Designing a divided wall column

A refiner plans the application of divided wall column technology for greater energy efficiency in the distillation of multi-component reformate

SAŠA POLOVINA, SREĆKO HERCEG and ANA GRANIĆ ŠARAC
INA

Two of the most important tasks facing refineries today are to increase conversion depth and to increase energy efficiency. To achieve deep conversion, refineries have started to invest in heavy residue processing. Increased energy efficiency is achieved by better utilisation of all energy sources and by increasing heat recovery in refinery processes.

Distillation is the most common and important physical separation method. Therefore it is not surprising that attention is dedicated to increasing energy efficiency in this process. Distillation is also a process that is responsible for a large percentage of the world's energy consumption. Due to its significant energy inefficiencies, some call it the 'energy dark horse'. It is known that distillation represents about 95% of all separations in the chemical industry and that, in the US alone, about 40 000 distillation columns exist in all industries.

Distillation is a thermal separation process, where reboiling part of the bottom liquid usually provides the heat required for the separation process. Condensation of overhead vapour is used to take a certain amount of heat to gain the desired product. With reboiling and condensation, a certain amount of energy is lost.

The minimum amount of energy required for a reversible distillation process ($Q_{min.}$) can be described with Equation 1:

$$Q_{min.} = \Delta S\, T_c / (1-T_c/T_h)$$

where ΔS = change in entropy
T_c = high temperature in the reboiler
T_h = low temperature in condenser.

According to the second law of thermodynamics, change in entropy can be defined as:

$$\Delta S = -R \Sigma X_i \ln (X_i)$$

where R = universal gas constant
X_i = molar fraction of component i.

The energy usage Q in conventional distillation arrangements is typically more than 50% higher than $Q_{min.}$ due to inevitable entropy losses, in particular due to remixing within column sections and interconnections and temperature differences in heat exchangers.[1]

Energy efficiency in conventional distillation (that is, classic distillation with one feed and two products) has been increased significantly in the last 20 years as a result of:

- Better process control systems
- Utilisation of high efficiency internals
- Using side heat exchangers enabling use of cheaper utilities
- Heat integration
- Heat pump assisted distillation.

However, the only way to minimise energy losses that are caused by a remixing effect in the separation of multi-component mixtures is to apply the principle devised by Petlyuk.[2] According to this, in every column in the sequence, only the easiest separation can be performed. This is the principle on which so-called divided wall columns (DWC) are based. Separation is achieved by using a vertical partition wall that divides mainly the central part

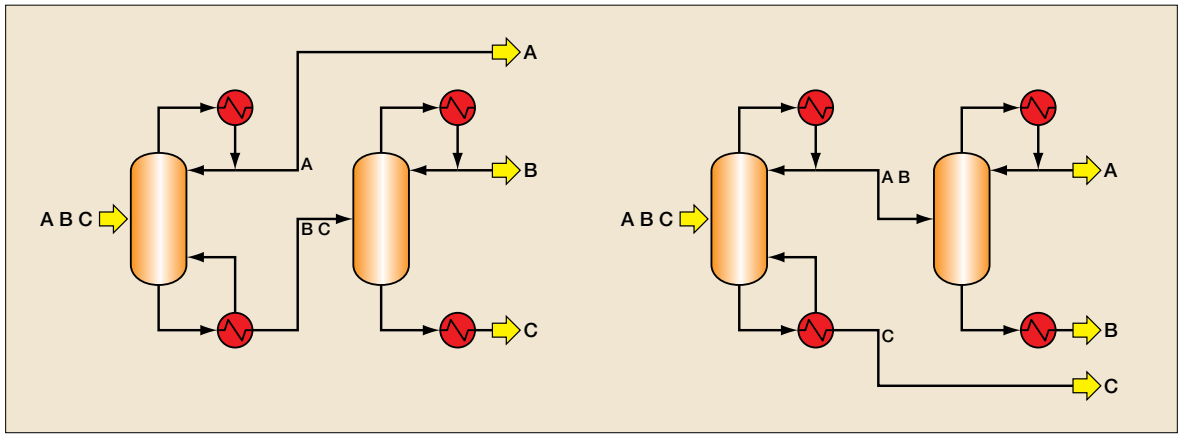


Figure 1 Direct and indirect sequence of distillation columns for the separation of a ternary mixture[3]

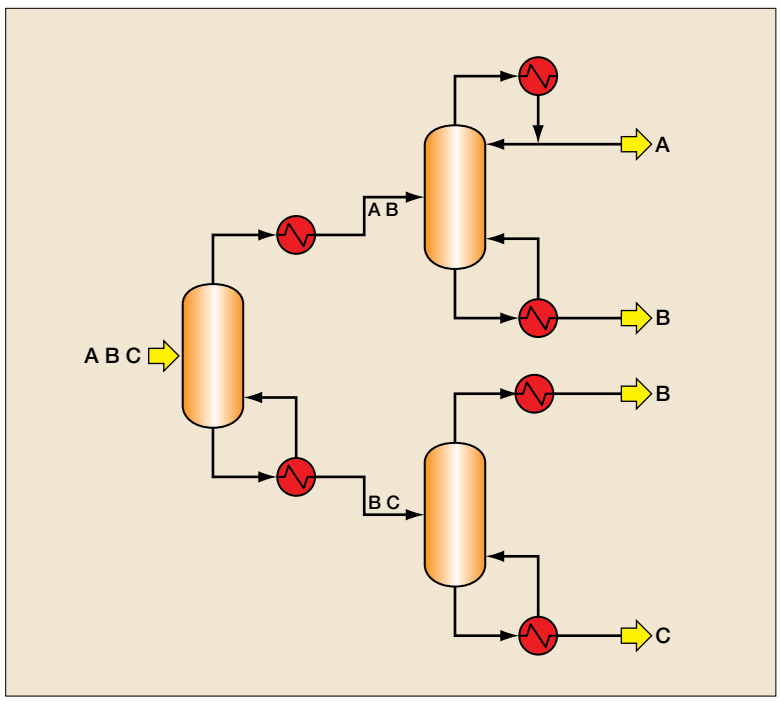


Figure 2 The so-called prefractionator sequence for separation of three products[3]

of the column into a prefractionator and main column sections. In each section, two components with the greatest difference in volatility are separated, while others are allowed to distribute.[3] The biggest application for divided wall columns was separation of mixtures of three products. Now, after the introduction of non-welded wall technology it is possible to use this principle for separation of multi-component mixtures consisting of four or more products.

