

# Distributed distillation with heat integration

Increased feed capacity and/or reduced energy cost are often possible in the development or modification of a distillation sequence, using a process with proven heat integration concepts, such as side reboiler or pump-around loop

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Distributed distillation has been practised, to a limited extent, in natural gas processing, olefin production and refinery units, for over 30 years. Its utilisation can allow a 10 to 30 per cent reduction in net separation energy usage at the cost of adding more equipment. However, distributed distillation concepts have received little attention in textbooks or industry focus publications, while a steady progression of related academic papers have been published.

New requirements for motor gasoline and diesel fuel will call for modifications to many existing refinery units. Producing products containing less aromatics, sulphur, and/or olefins are projected to require increased hydrotreating severity, reactor revamps, and building new hydrotreating reactors.

High vapour pressure  $C_5$  olefins are more reactive in terms of smog formation than light iso or normal paraffins such as butane and iso-pentane also commonly present in motor gasoline. In high smog areas, removal of  $C_5$  olefins from motor gasoline is desirable and may be required.

Energy intensive distillation systems are fundamental to many processing units in a refinery. Natural gas processing for NGL recovery, LPG processing and many petrochemical units, also use a sequence of energy intensive distillation steps. Production of ethylene and propylene by steam cracking, and the manufacture of alpha olefins from ethylene, are classic examples of units using a sequence of distillation towers.

Environmental incentives to reduce greenhouse gas emissions or new fuel taxes may provide major incentives for improved energy efficiency and reducing furnace heat duties. Distillation towers are major energy users in many process units. Improving the energy efficiency of traditional process units is critical to reducing operating cost, fuel usage and related flue gas carbon dioxide production. The economics of energy related modifications are frequently

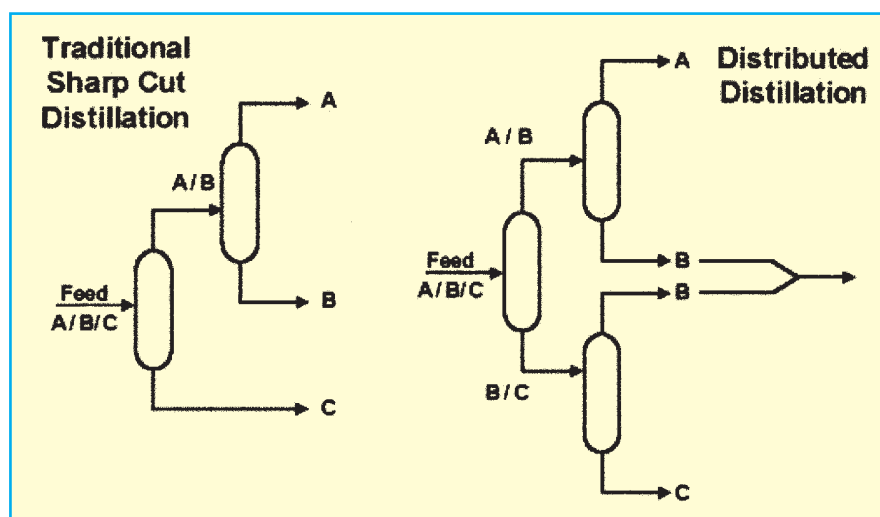


Figure 1 Three-component separation system

improved when a revamp combines a capacity expansion or a fundamental design change.

Figure 1 shows a traditional three-component separation sequence using two towers and a three-tower distributed distillation system for the separation of a three-component mixture of A, B and C. Adding a tower to create a distributed distillation system is sometimes ideal for a separation train capacity revamp. A revamp based on installing a new tower in an operating unit will typically minimise the shutdown time and modifications to the existing tower systems.

A distributed distillation system will frequently increase opportunities for heat integration by shifting heat duty temperature levels, or producing different purity product streams. A new tower design may result in improved heat integration by allowing an energy use cascade. Condensing the overhead vapour from a distillation tower in a heat exchanger providing reboil heat input to a second tower is a classic energy saving concept.

A distillation energy cascade to a side reboiler or feed heater is sometimes an option when temperature levels do not allow heat input to a tower bottom

reboiler. A distillation tower pump-around loop is an example of an energy cascade where reflux is provided to a tower and reboil or feed heating is provided to a different system.

In this article, distributed distillation is viewed as a separation sequence that provides one or more parallel flow paths for the production of a boiling fraction or component. Some authors have used the term Progressive Separation to describe one type of distributed distillation. The term is also used to describe a proprietary crude oil separation scheme promoted by Elf Aquitaine and Technip [US Patent 4664785; Process for distillation of petroleum by progressive separations; 1987].

