Flash Zone Optimization of Benzene-Toluene-Xylene Fractionation Unit

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Abstract
Optimization of a typical fractionation unit and reduction of energy consumption has been carried out only by modifying the implementation and change in the flash zone of a tower. Previous studies have economized such issue by other methods which require additional equipments, hence reducing the cost effectiveness. It is theoretically illustrated and represented by simulation means that an insertion of the vapour product of the flash drum into upper stages of the distillation column separately instead of common procedure of combination with the liquid exiting the bottom of the flash drum can lead to energy conservation in reboiler duty, increased L/G of the top section of the column and to a reduction in the upper section diameter. A typical example of a BTX plant is used by provision of Hysys™ as a simulation.
Keywords: Optimization, Distillation, Flash Zone, Energy Consumption, Simulation

1. Introduction
Separation processes in many industries are by far the most dominating energy users to an extent that the operating costs of chemical productive processes are highly influenced by the downstream separation units. Fifty to eighty percent of the whole process operating cost can come from the separation trains [1], mainly distillation. Great effort and scrutiny research has been put to place on distillation which as a result has tremendously improved our understanding and our designs for enhanced tower capacity and efficiency.

Fractionation units exhibit such a vital role that they are addressed as the heart of any chemical plant and one of the most complicated operations among separation processes [2]. However the drawback in spite of their indispensable existence is the large energy consumption hence new design methodologies for increasing overall energy efficiency are all aimed to reduce the energy requirements. Furthermore, the cost of a fractionation unit comprises of the capital cost of the column, determined largely by the number of trays and diameter of the tower, and the operating costs, determined by the steam and cooling water requirements [3] and since over half of a plant's energy demand for either process heat or cooling is used in full operational distillation columns [4], consequently, optimizing the process with maximum energy recovery through more engineering time given to the design phase and retrofitting older operating plants to save or recover even minuscule amount of energy used can significantly affect the economic efficiency of a plant.

Therefore, the search for an optimum solution has been on quest and heuristic methods as feed splitting [1], introducing small changes in feed composition [2] and optimizing preheat train network of exchangers [5] has recently gained attention amongst researches. Although these proposals yield significant energy saving, however majority of them require additional equipments which increases the investment costs or due to operational complexity, their application is limited only to few basic feedstocks or even require fundamental changes which may not be possible in current existing full operational towers. It is to this extent that a significant breakthrough in practice of optimization of distillation not only demands approval of a cost-benefit trade-off but also provision of the optimization goal at minimal infrastructural change.

Nevertheless what has been failed to recall during all past efforts is the probable revision of the outline, topology and simple re-streaming of the process without causing a cardinal change in the process. The approach adopted in this work is the first to focus on the possibility of an improved energy efficient design of new towers and optimization of existing units considering minimum expenditure, through re-streaming of the process and making use of light components from the pre-fractionation unit to conserve energy in the column.

2. Fractionation Unit

2.1. Process Description and Topology
In a simplified conventional fractionation unit, energy is added to the reboiler at the base of the column while it is partly recovered in the condenser at the top. The basic equipments required for a standard continuous industrial fractionation unit are simply shown in Fig.1. It contains a feed stream F, a top product stream or distillate (D) and a bottom-residue stream (B). Side products can also be withdrawn from the intermediate trays where products of