The main obstacle to greater expansion of this technology lies in the fact that there is no reliable method for calculation of distillation and dimensioning of the column, and therefore there is a huge delay in the application of this technology. Another reason for its poor adoption is potential issues in process control. As the divided wall column has greater numbers of degrees of freedom than a conventional distillation column, for proper and safe operation of the process it is necessary to harmonise significantly larger number of variables.

Development and application of divided wall columns

In conventional distillation technology, separation of multi-component mixtures of three products is carried out in at least two consecutive columns (the number of distillation columns required in series is equal to the number of products, minus one). **Figure 1** shows an example of two possible sequences for separating three products: the so-called direct and indirect sequences.

Figure 2 shows another possibility: a sequence of three columns. The first column in the series is a prefractionator or column that has no products, but is used for sharp separation of components with the great-

est difference in the relative volatility (of components A and C), while the middle-boiling component B is distributed in the top and bottom products of the prefractionator. The final separation is performed in two distillation columns. In one column, separation of components A and B is carried out, and in the other components C and B are to be separated. In this way, the middle-boiling component B is removed from the bottom of one column and from the top of the second column. This configuration, although it requires more equipment (one column and two heat exchangers) compared to the direct/indirect sequences, can lead to significant reduction in the amount of energy required to perform the same separation task, due to avoidance of energy losses related to remixing of component B.

If the reboiler of the top column and condenser of the bottom column are thermally coupled (see **Figure 2**), capital cost savings can be achieved. The first example of the described distillation sequence enabled reduction of energy and investment costs, by use of an additional evaporator (reboiler) and condenser located in the prefractionator.[4] Petlyuk described the design of fully thermally coupled distillation columns.[2] The reboiler and condenser in this configuration are located at the main column and the reboiler and condenser of the prefractionator are replaced with thermal coupling to the main column. In this way, the heat required for separation is provided with a single reboiler and condenser. **Figure 3** shows two sequences of distillation columns for separation of three products.[2,4]

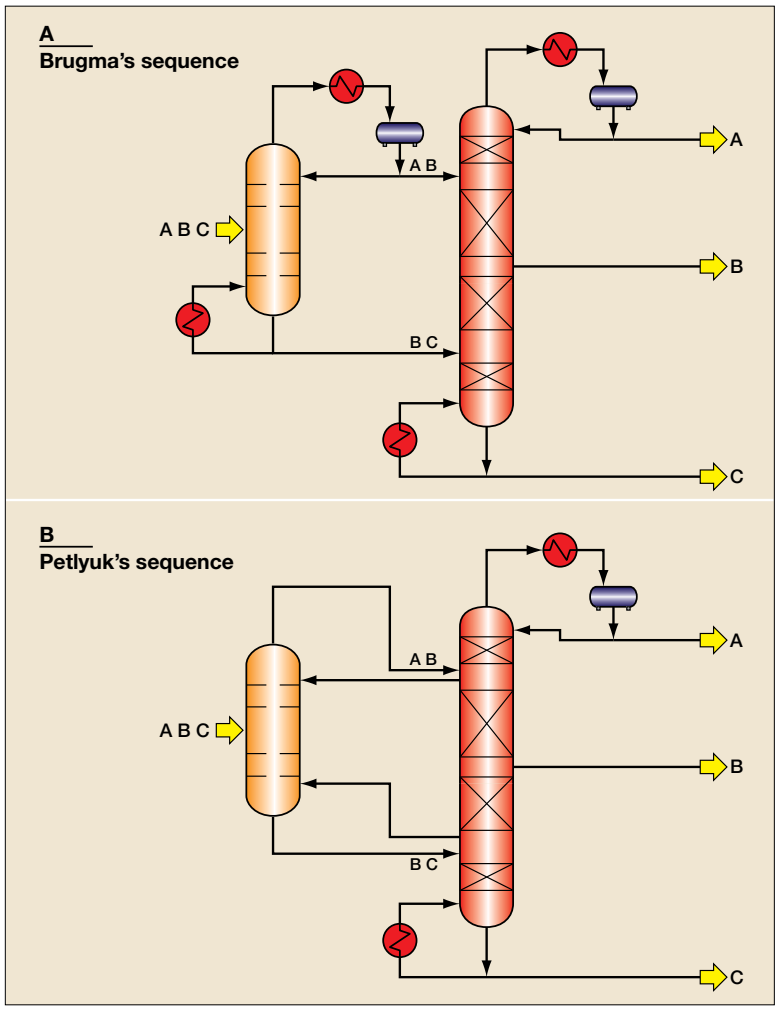


Figure 3 Brugma's sequence (a) and Petlyuk's sequence (b) of distillation columns for separation of three products[2,4]

A study in 1972 confirmed substantial energy savings, which can be achieved by usage of a thermal connection for the separation of components with close boiling points.[5] A later study of the separation of ternary mixtures[6] recommended operation of Petlyuk's pseudo-column if the concentration of component B (middle component) in a mixture is between 40% and 80%. A comparison of the optimal vapour flow in direct and indirect conventional distillation series and for Petlyuk's pseudo and normal design showed significant savings for the latter.[7] Underwood's equations were used for the calculation of minimum required vapour. A further study indicated that Petlyuk's design has five degrees of freedom, which was the basis for the further development of distillation technology.[8]