In this article, Progressive Separation is not used as a general process design description.

The basic concept in distributed distillation is to minimise the number of sharp separations used to fractionate a multi-component mixture. Within a simple distillation column or equilibrium flash drum, key and non-key components are separate, based on relative volatility.

Distillation of a multi-component mixture results in the development of a concentration profile of key and non-key

components or boiling cuts at each distillation stage.

A two tower system, as shown in Figure 1, separating components A, B and C does not utilise the concentration profile of the middle boiling component B within the first tower. The three-tower sequence, shown in the figure, allows the middle boiling component B in both the overhead and bottom products from the first tower while making a separation between A and C.

Using the concentration profile developed, the first tower in a downstream separation typically reduces total energy usage.

### Revamp potential

A two tower system like the one shown is typical of many existing process designs. Reducing the purity of one or both products from the first of the two towers allows operating at a lower reflux flow rate and reboiler heat input. This purity change then allows increased feed capacity while operating to an existing liquid or vapour traffic limit.

Installation of a new tower to create a distributed distillation system, as shown, is frequently an option to allow increased production without product purity changes. Using a new tower for the A/B/C or A/B or B/C separation also allows adding stages to reduce energy use. This new tower provides increased opportunities for heat integration to allow an energy cascade.

Using a distributed distillation design instead of adding one or more towers in parallel or series provides a cost effective capacity revamp and reduced operating cost per ton of product.

Many process units have distillation towers, designed years ago, with fewer equilibrium stages or trays than would be used for a new design. Other distillation towers may be operating with a different feed from design or higher purity product requirements, resulting in less than optimum capacity, energy usage or product purity.

Providing a higher purity feed to a distillation tower normally results in lower separation energy usage. A three-tower distributed distillation system produces a stream richer in component A and a stream richer in component C than a two-tower system. A new tower operating in this mode frequently allows increased feed capacity by shifting liquid and vapour traffic from the two existing towers.

Refinery and petrochemical units typically have a life cycle in excess of 25 years. During this time, process units typically experienced capacity expansions or modifications for feedstock changes or new product requirements. Distributed distillation is an approach to

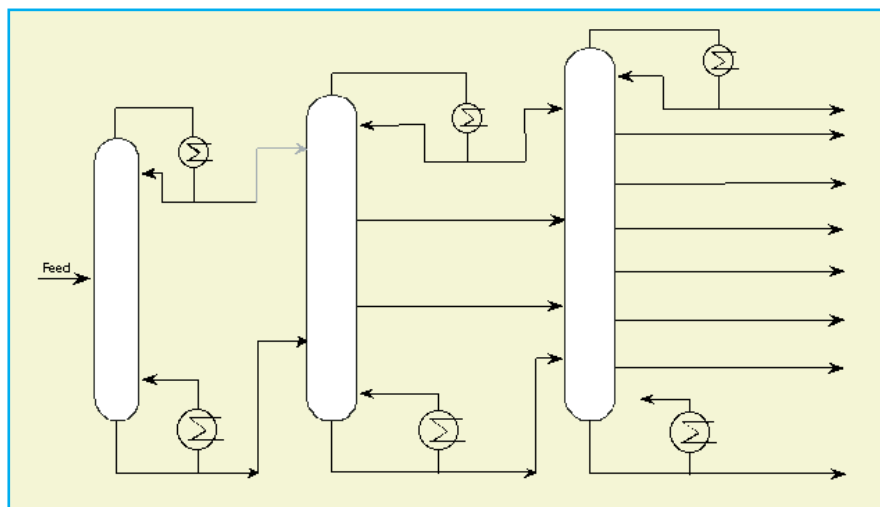


Figure 2 The Brugma distillation concept

consider when unit modifications are considered.

### Brugma distillation concept

The Brugma distillation concept shown in Figure 2 was an early approach to reducing energy usage by using reboiled towers in series [Patents Dutch (1937) and US (1943) issued to A J Brugma].

In the Brugma design, reboil heat input may be provided at three temperature levels and reflux condenser heat removed at three temperature levels. A Brugma system may be viewed as an early form of distributed distillation. The design uses the multi-component or boiling fraction composition profile developed in the upstream tower, making a rough separation.

A two-tower Brugma distillation system allows production of three or four boiling fractions. Processing a mixture of A, B and C in a two-tower Brugma distillation system results in the first tower producing streams rich in A/B and B/C, with the second tower making a side draw cut rich in component B. This system may be viewed as a variation of the three-tower system shown in Figure 1.

Producing component B as a side draw instead of using two towers presents a trade-off of equipment usage, energy usage, or purity of the B product.

