Feed-Splitting Technique in Cryogenic Distillation

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In distillation columns with condenser temperatures significantly below room temperature, such as in demethanizer towers, it is essential to minimize the expensive energy requirements of the refrigeration cycle that produces the tower reflux. A partial solution can be found by expanding the gaseous distillate to decrease its temperature; another solution is to use the cooler distillate to reduce the temperature of the tower feed. In this way, it has been shown that precooling only part of the feed brings about a significant decrease of its temperature, as well as decreases of the related minimum reflux ratio and finally the condenser duty. The fraction of the precooled feed has an optimum value that must be determined by means of a sensitivity analysis. By repeating the simulation of the tower and its auxiliaries, it is possible to find the optimal configuration, i.e., the configuration that minimizes the energy requirements at the condenser, by varying the feed-splitting ratio and by adjusting the tower geometry (i.e., identity of the feed trays) for each trial. A typical example to be applied in a demethanizer tower is provided here to illustrate the effect of feed splitting on the condenser and reboiler duties.

1. Introduction

Distillation is the most widely used operation in the chemical industry for the separation of mixtures of chemical species. It accounts for about 13% of the energy consumption of the industry as a whole and up to 23% of the energy use in specific sectors such as organic chemical production.[1]

Distillation is a heat-driven separation process in which the heat input at the reboiler (usually steam-heated) flows up the column and is expelled at a lower temperature in the overhead condenser. Most of the methods to improve the energy efficiency of distillation are focused on reducing the reboiler heat demand, as this is the main energy requirement.

In a previous article,[2] a process scheme was applied to reduce the reboiler heat duty of "warm" distillation columns, in which the column feed is preheated by the bottom product.

In this case, attention is focused on "cold" towers (i.e., towers having low temperatures in the condenser), where the main goal is to reduce the condenser heat duty, which is the main source of operating costs. This result is attained by partially precooling the feed(s) by the distillate, using a scheme symmetrical to that employed for warm columns.

2. Theory

The energy balance of a classical distillation tower with a condenser and a reboiler is given by

$$Q_C = Q_r + Fh_F - Dh_D - Bh_B \tag{1}$$

where Q_C is the condenser duty; Q_r is the reboiler duty; F, D, and B are the flow rates of the feed, distillate, and bottom product, respectively, and h_F, h_D, and h_B are the corresponding enthalpies.

In a cold tower, if the enthalpy of the feed is decreased by precooling with the distillate (Figure 1), two opposite effects can be achieved: (1) a decrease in the feed heat content, Fh_F,

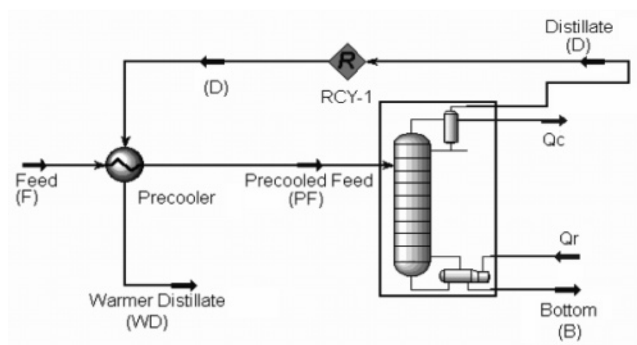

Figure 1. Distillation tower with complete feed precooling (the distillate is much colder than the feed).

which results in a decrease of Q_C, and (2) an increase of the reboiler duty that leads, in contrast, to an increase of Q_C.

The final result is still a decrease in the condenser duty, but the decrease is lower than the amount of heat transferred from the feed to the distillate.

This is evidenced clearly in a McCabe−Thiele diagram at minimum reflux conditions. By cooling the feed, the F point, representing the feed, moves upward, which results in a decrease of the slopes of the upper F−D working line (i.e., of the minimum reflux ratio) and the lower F−B working line (i.e., an increase of the minimum reboiler duty and, consequently, of Q_C according to eq 1).

The final result is positive, but it can be improved. The rationale of our proposal is that, by splitting the feed into proper proportions for two streams and by precooling only one stream (Figure 2), two positive effects can be obtained (see the corresponding McCabe−Thiele diagram in Figure 3): (1) the precooled stream, colder than before, would keep the minimum reflux ratio low, and (2) the warmer, not precooled, stream would keep the minimum reboiler duty at a low level.

