed speed (i.e., at a fixed CR) with satisfying the criterion mentioned above, we first need to detect the time instant at which the $\Delta T_B(= T_B - T_{nT})$ is maximum. Actually, the highest ΔT_B corresponds to the maximum compression ratio requirement and the implementation of this CR can only ensure the batch operation with $\Delta T_T \geq 20$°C.

The following expression can be used to calculate the CR:

$$CR = \frac{P_c}{P_i} = \left(\frac{T_{n_{TC}}}{T_{n_T}}\right)^{\mu/(\mu-1)} \tag{19}$$

Obviously, $T_{nTC} = T_B + 20$°C. Recall that P_i and P_c represent the inlet and outlet pressure of top vapour, respectively, with respect to the compressor.

In the variable speed VRBD configuration, an attempt is made to run the column with an exact ΔT_T of 20 °C aiming to avoid the compressor operation at a maximum CR throughout the whole batch operation. Accordingly, the CR must vary at every time step because of the variation of both reboiler and top vapour temperatures. The same expression (i.e., Eq. (19)) can be used to manipulate the CR dynamically.

(ii) constant Q_R: This operating criterion is applicable to both the fixed speed as well as variable speed VRBD column. Thus, there is no need of separate manipulation policies to fulfil the second criterion. In the VRBD scheme, the compressed vapour releases heat (Q_{CV}), which in turn leads to the reduction of external heat input to the still (Q_E). It implies:

$$Q_R = Q_{CV} + Q_E \tag{20}$$

Table 1
Modelling equations in CBD includes the MESH equations.

	Equation	Equation	Equation number
Reboiler (subscript '*B*'):			
	Total mole Balance	$\dot{m}_B = L_1 - V_B = -D$	1
	Component mole Balance	$\dot{m}_B \dot{x}_{B,j} = L_1 x_{1,j} - V_B y_{B,j}$	2
	Energy balance	$\dot{m}_B \dot{H}_B^L = Q_R + L_1 H_1^L - V_B H_B^V$	3
	Equilibrium	$y_{B,j} = k_{B,j} x_{B,j} = \frac{\gamma_{B,j} P_{B,j}^0}{P_T} x_{B,j}$	4
	Summation	$\sum_{j=1}^{C} x_{D,j} = 1;\ \sum_{j=1}^{C} y_{D,j} = 1$	5
Intermediate trays (subscript '*n*'):			
	Total mole balance	$\dot{m}_n = L_{n+1} + V_{n-1} - L_n - V_n$	6
	Component mole Balance	$\dot{m}_n \dot{x}_{n,j} = L_{n+1} x_{n+1,j} + V_{n-1} y_{n-1,j} - L_n x_{n,j} - V_n y_{n,j}$	7
	Energy balance	$\dot{m}_n \dot{H}_n^L = L_{n+1} H_{n+1}^L + V_{n-1} H_{n-1}^V - L_n H_n^L - V_n H_n^V$	8
	Equilibrium	$y_{n,j} = k_{n,j} x_{n,j} = \frac{\gamma_{n,j} P_{n,j}^0}{P_T} x_{n,j}$	9
	Summation	$\sum_{j=1}^{C} x_{D,j} = 1;\ \sum_{j=1}^{C} y_{D,j} = 1$	10
Condenser and reflux drum (subscript '*D*'):			
	Total mole balance	$\dot{m}_D = V_{n_T} - R - D$	11
	Component mole Balance	$\dot{m}_D \dot{x}_{D,j} = V_{n_T} y_{n_T,j} - (R + D) x_{D,j}$	12
	Equilibrium	$y_{D,j} = k_{D,j} x_{D,j} = \frac{\gamma_{D,j} P_{D,j}^0}{P_T} x_{D,j}$	13
	Summation	$\sum_{j=1}^{C} x_{D,j} = 1;\ \sum_{j=1}^{C} y_{D,j} = 1$	14

Because of the unsteady state characteristics of batch column, the Q_{CV} should vary with time. Under this circumstance, there are two possibilities arise as: $Q_{CV} > Q_R$ and $Q_{CV} < Q_R$.