Each of the towers shown in Figure 1, could require an overhead condenser, reflux drum, reflux pump set and reboiler. Using a

two-tower Brugma style system for separation of the A/B/C mixture has the potential to reduce the number of equipment items and the capital investment relative to a three-tower system.

In both the two-tower Brugma and three-tower systems the potential energy cost saving of first producing streams rich in A/B and B/C is possible. Producing product B as a side stream reduces the degrees of freedom available for purity control of three products.

### Petlyuk distillation concept

Figure 3 shows the Petlyuk distillation concept for processing an A/B/C mixture [Petlyuk et al; Thermodynamically optimal method for separating multicomponent mixtures; *International Chemical Engineering*, July 1965]. In a classic Petlyuk design, the first tower is supplied with reflux and

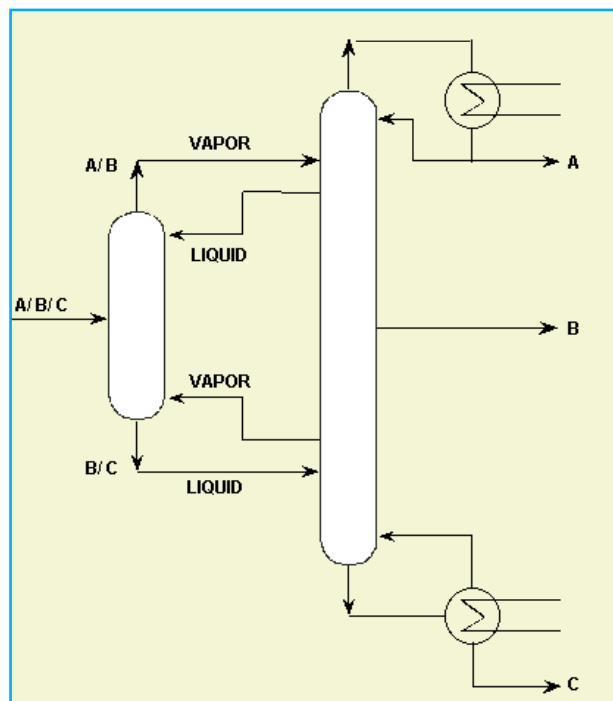


Figure 3 Petlyuk thermally coupled scheme

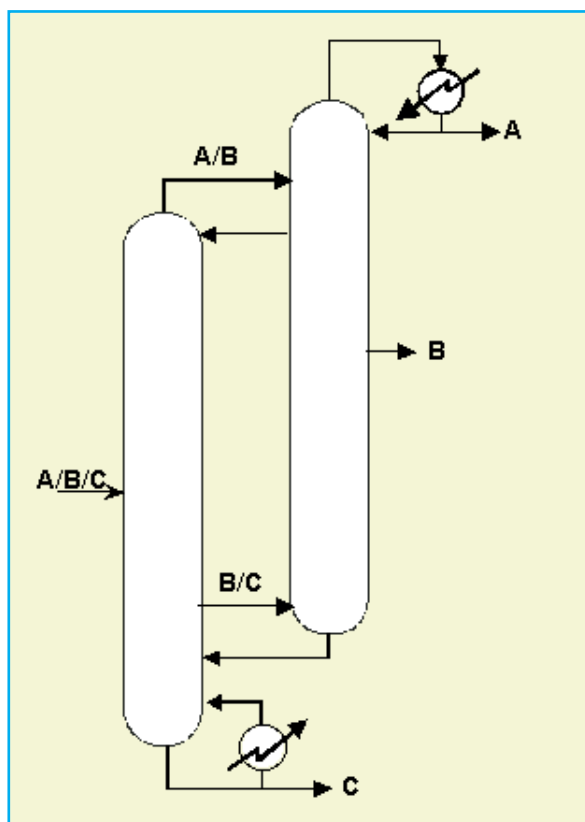


Figure 4 Modified Petlyuk structure for vapour flow control

stripping vapour provided by the downstream tower. As shown, the first tower produces streams rich in A/B and B/C and the second tower produces a side draw stream rich in the middle boiling component B.

The Petlyuk design has eliminated the condenser and reboiler associated with the first tower in the Brugma style system. Petlyuk's design results in all reboil heat input being supplied at the higher temperature required for boiling component C, and heat is rejected at the condensing temperature of component A. A fully thermally coupled system, such as shown in Figure 3, uses less reboil heat input than traditional separation configurations.

A fundamental issue in applying the Petlyuk concept is control of stripping and reflux rates within parallel tower sections operating at approximately the same pressure. Figure 3 does not show provisions for regulation of stripping vapour and reflux flow within either tower, as several control and design concepts exist.

