# Vacuum unit design effect on operating variables

**To revamp vacuum units, process modelling and equipment design know-how are needed, and the understanding of connected equipment performance can lead to higher gasoil quality and yields, with fewer unscheduled shutdowns**

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Process modelling errors and failure to design vacuum unit equipment as an integrated system has caused yield loss, poor gasoil quality, and unscheduled shutdowns. The vacuum unit charge pump, fired heater, transfer line, column internals, and ejector system must be evaluated and designed together so that operating temperature and pressure can be optimised to meet economic goals. Revamps need to push major equipment to its intrinsic limits to minimise investment.

Real equipment performance should be the basis of a revamp, not office-based assumptions or cursory reviews of the original equipment manufacturer's data sheets. Even though vacuum unit equipment is often highly constrained by existing equipment, an experienced revamp engineer can often manipulate heater outlet temperature, column flash zone pressure, coil steam injection, or vacuum bottoms stripping to achieve revamp yield and reliability targets.

The challenge is to accomplish these objectives while meeting stringent investment criteria or Capex restrictions. Vacuum unit heater-inlet-through-ejector-outlet (Figure 1) must be considered a single system when a practical, operable and cost-effective revamp is to be implemented. A successful revamp will meet all these standards.

Revamps should always start with a thorough test run gathering all necessary data on current unit and equipment performance. Next, the field data is used to calibrate a baseline process model and quantify current equipment performance. The calibrated model will be the basis of the revamp calculations. Only then can the revamp engineer apply equipment design know-how to identify all under-utilised equipment and exploit it to minimise investment.

Without this approach during the conceptual process design (CPD), all major cost bottlenecks may not be fully defined. This leads to scope growth and cost escalation during later stages of engineering.

Computer models are invaluable tools used during CPD to establish bottlenecks and define scope. But the process flow sheet models must represent true operation and account for the many equipment non-ideal situations encountered in a refinery vacuum unit. If not, inaccurate calculated process stream data will result in short run-lengths and low gasoil product yields.

One such example is the influence of transfer line phase separation on predicted heater outlet temperature and the calculated wash section liquid flow rates. Unlike some theoretical modelling concerns that have few real penalties, failure to account for transfer line phase separation will lower the gasoil product yield and result in a very low wash oil flow rate.

Wash bed packing requires adequate wetting in the middle of the bed to avoid stagnant zones where coking is initiated. In reality, accounting for transfer line phase separation raises the wash oil flow rate by 200–300 per cent over conventional modelling practices that assume liquid and vapour leaving the transfer line are in equilibrium. This one oversight alone has caused many unscheduled shutdowns after less than a year of operation. Process modelling plays a significant role in meeting overall refinery economics.

Rigorous equipment models need to account for the various designs that are encountered when revamping. Many vacuum unit fired heaters, transfer lines, columns and ejectors were designed and
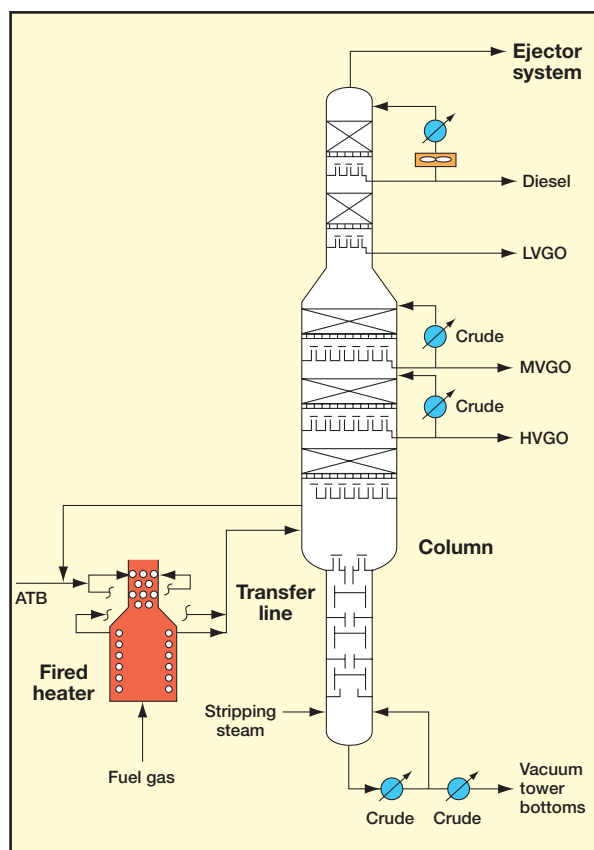


**Figure 1** *Single system, ejector through heater inlet*

built based on a low capital investment strategy and not best practices methodology. Poorly designed equipment may operate well at low temperature and high pressure, but when severity is increased, the run-length or yield expectations are not met.