In this case, the term Fh_F in the tower heat balance splits into two terms, $F_1h_{F1} + F_2h_{F2}$.

Furthermore, it can be observed that

$$F_1h_{F1} + F_2h_{F2} = Fh_F - Q_{PRE} \tag{2}$$

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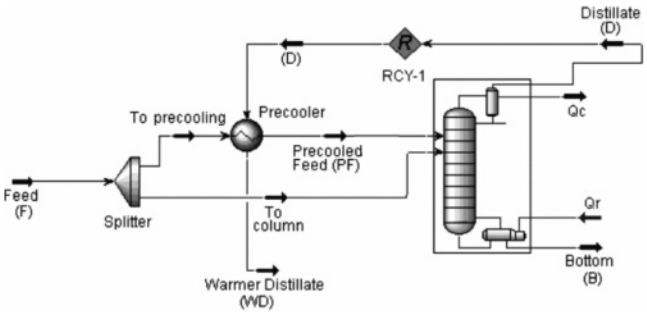

Figure 2. Plant configuration resulting from the application of the feed-splitting approach.

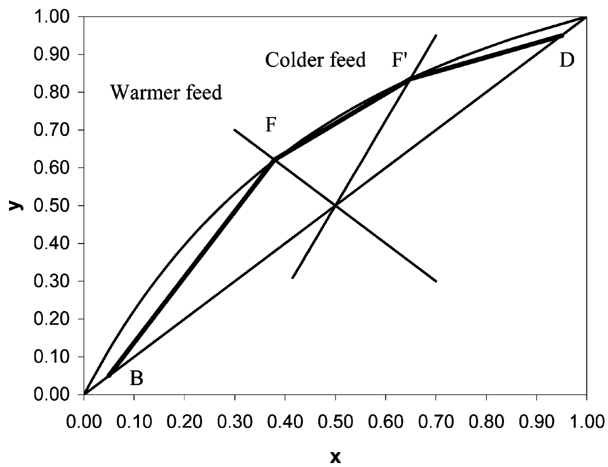

Figure 3. McCabe−Thiele diagram (at the minimum reflux ratio) for a distillation tower with two feed streams at different temperatures.

where Q_{PRE} is the duty at the precooler, that is, the heat exchanged between the distillate and the precooled stream.

A more complex mechanism is presented here that determines the best ratio of the flow rates for the two streams (i.e., the minimum condenser duty).

(1) By gradually decreasing the flow rate of the part of the feed that is cooled by the distillate, the respective temperature and, consequently, the minimum reflux ratio and the condenser duty decrease. Now, the minimum condenser duty is controlled by the temperature of the cold feed and decreases with the minimum reflux ratio according to the heat balance for the upper section of the tower (from the top to the “cold-feed” feed tray)

$$\begin{aligned} Q_{Cmin} &= V_{min}h_V - L_{min}h_L - Dh_D \\ &= D[R_{min}(h_V - h_L) + (h_V - h_L)] \end{aligned} \tag{3}$$

where Q_{Cmin} is the condenser duty at minimum reflux conditions; V and L are the flow rates of the vapor and liquid phases, respectively, of the cold-feed feed tray; D is the flow rate of the distillate; h_V, h_L, and h_D are the enthalpies of the vapor, liquid, and distillate, respectively; and R_{min} is the minimum reflux ratio. It should be noted that, for the minimum reflux conditions, V and L are in equilibrium. If the flow rate of the cold feed is high enough, the cooling action stemming from the distillate can be transferred completely to the feed. In this case, the total heat content of the two feeds remains constant, and according to eq 1, the minimum reboiler duty decreases by the same amount as the minimum condenser duty.

(2) A further decrease of the flow rate of the upper feed cannot recover any additional energy available from the distillate. The temperature of the cold feed still decreases, but at a lower rate, and the same happens to the minimum reflux ratio and to the condenser duty.

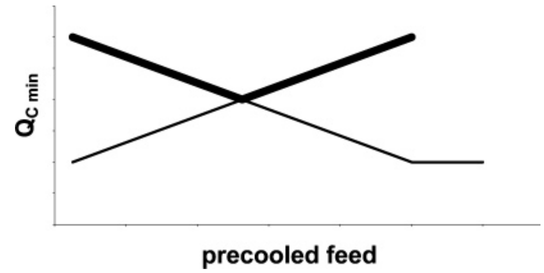

Figure 4. Minimum condenser duty as function of the feed-splitting ratio.

