Vacuum unit performance

Inaccurate feed characterisation and process modelling errors are major contributors to poor performance in a vacuum unit as refiners switch to heavier crude types

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Refinery grassroots or revamp vacuum units frequently fail to meet expected product yield, product quality or run length targets. Real performance versus design is often 3-6 wt% lower in vacuum gas oil (VGO) product yield on whole crude, with much higher VGO metals and microcarbon, and 12 to 24-month run lengths instead of 48 to 60-month targets. Lost profitability can be tens or hundreds of millions of US dollars per year. Designers blame a low VGO yield, poor VGO quality or short run lengths on crude blend, refiner operations, equipment suppliers' errors or numerous other perceived causes; rarely is it attributed to a lack of know-how by the designer. In this age of easy-to-use computer simulations, there is a belief that even an inexperienced engineer will be able to design a successful vacuum unit if appropriately sophisticated software is utilised. Experience proves otherwise. Low VGO product yield and poor vacuum unit reliability are becoming more common even though the global refining industry is becoming more competitive.

Today, most design engineers are experts in running process simulations and equipment models, yet almost none have validated them by comparing model output to actual measured performance. Many designers rarely have first-hand experience of the results of their work. The office-based approach presumes model results represent actual unit performance. Nowhere in the refinery can this assumption be more catastrophic than in the vacuum column simulation.

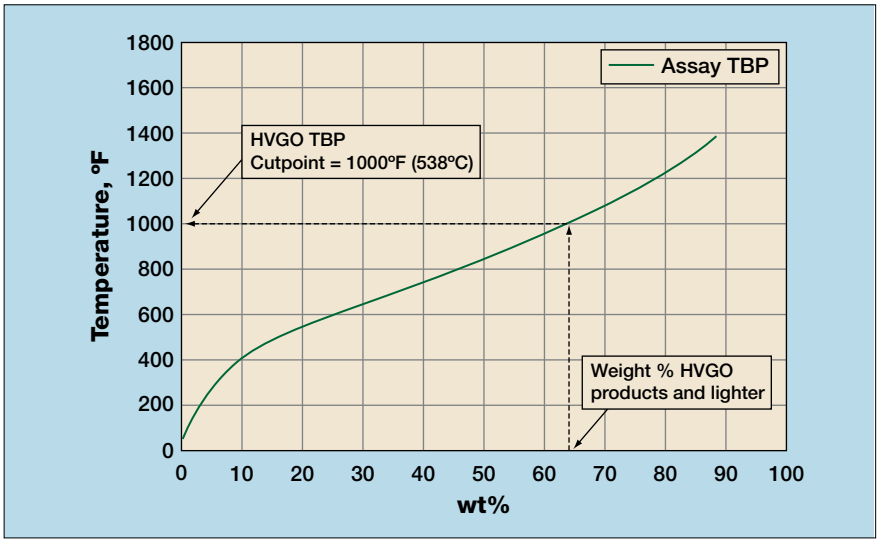

Figure 1 TBP cutpoint

It is no surprise that basic knowledge is critical for successful vacuum unit design, including appropriate analytical tests for accurate feed characterisation, and the influence of process simulation structure is overlooked when performing a grassroots or revamp design. This article addresses fundamental feed characterisation and process simulation principles, which are part of the basic know-how required to design a vacuum unit. Faulty equipment modelling and design are also contributing factors to under-performance in a vacuum unit, but they are not part of this discussion.

Feed characterisation

Accurate feed characterisation is essential to predict VGO product yield and to meet run length targets. Without knowing the true amount of VGO product in the feed, how is it possible to predict its yield? VGO product yield is often specified as true boiling point (TBP) cutpoint. The TBP curve plots the weight or volume per cent distilled of whole crude on the X-axis against TBP temperature on the Y-axis. The TBP temperatures are produced from a standard laboratory test or series of tests. VGO product TBP cutpoint is simply the temperature taken from the curve corresponding to the percentage of total products lighter than the vacuum residue produced (see Figure 1). Starting with inaccurate TBP data or, worse, using laboratory tests such as ASTM D1160 data (that must be converted to TBP temperature) ensures unit performance will not meet expectations. Even today, many vacuum units are designed and product specification guarantees based on ASTM D1160 distillations. This laboratory test should never be used to design or monitor a vacuum unit. Unfortunately, few designers are aware of proper feed characterisation techniques, including the use of