Most applications of the classic Petlyuk design will require controlling both stripping vapour flows between the two separation zones operating in parallel. Unless external energy is provided, the pressure drop available vapour transfer between towers is very small and determined by the number of trays and tray pressure drops in the two towers. US Patent 4230533 illustrates one approach

to liquid and vapour flow control.

If the upstream tower in Figure 3 is short and elevated, then gravity flow with traditional flow meters and control valves may be used. If gravity flow is not possible or cost effective, then one or more pumps may be used. Providing stripping vapour to the first tower at a slightly higher pressure than the downstream tower, allows direct stripping flow measurement and regulation.

An ejector, also known as a jet compressor, could be used to boost the stripping gas pressure if a suitable ejector motive gas is available. This motive gas could be steam or a hydrocarbon vapour, depending on specific design issues.

A single stage compressor with a compression ratio of about 1.1 could be used instead of an ejector. Single impeller compressors, similar in appearance to a single stage pump, have been used in light hydrocarbon services for over 25 years.

### Petlyuk design variations

One variation of the Petlyuk concept is shown in Figure 4. These design variations may provide control and cost advantages over the classic design while providing good energy efficiency [Agrawal and Fidkowski; More operable arrangements of fully thermally coupled distillation columns; *AIChE Journal*, Nov 1998; and *AIChE Spring National Meeting*, 1999].

The modified design in Figure 4 uses two sequential towers for improved control of stripping vapour flow by allowing the first tower to operate at a much higher pressure than the second tower producing streams rich in A and B. In this design, conventional pumps provide two reflux flows from the second to first tower.

Avoiding the use of an ejector or compressor for vapour transfer may allow reduced investment and reduced operating cost for some designs. US Patent 5953936 provides a summary of prior art and illustrates variations of the Petlyuk concept.

Another variation of the Petlyuk concept uses a side reboiler located above an internal head in the second distillation tower. This internal head allows stripping vapour flow through a control valve into the first tower or upper sec-

tion of the second tower. The liquid level above the internal head provides the pressure required for gravity flow from the upper section around the internal head through a seal loop or control valve. Use of a side reboiler potentially allows heat input to have a cooler temperature than the bottom reboiler heat source.

Petlyuk concept variations using a single tower with a vertical internal partition instead of two towers are described in several patents. US Patent 5755933 provides a starting place to review prior patent art and related publications. Using a tower within a tower instead of a vertical internal tower partition to produce one or more side stream products is also described in patents. US Patents 3844898 and 4615770 provide a starting point to review this prior art.

### Prefractionation

Installation of a reboiler on the first tower shown in Figure 3, to generate stripping vapour instead of using vapour from the downstream tower, creates a prefractionation tower system. This prefractionation tower would provide a capacity or reboiler energy incentive similar to the first tower in a Brugma or the three-tower distributed distillation system shown in Figure 1, but with heat removal at the condensing temperature of component A.

Using reflux supplied from the downstream tower using a pump eliminates the need for one reflux condenser and one reflux drum used in the Brugma or the three-tower distributed distillation system shown in Figures 1 and 2. A prefractionation tower design also may allow heat boiling the B/C mixture to be supplied at a lower temperature than heat used for boiling component C.

### Pinch technology

Academic and industry-focused publications have extensively presented capacity and energy benefits of concepts such as a pump around loop, side reboiler, side stripper, and a prefractionator tower. These concepts are frequently presented in terms of one or two distillation towers.

In the past 15 years, numerous articles have been published on pinch technology as a systematic tool for studying process energy usage and reducing utility (import) energy usage. A key input into a process case study based on pinch technology is the definition of each heating and cooling duty and the related temperature levels.

Application of pinch technology allows a logical approach for the review of a process unit or total plant heating and cooling loads. Based on this review either specific energy duties or temperature

levels will be selected for consideration of process modifications.

Both distributed distillation and traditional heat integration concepts frequently allow shifting the temperature level of a specific energy duty or dividing a heat duty between two or more temperature levels. Distributed distillation is a powerful process design concept when used to develop energy and temperature level options for use with a classic pinch technology heat integration study.

The typical objective in a distributed distillation system is to reduce energy usage by shifting the purification steps closer to being thermodynamically reversible. Theoretically, in a completely distributed distillation tower sequence, the most volatile and least volatile components feeding each tower will be separated.

### Crude oil applications

Distributed distillation may be viewed as a direct extension of conventional process design practices, such as using a flash drum, a pre-stripper or a pre-fractionator for increased separation train capacity or reduced energy cost. These concepts have been applied to crude oil fractionation trains to increase capacity, reduce fired heat duty, or allow using a smaller diameter crude oil atmospheric tower.