In the case **(**$Q_{CV} > Q_R$**)**, the latent heat (λ) released by the top vapour in the still is more than the heat required for liquid reboiling. It is experienced that the use of this extra heat (i.e., $Q_{CV} - Q_R$) does not necessarily improve the batch processing, particularly when the setup is run with optimal Q_R. Rather, it prolongs the startup operation because of the reboiling of relatively heavier fraction by that extra heat. This fact clearly demonstrates the necessity of overhead vapour (V_{n_T}) splitting so that a fraction of it ($V_{n_{TC}}$) can exactly alter the heat required by the still pot (i.e., Q_R) and the rest amount ($V_{n_{Ti}}$) can be directed to the overhead condenser. For this, the $V_{n_{TC}}$ can be calculated as:

$$V_{n_{TC}} = \frac{Q_R}{\lambda\,(\mathrm{at}\,T_{n_{TC}})} \tag{21}$$

Now, one can easily obtain the $V_{n_{Ti}}$ by subtracting $V_{n_{TC}}$ from V_{n_T}. It becomes clear that when $Q_{CV} > Q_R$, the VRBD requires to manipulate the overhead vapour splitting.

In the case of **(**$Q_{CV} < Q_R$**)**, the heat available from internal source is not adequate to run the column with the same dynamical features. Therefore, the balance of the heat requirements of the column (i.e., $Q_R - Q_{CV}$) is supplied by steam (an external heating medium) to the still. Obviously, in this case, the VRBD involves the manipulation of steam flow rate (m_S):

$$m_S\,\lambda_S = Q_E = Q_R - Q_{CV} \tag{22}$$

Where, the λ_S represents the latent heat of steam calculated at its temperature T_S.

It is worth noticing that this open-loop control law is formulated to compute a single manipulated variable for the fixed speed VRBD, i.e. either the vapor inflow rate to the compressor, $V_{n_{TC}}$ (when $Q_{CV} > Q_R$) or the auxiliary heat input to the still from an external source, Q_E (or m_S) (when $Q_{CV} < Q_R$). On the other hand, the variable speed scheme simultaneously adjusts the CR along with either the $V_{n_{TC}}$ or the Q_E. In present wok we have chosen variable speed than fixed speed because it gives more energy savings.

4. Results and discussion

In order to analyse the features of the proposed VRBD configuration, a simple example, separating a ternary mixture of hexanol, octanol and decanol is considered. We chose this system because it is a typical separation of compounds with wide boiling constituents. Initially the feed is charged in the reboiler, trays and reflux drum. the proposed VRBD column are simulated in MatLab environment with:

- a sampling instant of 0.0001 min,
- a tolerance limit of 10^{-4} for convergence

Table 2
Column parameters and specifications.

System	hexanol/octanol/decanol
Total feed charge, kmol	40.0
Feed composition (startup), mol fract	0.5/0.4/0.1
Tray holdup in each tray (startup), kmol	0.125
Reflux drum holdup, kmol	0.1
Murphree vapour-phase tray efficiency, %	75
Heat input to the still pot, kJ/min	4400.0
Distillate rate (fixed), kmol/min	0.064

The column consists of 11 trays numbered from bottom to top. Feed is filled in the reboiler, trays and reflux drum. The operating parameters and column specifications, summarized in Table 2.

The hexanol composition in distillate reaches the pre-specified concentration of 98% (at 13 min), the distillate withdrawal for hexanol (production 1 phase) continues till average distillate composition remains above the specified purity (98%). After the production 1 phase, slop cut 1 phase starts (26 min), this slop cut 1 phase continues till composition of octanol in distillate reaches 80% (436 min). As the composition of octanol in distillate reaches 80% the production 2 phase starts. This production 2 phase continues till the average composition of octanol in distillate reaches 80% (580 min). After the completion of production 2 phase, the column runs till the average composition of decanol in reboiler reaches 98% (at 625 min). The distillate composition profile is shown in Fig 3.