intermediate composition and properties not obtainable merely by mixing distillate or bottoms with feed are desired [6]. Design of the column is based on the theoretical consideration that heat and mass transfer from stage to stage (theoretical) are in equilibrium [7] and as vapour and liquid are brought into contact on trays (Stages), vapour flows up the column while the liquid counter-currently streams downwards. Other equipments, a part of the tray section, are a condenser (with its associated heat duty, Qc), where part of the condensate is returned to the top of the column to provide liquid flow above the feed point (to maintain reflux), and a reboiler (with its associated heat duty, Qr), where part of the liquid from the base of the column is vaporized and returned to provide vapour flow.
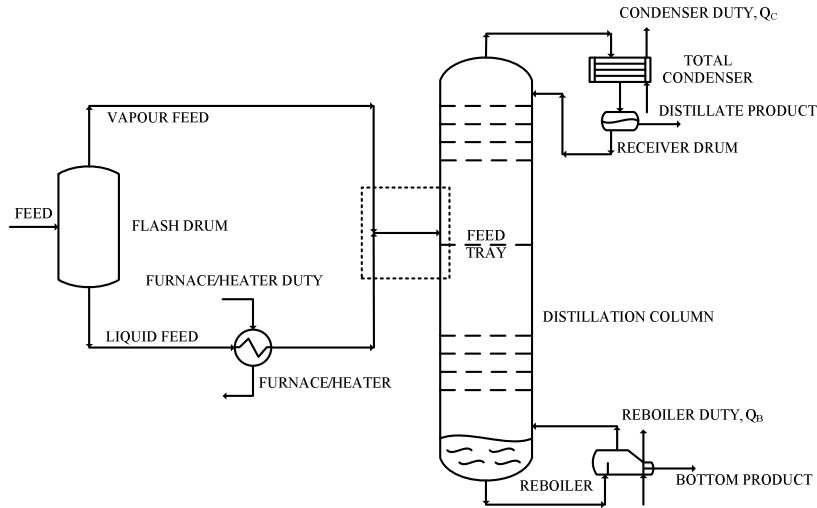

Fig.1 Common topology of a typical distillation unit

The conventional procedure for a simple separation can begin by introducing feed to a flash drum where light component(s) are vapourized and discharged off from the top and the liquid exiting the bottom. The bottom product will then be preheated either with the warm products of the tower using several heat exchangers or just by using a furnace (heater). The top product of the flash drum will meet up again with the preheated liquid upon entrance to the tower where they are inserted as one single stream.

Operating costs of industrial distillation towers mentioned above are usually dependent on the reboiler heat duty. Therefore, there has been a significant strive to reduce this amount during the past years. Several design alternatives can be proposed for more energy efficient considerations of operating distillation towers. Some of which requires additional capital investment like making use of feed/products exchangers for heat recovery, column revisions, better insulation or column control. In contrast to these changes, requiring small capital investment, there are other methods that will require more expenditure like addition of vapour recompression, heat pump or side heaters/coolers and condensers [8]. Methods as feed splitting and heating part of the feed have also been suggested by G. Soave and J.A. Feliu [1], through preheating by part of the feed and inserting the other part as cold to maintain low reflux ratio, however which requires additional equipments, but will yield a significant reduction in energy consumption of approximately 38%. It is to this extent that a significant breakthrough in practice demands approval of a cost-benefit trade-off.

What will be discussed here is basically modifying the basis of above implementation in order to reduce the investment costs and column heat duty, using minimum change in the existing topologies.

2.2. Proposed Soultion

As specified above, mentioned procedures will face several problems. The feed mixture will become more volatile, increasing the probability of vapour losses upon introducing the mixture to the tower. This is while the top product of the flash drum contains light end hydrocarbons which correlate more to the top of the column components, not to the middle or lower section (where feed is usually inserted). The usual topology will actually result in a higher reboiler heat duty as it requires additional elevation of light components to higher stage.

The proposal is to insert the top product of the flash drum in upper trays of the column separately instead of mixing it with the liquid exiting the bottom of the flash drum upon insertion to the distillation column (Fig.2).

Being composed of light hydrocarbons corresponding more to higher stages of the column, inserting this feed to upper trays results in an increase of purity of side stream products (if any exists) or column's main products (Distillate and Bottom). This is due to less contact of the vapour entering and the liquid falling down from upper

