High-performance trays alone do not guarantee performance improvements

A high-performance tray does not itself determine the effectiveness of the whole column. Other distillation equipment plays a vital role in a successful column revamp

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High-performance trays are widely used for projects to improve the performance of distillation columns. These types of trays can enhance column performance without complex modifications to the existing tray supports and column. However, utilising high-performance trays is not a guarantee of successful column performance. Other distillation equipment plays a vital role in a successful column revamp and overlooking these critical items often brings inferior results. These overlooked items are discussed and evaluated, and remedies are implemented in an actual hydrocracker revamp case study, which examines how targeted distillation column performance is achieved through careful analysis and design procedures.

Critical tray design issues
Since the early 1990s, high-performance trays have been applied to commercial distillation columns, especially where structured packing is not able to provide a noticeable performance gain. Most high-pressure distillation services where high liquid and low vapour traffic are formed inside the column are good applications for these types of trays. A simple change-out from conventional trays to high-performance trays can provide a noticeable capacity gain without the need for increasing the existing column diameter. Tray efficiency improvement can also be achieved in certain applications utilising some high-performance tray devices available on the market. In addition, column shell modifications, known as field construction hot work, are typically eliminated. This shortens the unit downtime offline, as field construction is reduced, thereby resulting in overall cost savings.

Due to the simplicity of the field fit-up and easily attainable performance gains, high-performance trays have become a common option for distillation column revamps. However, there are cases where the post-modification column performances do not meet targeted objectives in spite of the installation of high-performance trays. Since a large number of design/operation parameters influence the resulting column performance, it is not easy to generalise the typical root causes of inferior performance. Two major areas have often been observed in revamp failures: improper evaluation and selection of optional high-performance tray features; and inadequate or incorrect design of peripheral distillation components within the column.

A simple change-out from conventional trays to high-performance trays can provide a noticeable capacity gain

Optional high-performance tray features
Most high-performance trays in the industry have been designed with various performance-enhancing features to achieve the targeted column performance. These include specialised shaped downcomers, liquid inlet momentum breakers, tray inlet vapour/liquid contact initiation devices, and directional valves positioned in the tray periphery areas. A tray designer's knowledge and experience is critical for the correct selection of the appropriate performance-enhancing features. Keep in mind that while some features help to improve column performance in certain applications, these features can also cause performance losses or unit troubles in other applications. For example, fouling service applications require precise, careful selection of performance-enhancing features. Poor application know-how often causes improper optional features selection in the high-performance tray design.

The shape of valve units utilised on high-performance trays has been developed as a mechanism to improve vapour and liquid contact, minimise pressure drop and maintain desired fouling resistance. The low pressure drop valve units can reduce the overall distillation column pressure drop compared to conventional tray valve units. Lower column pressure drop can improve distillation column energy consumption and reduce the chance of thermal degradation. Therefore, many distillation column revamps aim for minimum column pressure drop through high-performance tray implementation. While this is an important tray design principle, what is often ignored is that low tray vapour velocity can down-
grade tray efficiency due to insufficient vapour/liquid contact volume. The tray designer should take this into consideration when optimising the tray pressure drop to prevent undesired efficiency loss.

Design of peripheral distillation components
The other common root cause of performance failure is poor design of the other distillation equipment in the column. Improperly designed peripheral distillation equipment can result in inferior performance even though the fractionation trays are designed correctly. It is essential to optimise the complete distillation equipment design, including fractionation trays and peripheral equipment, to ensure optimum column performance. The following discussion outlines the components often overlooked in distillation column design.

Chimney tray
The chimney tray is a common device for an intermediate stream draw-out of a distillation column, such as a refinery multi-product fractionator. A chimney tray can also be utilised as a transition device where the number of tray passes is changed in a particular section of the column. Since no vapour and liquid contact is created, liquid is collected and redistributed through a chimney tray. The chimney tray does not cause side draw product loss, which is a common problem when attempting to utilise a fractionation tray in direct side draw applications. In addition, a chimney tray provides better buffering against tray upset conditions and increases residence time for vapour component disengagement.