The first proposed application of the divided wall for separation of ternary mixtures

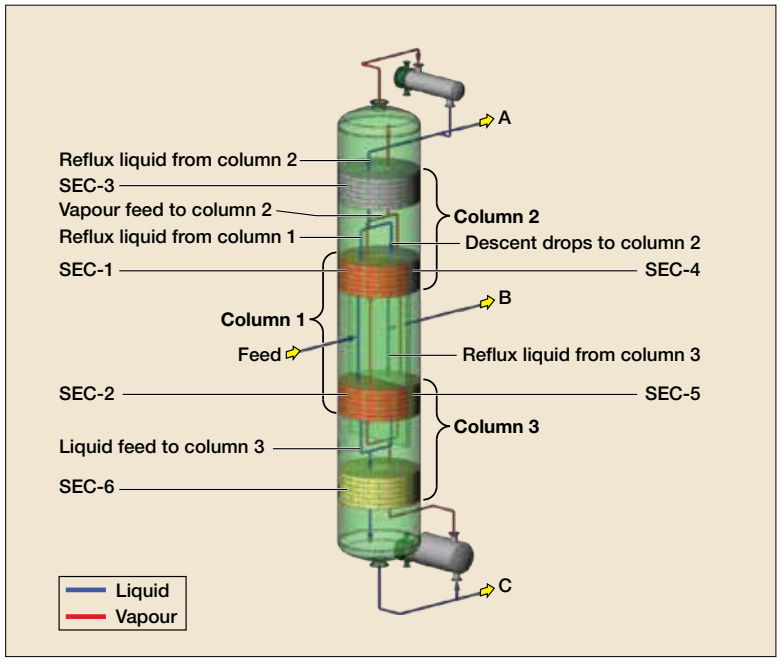


Figure 4 SPE divided wall column for the separation of a ternary mixture[11]

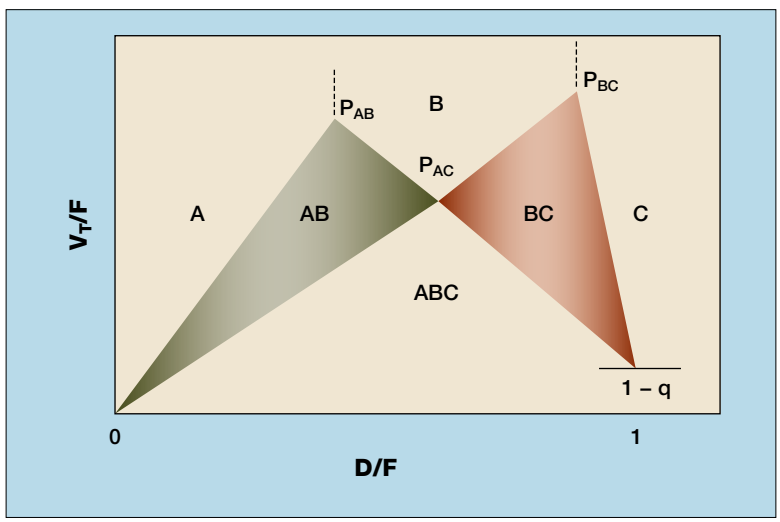


Figure 5 Vmin diagram for separation of ternary mixtures into three products (q indicates quality of feed; q = 0 – saturated vapour; q = 1 – saturated liquid)[12]

consisted of the main column in a single shell with three side products, where the feed input and three side products were separated by a dividing wall.[9] Later work enabled application of the dividing wall for separation in Petlyuk's series.[10] Using new technologies, conventional distillation separation of multi-component mixtures of three products in two distillation columns would be replaced by separation in a single divided wall distillation column (see **Figure 4**).[11]

In a divided wall column, the prefractionator as a separate unit has been abandoned. It is integrated into the shell of the head of the column and separated by a divided wall. Such a column is called a fully thermally coupled column. When the system is properly operated, the potential energy savings can be 20-40%, compared to conventional distillation systems. The first divided wall column was designed at BASF SE in Ludwigshafen, Germany.

Design basis

Design of divided wall columns is carried out in several steps:

- Shortcut calculation (Vmin diagram)
- Rigorous calculation and optimisation
- Column sizing – hydraulic design.

Shortcut calculation (Vmin diagram)

The purpose of a shortcut calculation is to provide data that will be the basis for rigorous calculation. One such method is the so-called Vmin diagram, which is a graphical representation of vapour and liquid flow rates required to perform all splits within a given distillation sequence. It is based on Underwood's equations, implying the following simplifications and conditions: infinite number of theoretical stages, constant relative volatilities and constant molar flow rates in a column. The required data for a Vmin diagram are: composition of the feed, relative volatilities, required key component splits, and thermal condition of the feed. **Figure 5** shows a Vmin diagram for the separation of ternary mixtures into three products. The diagram illustrates how the components of the ternary mixture in a conven-

tional distillation column, with an infinite number of equilibrium stages, are distributed between the top and bottom of the column depending on two defined degrees of freedom (D/F – net flow of product to the top per unit of feed; V/F – vapour flow rate above the feed).

For a ternary mixture (to obtain three products from multi-component mixtures), three points need to be formulated: the point of sharp split between components A and B, P_{AB}; the point of sharp separation of components A and C, P_{AC}; and the point of sharp separation of components B and C, P_{BC}. Initial values of liquid and vapour molar flow rates in a divided wall column can then be determined through relations between those peaks.

To calculate the minimal number of theoretical stages of the distillation process, Fenske's method is used, while the Gilliand correlation provides the relationship between reflux and number of required theoretical stages. Feed stage location is determined by the Kirkbride equation.

