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Economic feasibility of heat pumps in distillation to reduce energy use

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ABSTRACT

An *i*-butane/*n*-butane mixture was selected to analyze several distillation assisted heat pump processes when compared to conventional distillation. This conventional process, along with top vapour recompression, bottom flashing and absorption heat pumps, were simulated using the HYSYS software platform, in order to determine economically the best alternative.

Distillation with both top vapour recompression and bottom flashing heat pumps allows reduction of operation (energy) costs by 33% and 32%, respectively. This improves the economic potential (incorporating capital costs) by 9% and 10%, respectively. Due to the large steam consumption, when compared to the conventional case, the absorption heat pump is not suitable for this system.

1. Introduction

1.1. Background

Distillation is one of the most important separation methods both in chemical and petrochemical industries. It is estimated [1] that there are about 400,000 distillation columns in operation in the United States, which is about 90% of all separation processes in product recovery and purification. However, this process involves large energy consumption due to the heat to be supplied to the bottom reboiler, and the heat to be removed from the top condenser. Mix et al. [2] found that 60% of energy used by chemical industry was for distillation. For this reason, any way of reducing this energy consumption would provide a great benefit.

Although there are other alternatives [3], the introduction of a heat pump cycle to a distillation column has significant potential because with this system, the energy of the cold top stream is employed as energy supply for boiling the hot bottom stream.

The most popular heat pumps systems are the mechanical and absorption heat pumps. In the former, instead of using a separate condenser and reboiler, the top product can be compressed to a higher pressure and used to heat the bottom product, or the bottom product can be flashed in a valve and used to cool the top product. In absorption heat pumps, a separate closed loop fluid system (ammonia/water or lithium bromide/water are the most commonly employed) is used to transfer the heat up the temperature scale by means of heat of mixing. In these systems, the salt is used as the refrigerant and water as the absorbent.

Mechanical heat pumps systems have been applied to difficult separations. Fonyo et al. [4] used the heat pump concept in a C_4-splitter and found that in all cases the costs were lower when compared with conventional distillation. Quadri [5] optimised the design of a propylene/propane system using single and double stage vapour recompression systems, and, when Annakou and Mizsey [6] studied the same system, they found that when using either a single or double stage vapour recompression system, the annual costs could be reduced by 37%. Ferre et al. [7] applied a direct vapour recompression heat pump to an ethylbenzene/xylene separation and to an ethylbenzene/styrene separation; both cases reduced energy consumption.

An absorption heat pump was described by Davidson and Campagne [8] as an absorption refrigeration system redesigned for use at temperatures entirely above ambient. With this kind of heat pump, Tufano [9] estimated that a 40% energy saving could be reached. A recent application of absorption heat pumps is described by Aristov et al. [10]. Currently, absorption heat pumps are applied in several processes such as desalination [11,12], and Hektor and Berntsson [13] have applied a two stage compressor heat pump system in restoring the absorbent used to clean the flue gases in a pulp mill.

A comparison between both absorption and mechanical heat pumps was made by Fonyo and Benko [14]. They simulated five different processes and in all cases they found that distillation processes with a large temperature difference are suitable for absorption heat pumps, where mechanical heat pumps cannot be used.

There are also some distillation processes which employ both mechanical and absorption heat pumps which have been patented.

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Meszaros [15] patented a distillation plant with a vapour recompression heat pump which could be used successfully to separate a mixture of *i*-butane/*n*-butane/*c*/*t*-2-butene. With the mechanical heat pump patented by Meili [16], styrene can be separated from a mixture containing the more volatile ethylbenzene and small amounts of benzene and toluene. Erickson [17] patented an absorption heat pump system which could be added to an ethyl chloride/dichoroethylene separation system.

1.2. Objective

The objective of this work is to simulate an *i*-butane/*n*-butane distillation process to compare the costs of the conventional distillation with the cost of a heat pump distillation system. We chose this system because it is a typical separation of compounds with close boiling points, where the energy costs of the conventional process is high and the heat pump application is feasible.

Three different configurations are considered (vapour recompression, bottom flashing and absorption heat pump) to determine the best alternative to the conventional distillation. All the simulations were undertaken with the HYSYS model version 2004.2 under license from Aspentech [18].

2. HYSYS simulation of the distillation column systems

2.1. Property package for HYSYS

To develop this analysis, the Peng Robinson property package was used. This equation of state model is adequate to predict the equilibrium of light hydrocarbon mixtures as in the case of this work [19].

Concerning the ammonia/water system used for the absorption case, in the literature there are several studies [20,21] of thermodynamic equilibrium of this mixture with Peng Robinson model. They show that the Peng Robinson property package is suitable for this system.

2.2. Conventional column

The flow diagram of the conventional scheme is shown in Fig. 1. To compare the advantages of introducing a heat pump system into a conventional distillation column, 100 kmol/h of an equimolar binary mixture of *i*-butane/*n*-butane was fed to the column. The inlet stream was supplied as a saturated liquid at 710 kPa pressure; the mole fraction of *i*-butane in the top product was specified as 0.9, and the mole fraction of *n*-butane in bottom product was specified as 0.9. The column for this system was initially set up using the short-cut column design facility to obtain an initial estimate for the number of trays required and the reflux ratio needed in the column. The column was then simulated with the rigorous column facility which converged successfully.