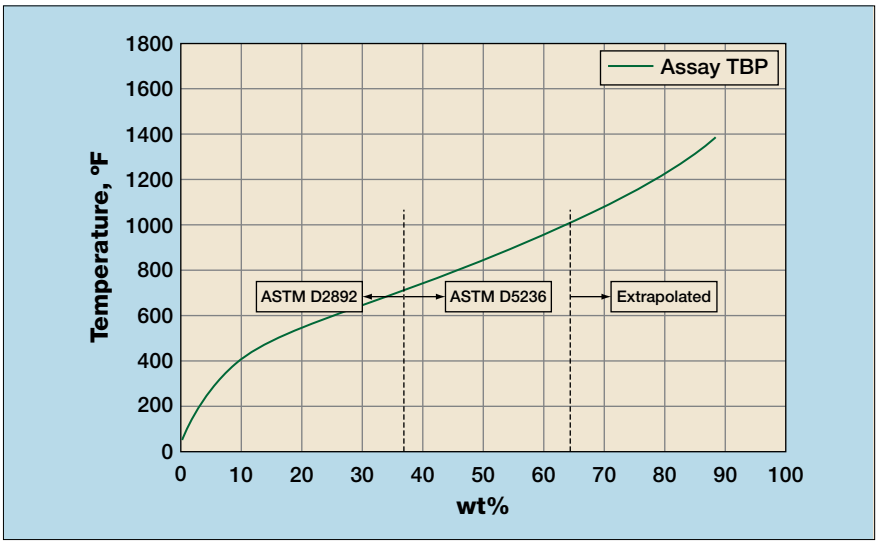

Figure 2 Crude assay TBP laboratory tests

a modern, high simulated temperature distillation (HTSD) ASTM D7169 chromatographic method.

TBP curve

Crude assay TBP curves are reported as a single curve, yet they are typically created from a series of tests (see Figure 2), including ASTM D2892 and D5236 batch distillation. TBP curves are used in process simulations, in conjunction with specific gravity curves, to generate pseudo-component properties. Psuedo-component properties are utilised by generalised correlations (GS, BK10) or equations of state (PR, SRK) to generate vapour equilibrium K-values. Understanding the individual test methods and how they influence the temperatures reported on the TBP curve is important. Since the ASTM D2892 and D5236 methods use completely different methods, and operate at both atmospheric and vacuum pressures, the TBP curve generated has inherent inaccuracies that influence the reported temperatures. In practice, as crude oil gets heavier (Maya, Venezuelan crude, Canadian bitumen, Arab Heavy, Marlim, and so on), the crude assay TBP curves generated by the standard ASTM D5236 become increasingly inaccurate above 800°F (427°C), leading to poor process modelling predictions of actual vacuum unit performance.

Most crude TBP curves are generated from ASTM D2892 and ASTM D5236 test methods. TBP curves generated by these test methods have a range of 70°F (21°C) to approximately 1000°F (538°C). Maximum reported TBP temperatures vary from 960°F (516°C) to 1050°F (566°C), depending on crude type and oil stability. Rarely is it possible to operate the ASTM D5236 test above 1000°F (538°C) atmospheric equivalent temperature (AET) with heavy crudes due to poor thermal stability. Only those crude TBP curves created with an ASTM D7169 allow characterisation beyond oil thermal cracking limits.

An ASTM D2892 test uses a distillation column with 15 theoretical stages of efficiency and operates at 5/1 to 2/1 reflux ratios to fractionate the 700°F (371°C) minus portion of the whole crude into individual cuts. These samples can be further analysed for such properties as specific gravity, PIONA, cetane index, freeze, and so on. Part of the ASTM D2892 operates at atmospheric pressure and part under vacuum to keep the batch pot temperature below the oil cracking temperature. Since the ASTM D2892 column has 15 stages and uses reflux, the individual cuts have only a small distillation overlap between adjacent cuts. On the other hand, the ASTM D5236 test is a single-stage flash operated at 2 mmHg absolute pressure or lower with no reflux to keep the pot temperature below the cracking limit. Since the ASTM D5236 does little fractionation, the reported initial and final cut temperatures

are higher than they would be if the samples were well fractionated such as the ASTM D2892. This lack of fractionation causes the 700°F (371°C) plus portion of the reported TBP temperature to be higher than a well-fractionated sample with little overlap. In fact, when plotting raw test measurements, there is a step change between the D2892 and D5236 portions of the TBP curve caused by the lack of fractionation typical of the difference in D2892 and D5236 test methods. The ASTM D5236 temperature is higher.