Table 1. Plant Feed

property	value	composition (mol/mol)	
T (°C)	−30	N_2	0.0100
P (bar)	60.00	CH_4	0.7500
F (kmol/h)	100	C_2H_6	0.1400
F (kg/h)	2177	C_3H_8	0.0600
$\alpha = V/F$	0.7195	i-C_4H_{10}	0.0100
		n-C_4H_{10}	0.0100
		i-C_5H_{12}	0.0100
		n-C_5H_{12}	0.0100

(3) Upon a further decrease of the fraction of cooled feed, a second value of the minimum reflux ratio prevails, which is determined from the minimum reboiler duty, which, in turn, is related to the temperature of the lower, warmer feed. The corresponding value of the minimum condenser duty is calculated from the tower heat balance, eq 1. The value of this duty increases with decreasing flow rate of the upper feed, because the total heat content of the feed (the term Fh_F in eq 1) increases.

(4) There is a value of the upper feed flow rate at which the two values of the minimum condenser duty are equal to a minimum value, corresponding to the optimal operating conditions (Figure 4).

3. Application

The above-described mechanism is applied to the simple case of a demethanizer tower (that has 30 theoretical trays and is operated at 30 bar), whose performance can be improved by the feed-splitting technique. The separation has to guarantee a distillate with 2% ethane and a bottom product with 1% methane.

The data on the feed to the system are reported in Table 1.

Now, different solutions of feed configuration can be chosen, taking into account that the liquid phase has a flow rate that is too low for good heat-exchange performance with the distillate.

In the first solution (solution A in Figure 5), the feed is split into two phases. The whole vapor phase is precooled by the distillate, expanded, and then fed to the column. The liquid phase is expanded and fed to the column.

In this configuration, it is necessary to find only the best feed trays. This can be done easily using HYSYS, release 3.2,[3,4] for the column simulation, by changing the position of each feed stream one at a time.

The column presents several feed configurations corresponding to a minimum condenser (and reboiler) duty. Then, the correct feed trays (that respect the column temperature profile) must be selected carefully. In these full-precooling conditions of the vapor-phase feed, the column requires a condenser duty of 96.26 MJ/h and a reboiler duty of 296.9 MJ/h.

The performance of this solution can be improved by precooling only a proper fraction of the vapor phase, as shown in solution B of Figure 6.

In this second solution, it is necessary to find the best feed trays and the best feed-splitting ratio (that is the percentage of the vapor phase to precool). This investigation was again

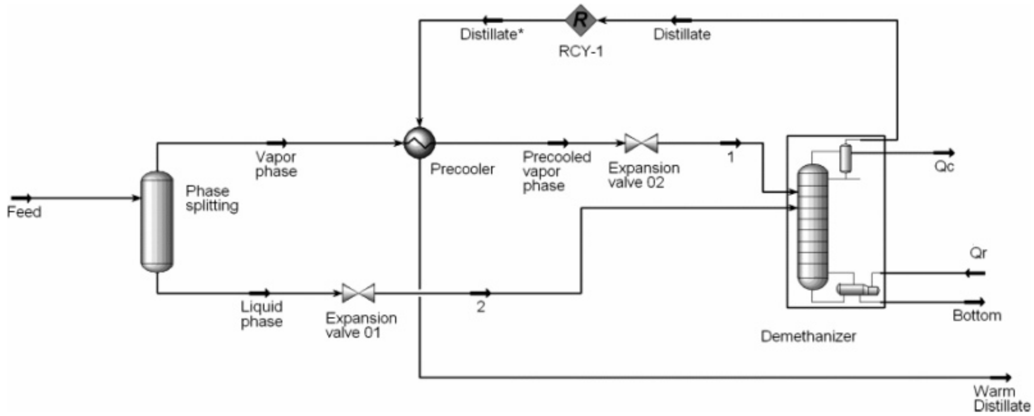

Figure 5. Solution A.