Some crude oil train process designs use only pre-flash drum to route light-ends to the atmospheric tower upper section or overhead condenser system. Other crude unit designs replace the pre-flash drum with a reboiled pre-stripper or full distillation tower system producing a liquid product with a lower specific gravity than light naphtha product from the crude oil atmospheric tower reflux drum.

The terms such as super-atmospheric tower or gasoline column are sometimes used to describe a tower provided with reflux to the top tray. The terms pre-flash stripper or pre-flash tower are frequently used to describe a separation step located upstream of a crude oil atmospheric tower operating without reflux.

The application of two towers upstream of the crude oil atmospheric tower in one modern 170 000 bpd refinery has been called Progressive Distillation [Rhodes; Environmentally advanced refinery nears start-up in Germany; *Oil and Gas Journal*, March 1997]. These towers operate at about 2 and 1 barg with the first tower producing a light naphtha with a 80–90°C boiling range and a second tower producing a medium naphtha with a 90–120°C boiling range.

This Progressive Distillation flowsheet applies distributed distillation creating a

variation of the crude oil separation scheme known as Progressive Separation.

Crude oil fractionation and the related heat exchanger networks have been extensively studied, using pinch technology, during the past 15 years. Pinch technology provides a rigorous method to study incentives and options to reduce energy usage. Its application for studying the use of a flash drum or tower upstream of a crude oil atmospheric tower is described by Trivedi, Andrawis and Habibullah in "Revamping crude units to increase capacity" (*PTQ*, Summer 1998).

Their article also considered use of a vacuum pre-flash tower upstream of a crude oil vacuum tower.

A refinery crude oil atmospheric tower may be considered as a type of complex distributed distillation where one tower makes several side stream products. The composition profile within the crude oil atmospheric tower is determined by the material balance at each draw off location, by heat input, and heat removal at various pump-around loops.

Side stream strippers are typically used to minimise light-ends in the side draw products. These side stream strippers may be viewed as being directly coupled or thermally coupled to the

crude oil atmospheric tower. The tower pump-around loops serve the same function as side condensers in adjusting the internal reflux flow and related tower composition profile for control of side draw purity.

### Hydrotreating issues

New requirements for production of lower sulphur motor gasoline and diesel fuel are projected to require new hydrotreating units. Optimisation of new and existing hydrotreating reactor feed streams may involve distillation train modifications. Sulphur compounds typically distribute within cracked and virgin material with a higher sulphur concentration in the higher boiling range fractions. Sulphur removal from cyclic hydrocarbons structures higher severity hydrotreating than paraffin or olefin compounds.

Motor gasoline boiling range fractions produced during high temperature cracking in an FCC, coker or olefin unit furnace coil will contain aromatics, olefins, diolefins and some sulphur compounds. Mild hydrotreating is typically used for the conversion of diolefins into olefins and removal of more reactive sulphur compounds.

Removal of diolefins is required for long term storage stability of motor gasoline produced by high temperature



cracking. Reactive sulphur compounds are removed to produce products passing a copper strip or similar corrosion test. High severity hydrotreating for near total olefins and sulphur removal is normally used only for C<sub>6</sub> or C<sub>6</sub>-C<sub>8</sub> fraction feeding an extractive distillation for aromatics recovery.

A three-tower distributed distillation system as shown in Figure 1 would allow hydrotreating of the lighter A/B cut at different conditions or using a different catalyst from that used for hydrotreating the heavier B/C cut. If a new hydrotreating unit is required for processing stream C shown in Figure 1, providing some incremental capacity for a B/C mixture could allow design optimisation through increased capacity or less modification of an existing hydrotreating unit processing A/B.

Separation train requirements for sulphur removal or other processing steps will determine the extent to which a distributed distillation sequence may be utilised. Amine and caustic (KOH or NaOH) treating in a liquid to liquid contacting system will remove a high percentage of sulphur compounds in the C<sub>3</sub> to C<sub>5</sub> boiling range. Distributed distillation could also be used to optimise hydrotreating investment by producing additional cuts with different boiling ranges and sulphur levels.

Frequently, mild hydrotreating a FCC gasoline C<sub>5</sub> fraction separately from the C<sub>6</sub>-C<sub>8</sub> or the C<sub>9</sub>+ fraction has several advantages. These could include less light olefins conversion with a smaller decrease in motor octane number (MON) while allowing higher severity reactor operation for sulphur removal from ring compounds.