Chimney overall open area and liquid hold-up volume parameters are usually considered in most industry-standard chimney tray designs. It is well known that a chimney tray also functions as a liquid collection device. On the other hand, it is not well known that chimney trays provide vapour distribution through the column. Vapour distribution is an important factor for column efficiency. Ignoring this factor can cause vapour maldistribution and result in decreased column performance. The vertical distance between the chimney hat and the tray above is critical to ensure proper vapour distribution. An insufficient distance can lead to poor vapour distribution. Constructing an accurate hypothetical angle of vapour distribution can help to anticipate vapour distribution performance through the chimney tray. A wide hypothetical angle between the chimney hat and the fractionating tray positioned above risks a higher chance of vapour maldistribution.

An insufficient peripheral vapour escape area between the chimney tray and the chimney hats negatively impacts chimney tray performance. Chimney trays are usually equipped with chimney hats to guard against process liquid bypassing from the raining-down liquid from above. The vapour escapes from the chimneys through the peripheral space between the hat and the top of the chimney. If this area is too narrow, excessive pressure drop and/or liquid entrainment can be generated and the column performance can be adversely affected. High energy consumption, oil cracking, thermal degradation and/or polymerisation are common symptoms observed in excessive high pressure drop distillation columns.

The individual chimney dimensions (length x width/diameter) and/or the chimney arrangement on the chimney tray deck also influence column performance. Improper chimney arrangements and dimensions can generate a substantial amount of liquid build-up on the chimney tray deck due to significant hydraulic gradient, eventually causing liquid overflow through the chimneys.2

In addition, evaluation of the chimney tray draw nozzle size should not be ignored. The draw nozzle size should be large enough, or the liquid head should be high enough, to generate the required driving force for momentum transfer. Excessive draw nozzle head loss can cause draw trouble, especially if a gravity flow line is utilised. If the withdrawn liquid is transported to a pump suction, the pump net pressure suction head (NPSH) needs to be verified. Increasing the draw nozzle size or the quantity of draw nozzles reduces nozzle head loss. Increasing the liquid head can be an alternate solution, but this can impact the overall column height should elevation space be limited. If these modifications are not feasible, installing a venting loop on the draw line and/or rearranging the piping geometries can help improve draw hydraulics.

Seal pan
The primary purpose of a seal pan in a distillation column is to prevent up-flowing vapour from bypassing through the downcomer. Overflowing liquid from the seal pan provides sealing for the tray downcomer. The seal pan is usually positioned near the reboiler return downcomer. The seal pan is also located in a column transition section where the number of tray passes changes. There are numerous industry guidelines and rules of thumb for design parameters. The vertical distance between the two-phase/vapour inlet nozzle and the column bottom liquid level is usually considered in most seal pan designs. The preferred orientation between the seal pan and the two-phase/vapour inlet nozzle is maintained in most standard seal pan designs. On the other hand, the horizontal distance between the seal pan and the two-phase/vapour inlet nozzle is often overlooked. If this distance is too close, the liquid...
component from the seal pan can be entrained by the two-phase/vapour inlet stream. These entrained liquid components will affect the column efficiency and/or capacity. When seal pans are located in a transition section where the number of tray passes is changed, liquid distribution between the seal pan and the next fractionating tray below is critical. Liquid maldistribution can reduce tray efficiency and/or capacity. The importance of proper liquid distribution cannot be stressed enough when a liquid inlet stream is introduced into a transition section. The seal pan lip throat should be sized correctly as it is designed to transfer the liquid to the inlet panel of the tray below. Otherwise, the liquid can splash into the below tray active area and, in turn, tray performance is adversely affected.

