

Customised tower design

To save project costs, the distillation towers for a new refinery were designed to meet process design objectives using best available technology

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The first new refinery in North America in over 30 years is located in Alberta, Canada, and is currently under construction. The feedstock is Athabasca bitumen, API 7-9, with a product slate of transportation-grade diesel, naphtha-based diluent for bitumen transport, and vacuum gas oil for further processing.

The project philosophy for the refinery's design and construction is to leverage technology, service and equipment suppliers early in the design process to streamline the engineering stage. Reducing the iterative process for technology and equipment selection, specification and detailed design can yield significant cost savings. These can be achieved by engaging and embedding vendors prior to and during the detailed engineering process. From past project experience, this can save the equivalent of the cost of the technology and equipment purchased, which can translate to tens of millions of dollars per vendor. Over 20 sole source agreements were initiated for this project, one of

them being for mass transfer internals for the entire refinery.

Koch-Glitsch performed preliminary validation simulations coupled with a design customisation review for all units for the new refinery in support of the project mandate. The purpose of the customisation review is to provide input at the early stages of the project with regard to vessel size and selection of mass transfer internals. Design improvements positively impacting separation performance and capacity can be factored into the tower dimension specifications through the proper selection of mass transfer equipment compared to the specified design by the various process licensors. This can reduce the overall cost of the decision-making process and actual execution of the work. The customisation effort differs from a cost-cutting review in that the distillation column is designed to meet requirements and does not suffer from a linear approach to costs, which may or may not be an optimal performance solution.

The customisation study also

provides the following future value:

- An understanding of the capabilities of current mass transfer technology and the relative capacity gain these devices provide
- A starting point for Phase II and Phase III grassroots tower sizing.

The selected engineering contractors expressed concern with the customisation approach. However, the recommended reductions in tower diameter provided a tangible design margin (typically 10%) in addition to leaving room, in some cases, for a further debottleneck with the use of even higher capacity equipment. The customisation study serves as a guide to understanding where available vessel cost savings can be captured due to excessively conservative sizing practices. The study uncovered that the overall average reduction in diameter for towers that could have smaller diameters across the entire plant is 20%. This reduction has resulted in substantial savings in vessel and tower internals, along with the savings from reduced engi-

neering hours, while still positioning the refinery for future rate increases.

Inside-Out design approach

The typical approach to grassroots mass transfer equipment is:

1. Engineering company or licensor performs simulation
2. Engineering company or licensor performs preliminary design calculations to generate vessel data sheets; vessel diameter is to be confirmed by equipment vendor.

An issue often arises when recommendations are made during step 2 in cases where the recommended diameter is smaller than that provided on the vessel data sheet. A conservative approach is taken to general equipment sizing, and often the engineering company or licensor is satisfied knowing that the specified diameter is sufficient rather than seeking to minimise cost by reducing the design margin or employing a different type of equipment technology. This occurs even though the preliminary design that set the vessel diameter often uses large safety factors, rule-of-thumb hydraulic calculations or design

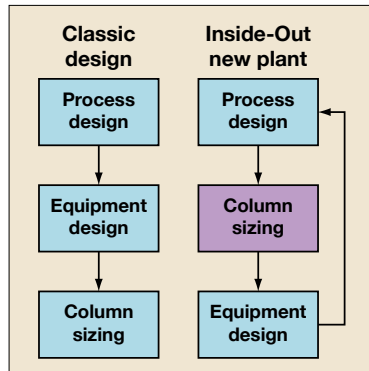


Figure 1 Inside-Out design approach

guidelines that do not incorporate the latest mass transfer technology and knowledge. In fact, many tower internals design manuals in use today were published in the 1980s and earlier, and pre-date the past 20 years of high-capacity tray and packing development.

The goal of the Inside-Out design approach¹ (see Figure 1) is to use the actual hydraulic calculations to optimise/refine/improve the tower simulation. This is done by accurately inputting the pressure drop at each stage (rather than an assumed constant pressure drop). At the same time, tray efficiencies are confirmed, and vessel diameter reductions can be found by matching the

best available equipment options to the application. Koch-Glitsch used the Inside-Out design approach in evaluating all of the internals for the new refinery.