It is now necessary to determine what is the maximum value of ΔT_B, based on which the operating CR can be found out. For this, we produce Fig 4. that demonstrates the temperature profile throughout the batch operation. The difference between temperatures of reboiler and top tray vapours $\Delta T_B (T_B - T_{NT})$, is the driving force for the compressor. As ΔT_B increases the compression ratio increases which increases the work done by the compressor. By

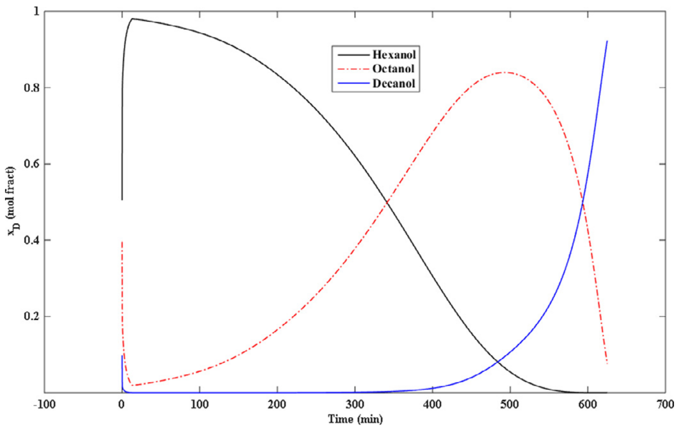

Fig 3. Distillate composition profile throughout batch operation.

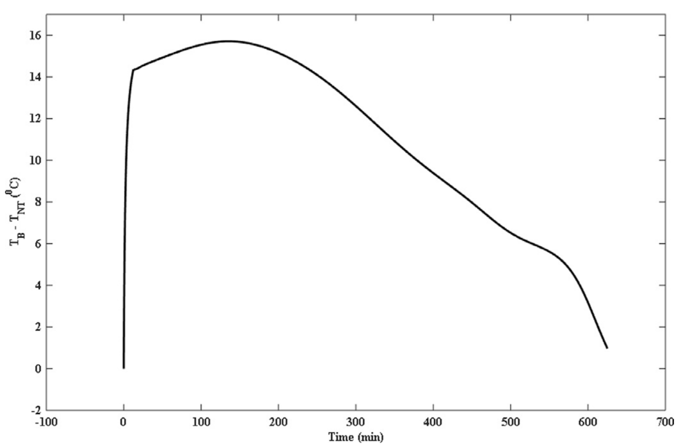

Fig 4. The temperature profile for batch column between reboiler and top tray.

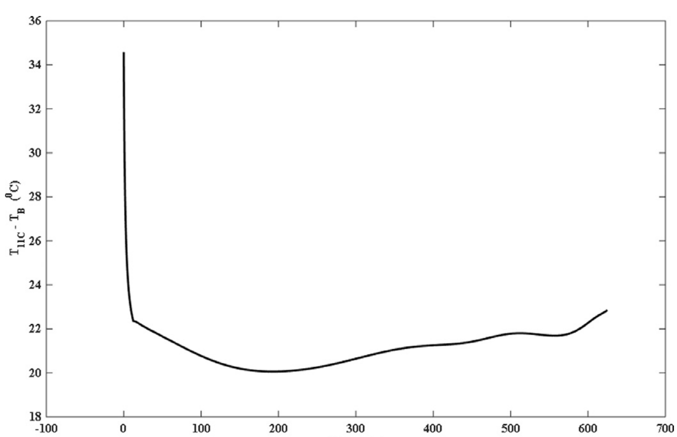

Fig 5. Variation of temperature difference between compressor outlet and reboiler in the VRBD at a fixed CR of 7.8.

sensitivity analysis found the highest value of ΔT_B in the process is 15.74 °C is shown in Fig 4. In subsequent work, have chosen variable speed CR than fixed CR because, the VRBD scheme is run with a variable speed compressor which changes its speed and compression ratio with every step time without working in highest compression ratio of the operation for the whole process.

By simulation analysis we have been fixed the Compression Ratio (C.R.) as 7.8 for SVRBD, which ensures that $(T_{11C} - T_B)$ $\Delta T_T \geq 20^0$C throughout the entire batch run. The Fig 5 shows the variation of ΔT_T for entire batch process.