Reboiler
The reboiler return line size and configuration should be carefully designed. In the case of a thermosyphon or forced circulation reboiler type, two-phase flow is usually formed at the reboiler return stream. Undesirable two-phase fluid regime at the reboiler return can cause adverse distillation column operation issues, including entrainment, flow instability, temperature and/or pressure fluctuation, hammering and pipe or equipment erosion. The importance of the reboiler return flow regime is emphasised when the reboiler return line is a 90-degree bend. In some cases, the thermosyphon reboiler is positioned at a lower elevation to obtain the required liquid driving head for the thermosyphon loop. This configuration requires a 90-degree bend reboiler return line. It is known that non-straight piping geometries increase the chance of hydraulic trouble at a two-phase flow regime. Selecting the optimum reboiler process stream flow configuration inside the distillation column is also important. Once-through, recirculation with a baffle and recirculation without a baffle are common process flow configurations in the bottom of a distillation column. Since each configuration has its own inherent strengths and weaknesses, the correct process flow configuration should be selected for problem-free operation.

Internal feed distributor
Feed distribution quality influences overall distillation column performance. Poor distribution quality increases a non-uniform concentration across a distillation column cross-sectional area and results in downgraded performance. The importance of feed distribution is emphasised when a distillation column has multi-pass trays and/or packing. The feed fluid should be matched to the adjacent tray liquid-to-vapour (L/V) ratio as closely as possible for optimal performance. If not, the imbalanced L/V ratio can reduce both the column efficiency and capacity. If the fluid is introduced as two-phase feed, the vapour discharge location of the feed distributor should be appropriately located far enough from the rectification section downflowing liquid; otherwise, the down-flowing liquid can be entrained by the vapour portion of the feed fluid, which results in reduced column performance.

Case study
Unit description and background
The following discussion is a case study of an actual hydrocracker fractionation unit revamp that demonstrates experienced evaluation and well thought out, proven design practices and procedures to fulfill the targeted performance.

A hydrocracking process is a heavy oil conversion process commonly implemented in modern petroleum refineries. It produces...
more middle boiling point range distillates compared to other heavy oil conversion technologies. Global diesel demand growth requires refiners to choose a hydrocracking process for the addition of conversion capacity.

Figure 1 shows the hydrocracker fractionation unit configuration in this discussion. The purpose of the hydrocracker unit is to convert gas oil boiling range materials to valuable light oils. These light oils are mainly middle distillate boiling point range products such as kerosene and diesel. The gas oil boiling range material feedstocks are produced from vacuum gas oil (from the vacuum distillation unit), heavy coker gas oil (from the delayed coker unit) and/or overflash stream (from the crude distillation unit). The hydrocracking reactor effluents are transported to high-, medium- and low-pressure flash drums for initial separation. After passing the three flash drums, the preheated reactor effluents, via heat exchanger trains, are introduced to the debutaniser feed flash drum. Flashed vapour directly fed to the debutaniser and the remaining liquid stream is introduced to the debutaniser after another preheating. At the debutaniser, off-gas and LPG boiling range materials are separated and produced as off-gas and overhead distillate streams, respectively. A furnace reboiler is utilised in this debutaniser. 

C, and higher boiling range materials are produced through the debutaniser bottom stream. This stream is preheated at the product fractionator charge furnace and fed to the product fractionator. This multiproduct fractionator separates the remaining reactor effluent into naphtha, kerosene, diesel and unconverted oil (UCO) boiling range materials. The kerosene and diesel side strippers are a part of the product fractionator. The kerosene side stripper utilises the reboiler, while the diesel side stripper uses superheated steam as the stripping media. Naphtha boiling range materials, which are produced at the fractionator overhead distillate, are further separated at the naphtha splitter in order to separate light and heavy naphtha streams. The waxy UCO stream withdrawn from the product fractionator bottom can be sold as a lube base oil plant feedstock.

When this original hydrocracker unit was commissioned, conventional valve trays were originally installed in all five distillation towers in the fractionation train: the debutaniser, product fractionator, kerosene side stripper, diesel side stripper and naphtha splitter. In 2004, high-performance trays and new feed distributors were installed in the kerosene side stripper, naphtha splitter and product fractionator wash section as part of the unit debottlenecking. These items were designed and supplied by a reputable industry tray supplier.