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trays. Eventually the rising vapour is less likely to experience condensation and be detected in the side stream products. Therefore the reboiler duty would decrease as no further boiling of light components would occur, consequently maintain the same purity of products with a lower reboiler duty.
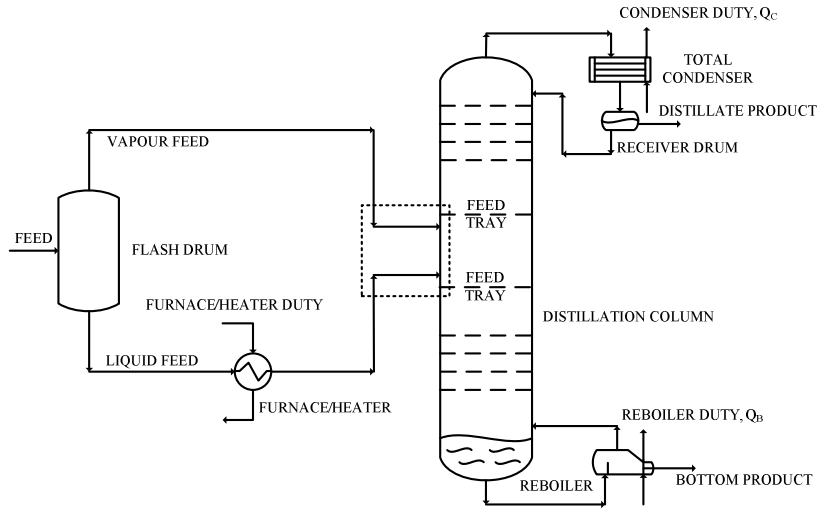

Fig.2. Proposed solution: inserting the vapour product of the flash drum in upper stages of the column separately

Another superiority of this solution is an increase of Liquid to Gas flows (L/G) between the two feed stages reducing the mid-up diameter of the column and further decreasing the investing costs. The ultimate result is and even more decrease of column diameter between the two feeds as shown in Fig. 3. The middle section diameter will be reduced following by an increase in the rectifying section as L/G is reduced even more (dashed lines).

$$\frac{L_m}{V_m} > \frac{L_r}{V_r} \qquad (1)$$

Therefore,

$$D_{BetweenFeeds} > D_{Rectifying Sectio} \qquad (2)$$
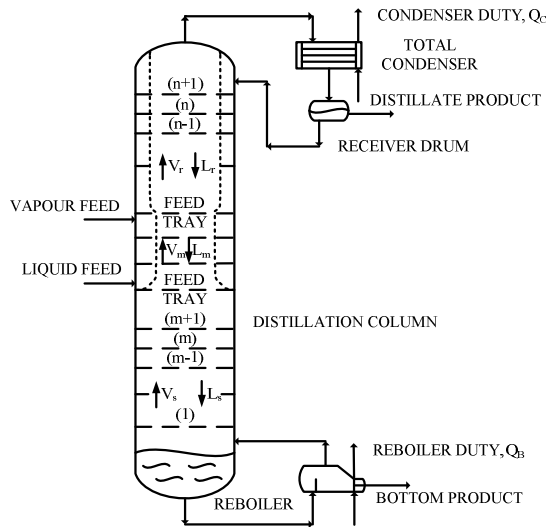

Fig.3. Detailed scheme of proposed topography; reducing mid-upper diameter

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However, if one also desires same pervious energy utilization in reboiler (or steam if available) and condenser can be considered, whereas in this case, higher purity of products will be achievable in compare to the basic common topology case. The first advantage of reducing reboiler energy consumption can be in an operating column while the second benefit of reducing investment costs can be made used of in the design phase of new distillation processes.