To determine optimal conditions, the top column pressure was varied between 500 and 1000 kPa, while maintaining the ratio R/R_{min} at 1.3. The lower pressure limit was 500 kPa because smaller values would significantly increase costs due to the requirement to use a refrigerant fluid, instead of air, for the coolers. In the same way, the ratio R/R_{min} was varied between 1.1 and 1.5 while keeping top column pressure at 700 kPa. In all simulations column pressure drop was kept constant at 20 kPa.

For a top product pressure of 700 kPa and a ratio R/R_{min} of 1.3, top column temperature is 52.24 °C and bottom column temperature is 63.65 °C. In this case, 33 theoretical stages are needed to reach the required separation. Table 1 shows how the number of theoretical stages varies with R/R_{min} and with column pressure.

In all simulations in this paper the feed is supplied at the same condition, and the product streams are required as saturated liquids at 700 kPa. Cooling is provided by air-cooled heat exchangers. Assuming an air temperature of 25 °C, the minimum process temperature is taken as 40 °C. No energy losses are assumed in these systems.

Klemola and Ilme [22] reported data from an industrial *i*-butane/*n*-butane fractionator working with an optimal column top pressure of 658.6 kPa. This is in accord with the value obtained in this work for the optimal top column pressure. As an example, in Table 2, temperature and pressure of the main streams of the process are shown, for the case when top column pressure is maintained at 700 kPa, and R/R_{min} at 1.3.

2.3. Distillation column with top vapour recompression heat pump

The flow diagram of the top vapour recompression scheme is shown in Fig. 2. The top column outlet stream is compressed with compressor [23] (K-100) to raise its temperature and promoting its energy content to be "more usable". When the top column pressure is 700 kPa and R/R_{min} 1.3, the temperature is increased from 52.6 to 88.5 °C and also the pressure is increased from 700 to 1540 kPa. The compressor polytropic efficiency was assumed to be 70%. After the compressor, the heat exchanger E-100 allows transfer of the energy of this stream to boil up the bottom column outlet stream. With the same top column pressure and reflux ratio

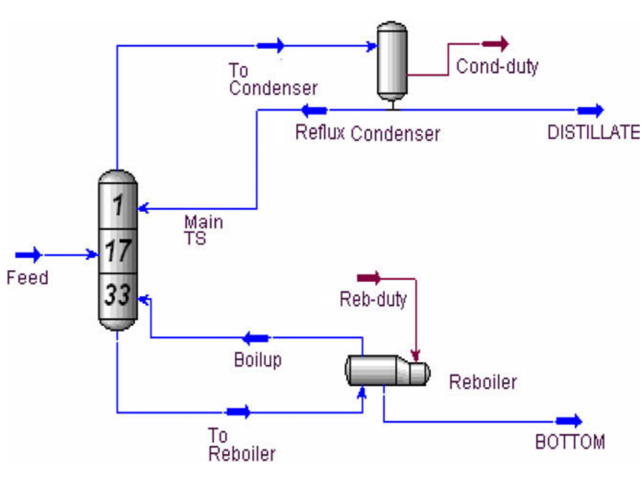

Fig. 1. HYSYS process flow diagram for the conventional column.

Table 1
Optimal theoretical stages and feed stages for different column pressures and reflux ratios

R/R_{min} = 1.3			Top pressure = 700 kPa		
Top pressure (kPa)	Theoretical stages	Feed stage	R/R_{min}	Theoretical stages	Feed stage
500/520	30	15	1.1	47	23
600/620	32	16	1.2	38	19
700/720	33	17	1.3	33	17
800/820	35	17	1.4	31	15
900/920	37	18	1.5	29	14
1000/1020	39	19			

Table 2
Conditions of the main streams of the conventional distilaltion process with column top pressure equal to 700 kPa and R/R_{min} ratio equal to 1.3

Stream	Temperature (°C)	Pressure (kPa)
Top product outlet stream	52.24	700
Bottom product outlet stream	63.65	720