In spite of the implementation of high-performance trays, this hydrocracker fractionation unit still faced two limitations. Kerosene yield at the product fractionator was lower than expected. Attempting to achieve a higher kerosene yield downgraded the kerosene product quality. The flash point value of the kerosene product was also decreased. The other limitation was observed in the naphtha splitter. The reflux ratio value of the naphtha splitter was higher than expected. The required degree of fractionation between the light and heavy naphtha could not be achieved at the lower reflux ratio.

**Process evaluation**

To resolve unexpected unit limitations, a dedicated process evaluation and overall system troubleshooting were conducted to pinpoint the root causes. A process simulation model was also developed as part of the process evaluation activities. The major purpose of the simulation modelling was to quantify internal vapour and liquid traffic values in the distillation columns. For reliable simulation modelling, operating data obtained from the unit test run were applied and advanced simulation topology was implemented to model the highly non-ideal sections, such as the product fractionator feed charge furnace, transfer line, and wash and bottom stripping sections. An interval of pseudo-components was adjusted to match the obtained laboratory distillation temperature span. **Tray efficiency** for each column fractionation section was quantified through extensive sensitivity analyses. All tray hydraulic capacities were evaluated, based on simulated column internal traffic values.

The simulation modelling identified poor tray efficiencies in the naphtha splitter stripping section. On the other hand, an evaluation of the existing installed high-performance trays and the rest of the conventional trays did not reach the capacity limitation point. The reboiler was suitable to provide the required performance, and the reboiler return size and geometries were acceptable as well. However, it was determined that the seal pan design was inadequate in the kerosene side stripper and naphtha splitter. In the kerosene side stripper...
stripper, the bottom stripping tray seal pan was too close to the reboiler return stream inlet nozzle. The horizontal distance between the nozzle end and the seal pan was 2.4in. The down-flowing liquid from the seal pan could be entrained by the high-velocity reboiler return stream. Figure 2 shows the pre-retrofit kerosene side stripper seal pan and reboiler return.

The transition design between the rectification and the stripping sections also had distribution issues in the naphtha splitter. The rectification and stripping sections were designed with one-pass and two-pass tray geometries, respectively. Feed steam was introduced between the rectification and stripping sections. Down-flowing liquid from the last rectification tray should be irrigated and mixed well with the introduced feed fluid at the top stripping tray inlet panels. Unfortunately, it was found that the rectification section seal pan throat was too wide and a significant amount of liquid was poured directly onto the top stripping tray active area. Liquid maldistribution occurred in the stripping section and caused poor tray efficiency. Figure 3 shows the pre-modification naphtha splitter transition section design. Like the kerosene stripper, the seal pans at the bottom tray were too close to the reboiler return nozzles.

It was also noted that the three chimney trays installed in the product fractionator were not properly designed. The hypothetical angle between the chimney hat and the tray above was too large to achieve adequate vapour distribution in the kerosene pumparound section trays. The chimney tray peripheral area between the top of the chimney and the chimney hat was not sufficient in this section. The wash section chimney tray also had an insufficient hat peripheral area.

Poor liquid and vapour distribution design was found in the diesel pumparound section. The existing trays in the diesel pumparound section are four-pass trays, and the top tray is configured with two off-centre-positioned downcomers. In this geometry, the diesel pumparound return distributor has to be fed to one centre and two side inlets. The split of liquid flow through the original distributor was not proportional to the internal L/V ratio from each pass. Each tray vapour passage open area was inconsistent between the diesel section chimney tray and the neighbouring trays. These inconsistencies caused an uneven L/V ratio in each active area and downgraded tray performance.