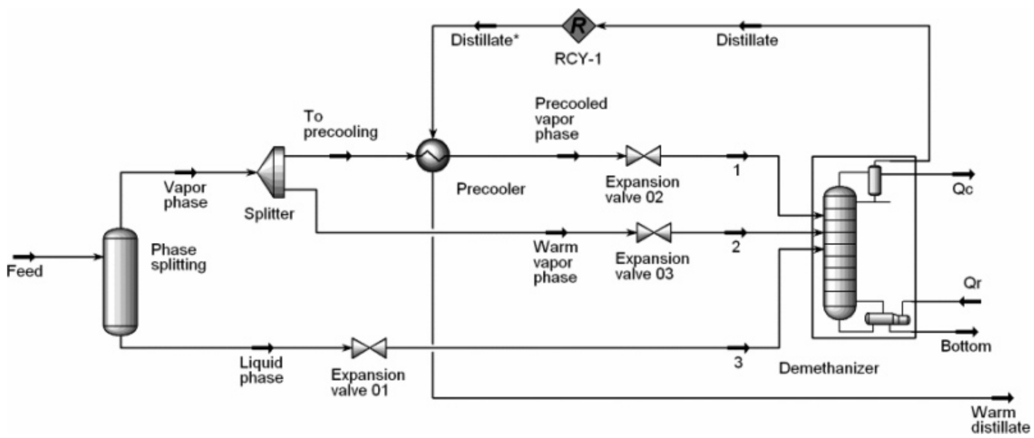

Figure 6. Solution B.

Table 2. Feed Configuration and Feed-Splitting Ratio Obtained by Optimization for Solution B

feed tray		property	value
stream 1	3	optimum feed split (%)	89
stream 2	15	Q_C (MJ/h)	−82.58
stream 3	22	Q_r (MJ/h)	281.5

conducted using HYSYS by simultaneously changing the feed-splitting ratio and the feed tray locations.

It should be noticed that, for all of the simulations of the precooler, the coefficient *UA*, i.e., the overall heat transfer coefficient multiplied by the exchange area, remains constant and equal to that of the solution A. Indeed, the theory of feed splitting was conceived to be easily applied to the revamping of an existing plant already using a "full-precooling" energy recovery system. Therefore, the same heat exchanger can be used for both full precooling and optimal partial precooling, without any additional fixed cost. At this level of approach, the coefficient *U* of the heat exchanger is also kept constant.

The study carried out to find the best feed-splitting ratio and the best feed trays for solution B leads to the results reported in Table 2.

For the best feed configuration, the condenser and reboiler duties as functions of the feed-splitting ratio are shown in Figure 7.

It can be observed that the condenser duty exhibits a minimum value (corresponding to 82.58 MJ/h) for the case of 89% of the vapor phase being subjected to precooling. Conversely, the reboiler duty remains more or less constant until the percentage of vapor phase subjected to precooling reaches the value of 89% (reboiler duty 281.5 MJ/h), and then it increases.

When applying the feed-splitting concept to the demethanizer column as discussed in this section, both the condenser duty

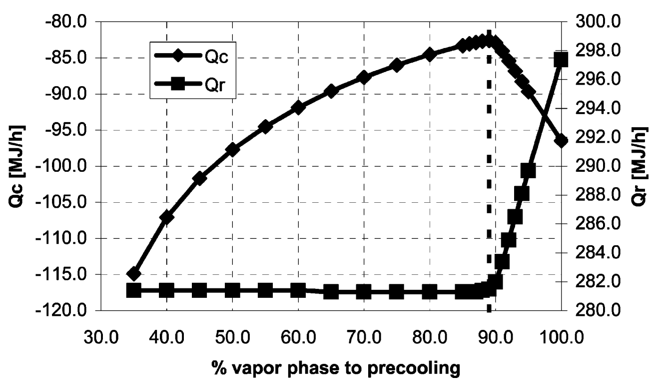

Figure 7. Feed-split optimization: Q_C and Q_r as a function of the percentage of the vapor phase subjected to precooling.

(main effect) and the reboiler duty (secondary effect) decrease under the same operating conditions [pressure, number of theoretical stages of the column, flow rate, and composition (i.e., temperature) of the distillate and bottom streams]. Furthermore, the application of feed splitting to an existing plant requires only the introduction of a bypass to the precooler and the replacement of the feed trays, which entails low fixed costs: this results in lower operating costs while the required investment is negligible. Therefore, this concept can be easily and efficiently used for the revamping of an existing plant.