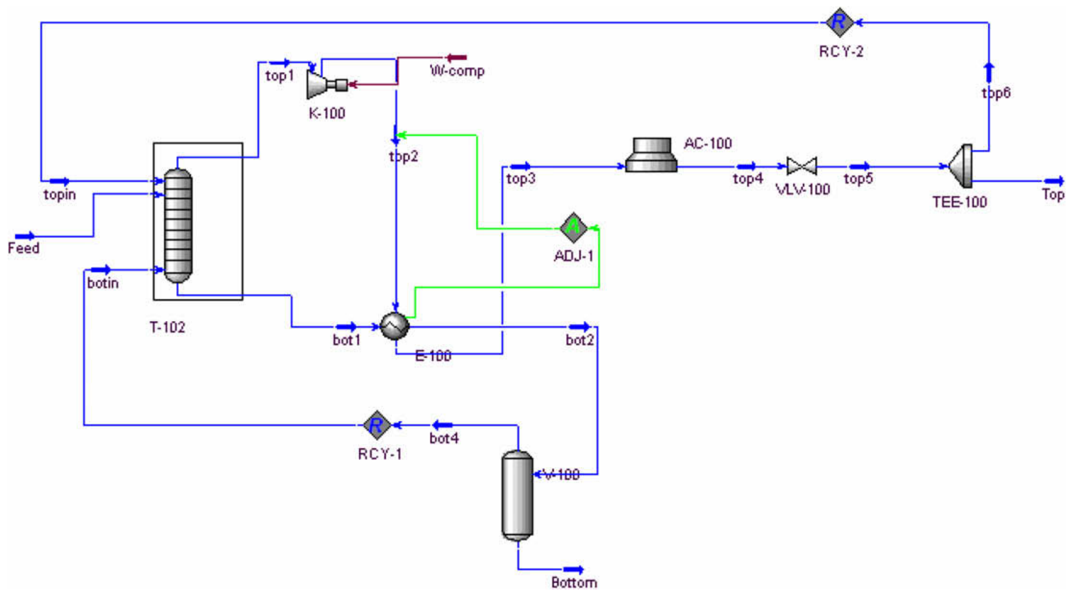

Fig. 2. HYSYS process flow diagram for the vapour recompression heat pump.

as before, the compressor outlet stream is condensed and cooled to 68.4 °C, while the bottom column outlet stream is partially boiled.

The adjust unit in HYSYS is used to calculate the outlet compressor pressure to obtain a minimum approach in the heat exchanger of 5 °C. This is a typical value for this kind of heat exchanger [24] and it is a compromise between operational and capital costs. Note that the following heat balance applies:

$$Q_C \cong Q_R \cong Q_{E\text{-}100} \quad (1)$$

$$Q_{K\text{-}100} \cong Q_{AC\text{-}100} \quad (2)$$

where Q_C is the conventional column condenser energy, Q_R is the conventional column reboiler energy and Q_i is the energy transferred in item *i*.

In spite of the exchange, the top column outlet stream must be further cooled before being recycled to the column. With the same top column pressure and reflux ratio as before, the top column outlet steam is air-cooled (AC-100 block) to 52.3 °C. This stream is then divided in two in TEE-100. One outlet stream is the final top product and the other one is recycled back to the column. After E-100 heat exchanger, the bottom column outlet stream is divided in V-100 flash drum. The vapour outlet stream is recycled back to the column, and the liquid outlet is the final bottom product stream.

As in the case of conventional distillation, the top column pressure was varied between 500 and 1000 kPa, while maintaining the ratio R/R_{min} at 1.3. In the same way, the ratio R/R_{min} was varied between 1.1 and 1.5 while keeping top column pressure at 700 kPa. Note that 'direct' heat exchange between the top and bottom streams is more thermodynamically efficient than via a separate heat pump fluid. Table 3 shows stream temperatures and pressure for the case when the top column pressure is 700 kPa, and R/R_{min} is 1.3.

Table 3
Conditions of the main streams of the top vapour recompression heat pump case with column top pressure equal to 700 kPa and R/R_{min} ratio equal to 1.3

Stream	Temperature (°C)	Pressure (kPa)
Top column outlet stream	52.64	700
Bottom column outlet stream	63.41	720
Compressor outlet stream	88.49	1540
Top product outlet stream	52.32	700
Bottom product outlet stream	63.71	720

2.4. Distillation column with bottom flashing heat pump

The flow diagram of the bottom flashing scheme is shown in Fig. 3. The bottom column outlet stream is expanded in VLV-100 valve to decrease its temperature and allow heat exchange with the top stream in E-100. When top column pressure is 700 kPa and R/R_{min} 1.3, the temperature of this stream is decreased from 63.5 to 37.0 °C and the pressure is decreased from 720 to 360 kPa. Heat exchanger E-100 enables boiling the bottom column outlet stream and top column outlet stream condensation. After the heat exchanger, the bottom stream must be recompressed to the column pressure in K-100 compressor. Hence its temperature is increased to 63.9 °C, and it must be slightly air-cooled before being recycled to the column (AC-100 block).

As in the case of conventional distillation, the top column pressure was varied between 500 and 1000 kPa, while maintaining the ratio R/R_{min} at 1.3. In the same way, the ratio R/R_{min} was varied between 1.1 and 1.5 while keeping the top column pressure at 700 kPa.

Table 4 shows stream temperatures and pressures for the case when top column pressure is maintained at 700 kPa, and R/R_{min} is 1.3.

2.5. Distillation column with absorption heat pump

The flow diagram of the absorption scheme is shown in Fig. 4. An ammonia/water refrigeration cycle is used both to boil up the bottom column outlet stream and to condense the top column outlet stream.