Many poorly designed vacuum heaters have been built because the selection criteria were based mainly on low price. Low-cost heaters have high heat flux imbalances among the passes that significantly reduce the maximum reliable heater firing.

Often, engineering approaches are based solely on superficial evaluations such as average radiant heat flux or vendor data sheet review. These methods minimise engineering cost but often

make incorrect conclusions about future equipment performance. When equipment is pushed, fundamental design errors become apparent. Thorough heater analysis will identify heat flux imbalances allowing the experienced engineer to exploit under-utilised capacity. Potential minimum capital design changes, such as external radiant section jumper-overs can be used to correct the heat flux imbalances.

During CPD all under-performing equipment needs to be identified so that investment can be focused where it is really needed. Manipulating operating temperature and pressure to produce a reliable, operable, and cost effective revamp requires full utilisation of all existing equipment.

## Critical operating variables

Operating temperature and pressure sets gasoil yield and unit run-length. The temperature required to meet the gasoil product yield target is determined by the heater outlet oil partial pressure, column flash zone pressure, and stripping section efficiency. Heater outlet temperature is limited by the rate of coke formation inside the heater tubes or column internals.

Some refiners target a one-year run between decokings; therefore, high coke formation rates are acceptable. Yet, today, many refiners have set four to five-year run-length and higher HVGO product TBP cutpoints as goals. Increasing heater outlet temperature and run-length requires all equipment design optimised together to keep the rate of coke formation low. Only then can operating pressure and temperature be optimised to maximise performance within the investment guidelines and the real equipment performance.

Process and equipment design influences operating temperature and pressure. For instance, when processing heavy Venezuelan crude oils, a best design practice would be to install a stripping section or improve existing stripping section efficiency to reduce the heater outlet temperature. Equipment design oversights, such as poor stripping section tray design, can increase heater outlet temperature by 10–20°F.

After a recent revamp, poor stripping section tray designs increased the heater outlet temperature from 785°F to 800°F to meet the revamp product yield target. Heater run-length was one year versus an anticipated three to four years. Details, such as optimum tray design, are important when revamping. An under-performing ejector system can easily cause 10–15°F higher heater outlet temperature. Operating pressure and temperature are heavily dependent on how well the equipment works.
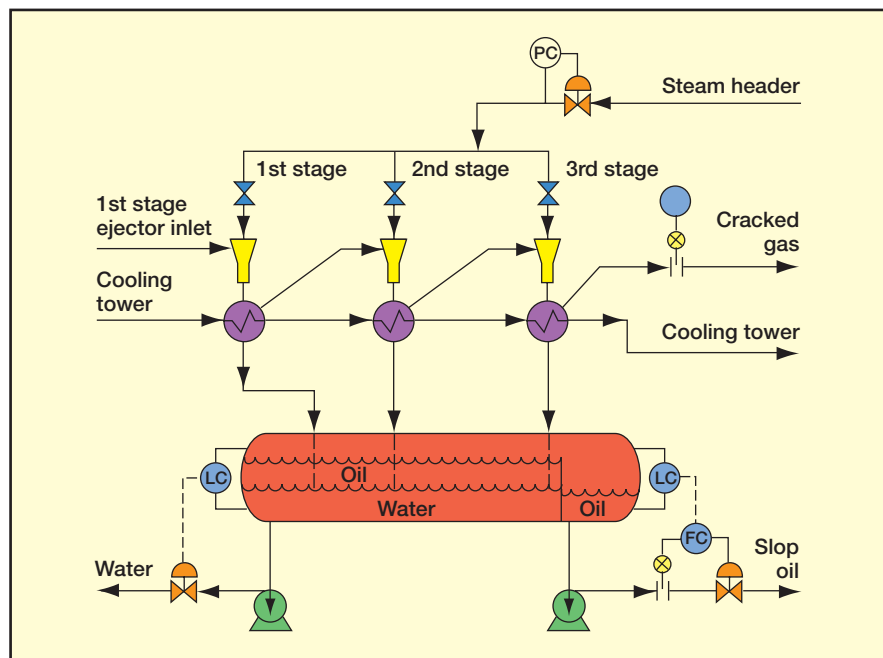