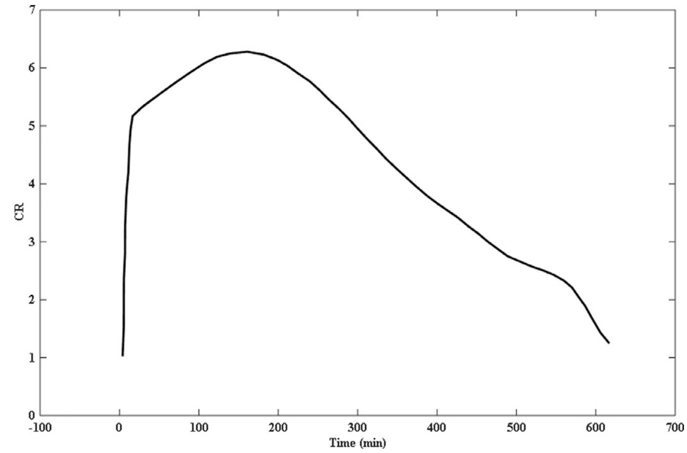

Fig 6. Compression ratio (CR) profile for batch column.

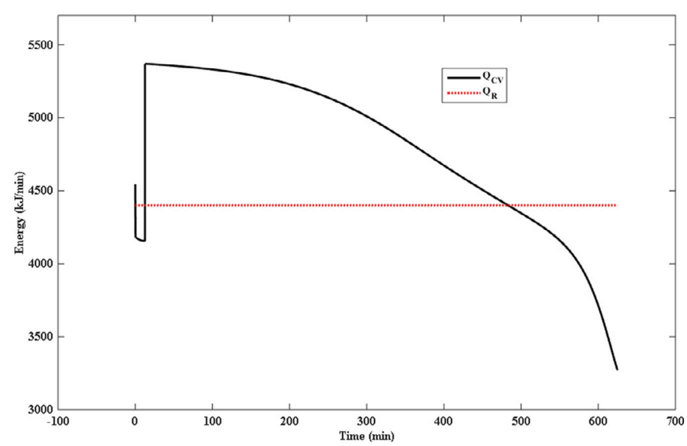

Fig 7. Energy available in the vapour outlet for SVRBD and DVRBD.

In SVRBD and DVRBD the maximum compressor ratio obtained are 7.8 and 2.79 respectively by sensitivity analysis. The work done by the compressor in DVRBD is less than the SVRBD scheme. The compression ratio of both VRBD is manipulated at each time instant such that there is a minimum temperature difference of 20^0C is maintained between the compressed vapour and reboiler. The CR profile is shown in Fig 6. for VRBD scheme. This temperature difference is to facilitate complete condensation of compressed vapour in the reboiler.

4.1. Energy savings

From Fig 7, energy available by the compressed vapour in most of the time instant of the process is higher than the constant reboiler duty. Supplying extra heat to the column results in increase in batch time. The extra energy of compressed vapour leads to the generation of more high boiling constituent vapours in distillate and change the dynamics of CBD. This will cause difficulty in the comparative study between CBD and VRBDs.

If the compressed vapour has higher energy than required in the reboiler ($Q_{CV} > 4400$ kJ/min), then the top tray vapour coming out from the column is splitted into two parts. The required amount of vapour is permitted to the compressor which gives the energy of 4400 kJ/min to the reboiler and remaining vapour is condensed in top condenser. The amount of vapour required to send to compressor is obtained by dividing reboiler duty with latent heat of compressed vapour required in that time step. If the compressed vapour has the required energy to run the reboiler then steam is not used in that step. But if compressed vapour fail to

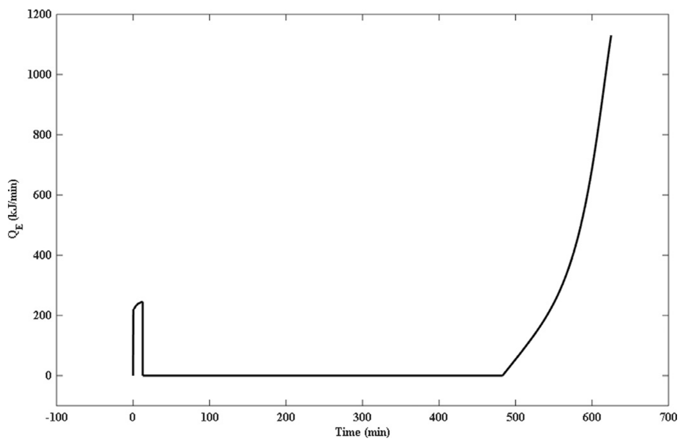

Fig 8. External heat input profile.

provide the whole heat for the reboiler ($Q_{CV} < 4400$ kJ/min), steam is used, with the compressed vapour to fulfil the reboiler requirement. The compressors used are variable speed compressors which change their speed and compression ratio at every time step such that a minimal temperature difference between reboiler and compressed vapours are maintained for complete condensation of compressed vapour in reboiler.