The absorption heat pump comprises a regenerator (T-103), a condenser (T-103 condenser + E-103), an expansion valve (VLV-100), an evaporator (E-100), an absorber (E-102) and a pump (P-100). The absorption fluid consists of a mixture of ammonia and water. The regenerator separates the ammonia and water so the top product is almost pure ammonia, which acts as refrigerant, and the bottom product is an ammonia/water mixture which is recycled to the absorber. The regenerator has been simulated with a distillation column model so the duty taken out of the column

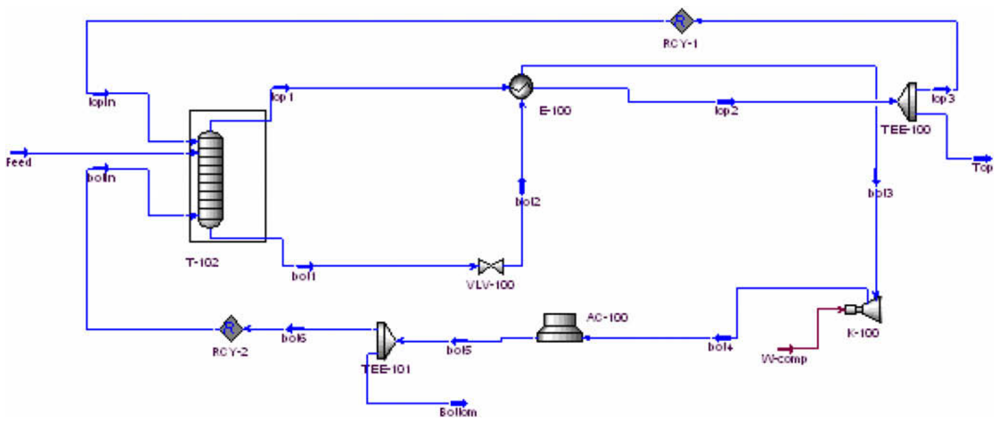

Fig. 3. HYSYS process flow diagram for the bottom flashing heat pump.

Table 4
Conditions of the main streams of the bottom flashing heat pump case with column top pressure equal to 700 kPa and R/R_{min} ratio equal to 1.3

Stream	Temperature (°C)	Pressure (kPa)
Top column outlet stream	52.60	700
Bottom column outlet stream	63.51	720
Valve outlet stream	37.03	360
Top product outlet stream	52.29	700
Bottom product outlet stream	63.83	720

condenser is used as a heat source in E-103 ensuring there is sufficient temperature difference. The energy is supplied to the regenerator as saturated steam.

The top stream is split into two streams. One of them is condensed with an external heat source and the other one is condensed by giving its heat to the evaporator (E-100) of the absorption cycle. After that, the two condensed streams are joined again. Part of the condensate is the final top product and part of it is recycled again to the column. The bottom product stream is boiled in a first step by taking heat from the condenser (E-103) of the absorption cycle, and in a second step by taking heat from the absorber (E-102).

In this case, only R/R_{min} equal to 1.3 and top pressure equal to 700 kPa case was simulated, because the use of the absorption heat pump was found to be less economical than the conventional distillation.

3. Economic evaluation of all alternatives

3.1. Economic analysis-procedure

The economic evaluation of each system was carried out by estimating the economic potential (EP), in terms of a combination of capital and operational costs, according to Jones [25] expression:

$$EP = C_v + C_f + (i_r + i_m) \cdot FC \tag{3}$$

where C_v is the process variable costs, in this case, related to utility consumption per year; C_f is the annual fixed costs; FC is the fixed capital investment; i_r is fixed capital recovery rate applied to FC, and i_m is the minimum acceptable rate of return on FC.

In this work, see also [26], C_f was assumed to 10% of FC, and $i_r + i_m$ was assumed to be 25% of FC; hence Eq. (3) can be rewritten:

$$EP = C_v + (0.35) \cdot FC \tag{4}$$

All of the economic data for the analysis was obtained from Ulrich [27] and Matches [28] webpage. All monetary values are given in 2007 US dollars.

To calculate the capital costs, the number of real stages was estimated from the number of theoretical stages which appears in Table 1, using an overall Murphee efficiency of 0.7 in all cases. Finally, the vessel height was obtained by considering a tray spacing of 0.6 m and the vessel diameter and wall thickness were determined depending on column pressure and reflux ratio. When varying the top column pressure from 500 to 1000 kPa, while maintaining R/R_{min} ratio at 1.3, the column diameter increases from 1.14 to 1.21 m. When varying R/R_{min} ratio from 1.1 to 1.5, while maintaining top column pressure at 700 kPa, column diameter increases from 1.1 to 1.25 m. The wall thickness was estimated as 0.025 m [29]. It does not change significantly with both column diameter and reflux ratio. All the coolers were considered to be air-coolers, and the fan power was determined according the method proposed by Lerner [30].

Operational costs were calculated by determining electrical power consumption of the pumps, compressors and fans, and the steam required for heating.

Table 5 shows the economic potential for all systems maintaining the top column pressure at 700 kPa and the ratio R/R_{min} at 1.3. Appendix A shows the detailed calculation procedure for the conventional system.