The authors have been using TBP curves generated with high-temperature simulated distillations (HTSD) since 1994 to characterise crude oils and diesel and heavier products. The HTSD results are input to the process simulation directly as TBP by weight. More than 50 vacuum units have been field tested to confirm the validity of using HTSD (ASTM D7169) as a TBP curve to model crude and vacuum unit operation. The field work included extensive use of the HTSD to characterise the whole crude and to characterise all the product streams heavier than diesel. These field test runs also included data consistency checks, such as synthesising the whole crude from the corrected material balance and individual streams HTSD distillations and specific gravities. The synthesised TBP distillation and whole crude HTSD TBP distillation were compared to ensure data consistencies prior to input to the process model. Prior to the HTSD method being recently accepted by the ASTM, HTSDs were used by only a few refiners to characterise crude oil.

ASTM D7169 high-temperature simulation distillations, from a properly calibrated and well-maintained machine, produce a TBP curve with lower temperatures in the 700°F (371°C) plus portion of the curve compared to a TBP curve generated with ASTM D2892 and D5236. This difference in TBP curves generated by HTSD versus conventional methods is generally more pronounced with heavy, low-API gravity crude oil (see Figure 3).

Characterising vacuum unit feed with ASTM D7169 has proven much better at predicting VGO yields and matching actual vacuum unit operating conditions than conventional TBP crude assays or other tests such as D1160 or D2887 simulated distillation that has a limit of 1000°F (538°C). Since ASTM D7169 is a chromatograph test it does not have the same inherent problem of limiting pot temperatures because of oil cracking as physical distillation tests and can characterise the oil up to a temperature of 1364°F (740°C). D7169 was specifically designed to characterise heavy oils such as crude oil and resids. This test expands the capability of the D2887 test, which has an EP limit of 1000°F (538°C). Field measurements and process modelling show that ASTM D7169 produces a more accurate TBP for the 800°F (427°C) plus portion of the curve than ASTM D5236.

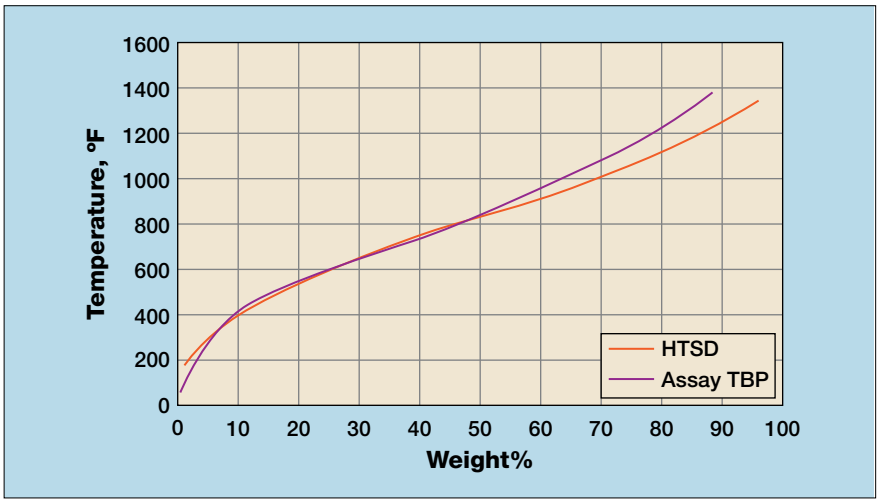

Figure 3 TBP vs HTSD comparison for 20 °API heavy crude

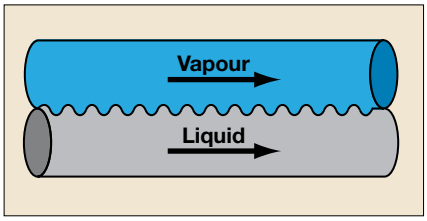

Figure 4 Transfer line stratified flow

Transfer line non-idealities

After feed is characterised accurately, the process simulation must be structured to ensure it correctly represents the actual transfer line operation. The transfer line does not operate ideally. Process and equipment simulations must be modified to take into account non-idealities such as phase separation and critical velocity constraints.