#### Alkylation, isomerisation and reformer units

Increased isomerisation unit and alkylation unit capacity are options to produce gasoline meeting new environmental objectives. Adding an isomerisation reactor and one or more distillation towers with increased heat integration is an option for increased gasoline production and obtaining a higher motor octane number.

Providing a sulphuric acid alkylation unit with feed streams containing different olefin types or olefin concentrations is an option for increased unit capacity or product value through optimisation of reactor conditions and acid concentration. Improvement amylene alkylation is possible by removing cyclopentene from the alkylation feed by distillation and mild selective hydrogenation to remove the diolefins [Peterson et al; A cheaper way of alkylating propylene and amylenes; *PTQ*, Autumn 1999].

An alkylation unit will frequently

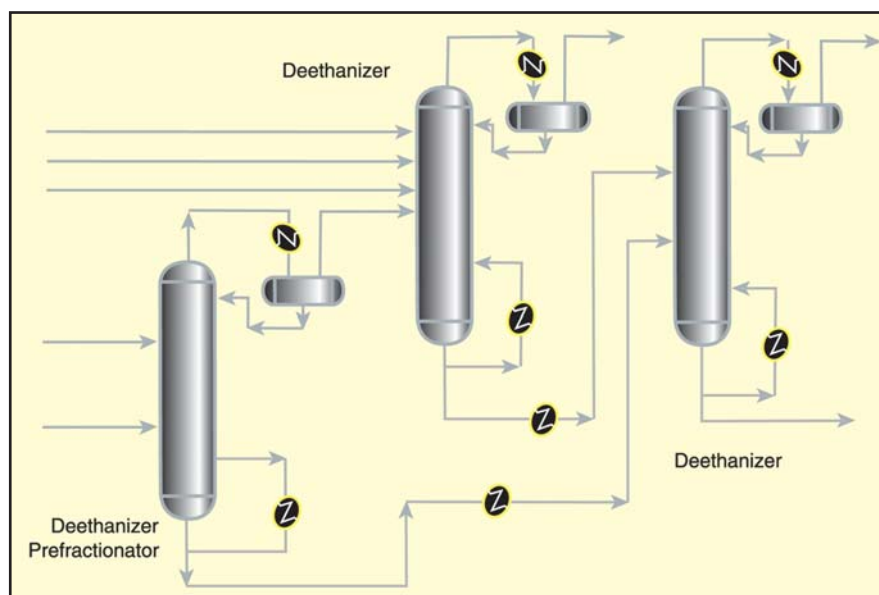


Figure 5 Demethaniser prefractionator flow scheme

have a deisobutaniser (DIB) tower producing an isobutane overhead product, a normal butane side draw product and a bottom C<sub>6</sub>+ (alkylate) product. Many DIB tower designs use a side reboiler to minimise the amount of heat duty supplied at the tower bottom temperature.

Installation of a DIB tower prefractionator similar to the concepts shown in Figures 1 and 2 could allow revamp for increased capacity or reduced energy cost by shifting temperature levels for reboil heat input. Installation of a side stripper at the normal butane draw location is an option to reduce the amount of isobutane in the C<sub>4</sub> recycle to an isomerisation unit.

The side stream stripper could also be used to increase the amount of lower temperature heat duty supplied to the DIB tower.

Reformer feed fractionation is currently practised in some refineries to reduce benzene production by reducing the amount of the benzene precursors feeding the reformer reactor. A distributed distillation design could be ideal for producing streams rich in isopentane, normal pentane, C<sub>6</sub> paraffins, and C<sub>7</sub>+ for use as a reformer feed. Isomerisation of C<sub>5</sub> and C<sub>6</sub> rich fractions in separate reactors allows recycle optimisation for increased capacity and obtaining higher motor octane number blending streams.

Some fixed bed reformer reactors have been converted into isomerisation units to reduce the amount of aromatics in blending stocks used for motor gasoline.

#### LPG and NGL processing

The combination of multiple effect distillation and distributive distillation for NGL fractionation of propane, isobutane, normal butane, and gasoline components is discussed in US Patent 5806339. This patent illustrates heat

integration concepts for increased thermodynamic efficiency through the use of partial cascade reboiling of columns with butane being distributed in two separate columns.

Variations of this patent concept could have applications in an alkylation unit or an isomerisation unit processing a straight run C<sub>5</sub>/C<sub>6</sub> mixture. US Patents 5755115 and 5791161 illustrate uses of a side reboiler for an NGL unit.

#### Olefin production

Figure 5 shows an application of distributed distillation in an olefin unit with multiple liquid feed streams produced using a sequence of refrigeration temperature levels to condense a cracked gas mixture.