4. Industrial Case

In an industrial example, the "feed-splitting" approach is applied to a demethanizer column of a "cold box" of an ethylene plant.

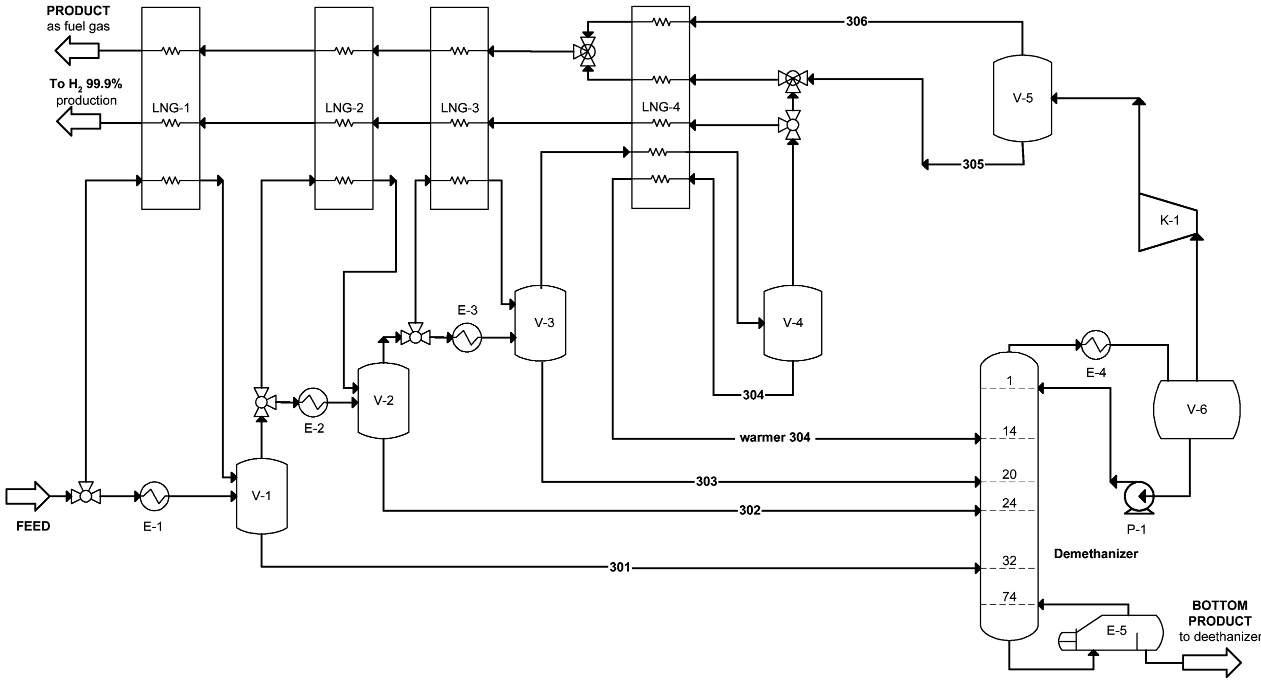

Figure 8. Base-case flowsheet after optimization.

Table 3. Plant Feed

property	value	composition (mol/mol)			
T (°C)	15	H_2	0.200	C_3H_4	0.006
P (bar)	43.60	CO	0.001	C_3H_6	0.076
F (kmol/h)	4865	CO_2	0.000	C_3H_8	0.002
F (kg/h)	104890	H_2S	0.000	C_4H_6	0.015
$\alpha = V/F$	1.0000	CH_4	0.274	C_4H_8	0.009
		C_2H_2	0.010	C_4H_{10}	0.001
		C_2H_4	0.357	C_{5+}	0.003
		C_2H_6	0.046		

4.1. Base Case. The configuration of the base case is shown in Figure 8. It can be observed that, before the demethanizer column, the feed (whose operating conditions are reported in Table 3) is split into four streams by a succession of partial condensations and phase splits. The condensations are performed by using refrigeration cycles and product/feed heat exchangers.

From the plant, three product streams (the first two at low ethylene content) are obtained: the stream product as fuel gas (1209 kmol/h) made up of 93.2% methane and 6.5% hydrogen; the stream to 99.9% H_2 production (1104 kmol/ h) made up of 80.9% hydrogen and 18.7% methane, and the stream bottom product (2552 kmol/h) made up of 68.0% ethylene and other heavier hydrocarbons.