3. Case Study

3.1. Process Description

Since simulation and optimization requires a complete comprehension of the process, a brief description of the process unit is necessary and in order to give proof of the proposed solution, an industrial example of a BTX fractionation plant will be considered as in Fig. 4. Operating conditions of the process and feed composition are also given in Table 1 and 2. An industrial feed is also composed of non-aromatic and C6 to C10 compounds but have been neglected due to low values. The feed enters the flash drum, where due to sudden change in volume, a portion of the feed is vaporized while the rest exits the bottom following by entering the heat exchanger or a furnace. By acquiring the required heat and formation of yet another vapour phase, the feed exits the heater and mixes with the vapour product of the flash drum where the mixture is then introduced to the flash zone of the atmospheric distillation column (as considered in a common separation process). Typical columns in the regarding process contains 75 trays, a condenser and a reboiler. Further addition of trays will not change the duties significantly and this value is obtained based on an optimum condition. Hydrocarbon vapours escalades to tray number 14 and the liquids shed on the 15. For separating the heavy component present in the bottom of the atmospheric distillation column; that is toluene, a reboiler is used under the last tray (bottom of the column). Distillate stream is cooled down using a total condenser and partially returned to maintain reflux. Specifications determined for this separation are an achievement of 99.9% purity of benzene and a flow rate production of 25,750 kg/hr at distillate or 24,250 kg/hr at residue respectively in the benzene column.
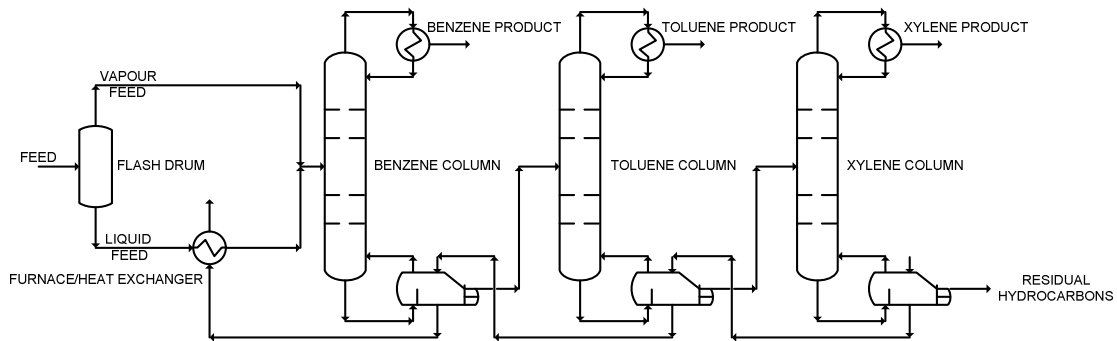

Fig.4. Typical fractionation unit of a BTX plant (Common topology of Pre-fractionation unit)

Table 1. Feed composition

Component	Mole Fraction (%)
Benzene	57
Toluene	27
Ethylbenzene	9.5
p-Xylene	1.5
m-Xylene	3.5
o-Xylene	1.5

Table 2. Operating conditions

Variable	Mole Fraction (%)
Feed Flow Rate (kg/hr)	0.67
Feed Temperature (^{o}C)	0.33
Feed Pressure (bar)	12
Column Top Pressure (bar)	1
Column Bottom Pressure (bar)	1.5

The problem that needs to be solved is how to take full advantage of the flash drum vapour product; a vapour mainly consisting of the light component (Benzene) and how to maximize its effects. The proposal presented in this work is to insert the mentioned vapour stream on a stage corresponding to similar tower composition, and exerting minimum reboiler energy consumption. Another objective of the vapour fraction would be to increase L/G in order to minimize the diameter and therefore reduce the capital cost of a tower. There is an optimum tray which corresponds to above mentioned advantages. The problem that needs to be solved is how to take full

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advantage of the flash drum vapour product; a vapour mainly consisting of the light component (Benzene) and how to maximize its effects. The proposal presented in this work is to insert the mentioned vapour stream on a stage corresponding to similar tower composition, and exerting minimum reboiler energy consumption. Another objective of the vapour fraction would be to increase L/G in order to minimize the diameter and therefore reduce the capital cost of a tower. There is an optimum tray which corresponds to above mentioned advantages.

3.2 Optimizing Feed Stage
The ideal feed point can be determined by graphical, shortcut, or rigorous techniques. Commercial simulations often incorporate search techniques [9] for seeking the optimum feed stage. These are usually based on minimizing an objective function. Small variations are introduced to the feed point, and their effect on top and bottom compositions is estimated by a rigorous or shortcut method. The results are substituted into the objective function, and the next trial begins with the feed entering a stage for which the objective function is lower.