**Figure 2** *Cracked gas influence on ejector capacity*

Baseline and revamp process model's results need to reflect the existing equipment design including all shortcomings. Sometimes it is necessary to tolerate the consequences of the existing equipment, such as high transfer line pressure drop, because the cost to fix it is very high. However, understanding the influence of all major equipment on temperature and pressure allows the experienced designer to consider practical alternatives such as improved stripping or adding a stripping section to the column.

Thorough engineering calculations need to be done early in CPD. Original equipment manufacturer's data sheets should not be used as the sole basis for process modelling and ultimate equipment performance. Rarely will original equipment data sheets reflect actual field measured performance. In practice, process models should be constrained by the existing or revamped equipment design. The process model may show that increasing heater outlet temperature would raise gasoil yield; however, the model needs to represent actual equipment design. For instance, an ejector system (Figure 2) may not be capable of handling higher cracked gas production resulting from a 10–15°F increased heater outlet temperature. Thus, increasing heater outlet temperature would raise column flash zone pressure and in some instances gasoil yield actually goes down.

Many times, grassroots process design methods have been used for revamps. Several equipment design specialists working independently have evaluated the major equipment. Heater specialists performing rigorous modelling of only the heater have determined that injecting coil steam would allow higher heater outlet temperature. Often these specialists have assumed the heater outlet pressure would be the same as the original data sheets indicated because no transfer line calculations have been done. Generally these equipment specialists have little understanding of connected equipment performance.

Vacuum unit heater and transfer lines must be evaluated as a single system. The model used must be capable of rigorous heater tube-by-tube calculation and simultaneous hydraulic calculations to account for the influence of revamp conditions on temperature and pressure. While coil steam injection would reduce the rate of coke formation in the heater and permit higher operating temperature, it also would raise transfer line pressure drop.

Quantifying coil steam impact on gasoil yield would require the whole system to be modelled as a single entity so that all temperature and pressure effects are determined.

## Integrated equipment systems

Once the process flow scheme is selected, major equipment design and capital investment constraints will determine what temperature and pressure changes are feasible. Atmospheric tower bottoms (ATB) is pumped from the crude column to the vacuum column flash zone. These pumps must supply enough differential pressure to overcome the system pressure drop and allow enough control valve pressure drop to control the individual pass flow rates.

Heater pressure drop is a function of the tube size, tube length, coil steam
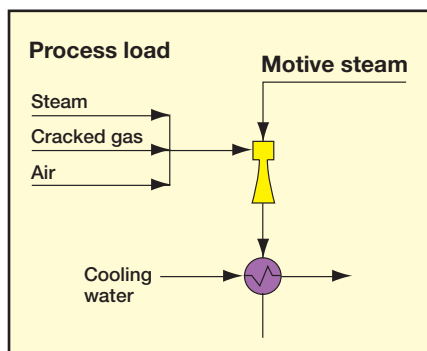
**Figure 3** *Ejector process gas load*

rate, heater tubes' coke layer thickness, and transfer line pressure drop. A low-pressure drop transfer line will be 2.5–3.0psi. Others will have a pressure drop of 14psi or more with a few vacuum units' heater outlet pressures operating above atmospheric pressure.

Transfer line pressure drop will govern the HVGO product yield for a fixed heater outlet temperature, flash zone pressure, and stripping section efficiency. Column flash zone pressure is fixed by ejector system performance and pressure drop through the column internals. Heater outlet temperature, and coil and stripping steam rates strongly influence flash zone operating pressure through their affect on ejector system process gas load.