3.2. Conventional distillation

In order to determine the optimum reflux ratio and the optimum column pressure the variation of economic potential with these two critical parameters was calculated. Fig. 5a shows the variation of economic potential with reflux ratio, while maintaining column top pressure at 700 kPa. Fig. 5b shows the variation of economic potential with column pressure, while maintaining the ratio R/R_{min} equal to 1.3.

For a consistent economic evaluation, the feed and product streams have constant conditions of pressure, vapour fraction and composition. In the cases when column working pressure is lower than these values, in order to evaluate the economic potential, one pump to compress the top stream and another one to compress the bottom stream, were added. In the case when column pressure is higher, a pump to compress the feed stream was added.

As can be seen in Fig. 5a, the capital costs decrease as the ratio R/R_{min} increases. The costs of both the compressor and the kettle reboiler increase with the ratio R/R_{min} but the cost of the vessel is lower when the ratio R/R_{min} increases because fewer theoretical stages are needed. The second factor is the most important in this case. The operational costs increase with R/R_{min} ratio because the flows in the column are higher so the energy consumption is higher too. Evaluating the economic potential there is an optimum when R/R_{min} ratio is about 1.3.

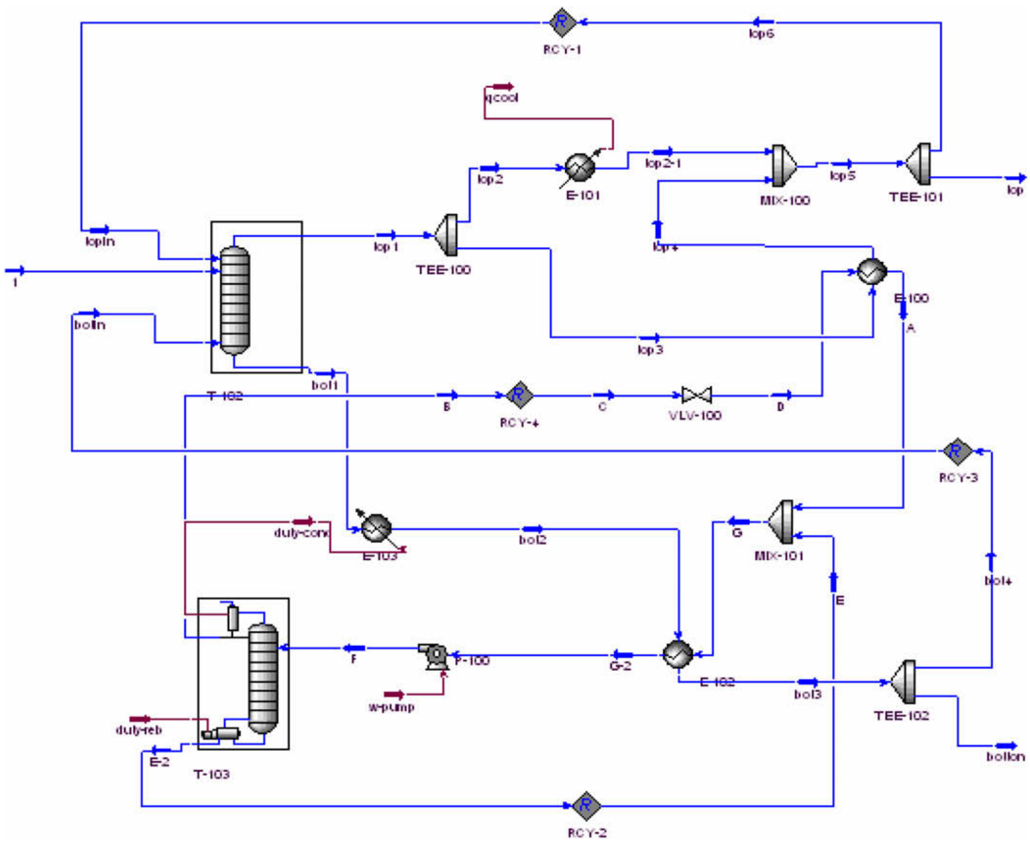

Fig. 4. HYSYS process flow diagram for the absorption heat pump.

Table 5
Capital ($) and operational ($/year) costs for the three systems, tested with column top pressure equal to 700 kPa and R/R_{min} ratio equal to 1.3

	Conventional distillation	Top vapour recompression	Bottom flashing
Capital costs			
Column	565,100	565,100	565,100
Condenser	53,700		
Reboiler	69,800		
Compressor		116,800	111,600
Heat exchanger		106,000	98,200
Cooler		19,900	2700
Operational costs			
Steam	183,200		
Electricity	74,400	189,600	176,300
Economic potential	498,700	497,100	448,500

Fig. 5b shows that capital costs increase as column pressure increases, due mainly to the increase in the tray and vessel cost. At the same time, operational costs decrease as column pressure increases due mainly to the overall lower energy consumption. According to the economic potential, the optimum working top column pressure is 700 kPa.