To obtain the above-mentioned products, the distillation column must guarantee a distillate with 0.1% ethylene and a bottom product with 0.001% methane. Furthermore, to avoid a large loss of ethylene with the stream at high hydrogen content, the temperature of vessel V-4 must be −133.4 °C (with a pressure of 42 bar). In this way, the product for 99.9% H_2 production contains 0.1% ethylene.

The column has 74 real trays with a global efficiency of 76.5% (57 ideal trays), an overhead condenser pressure of 40.7 bar, and a reboiler pressure of 42 bar.

With the optimal feed tray configuration determined using HYSYS, reported in Table 4 and also presented in Figure 8, the demethanizer column requires a reboiler duty of 5103.34 $kJ/kmol_{FEED}$ and a condenser duty of 90.86 $kJ/kmol_{FEED}$.

4.2. Application of the Feed Split. The application of a feed split to this plant requires a special technique.

Table 4. Optimal Feed Configuration for the Base Case

feed tray		duty	value ($kJ/kmol_{FEED}$)
warmer 304	14	Q_C	690.86
303	20	Q_r	5103.34
302	24		
301	32		

A proper fraction of the vapor phases coming from vessel V-3 is fed directly to the column without passing through heat exchanger LNG-4 and vessel V-4. To maintain a constant temperature in vessel V-4, it is necessary to split both streams 304 and 306. Therefore, the proper fraction of the liquid phase coming from vessel V-4 is fed directly to the column and a proper fraction of stream 306 must bypass heat exchanger LNG-4.

It is possible to determine the optimal feed-split ratio for the vapor phase coming from vessel V-3, and the optimal feed tray configuration for the new case is reported in Figure 9.

It can be observed that, for the case of 89.2% of stream 303-vap using precooling (stream 304 is completely fed to the column without passing LNG-4, and stream 306 does not bypass LNG-4), the condenser duty is minimum and equal to 618.35 $kJ/kmol_{FEED}$, with a reboiler duty of 5044.23 $kJ/kmol_{FEED}$, as shown in Figure 10.

Table 5 provides a summary of the results obtained after optimization by applying the feed-splitting technique with HYSYS.

Otherwise, as a result of the application of the feed split, the two products with low ethylene content change their flow rate and composition: (a) the stream product as fuel gas (1322 kmol/h) is now made up of 86.4% methane and 13.4% hydrogen, and (b) the stream to 99.9% H_2 production (991 kmol/h) is still made up of 80.9% hydrogen and 18.7% methane. Indeed, the fraction of stream 306-vap fed directly to the column takes with it some hydrogen that, in the base case, was separated in vessel V-4. Because of the lighter distillate, the overhead condenser temperature decreases from −90.2 to −94.4 °C.