Mass transfer internals

A benefit of the Inside-Out approach is the ability to incorporate equipment characteristics, such as pressure drop and separation efficiency, into the simulation. For the customisation study, high-performance equipment, such as Superfrac trays and Proflux severe service grid, were used.

Superfrac technology

The Superfrac tray comprises a suite of features that together form a high-performance, single-pass, cross-flow tray that has the highest combined capacity and efficiency of all cross-flow trays tested at FRI. The features that can make up a Superfrac tray are:

- High-capacity and high-efficiency valves available in different sizes
- Vapour tunnel or truncated vapour tunnel downcomers with various downcomer outlet shapes to maximise tray capacity and efficiency
- Inlet weir and bubble promoters
- Push valves and other directional devices
- Multi-pass arrangements
- Special features to deal with fouling
- Mechanical installations to simplify installation.

Depending on the process situation for the specific column, one or all of the above features can be incorporated to create a design specific to each vessel evaluated. As an

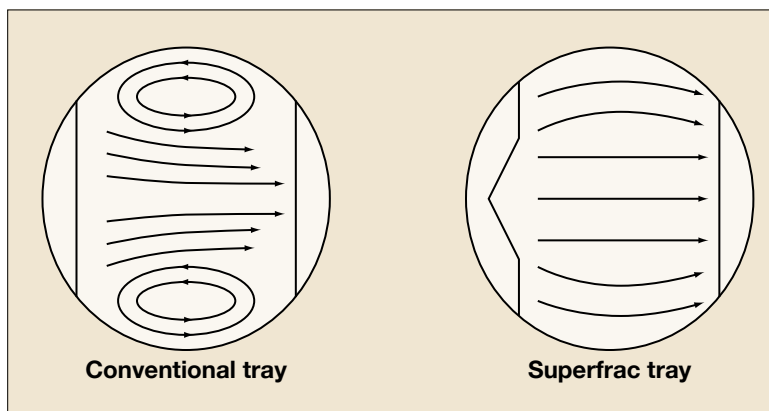


Figure 2 Liquid flow patterns on different tray decks

example (see Figure 2), the Superfrac tray can be fitted with shaped inlet weirs and push valves around the perimeter of the tray to aid in providing a uniform liquid distribution across the active area and preventing the formation of stagnant areas around the tray perimeter. Stagnant zones are common when there is a high L/V ratio, such as occurs in a hydrocarbon steam stripper, so the use of the push valve should result in improved performance. This improved performance is included in the simulation that defines the internal liquid and vapour loadings that are used to size the column.

In general for the customisation study, the trays used have shown to have an efficiency gain of ~10% relative to conventional trays. In operation, this will translate to between 5-15% less reboiler duty (depending on the relative volatility of each system/column). This gain in efficiency, leading to a reduction in heat duty, provides a further lowering of the operating point of the column (due to lower loadings from less reflux and reboil), offering an additional margin in the tower design. Figure 3 demonstrates that the Superfrac tray shows both a higher capacity and a higher efficiency than similar high-performance trays when the trays are being pushed to their design limit.

Packings and grid

For situations where packings are typically considered, Flexipac HC high-performance structured packing and Proflux severe service grid packing

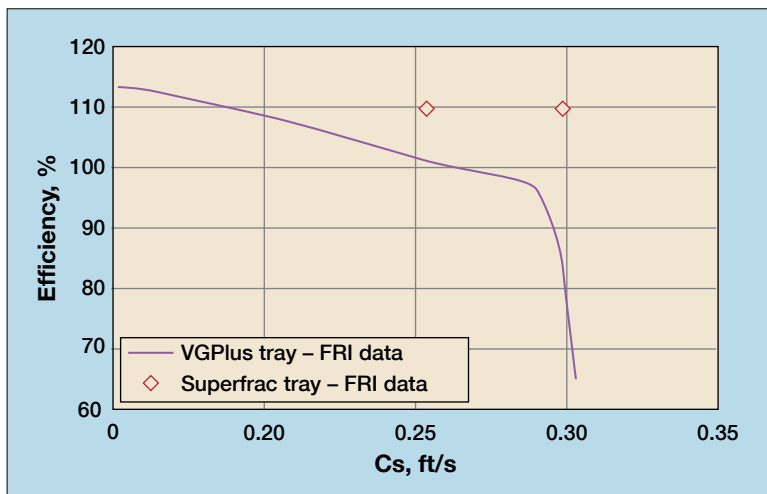


Figure 3 Performance² of Superfrac tray in FRI test (I-C₄/n-C₄, 165 psia, total reflux). VGPlus tray data taken from reference

were used in the customisation analysis. Proflux grid packing (see Figure 4) provides the following advantages over a conventional design of a combination of traditional grid and structured packing bed:

- Higher throughput capacity/less pressure drop per unit height
- Superior de-entrainment

performance at high throughput capacity

- Higher fouling resistance
- More robust structure (higher crush strength).

Proflux grid packing was used to reduce the diameter of the column while still providing some design margin.

Flexipac HC structured packing combines improved



Figure 4 Cross-sectional photo of Proflux severe service grid

Operating point analysis for specified equipment in atmospheric crude column

| Section | Specified equipment | Operating point, % |
|-----------------|-------------------------------------|--------------------|
| Top pumparound | Flexipac 3X structured packing | 63 |
| Upper frac zone | Trays 1-12 | 66 |
| AGO frac | Trays 13-15 | 66 |
| Lower frac zone | Flexipac 250Y HC structured packing | 49 |
| Stripping | Trays 16-27 | 41 |

Table 1

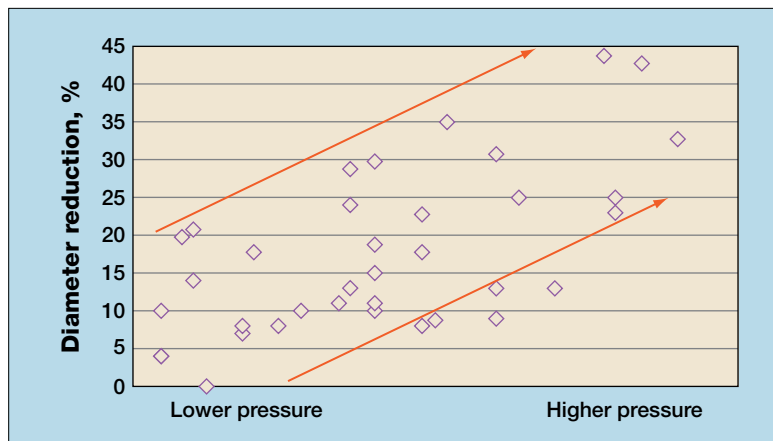


Figure 5 Diameter reduction summary

capacity and efficiency characteristics, resulting in lower pressure drop per theoretical stage than conventional structured packing. Similar in construction to conventional Flexipac structured packing, Flexipac HC has a modification in the geometry of the corrugation at the top and bottom for each packing layer. The change in geometry:

- Eliminates the abrupt change in flow direction of the liquid and vapour phases at the packing layer interface
- Eliminates the premature build-up of liquid
- Helps maintain the low pressure drop characteristics of structured packing throughout the efficient operating range of the packing.

The specific characteristics of the structured packing were

used in the customisation study to impact the column dimensions.

Customisation study results

Figure 5 provides a summary of the range of diameter reductions through the use of the Inside-Out design approach and the high-performance mass transfer internals characteristics used in the customisation study. The overall average tower reduction for the entire plant's 36 columns is approximately 20%, with a range between 0% and 45%. A possible observation to note from Figure 5 is that as the pressure increases, the potential for larger diameter reductions from initial licensor specifications increases. The noted higher diameter reductions at higher pressures are reasonable

as they are based on experience derived from extensive pilot work with tower sizing correlations at higher pressure, which have been enhanced and proven commercially by mass transfer research organisations (such as FRI) and Koch-Glitsch.

Example 1:

Crude unit atmospheric column

For an effective analysis of the entire tower, a simulation that adequately models the stage-to-stage liquid/vapour profile and the pressure drop stage-to-stage was developed. Based on a review of the vapour/liquid profile in the column, the specified column diameter had been set based on limiting the liquid flux in the top pumparound section. For the main section of the tower, a diameter of 4000 mm was specified with a 2900 mm stripping section. The actual operating point (approach to flood) of the specified equipment is relatively low for a crude column (see Table 1).