The warmest of feed stream contains a higher percentage of C<sub>3</sub> (propylene/propane mixture) and a low concentration of methane than colder feeds. The coldest of feed stream contains a high percentage of methane, some ethylene, and no C<sub>3</sub>. Providing two feed streams to the deethaniser, which are partially separated, allows a lower reflux ratio than a single feed deethaniser. Two-feed streams also allow more optimisation of feed vaporisation to minimise energy cost.

Installation of a demethaniser prefractionator, as shown in Figure 5, allows both an olefin unit capacity and energy saving revamp option. Some revamps add a stripper tower with a reboiler and no reflux condenser. This demethaniser pre-stripper provides an energy saving similar to adding a demethaniser side reboiler. A demethaniser prefractionator with a reboiler and reflux condenser may provide similar energy saving as a demethaniser with a side reboiler and side condenser. Use of parallel flow

paths for  $C_3$  processing in an olefin unit has been described by others as Progressive Separation and a distillation Split Flow Scheme [Kaiser and Picciotti; Better ethylene separation unit; *Hydrocarbon Processing*, Nov 1988; and Di Cintio et al; Separate ethylene efficiently; *Hydrocarbon Processing*, July 1991].

Several contractors design olefin units with a parallel  $C_3$  flow path.

During the past 10 years, six new olefin production units and one revamp project, designed by Stone & Webster, have used a distributed distillation design to route about 25 to 40 per cent of the unit  $C_3$  (propylene and propane) production around two distillation towers operating in series.

In these new olefin units, distributed distillation of the  $C_3$  fraction reduces the diameter of two towers, reduces reboiler heat duties, and results in a cooler process side temperature for one reboiler by reducing the  $C_3$  concentration to the reboiler.

Most of these olefin unit designs by the company have also provided parallel flow paths for  $C_2$  (ethylene and ethane) processing. This parallel  $C_2$  flow is a variation on the design shown in Figure 5, with the demethaniser bottom flow bypassing the deethaniser to become the top feed of a two feed  $C_2$  splitter. This top  $C_2$  splitter feed was a higher concentration of ethylene than the deethaniser overhead.

Providing two feed streams to the  $C_2$  splitter also allows reduced energy usage. Vaporising some or all of the demethaniser bottom flow provides another option to minimise energy cost. A description and flowsheet for this design has previously been published [Reid and Bush; Ethylene revamps using Advanced Recovery System; *PTQ*, Winter 1997/98].

In a grassroots olefin unit design, several reasons exist to consider installation of a demethaniser prefractionator feeding a demethaniser. A dual demethaniser system normally allows the prefractionator, operating at warmer temperature conditions than the demethaniser tower, to be built from lower cost carbon steel. The demethaniser with an overhead temperature approaching 100°C is frequently built from a 304 or higher grade stainless steel.

Another consideration is that a limited number of shops are able to build towers with a pressure rating of 34 atmospheres with a diameter greater than 3m.

Installation of a demethaniser pre-fractionator with a reflux condenser instead of a simple stripping tower allows shifting some of the total reflux condenser duty to a warmer and low

cost refrigeration level. Using two reboilers operating at different temperatures, in the dual tower system, also improves reboiler refrigeration recovery.

#### Number of equilibrium stages

Many older towers have fewer trays than could be justified in a new design using recent energy cost data. A classic energy revamp concept is to install more trays in an existing tower using a closer tray spacing or adding a tower in series. Providing more equilibrium stages allows a tower to be operated at a lower percentage above the theoretical minimum reflux ratio.

Installing more trays on a close spacing is a well documented revamp option for either increased feed capacity or reduced energy usage. Even with high capacity tray designs, a closer tray spacing will frequently result in a lower ultimate capacity for a fixed tower diameter. Installing a tower in series to provide more equilibrium stages is frequently not cost effective unless both a capacity increase and energy saving are obtained.

#### Side reboiler use

Frequently, the highest liquid and vapour traffic in a distillation tower is directly above the bottom reboiler. When tower capacity is limited by these bottom section trays, installation of a side reboiler sometimes allows both a capacity increase and energy cost reduction. Occasionally a side reboiler revamp will be justified solely for increase capacity.

A few process designs use a side reboiler to reduce the diameter of a new tower. This concept may result in a lower total installed cost or allow a higher capacity unit while using a shop instead of a field fabricated tower. Separations such as propylene from propane and isobutane from normal butane use a high reflux ratio, a large number of trays and large diameter towers.

In one unit, producing polymer grade propylene, a side reboiler maximised waste heat recovery and allowed use of a stripping tower that was 3.5m in diameter in series with a rectification tower of 4.7m in diameter.