Unit pressure profile from the ejector system through heater inlet needs to be calculated accurately to properly design the heater tube sizes. Pressure profile establishes the heater outlet temperature needed to meet the targeted gasoil yield. Consequently, pressure profile has an effect on the rate of coke lay-down in the heater. Pressure in the last two to three tube rows of the radiant section decreases rapidly and oil vaporisation begins.

Accurate feed heating-cooling curves are essential to properly size the outlet tubes and size transition locations. Incorrect sizing of the last two to three tube rows can result in low oil vaporisation in the larger tubes, high oil residence time, and rapid coking. Most heater vendors will only transition the tube sizes at the bends to minimise cost. Transition tubes often have both high oil film temperature due to lower oil mass flux rate and high oil residence time caused by the increase in tube size.

The key to vacuum heater design is to ensure that the peak oil film temperature and oil residence time do not increase the rate of cracking above the ejector capacity or result in rapid coke lay-down inside the tubes. Rigorous heater modelling will include tube-by-tube analysis of oil residence time and peak oil film temperature. Both of these determine the rate of cracking inside the

heater tubes, which sets the amount of cracked gas flowing to the ejectors.

Heater cracked-gas production is highly dependent on oil stability and heater design. Cracked gas rates exceeding 0.2–0.3 wt% of feed cause rapid coke lay down in the heater tubes.

Increasing heater outlet temperature makes more cracked gas. Ejector system suction pressure increases as the process gas load (Figure 3) to the ejector increases. Conversely, lower ejector gas load reduces operating pressure. Lower column flash zone pressure raises the gasoil yield for a given heater outlet temperature and pressure. Hence, heater and ejector system performance are integrally linked. It is not possible to change one without affecting the other.

Ejectors capacity is determined by its performance curves. Certified ejector curves show process gas load in steam or dry air equivalents on the X-axis versus suction pressure in mmHg absolute on the Y-axis. Process gas load is set by coil and stripping steam rates and cracked gas production in the heater. Increasing coil steam allows higher outlet temperature for a given amount of cracked gas, but the steam also raises the first and second stage ejector process gas load. This will increase flash zone pressure without properly sized ejectors. Coil steam also raises pressure drop across the heater. Therefore, the charge pumps must also be capable of pumping the ATB through the heater.

Manipulating pressure and temperature to the limits of the major equipment is fundamental to any vacuum unit revamp. Rigorous modelling of the charge system, fired heater, transfer line, column internals, and ejector system together as a single system is necessary to understand product yield and run-length consequences of all potential equipment modifications. Finding cost effective modifications that allow the temperature and pressure profile to be manipulated within investment constraints is the revamp engineer's goals.

## Equipment design
Major equipment needs to work together and the individual equipment capacities should be balanced while optimising pressure or temperature. An ejector system that can reduce the column operating pressure well below the maximum capacity factor will cause black HVGO product. Therefore, the ejector capacity will not be useful. Major equipment design considerations and their influence on temperature and pressure are covered in the following sections.

### ATB charge pumps
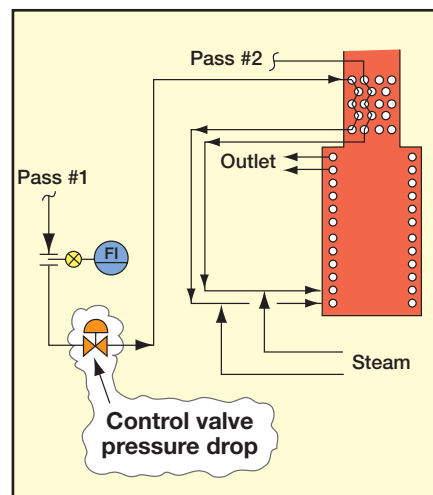Charge pump hydraulics plays a significant role in the heater design. The ATB pump supplies the differential pressure



**Figure 4** *Feed pump hydraulics: coil steam injection*

necessary to pump the oil through the heater and transfer line to the column flash zone. Heater pressure drop should be a consequence of the convection and radiant sections design rather than a significant specification. Yet many designers incorrectly place emphasis on limiting the heater pressure drop in the bid specifications (Figure 4). Low-pressure drop heaters have chronic coking problems because oil residence time and peak oil film temperatures are high. Replacing the charge pumps is a small capital cost relative to the losses of an unscheduled shutdown.