Modification of distillation equipment
Based on the aforementioned process evaluation and root cause analysis, the following modifications to distillation equipment were applied to the distillation columns. A segregation baffle was designed and installed between the seal pan and the reboiler return stream inlet nozzle for the kerosene side stripper and the naphtha splitter bottom section. As Figure 4 shows, this baffle segregates down-flowing liquid from the up-flowing reboiler return vapour portion and eliminates the chance of liquid entrainment. The bottom stripping tray downcomer height was extended to increase the distance between the seal pan and the reboiler return nozzle for the kerosene side stripper. Sufficient vertical distance between the extended seal pan and the bottom liquid level was maintained. In addition, it was determined that the original high-performance trays installed in the kerosene side stripper were not suitable to handle the required vapour/liquid traffic. Newly
designed GT-Optim high-performance trays were installed for optimum kerosene stripping performance. To correct the distribution between the rectification and stripping sections of the naphtha splitter, the rectification section pan throat width was reduced to match the inlet panel width of the top stripping tray. However, simple throat width reduction increases downcomer liquid backup height and limits rectification tray capacity. To keep adequate downcomer backup height and correct liquid distribution, a diverging pan was installed at the rectification section seal pan. This diverging pan collects and guides the seal pan liquid to the top stripping tray inlet panel without throat width reduction. To guide the seal pan liquid to the top stripping tray inlet panel, vertically positioned guide baffles were also added. Figure 5 shows the installed diverging pan and vertical guide baffle in between the rectification and stripping sections.

A redesigned diesel pumparound return distributor was implemented in the product fractionator. Flow-controlling orifices were implemented in the design to split liquid flow to match the tray pass liquid ratio. The differential head of the diesel pumparound pump was verified to accommodate the additional pressure drop generated through the new diesel pumparound return distributor. New chimney trays were designed and installed in the kerosene pumparound, the diesel pumparound and the wash sections. To improve vapour distribution, new chimneys were designed with a narrow hypothetical angle. The hat peripheral area was increased to avoid unnecessary pressure drop generation. Each chimney tray open area now matches the neighbouring tray open area ratio.

The overall modification of the hydrocracker targeted debottlenecking of the kerosene yield and naphtha splitter efficiency limitations, taking into account the required flexibilities for future operating modes. As mentioned earlier, UCO, the hydrocracker bottom product, can be sold as a lube base oil plant feedstock. When the hydrocracker operation is switched to maximum fuel production mode, maximum distillate lifting at the hydrocracker product fractionator is required. The process study for the maximum fuel production mode showed that extra distillation equipment modifications were necessary to handle the required column internal traffic in the product fractionator. New GT-Optim high-performance trays were installed in the naphtha/kerosene fractionation section, the kerosene/diesel fractionation section and the two pumparound sections. To improve liquid distribution across the active areas, a specially designed inlet weir was implemented.

The process study also revealed that the kerosene and diesel pumparound chimney tray draw nozzle head losses were too high to maintain self-venting in the maximum fuel production mode case. To ensure problem-free hydraulics between the product fractionator and side strippers, rigorous line hydraulic evaluations were conducted. Actual pipe geometries gathered from pipe isometric drawings and actual control valve data, including flow coefficient and...
pressure drop, were utilised. Figure 6 shows the kerosene/diesel gravity line configuration. The line hydraulic calculations verified that the existing configurations and chimney tray liquid heads were acceptable for maximum fuel production mode cases.

Post-modification performance
The pre- and post-retrofit performances are summarised and compared in Table 1. Since the kerosene side stripper is part of the product main fractionator, two column performances are combined and described in the product fractionator category. For the post-retrofit performance test run, the hydrocracker operating mode with the highest kerosene yield was selected to verify maximum kerosene yield. Table 1 shows that the kerosene yield is substantially increased, and the kerosene 5% distillation temperature and flash point indicate improved kerosene product quality at the same time. In addition, the energy consumption of the kerosene side stripper reboiler is reduced.

Although both of the product fractionator and naphtha splitter reflux ratios are significantly reduced, adjacent product fractionation efficiencies are improved or unchanged in both columns. These performance improvements can be translated as column efficiency improvements. Reduced reflux ratios without sacrificing product qualities eventually contributed to energy savings in this hydrocracker operation.

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