The external energy requirement is calculated by the following equation:

$$Q_E = (Q_R - Q_{cv}) \tag{23}$$

The external energy requirement from the source for the SVRBD is shown in Fig 8. As suggested by [19], it is logical to assume that 3 kW of thermal energy is needed to produce 1 kW of electrical power. Actually, the factor of 3 is determined by taking into account the cost of electricity etc. Accordingly, the total heat consumption in a VRBD (Q_{Cons}^{VRBD}) is estimated as:

The total energy consumption for VRBD is given by

$$Q_{cons}^{VRBD} = Q_E + 3Q_{Comp} \tag{24}$$

The factor three is for converting the compressor duty in to thermal energy which produces an equivalent amount of electrical power. It is determined empirically taking energy cost of electricity [12].

$$Energy\ Savings = \frac{\left(Q_{cons}^{CBD} - Q_{cons}^{VRBD}\right)}{Q_{cons}^{CBD}} \times 100 \tag{25}$$

The energy consumption for SVRBD, DVRBD and CBD obtained are 1.3729×10^6 kJ, 1.311×10^6 kJ and 2.75×10^6 kJ, respectively. Accordingly, with reference to the CBD, the heat integrated scheme SVRBD secures a significant energy savings of 50% and DVRBD 52%. It is observed that the reboiler heat demand can be met simply by the internal energy source and significant amount of external energy can be reduced.

4.2. Cost savings

Cost saving analysis of vapour recompressed batch distillation is done on yearly basis by total annualized cost. Total annualized cost (TAC) is calculated as (without interest rate),

$$\text{TAC}\left(\frac{\$}{\text{year}}\right) = \text{Operating cost}\left(\frac{\$}{\text{year}}\right) + \frac{\text{Capital cost}(\$)}{\text{Payback period}(\text{year})} \tag{26}$$

Here the capital investment include the costs of equipment's (i.e., distillation column, heat exchangers and compressor) whereas operating cost include all the utilities required for the plant (i.e., cooling water and steam to provide reboiler heat duty).

Cost estimating formula and parameter values [18] and Cost of utilities and index [15] are avaialble as Tables 3 and 4 in the supplementary material. The operating cost of the compressor is calculated as suggested by [18] based on the bhp (= hp/0.8) and a motor efficiency of 0.6. Here we assume the compressor efficiency of 0.8.

Comparison of capital costs and operating cost for TAC savings is shown in Table 5. Comparing the results, SVRBD secures a savings in TAC of 10.03% and DVRBD secures a savings in TAC of 12.21%.

Table 5
Comparison of capital and operating costs.

	CBD	SVRBD	DVRBD
Total capital cost($)	157,664.2	223,643.5	218,305.5
Total operating costs($/year)	21,674.54	11,318.53	11,037.1
TAC ($/year)	37,440.96	33,682.81	32,867.66
TAC savings (%)		10.03	12.21

5. Conclusions

By simulating a base case of ternary wide boiling mixture, hexanol/octanol/decanol in a batch distillation using vapour recompression scheme with a electrically driven single stage variable speed compressor and double stage variable speed compressors, the feasibility of VRBD is studied. Comparing with the conventional distillation column on energy savings and cost analysis, the potential of VRBD for separating ternary wide boiling constituents is analysed. The SVRBD scheme is effective to obtain a energy savings of 50% and TAC 10.03%.

The temperature difference between reboiler and top tray temperatures are very high, higher compression ratio are required which increases the compressor duty and there by energy consumed by the system. In order to reduce the compressor duty and protect compressor from wear and tear DVRBD scheme is introduced and this secures a energy savings of 52% and TAC savings of 12.21%

Supplementary materials

Supplementary material associated with this article can be found, in the online version, at doi:10.1016/j.reffit.2017.04.007.

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