Pressure drop through a well-designed low residence time heater will be approximately 120psi when clean. When fouled, this same heater will have 200psi pressure drop, which would require an ATB pump discharge pressure of approximately 275–300psig. Pressure drop is a product of oil mass flux rate, coil steam rate, and the steam injection location.

Low-mass flux heaters that operate with no coil steam have only 30psi pressure drop when clean. While low-pressure saves pump size and power consumption and possibly pump replacement, these heaters also have radiant section oil residence time over 120 seconds and high peak oil film temperature.

Heaters designed with coil steam have higher pressure drop. Steam can be introduced into the radiant section, convection-radiant crossover tubes, front of the heater, or multiple locations. The injection location and amount of steam determine pressure drop. Ideally, the injection location is chosen based on the peak oil film temperature and residence time in the individual heater tubes. Heaters designed at high oil mass flux (>400lb/sec-ft[2]) and with coil steam have lower oil residence time and peak oil film temperatures.
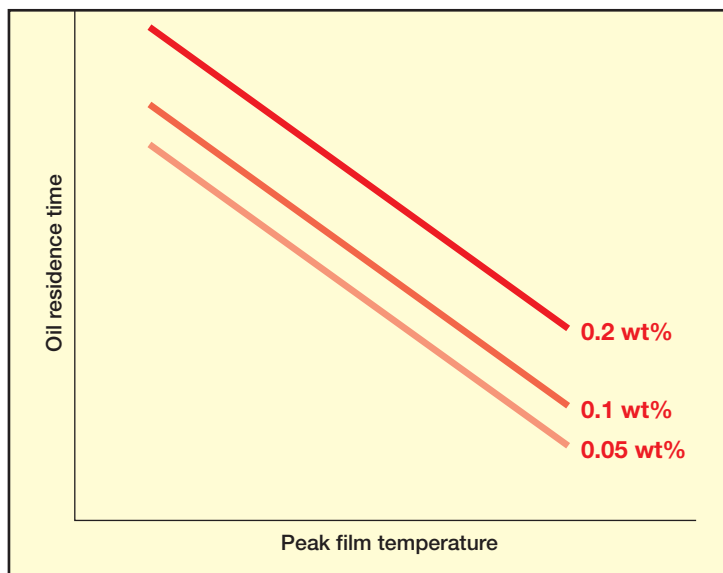
**Figure 5** *Cracked gas production: oil film temperature vs residence time*



**Figure 6** *Minimised pressure drop: multiple pressure locations*

These heaters can operate at average heat flux rates above $10\,000\,Btu/hr\text{-}ft^2\text{-}°F$ without excessive coking. Therefore, the ATB pump is a critical component of meeting both the HVGO product cutpoint and run-length.

**Fired heater**
The fired heater supplies the energy needed to raise the oil temperature high enough to meet the gasoil yield target and supply sufficient wash oil flow to keep the wash section packing from coking. Heater outlet temperature and pressure are important variables. Heater outlet pressure depends on the transfer line design. The maximum allowable oil temperature is based on the rate of coke laydown inside the tubes and the targeted heater run-length.

Coke insulates the oil inside the tube from the hot flue gas. Therefore, the outside tube metal temperatures (TMT) must increase to transfer heat through the insulating coke layer. Tube metallurgy, tube thickness, and tube life will determine the maximum TMT before the heater must be decoked.

Targeted heater run-length will determine the acceptable rate of coke formation. Coking rate can only be inferred by measuring the TMT with either skin thermocouples or infrared thermography. When coke is produced, so is cracked gas, and it can be directly measured from the ejector hotwell. As coke formation increases, so does the amount of cracked gas. Gas leaving the hot well will be primarily cracked gas unless the unit has excessive air leakage.