The diameter was set in a way that allows for potentially up to a 50% increase in throughput, with only minor equipment changes, such as redesigned tray active area panels and new distributors for higher rates. Using high-capacity trays in the grassroots design will reduce the diameter and still leave ample room for a further rate increase. The customised tower diameter was 3500 mm with a 2400 mm stripping section.

The overall operating point of high-performance internals (see Table 2) would not be substantially higher than the grassroots design and would still allow for a further increase

in throughput without requiring a new vessel.

The operating point of the specified packing in the top pumparound section is only 63% capacity. Due to the relatively high liquid flux in this tower for this plant, a grid style of packing is best suited to this section. Using a combination of Flexigrid type 2 and type 3 severe service grid packing, the diameter of the upper section of the column could be reduced from 4000 mm to 3500 mm (14% reduction), with the packing operating point increasing only to 69% capacity. It is not unusual for grid packings to operate at up to 90% capacity in heat transfer sections post-revamp; therefore, there is still room for an additional throughput increase of up to 30% ($90/69 = 1.30$). As a trade-off, the height of the packed bed would need to increase by approximately 1000 mm to account for the differential in heat transfer coefficient. The cost/benefit analysis that was performed suggested the tower diameter reduction and change in internals type substantially offset the slight increase in height.

**Example 2:
Crude unit vacuum column**

The 7600 mm specified vacuum column flash/wash bed/HVGO pumparound zones have been sized to have a $C_v = 0.30$ ft/s in the flash zone and 0.35 ft/s into the HVGO pumparound. Generally, Koch-Glitsch is aligned with this design philosophy for grassroots oil sands vacuum fractionators, where substantial increases in throughput are expected. A key objective of the plant design

| Operating point analysis for recommended equipment in atmospheric crude column | | |
|--|---------------------------------------|--|
| Section | Specified equipment | Operating point, % |
| Top pumparound | Proflux severe service grid | 69 (requires bed height to increase by 1m) |
| | Flexigrid severe service grid packing | |
| Upper frac zone | Superfrac trays 1-12 | 71 |
| AGO frac | Superfrac trays 13-15 | 70 |
| Lower frac zone | Flexipac 250Y HC structured packing | 64 |
| Stripping | Superfrac trays 16-27 | 64 |

Table 2

was not to build in excess design margin beyond 10-15%, so the column diameter in the flash zone can be reduced from 7600 mm to 6900 mm, a 9% reduction that results in a C_v of 0.4 ft/s. The HVGO pumparound diameter can also be reduced to 6900 mm by using Proflux grid and Flexigrid grid packing at the bottom of the bed where the vapour load is highest. The operating point of the packing will only be at 76% of capacity at the reduced diameter. Heat transfer sections typically operate around 90% capacity at revamp conditions.

Example 3: Hydrocracker product fractionator

Since hydrocracker product fractionators typically use trays that have extremely long flow path lengths, the same degree of efficiency increase as compared to a medium-pressure light hydrocarbon distillation (which relies on

short flow path lengths due to high liquid load) is not expected for high-performance trays. Yet, for this analysis, the Koch-Glitsch simulation still incorporates additional separation efficiency (see Table 3). The suggested increase to the theoretical stage count is based on a conservative expectation of efficiency increase with high-performance trays.

In addition to the use of the Superfrac tray, it was recommended to convert the pumparound sections from the conventional trays specified to a combination of Flexigrid grid packing, Proflux grid and Flexipac structured packing. The use of packing will minimise vessel diameter and also give significant pressure drop savings. In addition to the packing, a trough distributor or spray header is required to meter the pumparound return liquid to the packing. The combined height of the kero and diesel pumparound sections with packing match the total height allocated in the original vessel specification; however, the allocation of height will be reversed (more height is needed in the diesel pumparound due to a lower driving force).