Rigorous calculation

To obtain a basis for hydraulic design and sizing – that is, to obtain the final values of the design parameters, a rigorous simulation using commercial process simulators is needed. Such simulators do not incorporate a DWC model, so it is necessary to build up a thermodynamically equivalent sequence of conventional columns. These models are applicable if heat transfer through the dividing partition can be neglected.

The number of stages is set to $2.5N_{min}$ (where N_{min} is the number of equilibrium stages at total reflux) as calculated by the Fenske equation in shortcut simulation. The flow rates of all streams are set to match the values obtained from the Vmin diagram.

Sizing – hydraulic design

After simulation, the process parameters obtained are used for sizing. A divided wall column can be equipped as a conventional distillation column, with various types of trays or packings. The difference compared to sizing a conventional column is in the regulation of vapour split across the partition. The liquid split is easily regulated in operation, while the vapour flow on both sides of the wall depends primarily on resistance to flow, which is provided by internal elements of the column, inclination of the wall, and by the wall itself. By proper sizing, equal pressure drop on both sides of the wall is achieved, meaning that vapour flow rates in operation will be the same as in design.

For the assessment of the pressure drop in structured packings, the Delft model is used. Its biggest advantage over other models is that it does not require any special, experimentally determined parameters for packings, only their geometric characteristics.

Design of reformate fractionation for a divided wall column

Operation of INA Rijeka refinery's existing plant (a splitter) for fractionation of reformate with a capacity of 80 t/h was simulated as the basis of the design. The process was modelled with ChemCAD process simulator software. For the modelling of vapour-liquid equilibrium, the Soave-Redlich-Kwong (SRK) equation was selected as one of the most frequently used models for hydrocarbon mixtures.

Simulation of the base case

Given the increasingly stringent requirements for the content of benzene in motor gasoline, reformate fractionation plant has become a strategically important process in refining. Reducing the benzene content of heavy reformate enables the use of larger quantities of heavy reformate for gasoline blending.

By separating the C_5-C_6 fractions of the products from catalytic reforming (reformate), the concentration of aromatic components in heavy reformate is higher, which leads to an increase in RON. Therefore, the requirement for RON from the catalytic reformer (a semi-regenerative reformer) is smaller, which means less severe working conditions, thereby significantly extending the length of the catalytic cycle of the plant.

As the feed for reformate fractionation is composed of many different types of molecules, for the purposes of calculation, the number was reduced to 15 main components. **Table 1** shows the chemical composition of such a mixture. The first numeric data row in **Table 1** shows the chemical composition of reformate for simulation.

Main parameters, composition of feed, quantity and compositions of products obtained by simulation in base case (BRC – benzene rich cut)

Stream name	Reformate	Light reformate	BRC	Heavy reformate
Temperature, °C	115.1569	40	40	143.2143
Pressure, bar	2.8	2.7	1.7	1.7
Enthalpy, MW	-13.472	-13.69	-0.18206	-0.13731
Total flow, kmol/hour	852.8956	269.9423	72.3152	510.8459
Total flow, kg/hour	80000.0367	21498.9317	5993.9609	52524.0687
Mass percentage, %				
i-pentane	6.300794	23.445995	0	0
n-pentane	4.770194	17.750454	0	0
2-methylpentane	5.572594	20.736268	0	0
3-methylpentane	4.346296	16.173041	0.000001	0
n-hexane	5.371594	19.980904	0.000204	0
benzene	6.970397	1.904563	73.879582	1.434139
2,4-dimethylpentane	0.41	0.008776	5.426328	0.004345
2-methylhexane	2.241398	0.000002	19.839668	1.154297
n-heptane	2.049998	0	0.850597	3.025316
toluene	23.890366	0	0.003621	36.387306
ethylbenzene	4.264195	0	0	6.494849
p-xylene	5.507694	0	0	8.388839
m-xylene	12.395789	0	0	18.88018
o-xylene	6.341793	0	0	9.65927
1,2,4,5-tetramethylbenzene	9.566893	0	0	14.571458

Table 1

Inlet parameters and results of rigorous calculation for the reformate fractionation plant (capacity of 80 t/h)

Inlet parameters		**Results**	
Inlet column temperature, °C		Evaporator heat duty, MW	
322-C-001	115	322-H-001	13.6
322-C-002	154	322-H-002	3.7
		Total	17.1
Top column temperature, °C		Condenser heat duty, MW	
322-C-001	84	322-C-001	-13.4
322-C-002	97	322-C-002	-3.4
		Total	-16.8
Bottom column temperature, °C		Benzene content, % m/m	
322-C-001	154	Light reformate	1.9
322-C-002	143	Benzene rich cut	73.8
Heavy reformate	1.4		
Top column pressure, bar			
322-C-001	2.7		
322-C-002	1.7		
Reflux ratio			
322-C-001	4.3		
322-C-002	4.5		

Table 2

The distillation products have to meet the following requirements:

- Benzene content in light reformate must not be higher than 2 wt%
- n-hexane must end up in light reformate as much as possible
- The benzene content of the benzene fraction must be 50-75 wt%
- The maximum benzene content of heavy reformate is 1.5 wt% and the process is optimal if the benzene content is between 0.5-1 wt% by weight.

Operating modes used in the daily operation of the existing reformate fractionation plant (splitter) were utilised for defining operating conditions. Also, the total number of trays (equilibrium stages) and feeding points (tray ordinal number) were fixed according to the existing plant operation. These data were also taken as unchangeable (constant) throughout calculation. Thus, the calculation of a column for separation of light reformate and heavy reformate (322-C-001) was carried out with 77 trays and with feed entering the 30th tray, while the calculation of a column for separation of benzene reach cut and heavy reformate (322-C-002) was carried out with 52 trays and with feed entering the 22nd tray.