Low cracked gas rate will be 0.05 wt% of feed. High cracked gas rate will be 0.2–0.3 wt%. Once the gas rate exceeds these values, short heater run-lengths should be expected.
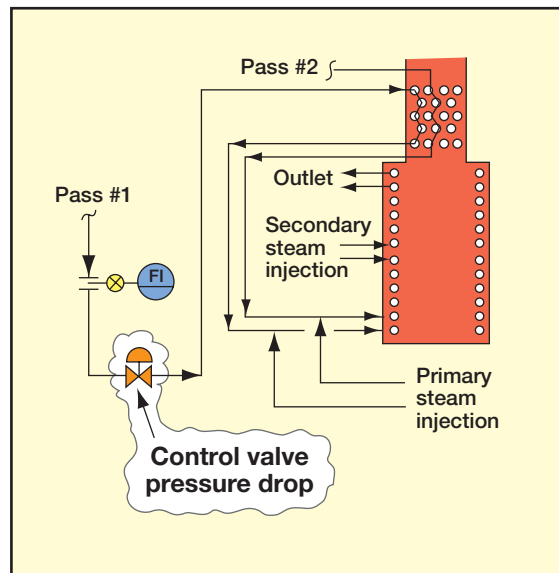
The key to vacuum heater design is being able to determine the time and temperature relation in each tube and the quantity of gas the ejector must be capable of handling. Rigorous tube-by-tube rating allows the designer to evaluate oil residence time and peak oil film temperature in each tube (Figure 5). Coke forms because the oil film inside the heater tube reaches temperature where the oil is no longer thermally stable. High temperature will not cause excessive coking at low residence time.

Coil steam lowers oil residence time. The revamp engineer must decide where to inject the coil steam and how much steam is needed to suppress coking while balancing feed pump hydraulic limits. Picking the wrong location will cause coking in the tubes upstream of the coil steam injection point or high-pressure drop which limits charge rate (Figure 6). The influence of heat flux imbalances and other nonideal conditions should not be overlooked because many low-cost heaters have very poor radiant section heat distribution. Heaters designed with the individual tube passes stacked have poor heat distribution.

Thorough evaluation of the heater and the transfer line is necessary to establish gasoil product yield and heater run-length. A rigorous heater and transfer line model will calculate pressure drop and limit operating pressure to the two-phase critical velocity limit. These models must make the calculations backwards from the flash zone through the heater.

After each run, peak oil film temperatures and oil residence times in each tube are evaluated to determine if rapid coke lay down will occur. If high coking rates are predicted, then heater outlet temperature or oil residence must be reduced to prevent short heater run-lengths.
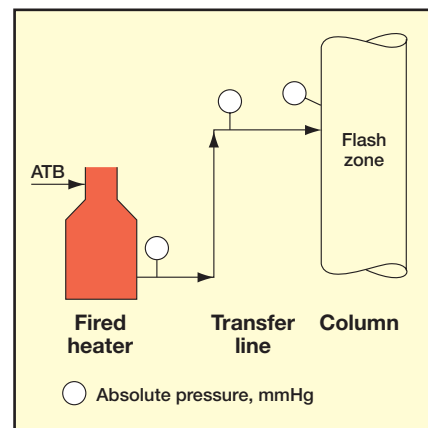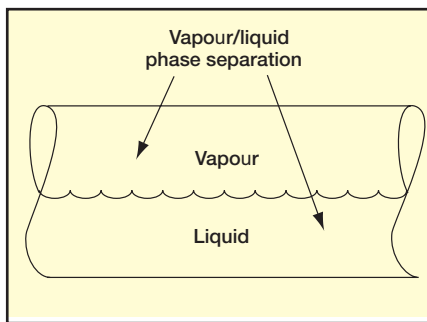
**Transfer Line**
Heater outlet (Figure 7) pressure is set by the transfer line design. Transfer line pressure drop and two-phase critical velocity both influence this pressure. Pressure drop is based on transfer line configuration. Two-phase critical velocity is much lower than the sonic velocity of the gas phase alone. Hence, many designers will calculate a low percentage of sonic velocity and predict a much lower heater outlet pressure than would be feasible. Calculation methods for critical velocity are complex.

Accurate field measurements of many transfer line pressures and temperatures and subsequent calculations using a model with critical velocity calculations have confirmed that critical velocity is real. Therefore, heater outlet pressure is a function of both hydraulic losses and two-phase critical velocity.