3.3. Top vapour recompression heat pump

A similar economic analysis was performed in this case. Fig. 6a shows the variation of economic potential with reflux ratio, while maintaining column top pressure at 700 kPa. Fig. 6b shows the variation of economic potential with column pressure, while maintaining the ratio R/R_{min} equal to 1.3.

As can be seen in Fig. 6a, the optimum R/R_{min} is again about 1.3. In the heat pump case, capital costs are higher than in the conventional distillation column because of additional items, namely an expensive compressor and an air-cooler. Nevertheless, the operation costs are lower in the heat pump case primarily because of the reduction in steam usage as expected. The economic potential is similar in both the conventional and heat pump cases but the heat pump provides an annual energy saving of 24%.

While maintaining the ratio R/R_{min} equal to 1.3, Fig. 6b shows that the optimum column pressure is located at the lower limit. Pressure cannot be decreased from this value because it would be necessary, then, to use a refrigerant, instead of air, for cooling so the process would not be economical. Working at the lower limit pressure and with a ratio R/R_{min} equal to 1.3, an energy saving of up 33% can be achieved as well as reducing the economic potential by 9%. The energy saving value is close to the one obtained by Annakou and Mizsey [6] in the separation of a propylene/propane mixture (37%) and the one obtained by Fonyo et al [4] in a C_4-splitter (42%).

Simple Payback Period can be calculated as additional capital costs divided by savings per year. In this case this value is 1.25 years, which is close to the value (1.7 years) obtained by Ferre et al [7] in ethylbenzene/styrene separation.

3.4. Bottom flashing heat pump

The same economic analysis was performed in this case. Fig. 7a shows the variation of economic potential with reflux ratio, while maintaining column top pressure at 700 kPa. Fig. 7b shows the variation of economic potential with column pressure, while maintaining the ratio R/R_{min} equal to 1.3.

The capital costs decrease as the reflux ratio increases, as in the top vapour recompression heat pump, and the order of magnitude is similar to the top vapour recompression heat pump. The opera-

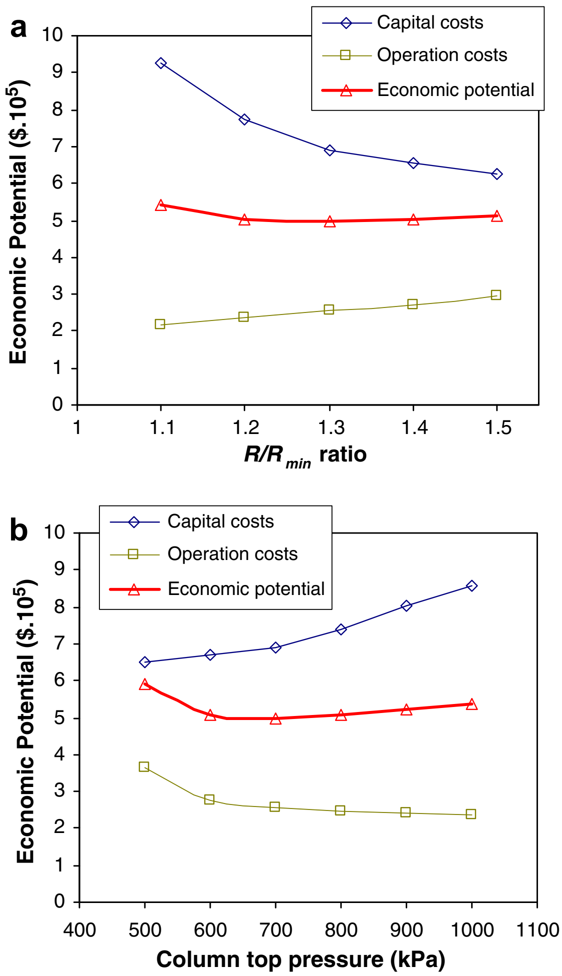

Fig. 5. (a) Variation of economic potential with column reflux ratio for the conventional distillation. (b) Variation of economic potential with column pressure for the conventional distillation.