Table 4 summarises the customised vessel configuration

| Simulation increase stage count for product fractionator | |
|--|---|
| Section | Additional stages to add to simulation with Superfrac trays |
| Top frac/swing Kero | 1 |
| Kero diesel frac | 1 |
| Diesel frac | 1 |
| Stripping | 0 |

Table 3

Specified internals for product fractionators and jet flood performance

| Section | Tray spacing, inches | Flood, % | Equipment |
|------------------|----------------------|----------|---|
| Top frac. | 24 | 74 | 2-pass Superfrac tray |
| Swing Kero frac | 24 | 65 | 2-pass Superfrac tray |
| Kero P/A | 58.5 (bed) | 73 | Proflux severe service grid, Flexigrig grid packing & Flexipac structured packing |
| Kero diesel frac | 24 | 77 | 2-pass Superfrac tray |
| Diesel P/A | 193 (bed) | 74 | Proflux severe service grid, Flexigrig grid packing & Flexipac structured packing |
| UCO frac (wash) | 24 | 79 | 2-pass Superfrac tray |
| UCO stripping | 24 | 77 | 2-pass Superfrac tray |

Table 4

versus a column with strictly conventional trays specified. The new diameter from feed to overhead is set at 4500 mm from 5000 mm, and 2700 mm is used for the stripping section from 3000 mm (an 11% tower reduction along with pressure drop savings).

Example 4: Swing cut stripper

The vessel diameter can be reduced from 1000 mm to 900 mm by using the Superfrac

tray. The original simulation was based on an 11 kPa pressure drop across the trays. The actual pressure drop will be closer to 6.5 kPa, which, when re-input into the simulation, will modify the tray loadings on the column, allowing for a diameter sizing adjustment.

For all the towers in the customisation analysis, all the nozzle sizes and feed distribution devices were reviewed based on the maximum flows in

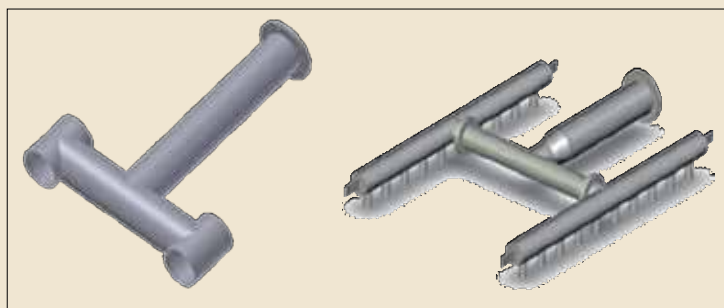
the simulation that can be experienced during the operation of the plant between shutdowns. Table 5 provides an example of the nozzle analysis performed. In quite a few cases, revised feed distribution devices were recommended due to the specific flow regimes calculated.

Example 5: Lean oil absorber

The specified 1829 mm vessel diameter has a maximum operating point of 73% jet flood at 110% of the design rates. The application of the Superfrac tray with Minivalve VG-0 valve will result in a significant reduction in diameter to 1372 mm, a 33% reduction. The operating point of the customised design is 80% jet flood at 110% of the design rates; therefore, additional margin is available while still meeting the 30 kPa pressure drop specification.

Example of a nozzle and feed device comparison

| Nozzle/name | Licensor nozzle diam., inches | Phase | Evaluation | Recommendations |
|----------------|-------------------------------|---------|--|--|
| N1/strip steam | 4 | Vapour | Velocity ~76 ft/s | No changes |
| N2/hot feed | 8 | 2-phase | $ru^2 \sim 800 \text{ lb/ft}^2$ Slug flow regime per Baker chart ⁴ | Increase diameter to 10 to move away from slug flow regime |
| | | | Where $r = \text{density}$ $u = \text{velocity}$ | |
| N3/cold feed | 16 | 2-phase | $ru^2 \sim 3039 \text{ lb/ft}^2$ | Licensor device is not considered optimal as this is a two-phase feed. Recommend using an H-pipe with orifices to obtain necessary distribution. Licensor device is not considered optimal as this is a two-phase feed. Recommend using an H-pipe with orifices for improved separation and distribution. |