Prior to this, column's composition profile of applied mixture will be given to obtain an initial estimation of the composition corresponding stage to that of the vapour feed exiting the top of the flash drum (Fig. 3). The composition of the vapour feed has been given in Table 1.
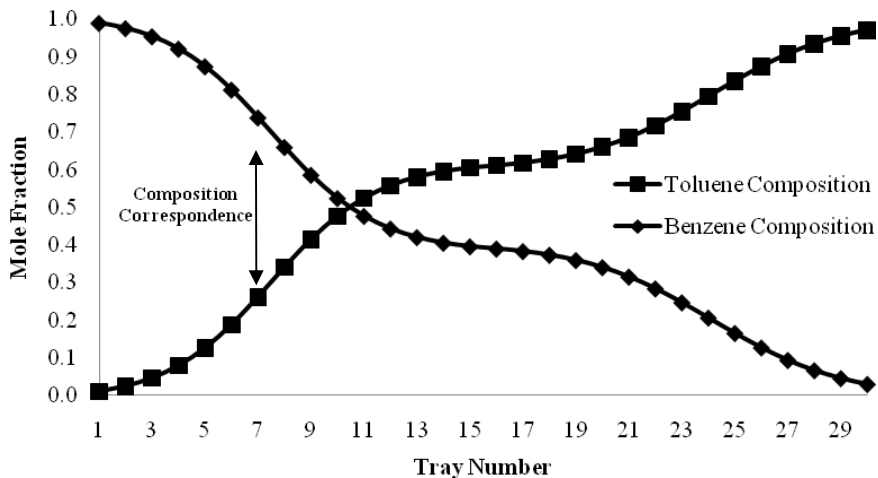

Fig.5. Benzene-Toluene composition profile along the tower

Table 1. Benzene-Toluene composition of the flash drum streams

Component	Vapour Stream Composition (Mole Fraction)	Liquid Stream Composition (Mole Fraction)
Benzene	0.67	0.46
Toluene	0.33	0.54

As seen, vapour stream composition of the flash drum is in correspondence with the mixture composition in the column at proximity of stage 8. Therefore an early standpoint would be to insert the vapour feed at this spot.

However relying only on composition similarity would not be adequate for any act on modifying the topology, hence other means for further confirmation and finding minimum reboiler duty reduction is vital. Consequently, results form a computer simulation can be plotted to determine the optimum feed stage and examine the precision of above approximation. Simulation runs are performed at several different feed points, keeping the material balance, reflux ratio, and total number of stages constant. For this case, the purity of the products will be held as a predetermined specification for the software to observe the variations in reboiler and condenser duty by change in the feed stage. A steady state simulator like HysysTM [10] [11] is the best tool to find such an optimum tray. Before initiation of finding the optimum tray one must find the optimum for the base case (Fig. 1) and that is to insert both feed at stage 15, yielding the minimum of both reboiler and condenser duties.

Now for finding the desired tray, after defining the feed using given information, represented equipments and completing the simulation using the constituent process conditions brought in the previous section, several vapour feed inlet stage runs has been performed and results have been gathered for reboiler duties shown graphically in Fig. 3. Vapour feed insertion to stages higher than tray 5 will give such high values that will not offer a convenient deducible picture and therefore have been neglected in the figure. However for the record, values of inserting this feed at further stages than 5 have been tabulated in Table 2.
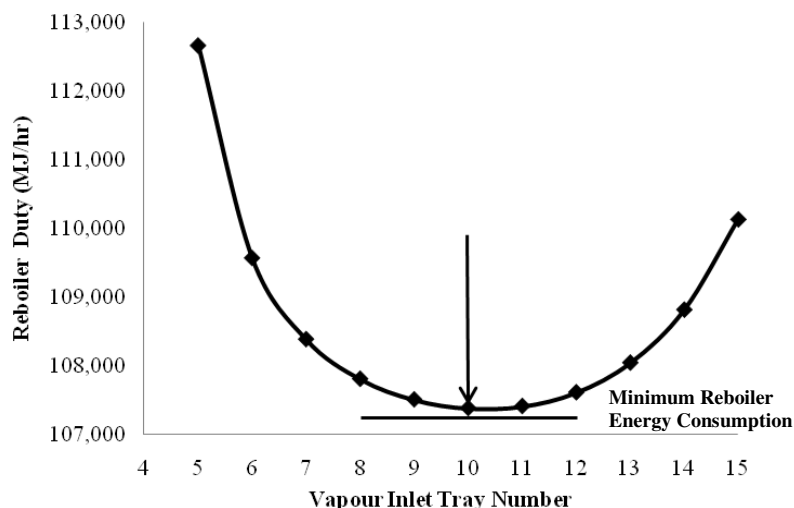
Fig.6. Results of optimization procedure using different runs of vapour feed inlet stage