Transfer lines usually have long horizontal runs with line diameters of 36-84in or larger 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 (see Figure 4). In the stratified flow regime, liquid and vapour 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.

Calculating the transfer line pressure profile requires the use of a single model linking together the vacuum column flash zone, transfer line and vacuum heater. The model must identify the location and pressure in the transfer line where liquid and vapour separate and are no longer in equilibrium. The hydraulic tool used to calculate the transfer line pressure profile must properly calculate critical two-phase mass velocities and adjust the pressure so that critical velocity is not exceeded. Few hydraulic models can accurately make the critical mass velocity calculation, hence

estimating the point where phase separation occurs is problematic.

The transfer line pressure profile and heater outlet pressure (see Figure 5) are set by the transfer line design. The transfer line pressure drop and two-phase critical velocity both influence this pressure. The pressure drop is based on a transfer line configuration. Two-phase critical velocity is much lower than the sonic velocity of the gas phase alone. Hence, many designers calculate a low percentage of sonic velocity and predict a much lower heater outlet pressure than is feasible by incorrectly using the sonic gas velocity instead of the critical velocity based on a two-phase flow. Calculation methods for critical two-phase velocity are complex but essential for accurately predicting vacuum unit performance.

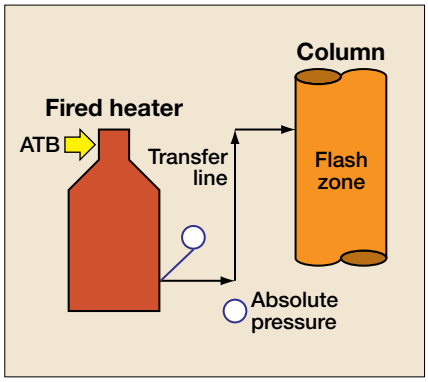

Figure 5 Heater outlet pressure

Process model structure

The majority of the vacuum unit designs that fail to meet VGO cutpoint or run length objectives are a result of a failure to acknowledge the superheated vapour feed to the wash bed. Incorrect model structure over-predicts VGO product yield by 3-6 wt% on whole crude and under-predicts the wash oil flow rate needed to prevent coking by as much as 300%. Most designers assume the transfer line liquid and vapour entering the vacuum column flash zone are in equilibrium. Vapour and liquid in the large portion of the transfer line are not in equilibrium. Liquid and vapour phases are stratified, with vapour moving in the top of the line and liquid on the bottom of the large-diameter pipe. The process simulation needs to account for phase separation in the transfer line and the associated superheated vapour temperature entering the wash section.

Figure 6 shows the proper simulation structure to account for transfer line phase separation and non-equilibrium. This technique, while only an approximation, more accurately predicts VGO product yield and vacuum residue yield, and better estimates the wash flow rate needed to avoid coking the wash section packing. The transfer line pressure profile from the heater outlet to the column can only be calculated accurately with a hydraulic model that uses critical flow limiting algorithms.