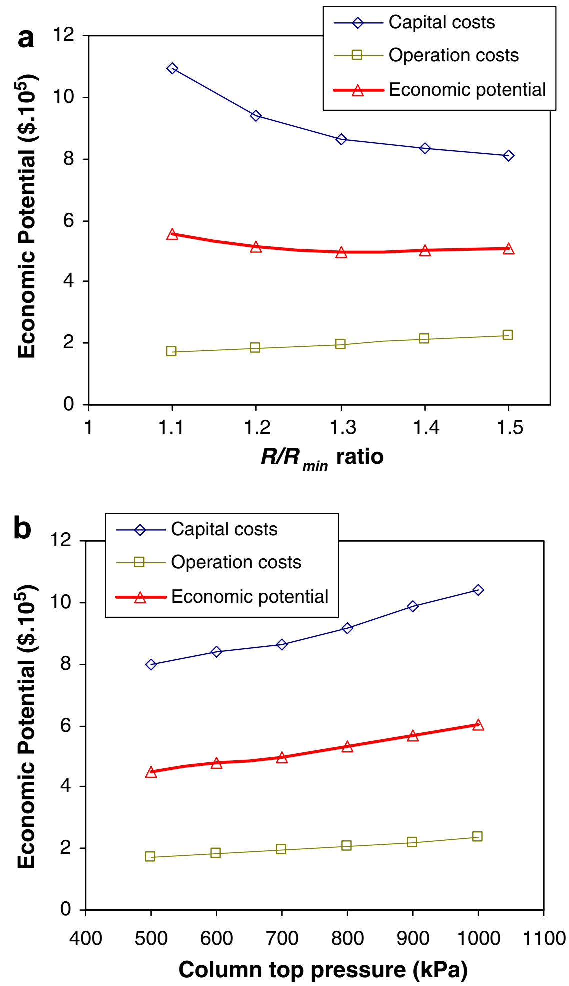

Fig. 6. (a) Variation of economic potential with column reflux ratio for the vapour recompression heat pump. (b) Variation of economic potential with column pressure for the vapour recompression heat pump.

tional costs decrease until the ratio R/R_{min} is equal to 1.3 and then they increase. This is due to the energy consumption in the compressor K-100, which depends on the valve VLV-100 outlet pressure. This pressure is fixed so that the vapour fraction of the compressor inlet stream is equal to one. Finally, the lowest economic potential is reached when the ratio R/R_{min} is about 1.3.

Fig. 7b shows that capital costs increase with column pressure, as in the top vapour recompression heat pump. This is due to the increase in the number of theoretical stages needed. The effect on operational costs is the same as in the case of varying reflux ratio: they decrease until top column pressure is about 700 kPa and then they increase again. For this reason, the lowest economic potential is reached when top column pressure is 700 kPa.

At optimal working conditions (R/R_{min} ratio equal to 1.3 and top column pressure equal to 700 kPa), an energy saving up of 32% can be achieved as well as reducing the economic potential by 10%. The energy saving value is close to the one obtained by Annakou and Mizsey [6] in the separation of a propylene/propane mixture (37%) and the one obtained by Fonyo et al. [4] in a C_4-splitter (27%). In this case, the Simple Payback Period is 1.09 years.

3.5. Absorption heat pump

For this case, just a simulation with a ratio R/R_{min} equal to 1.3 and top column pressure equal to 700 kPa was performed, because the economic analysis showed that, although the capital costs were close to the conventional distillation ones, the operational costs were higher when compared to conventional case, due to the large steam consumption. Fonyo and Benko [11] state that absorption heat pumps are more suitable for systems with large temperature difference where mechanical heat pumps cannot be applied. For the system studied in this paper, which a small temperature difference of about 10 °C, a mechanical heat pump is more suitable than an absorption one. Table 6 shows the capital and operational costs of absorption heat pump system, when compared with conventional distillation system.

4. General considerations

Heat pumps systems have been demonstrated to be economically feasible due to the energy savings that can be achieved. But this is not the only advantage of these systems because, by using a heat pump, the CO_2 footprint could be reduced. Hence, there is an environmental advantage which could become more important if CO_2 quotas are enforced by governments.

One disadvantage of these systems is that they involve greater complexity when compared with conventional distillation systems, due to the presence of the compressor and the cooler. Another alternative to heat pump systems could be improving heat integration. Depending on the plant, there may be waste heat available from another process for the reboiler.

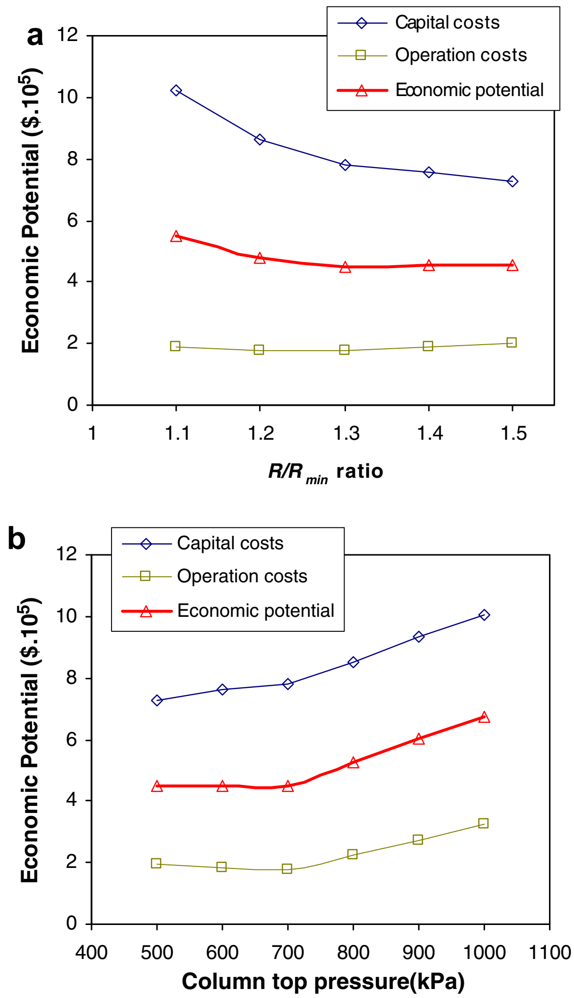

Fig. 7. (a) Variation of economic potential with column reflux ratio for the bottom flashing heat pump. (b) Variation of economic potential with column pressure for the bottom flashing heat pump.