The composition of the feed, quantity and composition of the products obtained by simulation in the base case at the capacity of 80 t/h are shown in **Table 1**. The data obtained by strict calculation are in accordance with actual process data and the concentrations of benzene in the products are below standard limits. (Due to the large number of compounds in the products of reformate fractionation, daily analysis and quality control is carried out only for controlling the benzene content of products.) Main values and process parameters used for the calculation, as well as the number of trays and feed

point, are in accordance with real process conditions.

In **Table 2**, inlet process parameters and the results of rigorous calculation at a plant capacity of 80t/h are shown.

DWC design

For calculation purposes, the feed composition was reduced to 15 components, which were arranged in order of increasing boiling point. **Table 3** shows: the composition as mole fractions of the mixture that was used for the calculation of the divided wall column at 80 t/h; distribution coefficients at the feed supply tray; and the desired column products (D – distillate, B – bottom, S – average product). The grouping of components by resulting products corresponds to the situation in the base case.

In the first phase of calculation, a shortcut calculation is performed. In this step, each section is observed as a separate conventional distillation column, where separation takes place. For each product, a key distillation component has to be selected. These 'key components' are for light reformate n-hexane, benzene for the benzene fraction, and n-heptane for heavy reformate.

In the first column, separation of components with the greatest difference in relative volatility (n-hexane and n-heptane) takes place. For the calculation, distillate product should contain 99% n-hexane in the feed, and 99.5% n-heptane in the feed should be recovered at the bottom of this column. The minimum number of equilibrium stages required for the separation is also fixed.

In the second column separation takes place between the key component of light reformate, n-hexane, and the key component of the middle product, benzene. Distillate should contain 60% n-hexane in the feed to this column, and 95% benzene in the feed should remain in the bottom product of the column.

For the third column shortcut distillation, a calculation for benzene, the key component of the middle product, and n-heptane, the key component of the heavy reformate, is performed. We require a top product containing 90% benzene in the feed and a bottom product with 99.2% n-heptane in the feed to the column.

In this calculation, basic parameters for the second phase of the calculation (rigorous calculation) were generated. For the purpose of rigorous calculation, a divided wall column is regarded as three conventional distillation columns. The result of rigorous calculation is the minimum vapour flow required for the separation. This minimum vapour flow is then needed for the creation of a Vmin diagram, the third phase of calculation.

Chemical composition of the feed, distribution coefficient at the feed supply tray and the desired column distribution by components

Component	Mole fraction	Distribution coef. at a feed supply tray	Product
i-pentane	0.081913	3.289	D
n-pentane	0.062015	2.75	D
2-methylpentane	0.060654	1.578	D
3-methylpentane	0.047307	1.424	D
n-hexane	**0.058466**	**1.283**	**D**
benzene	**0.083699**	**0.914**	**S**
2,4-dimethylpentane	0.003838	0.986	S
2-methylhexane	0.020981	0.762	S
n-heptane	**0.019189**	**0.600**	**B**
toluene	0.2432	0.402	B
ethylbenzene	0.037674	0.196	B
p-xylene	0.04866	0.182	B
m-xylene	0.109516	0.178	B
o-xylene	0.05603	0.158	B
1,2,4,5-tetramethylbenzene	0.066857	0.03	B

'Key components' shown in bold

Table 3

Characteristic points of a Vmin diagram

Specified in the calculation

Section	Point	Required separation	
C_1	P_{AC}	r(C,D)=0.005	r(A/B)=0.01
$C_{2,1}$	P_{AB}	r(B,D)=0.05	r(A/B)=0.4
$C_{2,2}$	P_{BC}	r(C,D)=0.08	r(B/B)=0.1

Results

V_T/F	D/F	N	N_f	N_{min}
0.62	0.35	64	34	15
1.03	0.29	64	47	16
0.78	0.4	67	45	17

(C_1 – prefractionator; $C_{2,1}$ – main column above side product; $C_{2,2}$ – main column below side product; F – feed; V_T – vapour molar flow through column; D – distillate molar flow through column)

Table 4

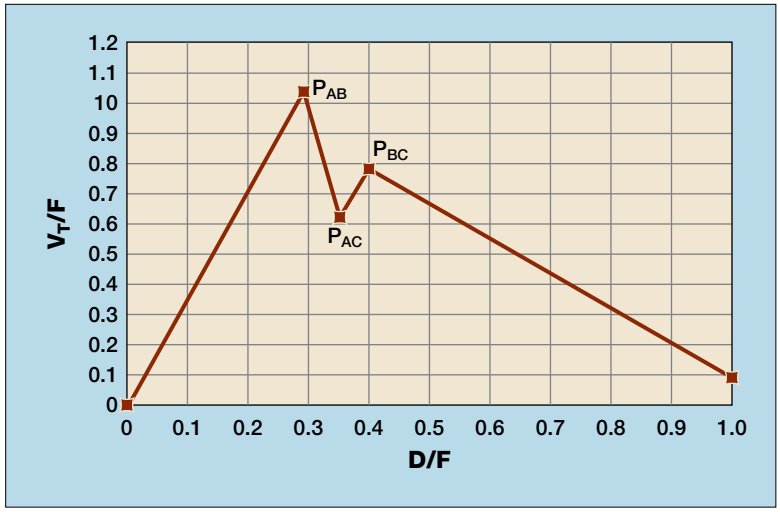


Figure 6 Vmin diagram for the separation of three key products of reformate in a divided wall column