A revamp reducing tray traffic in a tower zone normally requires adding more equilibrium stages or operating with a higher total amount of heat input and removal to obtain the same product separation as before. In terms of the classic McCabe-Thiele diagram, adding either a side condenser or side reboiler results in adding a new operating line. Typically, one or more of the process operating lines are moved closer to the equilibrium curve.

The primary benefit of using a side condenser and/or side reboiler is

reduced energy cost through shifting some of the separation energy to a lower cost utility temperature level. Using either or both a single side condenser or a side reboiler normally does not result in a net reduction in total separation heat duty, if both process designs have a similar stage count and location optimisation.

A net reduction in total separation heat duty may be possible when addition of an intermediate heat exchanger is combined with additional separation stages to allow a closer approach to the theoretical minimum boil-up and/or reflux ratio.

As the amount of heat input through a side reboiler is increased in a tower with a fixed number of stages, frequently the total heat input required for a constant feed flow rate and composition will increase. Occasionally, as heat duty is shifted from the bottom to side reboiler no change in reflux rate and total heat input will be required to obtain constant top and bottom product purity.

If a tower feed location or number of stages is not optimum for current operating conditions, a shift in tower traffic and McCabe-Thiele diagram operating line sometimes does not result in higher energy usage.

Using a side condenser, pump around, and/or a side reboiler typically results in the McCabe-Thiele diagram process operating lines being closer to the equilibrium curve. Designing a tower system with these process operating lines closer to the equilibrium curve typically requires using more trays, more heat exchanger surface area, high quality vapour liquid equilibrium data, and past knowledge of tray efficiency.

A successful process design utilising a side reboiler must be based on a simulation that correctly estimates both tower equilibrium stages and the physical tray temperature profile. A robust process design should consider changes in the tower temperature profile with changing conditions such as feed composition and operating pressure that would change the side reboiler temperature approach.

Occasionally, addition of a side reboiler for increased feed capacity will be combined with installing more trays on a reduced tray spacing between the side and bottom reboilers. Reduced tray traffic below a side reboiler will frequently allow using a reduced tray spacing to obtain more equilibrium stages. Adding a few stages below the new side reboiler will typically allow tower operation at the prior condenser heat duty and total reboil heat duty while reboil vapour is

produced at two stage locations. The return on investment for reducing energy usage in single distillation tower will frequently not justify a revamp project. Combination addition of side reboiler with a tray revamp or addition of a tower in series sometimes provides both a small capacity increase and reduced operating energy cost. Adding a tower to create a distributed distillation system can sometimes provide both a large capacity increase and new options for reducing total energy cost.

### Side reboiler revamp

Installation of a side reboiler to an existing tower will typically require tower shell welding for addition of liquid supply and vapour return nozzles. Occasionally, a manway cover may be modified to avoid tower shell welding. Whenever possible, side reboiler nozzles should be added using a "hot tap" method to minimise revamp cost and time.

A key advantage to using a hot tap method is that total tower hydrotesting is frequently not required. This method is based on first welding a nozzle pad to the tower, the nozzle pad welding is then hydrotested, an isolation valve is added to the new nozzle, then a hole is cut in the tower using the hot tap machine. In some tower services, the hot tap work may be done prior to a unit shutdown or during preparations for tower entry.

Frequently, a chimney tray is installed to provide a liquid draw to the side reboiler. Returning the vapour or two phase flow from the side reboiler to a chimney tray provides good liquid and vapour distribution for multi-pass tray. Installing of a chimney tray or a pan for liquid collection frequently requires removal of one or two trays.

Occasionally, a tower manway may be utilised for both the liquid supply and reboiler return nozzles by replacing or modifying the conventional manway blind or dome head. This requires adding a liquid flow path to feed a draw-off box welded to the new manway blind or installing a chimney tray. The reboiler return flow may or may not need a distribution header within the tower depending on the return flow liquid content and type of trays being used.

If a return flow distribution header is used, it may be a simple pipe with slots attached to the manway blind and support at the other end with a spring attached to a tray ring.

### Conclusion

Distributed distillation is one of several basic tools in development or modi-

fication of a distillation sequence. Pinch technology concepts frequently show where improved heat integration is desirable. Using one or more side reboilers is frequently a good option for low level heat recovery.

Using a circulating hot oil or hot water system is frequently ideal for linking heat sources and heat sinks.

Increased application of traditional distillation energy saving concepts with proven heat integration concepts can reduce fuel consumption. Less fuel usage results in lower CO<sub>2</sub> emissions. Small reductions in fuel usage or utility

cost saving that requiring a large investment or increased shutdown time are frequently difficult to justify for older hydrocarbon processing plants.

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