Table 2. Values of vapour feed insertion at stages above tray 5

Tray No.	Condenser Duty (MJ/hr)	Reboiler Duty (MJ/hr)
Stage 1	567,000	509,200
Stage 2	315,000	256,000
Stage 3	219,000	161,000
Stage 4	182,000	124,000

As shown above, the optimum stage for this case yielding the minimum reboiler duty has been found on stage 10 (Fig.4). The simulated results having both feed inserted at one stage (Tray 15), will be considered as the base case, to compare to the alternative proposal of separate insertion of the feeds. Table 3 summarizes the result of such optimization performed in this work. Furthermore, the result is in concordance with the initial estimation as in Fig. 3 that at proximity of stage 8, minimum reboiler duty shall be achieved, now here stage at 10.

Table 3. Results of the comparative study

Case study name	Condenser Duty (MJ/hr)	Reboiler Duty (MJ/hr)
Base case (Fig.1)	168,000	110,000
Proposed Solution (Fig. 2)	166,000	107,000

By inserting the vapour feed to upper section of the column, it is possible to decrease the energy demand of the condenser down by 2000 MJ/hr, and at the same time, decrease further the reboiler duty by 3000 MJ/hr. Considering this optimum feed stage, it is possible to save 2.7% of reboiler duty and about 1.2% of condenser duty with respect to the base case. Needless to mention that although the amount might not be significant, however one must bear in mind that this energy saving has been obtained at no additional cost of implementing any sort of additional equipment or difficult expensive rerouting of the streams.

Finally, considering the fact that by insertion of vapour feed to upper stages, L/G will be increased in the upper section of the column, one can expect to design new columns having lower mid-upper section dimension than for using the ordinary topology, thus reducing capital or investment costs of new designs. For the example shown here on, the average L/G at 3 different sections of the column are given in Table 4.

Table 4. Liquid to Gas flows at different sections of the column

Column Section	$(L/G)_{ave.}$
Rectifying Section	0.65
Middle Section (Between the Liquid and Vapour Feed)	0.75
Stripping Section	1.53

4. Case Study

The decrease of L/G from the middle section to the rectifying part, results in an additional increase of the column diameter (however still lower than the stripping section) confirms the configuration given in Fig. 3.

Such proposal will yield different energy savings for different mixtures as have also been examined (not shown in this work), and it is very dependent on the initial feed vapour phase fraction. It is also possible to optimize further the energy consumption by utilizing other proposals as feed splitting [1], however accurate simulations must be carried out to be ascertain of the results. Although, one must not forget that an increase in capital and operating cost caused by any sort of improvements can be compensated by better sells due to the product quality improvement, but the determining key factor to the reinstate in these improvements is a comprehensive knowledge of the capital cost of the column, operating cost associated with utility requirement and cost of products for different quality, which by then optimal distillation column designs may be obtained.

With respect to the common topology of same stage feed insertion in distillation columns, inserting the vapour feed on a separate stage provides the following results:

1. Splitting the feed into two feeds of liquid and vapor upon vaporization in the flash drum maintains a simpler control of the process.
2. Decrease the energy demand of the condenser and reboiler to an extent that finding and optimized stage can achieve an energy conservation of 2.7% and 1.2% respectively.
3. Having L/G increased, same purity of products can be achieved with a smaller diameter tower. In fact, the fixed (capital) cost of a tower is reduced.
4. Since the flash drum vapour product contains component corresponding more to the upper stages of the distillation column, inserting this feed to the upper trays results in an increase of products purity with the same condenser or reboiler energy consumption.
5. Such topology requires negligible change in existing units for the introduction of vapour feed to an upper tray.
6. Further runs of Hysys^{TM} or another steady state simulator is required for other mixtures.

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