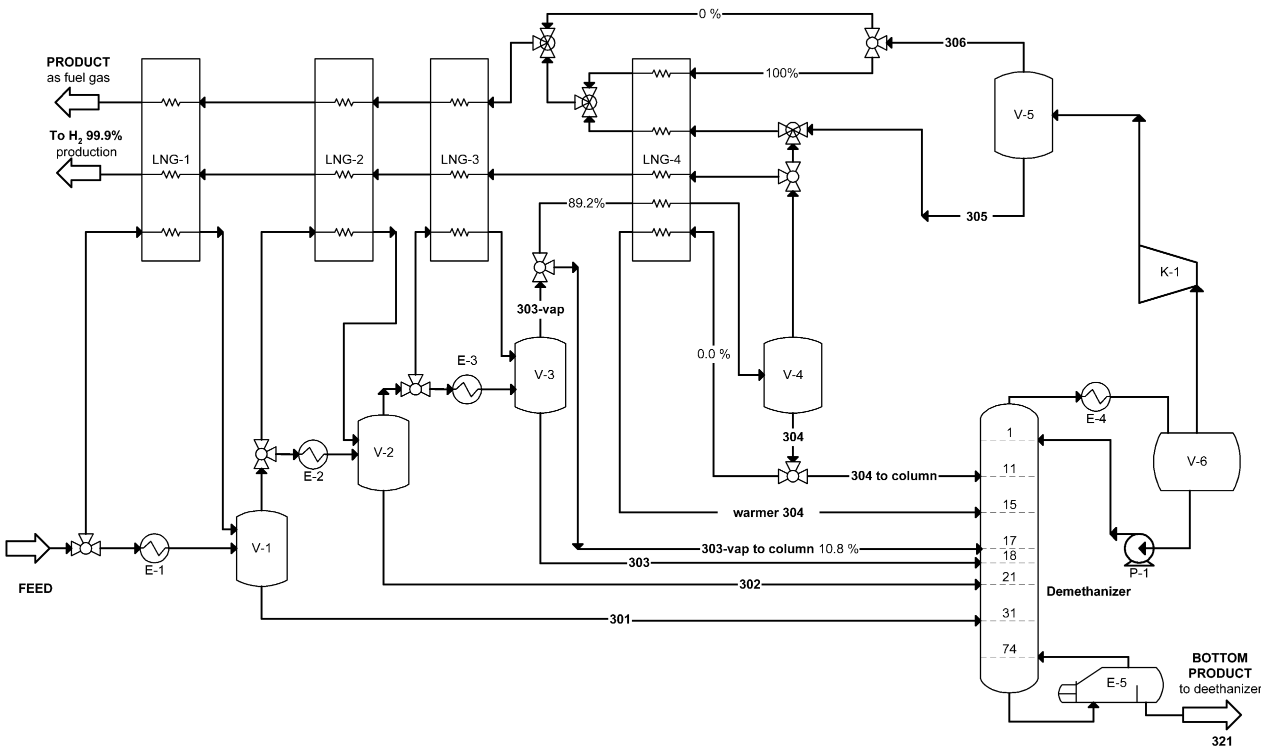

Figure 9. Feed-split configuration after optimization.

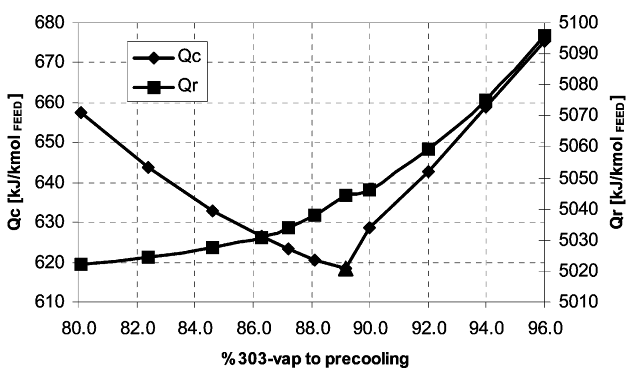

Figure 10. Feed-split optimization: Q_C and Q_r as a function of the percentage of stream 303-vap subjected to precooling (the triangle indicates the minimum value for Q_C).

Table 5. Optimum Feed Configuration with a Feed Split Applied

feed tray		duty	value (kJ/kmol$_{FEED}$)
304 to column	11	Q_C	618.35
warmer 304	15	Q_r	5044.23
303-vap to column	17		
303	18		
302	21		
301	31		

Table 6. Performance Comparison for the Industrial Case

	303-vap to precooling (%)	Q_C (kJ/kmol$_{FEED}$)	Q_r (kJ/kmol$_{FEED}$)	ΔQ_C (%)	ΔQ_r (%)
base case	100	690.86	5130.34	0	0
feed split	89.2	618.35	5044.23	**−10.50**	**−1.68**

5. Conclusions

Considering the data reported in Table 6, the advantages of applying the feed-splitting concept to an industrial plant are highlighted clearly: the outcome is a decrease of both the condenser duty (main effect) and the reboiler duty (secondary effect).

However, even if the feed split leads to energy savings, it is necessary to evaluate the incidence of the loss of 91.4 kmol/h (−10.2%) of hydrogen in the product. A balance between the loss of revenue resulting from the decrease in hydrogen production,[6] the decrease in costs from the diminished energy consumption at the condenser, and the revenue from the increased production of fuel gas because of the hydrogen addition[7] shows that the new process scheme is profitable.

The recovery of ethylene in the bottom product does not change.

Another advantage derived from the application of feed-splitting to this plant is the elimination of a heat exchanger (stream 304 is completely fed to the column) from LNG-4. No changes occur for the other LNGs.

Furthermore, an increased distillate flow rate can provide more electricity in expander K-1.

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