Table 6
Capital ($) and operational ($/year) costs for the absorption system, when compared with conventional distillation systems, with column top pressure equal to 700 kPa and R/R_{min} ratio equal to 1.3

	Capital costs	Operational costs	Economic potential
Conventional distillation	688,600	257,600	498,700
Absorption heat pump	697,426	354,361	598,460

Future work will concentrate on the study of feasibility of a heat pump in a distillation system where refrigeration is required for the condenser, because this is intrinsically employing a heat pump for the refrigeration and the economic effects would be different.

5. Conclusion

This study describes the simulation of a conventional *i*-butane/*n*-butane distillation process and then with three heat pumps systems incorporated, using the HYSYS computer software. The simulations are employed to assess the economics. Potential energy savings are shown by incorporating heat pumps for distillation of mixtures with close boiling points.

For the case studied, with a top vapour recompression heat pump, the capital costs are almost the same as in the conventional distillation process, but the energy costs can be reduced by about 33%. This involves an economic potential reduction of 9% and a Simple Payback Period of 1.25 years. With a bottom flashing heat pump, the capital costs are very close to the conventional distillation ones, but energy costs are reduced by 32%. In this case, the economic potential is reduced by 10% and the Simple Payback Period is 1.09 years.

The values obtained for energy savings and Simple Payback Period are similar to those from literature for propylene/propane and ethylbenzene/styrene separations. Significant savings should also be possible in processes with high energy consumption, such as separation of *p*-xylene from *m*-xylene and *o*-xylene, and separation of iso-pentane from *n*-pentane. An absorption heat pump is not suitable for this system.

Acknowledgement

All the simulations were performed with HYSYS software from Aspentech.

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Appendix A. Calculations for conventional distillation

For the conventional distillation column, with a top column pressure of 700 kPa and a R/R_{min} ratio of 1.3 the capital costs are calculated as follows:

A.1. Distillation column

For a top column pressure of 700 kPa and a relation R/R_{min} of 1.3, 33 theoretical stages are needed. If we consider a Murphee efficiency of 0.7, 47 real stages are needed. By considering a tray space of 0.6 m, column height can be determined and taking into account that column diameter is equal to 1.2 m, from Ulrich [21] figure 5.48, individual tray cost is estimated as $420. By actualizing prizes to 2007, the final tray cost is $638.

From Ulrich [21] figures 5.44, 5.45 and 5.46 and actualizing prices to 2007, the vessel cost is estimated as $352,100.

The total column cost is estimated as $565,100.

A.2. Reboiler

From HYSYS program, the reboiler duty is equal to 1981 kW and *UA* is equal to 94,058 kJ/°C h. From Coulson and Richarson [18], the overall transport coefficient, *U*, is taken as 150 J/s m^2 °C.

The area of the reboiler is calculated as *UA*/*U* and is equal to 174 m^2. Finally, directly from Matche web, the heat exchanger price is estimated, as a function of its area, as $69,800.

A.3. Condenser

From HYSYS program, the condenser duty is equal to 1981 kW and *UA* is equal to 327,223 kJ/°C h. From Coulson and Richarson [18], the overall transport coefficient, *U*, is taken as 500 J/s m^2 °C.

The area of the condenser is calculated as *UA*/*U* and is equal to 182 m^2. Finally, directly from Matche web, the heat exchanger price is estimated, as a function of its area, as $53,700.

A.4. Operational costs

From the reboiler duty and the heat of vaporization of steam (2148 kJ/kg), steam consumption is 3319 kg/h. By considering the price of low pressure steam as 6.9 $/tonne, the annual steam consumption involves $183,200.

From Lerner [23], the base power fan of an air-cooler is estimated as 0.595 kW/m^2. If the condenser area is 108 m^2, the base power of the fan is 108 kW. By considering the price of electricity as 8.6 cents/Kw h, the annual electricity consumption involves $74,400.

A.5. Economic potential

For the case analyzed, the capital costs are $688,600 and the annual operational costs are $257,700. Finally, the Economic Potential, by applying Eq. (4), is equal to $498,700.

Appendix B. Nomenclature

$R/R_{\min}$: relation between column reflux ratio and minimum reflux ratio
Q_C: conventional column condenser energy
Q_R: conventional column reboiler energy
Q_i: energy transferred in item i
EP: economic potential
C_v: process variable costs
C_f: annual fixed costs
FC: fixed capital investment
i_r: fixed capital recovery rate applied to FC
i_m: minimum acceptable rate of return on FC
A: heat exchanger area
U: overall heat transport coefficient

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