Transfer lines usually have long horizontal runs with line diameters of



**Figure 7** *Heater outlet pressure influence on product yields*

**Figure 8** *Process modelling: transfer line phase separation*

36–84in prior to entering the column flash zone. These large diameter horizontal runs cause the liquid and vapour phases to separate. Calculated phase regimes are either stratified or stratified wavy (Figure 8). Stratified phases cause the liquid and vapour to have poor mass and energy exchange across the interface. Hence, liquid and vapour contacting is poor. Thus, phase separation causes vapour to flow through the top of the horizontal portion of the transfer line.

This vapour becomes superheated due to pressure reduction as the vapour approaches the column flash zone. Vapour and liquid entering the flash zone of a vacuum column are not in phase equilibrium. Vacuum column wash section coking causes unscheduled shutdowns. There are several potential causes. However, failure to account for phase separation in the transfer line is one of the most common. Coked wash section packing has occurred in nearly 100 per cent of columns operating above 750°F where the wash rate was calculated assuming equilibrium.

Wash oil flow rate to the packing must be sufficient to keep the packing wetted, otherwise high oil residence time stagnant zones are created in the middle of the packed bed. Transfer line phase separation increases the required amount of wash oil flow rate. Superheated vapour entering the wash section vapourises a portion of the liquid flowing down through the packing.

Once the vapour superheat is removed, oil condensation occurs from the vapour. The bottom of the wash section is kept wetted by flash zone entrainment. The top of the packing is wetted by the wash oil flow rate. The wash oil flow rate must be high enough to ensure that the middle of the packing remains wetted. Otherwise coke forms in the middle of the packed bed. Bed depth has a significant influence on coking because as bed depth increases there is more middle section depth that must be wetted.

**Vacuum column**
Vacuum column diameter often drives the system pressure and temperature

optimisation. Exceeding vacuum column capacity causes high vacuum tower bottoms (VTB) entrainment into the HVGO product. Minimising column flash zone pressure reduces the heater outlet temperature. Higher column operating pressure allows more gasoil production without exceeding the column capacity factor, but it requires higher heater outlet temperature.

Lower column operating pressure reduces heater outlet temperature. However, lower column operating pressure requires a larger column cross-sectional area to produce a given amount of gasoil.

Vacuum column diameter is often one of the major limits that the revamp engineer must circumvent. Ejector system inlet pressure and heater outlet temperature must be optimised. Once the overhead pressure is selected, the pressure drop through the column determines the flash zone pressure. The flash zone pressure (oil partial pressure) and the heater outlet temperature and pressure determine the amount of oil vaporised.

Proper design of vacuum column wash sections, flash zone, and stripping sections are necessary to meet yield, quality, and reliability targets. Stripping section performance influence heater outlet temperature. Higher stripping steam rate and improved efficiency reduce the heater outlet temperature. Flash zone internals should separate much of the VTB from flash zone vapour and help distribute vapour to the wash section. Wash section internals include liquid and vapour distributors above and below the bed, respectively, and the packing. Packing type and bed depth determine efficiency.

Figure 9 shows the various parts of a vacuum column.
**Flash zone and wash section**
As flash zone operating temperature and capacity factor increase, design of these internals becomes increasingly important. Design mistakes have reduced HVGO product yield and caused coking and unscheduled shutdown of numerous vacuum units.
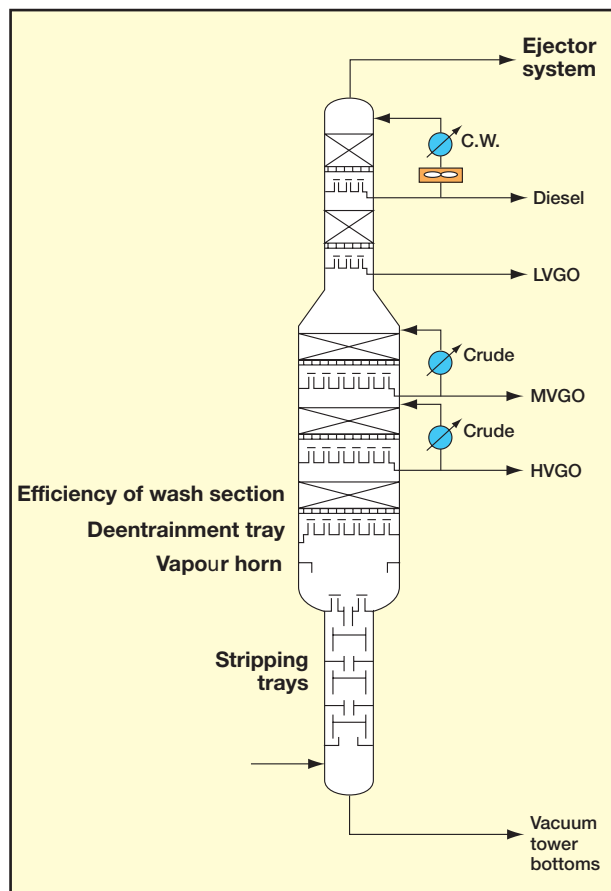