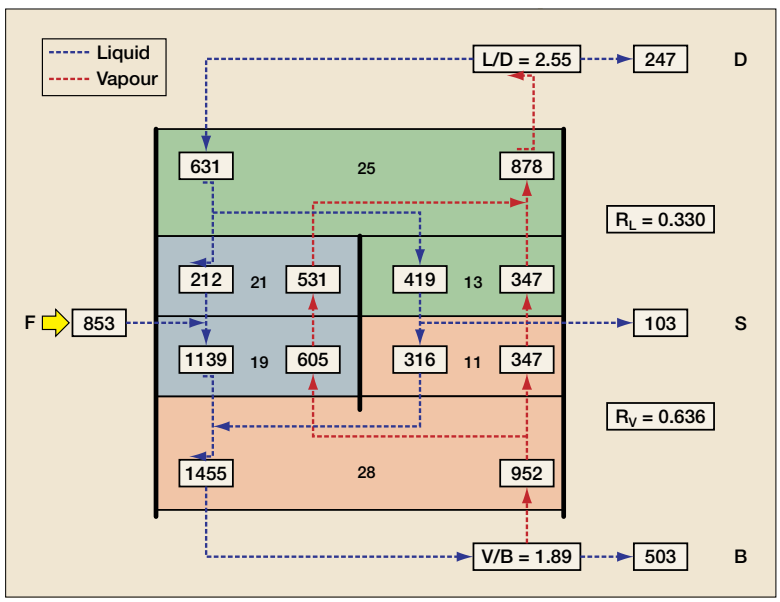


Figure 7 Vapour and liquid material balance, and distillate, middle product and bottom product yields in thermally connected columns (F, D, S, B – molar flows of feed, distillate, side product and bottom product; R_L – Liquid splitting over dividing wall; R_V – vapour splitting over dividing wall; L/D – reflux ratio; V/B – reboiling ratio)

Table 4 shows characteristic points of a Vmin diagram generated by rigorous calculation.

After determining the characteristic points, a Vmin diagram can be constructed (see **Figure 6**). This diagram gives a clear idea of the minimum energy needed for the required separation.

By creating this diagram, the required amount of vapour for the separation and the amount of product distillate can be determined. By defining the relationship between these two variables, other variables can be determined.

The highest point in the Vmin diagram (P_{AB}) represents the minimum amount of vapour required for the separation of n-hexane and benzene. The minimum amount of vapour is related to the minimum amount of energy required for separation in thermally connected columns.

Point P_{AC} represents a preferred point and shows the amount of vapour required for the separation of n-hexane and n-heptane in a prefractionator, for the given amount and purity of distillate product. The preferred point represents the optimal means of operating the distillation process for the required purity of the product.

After the Vmin diagram is constructed, material balance calculations for the three distillation columns should be performed. The balance must include liquid and vapour flows for each column (section). These flows will be the basis for sizing the divided wall column.

Figure 7 shows a vapour and liquid material balance, with distillate, middle product and bottom product yields in thermally connected columns with a capacity of 80 t/h.

A material balance was calculated from the balance equations, where one can start from an initial value of the required vapour flow rate from the evaporator, or from the top of section 2.1, and the amount of distillate produced. In both cases, the amount of vapour flow through the prefractionator was determined based on the Vmin diagram, so that

operation of the prefractionator was in the area of the preferred point. The rest of the vapour is sent to the top of the main column, bypassing the prefractionation section.

In the fourth phase of calculation, data from the Vmin diagram as well as data from a rigorous calculation for each column served as initial points of calculation for the reformate splitter. The capacity of the reformate splitter is 80 t/h and is a thermally completely connected column. The initial number of equilibrium stages was fixed at $N \geq 4N_{min}$. Feed flow rates were fixed at a value determined from the Vmin diagram and material balance. After convergence of simulation in the process simulator ChemCad, optimisation of the distributed liquid and vapour in the sections above or below the wall was performed.

The number of equilibrium stages, N, was set at $N \geq 2.5N_{min}$. Stages that do not take part in separation were eliminated. In this way, for the prefractionator section, 40 equilibrium stages were determined, and for the main column 80 equilibrium stages were obtained.

Table 5 shows the composition of feed and products determined by simulation of the reformate splitter as one thermally completely connected column.

Divided wall column sizing

For the reformate splitter at INA Rijeka refinery, column C-001 was presumed to be of sufficient size and was selected for simulation of a divided wall column. Column C-001 is 36 m high and has a diameter of 2.3 m at the widest point.

The interior of the column consists of structured fillers. The Delft model was used to estimate the pressure drop through the column.[12] For all sections, a height equivalent to a theoretical plate (HETP) value of 0.4 m was used.

Table 6 shows the required number of equilibrium stages and the equivalent height of a structured filler at the capacity of 80 t/h, and **Table 7 i**ndicates the basic dimensions and estimated pressure drop across the layers of the filler.

To reduce pressure drop through section $C_{1,2}$ and $C_{2,4}$, the height of a packing bed was decreased (compared with height of a packing bed shown in **Table 5**), and the height of a packing bed was increased in section $C_{2,1}$. Feed enters the prefractionator part of the DWC above tray 19; there are 21 stages in the upper rectification section and 19 in the lower stripping section. As a result of prefractionation, we obtain a mix of light reformate and BRC at the top of the rectification section and a mix

Feed and product parameters and composition determined by simulation of a reformate splitter as one thermally completely connected column

	Reformate	Light reformate	BRC	Heavy reformate
Temperature, °C	100	76.5658	108.5178	163.0543
Pressure, bar	2.7	2.7	2.7	2.7
Total flow, kmol/hour	852.8957	236.2961	103.0002	513.6421
Total flow, kg/hour	80000.0509	18606.8835	8528.6865	52868.0441
Mass percentage, %				
i-pentane	6.300794	27.0899	0.000009	0
n-pentane	4.770194	20.509313	0.000142	0
2-methylpentane	5.572594	22.758687	2.608568	0
3-methylpentan	4.346294	15.619758	6.675395	0.000001
n-hexane	5.371594	12.220876	23.712844	0.000019
benzene	6.970394	1.799891	55.30771	1.000001
2,4-dimethylpentane	0.41	0.00157	3.824329	0.002657
2-methylhexane	2.241398	0.000001	7.388397	2.202627
n-heptane	2.049998	0	0.43665	3.032277
toluene	23.89037	0	0.045967	36.144784
ethylbenzene	4.264196	0	0.000002	6.452656
p-xylene	5.507695	0	0.000001	8.334335
m-xylene	12.395791	0	0.000002	18.757503
o-xylene	6.341793	0	0	9.596482
1,2,4,5-tetramethylbenzene	9.566893	0	0	14.476661