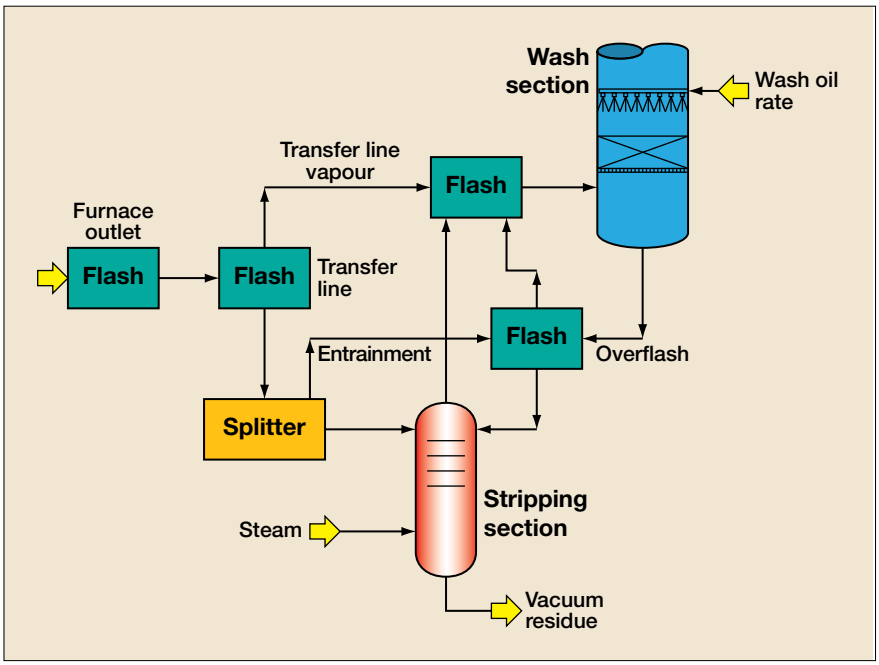

Figure 6 Process model structure

Example

In a recent example, the designer specified a dry vacuum unit (no heater coil steam and no stripping steam) to process 100% Maya crude oil while achieving a VGO product TBP cutpoint of 1050°F (566°C). When the unit started up, the VGO product yield was more than 4 wt% lower on whole crude than design. Table 1 compares the design basis equilibrium model and properly structured non-equilibrium models. The equilibrium model shows the calculated heater outlet temperature of 773°F (412°C) needed to meet a VGO product cutpoint of 1050°F (566°C) with a vacuum column flash zone pressure of 15 mmHg absolute pressure. Alternately, the non-equilibrium model predicts the VGO product cutpoint closer to a 985°F (529°C) cutpoint or 4 wt% lower VGO product on whole crude than design when the model is configured to properly account for vapour superheat. Unfortunately, many refiners experience these low yields after revamping or designing new vacuum units.

Vacuum column wash bed coking often sets the vacuum unit run length. Model structure has a major effect on the calculated wash oil flow rate needed to keep the bed

Design basis and actual VGO product yield

	Equilibrium model	Non-equilibrium model	_
VGO TBP cutpoint, °F (°C)	1050 (566)	985 (530)	- 65 (36)
Heater outlet temperature, °F (°C)	773 (412)	773 (412)	-
Flash zone pressure, mmHg absolute	15	15	-

Table 1

from drying out. The wash oil flow rate must be sufficient to keep the middle of the bed wetted. Dry-out or, more appropriately, a high oil residence time in the middle of the bed leads to coking. Once the bed begins to coke, the open area is reduced, leading to residue entrainment and high VGO product contaminants. The VGO product colour turns black. As the bed cokes, the heater outlet temperature must be reduced to lower the vapour flow rate to control the amount of VGO contaminants. Hence, VGO product yields actually decrease as the run length progresses or the unit must be shut down to remove the coked packing (see Figure 7).

Several refiners have experienced premature annual shutdowns before making the design changes needed to prevent them. These changes have included larger wash oil spray headers designed to handle a higher wash oil flow rate identified with a proper model structure. Often the increase in wash oil needed is two-and-a-half to three times that of an equilibrium model.

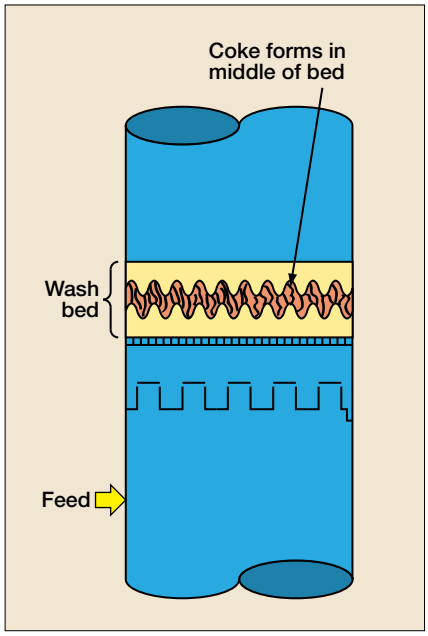

Figure 7 Coked wash bed

Vacuum unit design

Vacuum units need to meet VGO product yield and run length goals to be successful. Fundamental principles and know-how are needed, especially as refiners take advantage of heavier, lower-cost crude oils. Inaccurate feed characterisation and process modelling errors are but two of the major contributors to poor performance. Many equipment design considerations and features are also needed for a successful revamp.

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