**Figure 9** *Vacuum column design*

Vapour and liquid feed enters the column at velocities as high as 380–400ft/sec. The vapour phase contains small droplets of VTB that have been generated in the transfer line. Droplet size is too small to allow settling in the transfer line because the velocity is too high. Ideally, vapour leaving the horn should enter the wash section uniformly distributed across the column cross-section and it should be free of entrainment because entrainment contains high metals, concarbon and asphaltenes.

The wash zone removes entrained residue from the flash zone vapour and provides some fractionation of HVGO product. The fractionation requirements depend on crude oil type, gasoil yield target, and product quality. Typically, a wash zone consists of a packed bed, spray header or gravity flow trough type liquid distributors, and a collector tray below the packing. Coking in the wash zone occurs either in the packing itself or on the collector tray under the packed bed.

Revamps may produce high gasoil yields for a 6–18 month period; however, wash oil flow rate is not sufficient to operate for four to five years. The collector tray below the wash bed can also coke due to high liquid residence time. High liquid residence time results from either the tray

design or operating with high liquid level on the tray. Collector tray oil residence time must be kept to a minimum.

**Stripping section**

Vacuum column stripping reduces heater outlet temperature by vaporising a portion of the liquid entering the flash zone. Stripping sections help yield incremental gasoil by reducing the partial pressure of the oil. Properly designed trays will have high efficiency and minimise coking potential.

High liquid residence time on the stripping trays can cause coke formation. Some designers provide cold VTB quench to the top tray to reduce or eliminate coking. However, the quench reduces or eliminates the benefit of the stripping trays. Stripping sections must be sized properly; otherwise oil residence time will be high. Additionally, oversized trays have stagnant liquid zones and very low hole area. Properly designed stripping trays will not coke. Equipment design details are important.

**Ejector system**

An ejector is similar to a compressor; the motive steam supplies the energy to compress the process gas. Design motive steam rate increases as suction gas load goes up. The first stage ejectors must handle process steam, cracked gas, and air leakage. Air leakage is typically very low. The second stage ejector's process load is primarily cracked gas and some water vapour. The third stage suction load is mainly cracked gas. Ejector inlet pressure is dependent only on suction gas load unless the required ejector discharge pressure exceeds the maximum discharge pressure (MDP).

Revamps must optimise system pressure to balance the ejector system size and capacity against the existing system limits. Many revamps require coil steam to control cracking in the heater. Coil steam increases load on the first and second stage ejectors. As coke formation increases so does the amount of cracked gas. High heat flux and high outlet temperature vacuum heaters produce approximately 0.2–0.3 wt% cracked gas on feed. The revamped ejector system must be capable of handling the process gas load so that column flash zone pressure can be minimised based on maximum column capacity factor.

Ejector gas load, suction pressure, discharge pressure, and steam pressure all set the amount of motive steam required to meet the targeted process gas load. Existing ejector system inter-condenser capacity limits must be carefully considered during a revamp. Significantly higher inter-condenser duty can increase cooling water requirements beyond the existing cooling water system.

Higher process gas load increases the motive steam requirements and raises inter-condenser duty. Cooling tower modifications or a new cooling tower cost can be very high. Thorough ejector system evaluation will include condensing loads associated with coil steam injection and other process gas increases.

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