Table 5

Required number of equilibrium stages and the equivalent height of a structured filler at 80 t/h capacity

Section	Theoretical number of equilibrium stages	Height of a packing bed, m
$C_{1,1}$	21	8.4
$C_{1,2}$	19	7.6
$C_{2,1}$	25	10
$C_{2,2}$	13	5.2
$C_{2,3}$	11	4.4
$C_{2,4}$	28	11.2

Table 6

Heat duty of conventional separation sequence and DWC			
Parameters	**322-C-001**	**322-C-002**	**DWC**
Evaporators heat duty, MW	13.6	3.7	8.52
Annual cost of consumed refinery fuel gas, Euro	3 259 363	1 805 359	

Table 8

Condenser heat duty and related savings			
Parameters	**322-C-001**	**322-C-002**	**DWC**
Condenser heat duty, MW	-13.4	-3.4	-6.79
Annual cost of spent electrical energy, Euro	65 280	43 520	

Table 9

of heavy reformate and BRC at the bottom of the stripping section.

In the top section 2.1 of the main column, which has 26 theoretical stages, light reformate is received. Part of it returns to the top as reflux.

The main column central section above the BRC product outlet 2.2 has 13 theoretical stages, and the section below BRC outlet 2.3 has 11 theoretical stages

The bottom section 2.4 of the main column consists of 28 theoretical stages, and it has the largest liquid load in the column. Because separation in sections 2.1 and 2.4 requires a large number of theoretical equilibrium stages, the height of the structural filler has to be more than 10 m; these sections must be divided into two separate layers. The upper part of the main column consists of sections 2.1A and 2.1B containing structured fillers with a height of 5.2 m, while the lower part of the main column consists of sections 2.4A and 2.4B, containing structured fillers with a height of 5.4 m. For better distribution of liquid on the surface of the filler, eight dividers and seven liquid collectors must be installed in the column. **Figure 8** shows the dimensions of the divided wall column for separation of reformate at a capacity of 80 t/h.

The result of sizing is a divided wall column with a height of 56 m, which fully meets the required product quality.

Energy savings of a divided wall column

Energy savings are based on better utilisation of heat in the heat integrated column as well as on the elimination of a second column and accompanying equipment.

Table 8 shows heat duties of condensers and evaporators for the existing columns and for new DWC.

When we compare heat duties for evaporators in a conventional separation sequence with a DWC, we see that a DWC, for the same separation and the same product quality requires 8.78 MW less energy than the conventional sequence. The reason lies in a completely thermally linked distillation column and in complete elimination of one evaporator. **Table 8** also shows that by using a DWC, a refinery can save an annual €1 454 004 in fuel gas.

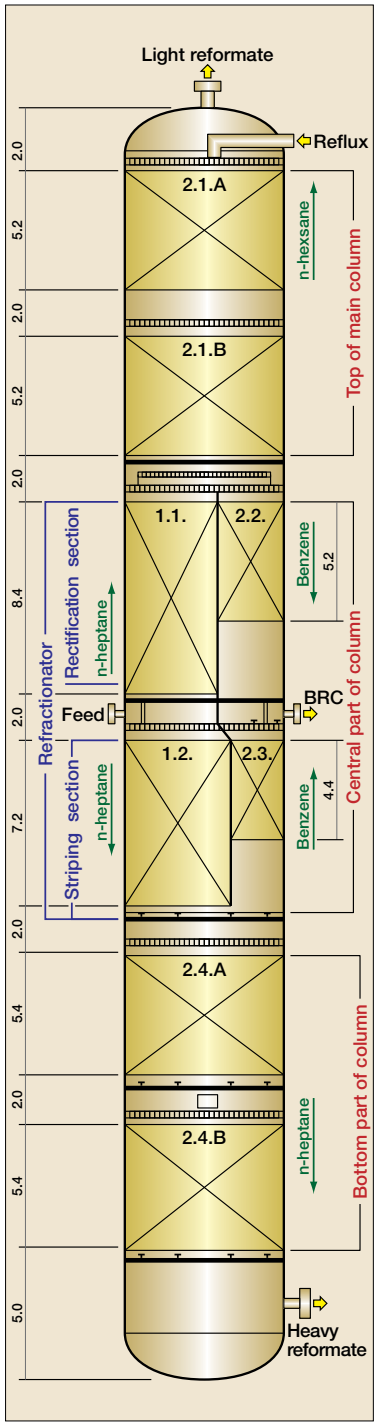


Figure 8 Sized divided wall column for reformate fractionation

Annual cost estimation is based on processing of 268 991 t/y of feed. For processing 1000 kg of feed in a conventional separation sequence, 28.85 kg of refinery fuel gas is needed, while for the same quantity of processed feed in a DWC, 15.98 kg of refinery fuel gas is needed. (The price of refinery fuel gas is €0.42/kg of feed.)

For calculation of a DWC, air coolers were used as the most economical heat exchangers. (Electrical energy costs €120/MWh.)[1] **Table 9** shows the condensers' heat duty as well as their electrical energy cost. The refinery can save an annual €21 760 by using a DWC instead of a conventional separation sequence, based on 268 991 t/y of feed processing.

For estimation of investment costs, Dejanovic's method was used.[13] Estimated investment costs include the costs of installing a new DWC shell, the internal architecture of the column, and installation of an evaporator and condenser. An existing plant for reformate fractionation was used as the basis for design. Investment cost estimates for a conventional separation sequence and a DWC were made.