Specified design for two-phase feed pipe

Koch-Glitsch recommended design

Table 5

Common causes of tower diameter over-design

| | |
|---|---|
| Over-conservative application of derating factors with modern flooding correlations | Some texts still apply conservative over-design practices for both vacuum and high-pressure hydrocarbon distillation. |
| Using a constant tray spacing across all trays | Often specifications come with a 24-inch tray spacing to simplify the vessel fabricators' work. High relative volatility distillation columns often have a wide variation in vapour flow across the column. Using multiple tray spacings to match the change in loads will lead to a smaller diameter. |
| Reluctance to use three-pass trays | Avoiding three-pass trays to use two-pass trays with large downcomer mouths may result in oversized tower diameters in the 8–11 ft range. |
| Over-specification of turndown requirement | For example: requesting 25% turndown when 50% is sufficient, moving valves would be needed rather than higher performance fixed valves. Trays may also require a higher pressure drop to limit weeping at turndown condition, leading to higher downcomer backup at maximum capacity, possibly increasing the tower diameter. |
| Re-applying safety margins at each stage of design | Conservatism in process simulation loads combined with unnecessary over-design factors or limitations on tray hydraulics increases column sizes with 30–40% safety margin. |
| Specification of unnecessary pressure drop limitations | Legacy pressure drop specifications do not factor in the geometries of modern-day trays. |

Table 6

Common causes of tray tower diameter over-design

By applying the Inside-Out design approach early in the process, the common causes of tray tower diameter over-design can be mitigated (see Table 6). With frequent advances and improvements in design criteria for tower internals, it can be difficult to keep up with the continuous refinements and innovations in the understanding and design of tower internals used to optimise column performance. Engineers at Koch-Glitsch have identified some of the common, recurring causes of tower

diameter over-design that can be rectified for a cost-effective distillation tower design (see Figure 6).

Conclusion

Significant project savings are realised through reductions in vessel diameter, use of state-of-the-art technology, and engaging knowledge and experience in the design of mass transfer internals prior to and during the detailed engineering process. For a new refinery with 36 columns using mass transfer internals, an average 20% reduction in tower diameter can be readily achieved. A

further article will provide an evaluation of the actual performance of the vessels compared to the design.

FLEXIGRID, FLEXIPAC, MAX-FRAC, MINIVALVE, PROVALVE, SUPERFRAC, HC and PROFLUX are marks of Koch-Glitsch, LP.

References

- 1 Remesat D, Inside-Out design approach, *Hydrocarbon Processing*, 2006.
- 2 Nieuwoudt I, et al, Best of both... Nieuwoudt I and Penciak J, Koch-Glitsch, USA, Explore how high performance trays can deliver the best capacity and efficiency, *Hydrocarbon Engineering*, June 2007.

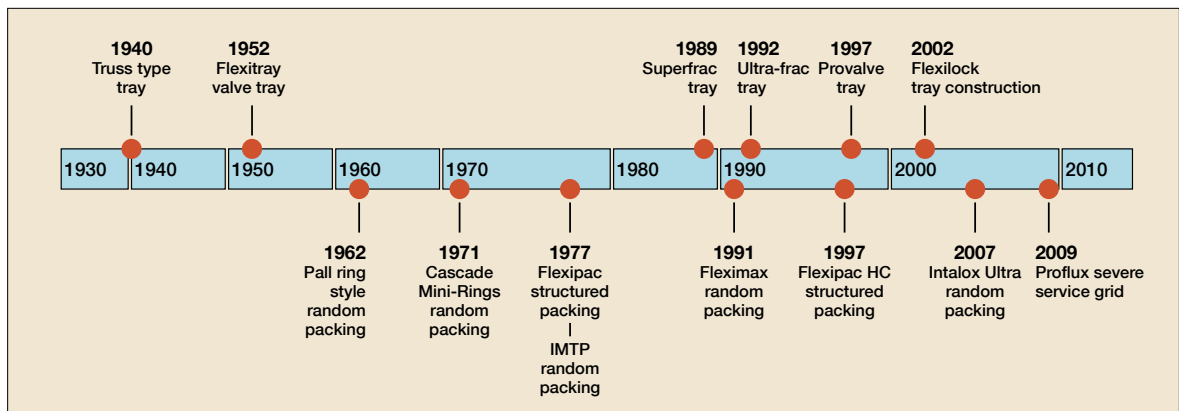


Figure 6 Mass transfer equipment development timeline

3 Customisation Report for Client, Confidential, 2012.

4 Baker O, Design of pipelines for the simultaneous flow of oil and gas, Fall Meeting of the Petroleum Branch of AIME, 19-21 October 1953, Dallas, Texas.

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