Initial calculations indicate that, regardless of the much higher investment costs of DWC equipment with structured fillers, the cost of a new DWC is no higher than the cost of the present columns (taking into account the accompanying evaporators and condenser-coolers in both cases). This is logical because reformate fractionation using DWC eliminates the investment costs of one column with accompanying equipment. All of the existing equipment (evaporators and coolers) can also be utilised for the DWC. The resulting reduction energy consumption of more than 50% would make investment in and operation of a DWC solution economically viable.

Conclusions

Using a calculation method with a Vmin diagram, a reformate splitter in INA Rijeka refinery was redesigned as a divided wall column. After simulation of the base case, a good match with actual process conditions was observed. Then a complete calculation methodology for a divided wall column was performed. Calculations include a basic shortcut calculation, thorough rigorous calculation, construction of Vmin diagram, thermal connection of columns and sizing of the unit as a divided wall column.

The Vmin method seems to provide fast and relatively easy estimates of operational flows, based on possible operational or investment savings.

Rigorous simulation of the base state of the unit, at a capacity of 80 t/h, showed a good match with actual process parameters, and the quantity and quality (for instance benzene content) of the resulting products correspond to the quantity and quality of the products produced by the actual reformate splitter unit.

In a process simulation using ChemCAD, separation of benzene and n-hexane proved to be the hardest task, which corresponds to engineering experience in the operation of the actual unit. An optimal mode at which a greater percentage of n-hexane would finish in light reformate, whilst keeping the content of benzene in light reformate below 2 wt%, was sought.

Design of this type of column enables operation of the reformate splitter as one column with three products, thus eliminating the work of the second column, one furnace for reboiling column bottoms, and a system for condensing top product.

Sizing of a DWC was based on the existing 322-C-001 column. A column height of 56 m (including internal architecture) would be sufficient to obtain product quality.

DWC would be about 20 m higher than the existing 322-C-001 column, but this does not exclude the possible redevelopment of an existing column to DWC. The position of the dividing wall, as well as flows through the column sections and the corresponding pressure drop through the column, were calculated for 322-C-001 column's diameter of 2.3 m, and the calculation showed that the diameter of the column is not a limiting factor for its use as a DWC. The required height of the column can be obtained by separating and redesigning with new parts a section of the existing shell.

Preliminary economic calculations indicate that more than 50% reduction in energy consumption would make investment and operation of a DWC economically viable.

Acknowledgement

We would like to thank Dr Igor Dejanović for his help and support in preparing this article, and University FKIT, Zagreb,

for the use of ChemCAD simulation software.

References

1 Halvorsen I J, Skogestad S, Energy efficient distillation, *Journal of Natural Gas Science and Engineering,* 2011, vol 3, no 4, 571-580.

2 Petlyuk F B, Platonov V M, Slavinskii D M, Thermodynamically optimal method for separating multicomponent mixtures, *International Chemical Engineering*, 1965, vol 5, no 3, 555-561.

3 Dejanović I, Razvoj metode dimenzioniranja kolona s razdjelnom stjenkom (Developing of a model for divided-wall columns' design), Doktorska disertacija, Sveućilište u Zagrebu, Fakultet kemijskog inženjerstva i tehnologije, Zagreb (in Croatian), 2010.

4 Brugma A J, Process and device for fractional distillation of liquid mixtures more particularly petroleum, US Patent 2295256, 1942.

5 Stupin W J, Lockhart F J, Thermally coupled distillation - a case history, *Chemical Engineering Progress*, 1972, vol 68, no 10, 71-72.

6 Tedder D W, Rudd D F, Parametric studies in industrial distillation. Part 1. Design comparisons, *AICHE Journal*, 1978, vol 24, no 2, 303-315.

7 Fidkowski Z, Królikowski L, Thermally coupled system of distillation columns: Optimization procedure, *AICHE Journal,* 1986, vol 32, no 4, 537-546.

8 Chavez C R, Seader J D, Wayburn T L, Multiple steady-state solutions for interlinked separation systems, *Industrial & Engineering Chemistry Fundamentals*, 1986, vol 25, no 4, 566-576.

9 Wright R O, Fractionation apparatus, US Patent 2471134, 1949.

10 Kaibel G, Distillation columns with vertical partitions, *Chemical Engineering & Technology*, 1987, vol 10, no 1, 92-98.

11 Sumiju Plant Engineering Co. 2013, viewed 1 Nov 2012, www.spe.shi.co.jp/en/distill1.html.

12 Rix A, Olujić, Ž, Pressure drop of internals for packed columns, *Chemical Engineering and Processing: Process intensification*, 2008, vol 47, no 9-10, 1520-1529.

13 Dejanović I, Matijašević L, Olujić Z, Divided wall column application for platformate splitter – a case study, Pierucci S, Buzzi Ferraris G (editors), 20th European Symposium on Computer Aided Process Engineering – ESCAPE20, Elsevier B.V., Ischia, Naples, Italy, 6-9 June 2010.

Saša Polovina is a Project Expert with INA's Belarus Naftan team. He was previously Area Manager at INA's Rijeka refinery and is now responsible for making detailed engineering project plans, as well as supervision, commissioning and start-up of process production plants. He holds an MSc in chemical engineering and technology from the University of Zagreb, Croatia.

Srećko Herceg is with INA's Belarus-Naftan team at JSC Naftan, Novopolotsk, Belarus. Projects he has been working on include construction of a low temperature isomerisation unit, revamp of a HDT diesel and naphtha unit, and construction of a sulphur recovery unit with a sour water stripping unit. He has been undertaking postgraduate study at the Faculty of Chemical Engineering and Technology at the University of Zagreb.

Ana Granić Šarac is a Process Engineer at INA's Rijeka refinery, providing technical support for processing plants including catalytic reforming, hydrogen management and purification, isomerisation, LPG/gasoline Merox, and gasoline hydrotreating. She graduated as a chemical engineer from the University of Zagreb's Faculty of Chemical Engineering and Technology.