
Distillation of the “Light Ends” from Crude Oil in Petroleum Processing

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Contents

Introduction	200
A Process Description of a “Light End” Unit	200
Developing the Material Balances for Light Ends Units	202
Light Ends Specifications Cases	203
Example Case	204
Calculating the Operating Conditions in Light Ends Towers	206
Calculating the Tower Top Pressure and Temperature for Totally Condensed Distillate Product	206
Calculating the Overhead Conditions for Partially Condensed Distillate and Product . . .	208
Calculating the Tower Bottom Conditions for Light Ends Towers	210
Calculating the Number of Trays in Light Ends Towers	210
The Short-Cut Method for Predicting Number of Theoretical Trays	211
An Example of the Underwood Equation Calculation	213
Condenser and Reboiler Duties	214
Calculating the Condenser Duty	214
Calculating the Reboiler Duty	215
Calculating the Overall Tower Heat Balance	215
Tower Loading and Sizing	216
Tower Loading and Overall Tower Diameter	217
Tray Spacing	218
Example Tower Sizing Calculations	223
Checks for Light Ends Tower Operation and Performance	225
Cold Feed	225
Hot Feed	226
Ideal Feed Condition	226
Entrainment	226
Downcomer Backup and Flooding	226
Low Tower Loading	227
Operating Close to Critical Conditions	227

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Abstract

The “light ends” unit is the only process in a refinery configuration that is designed to separate “almost” pure components from the crude oil. Its growth initially resulted from the need of those components such as butanes and propanes to satisfy a market of portable cooking fuel and industrial fuels. The need for modern gasolines has added to the demand for light ends to make high-octane blending components. This chapter describes the process and design of units to separate the light ends into useful products or into cuts for further processing in the refinery. An example design is provided.

Keywords

Light ends • Butane • Propane • LPG • Refinery

Introduction

The “light ends” unit is the only process in a refinery configuration that is designed to separate “almost” pure components from the crude oil. Its particular growth has resulted from the need for those components such as butanes and propanes to satisfy a market of portable cooking fuel and industrial fuels. That these products can be suitably compressed and stored in small, easily handled containers at ambient temperatures provided the market popularity for these products, suitably titled butane LPG and propane LPG. The term LPG refers to liquefied petroleum gas.

The introduction of “no lead” gasoline in the late 1960s established the need of octane sources additional to the aromatics provided by high severity catalytic reforming. A source of such high-octane additives is found in some isomers of butane and pentane. This added to the need for light ends processes which in many cases included the separation of isobutanes from the butane stream and also isopentanes from the light naphtha stream.

A Process Description of a “Light End” Unit

The “light ends” of crude oil are considered to be those fractions in the crude that have a boiling point below cyclohexane. The “light ends” distillation units however include the light naphtha cut separation, which is predominately pentanes and cyclopentanes, from heavy naphtha which contains the hexanes and heavier hydrocarbons necessary for the catalytic reformer feed. The feed to the “light ends” distillation process is usually the full range naphtha distillate from the atmospheric crude distillation unit overhead condensate drum. In many cases the distillates from stabilizing cracker and reformer products are added to the crude unit overhead distillate to be included in the “light ends” unit feed.

A typical process configuration for this unit is given in the flow diagram (Fig. 1).

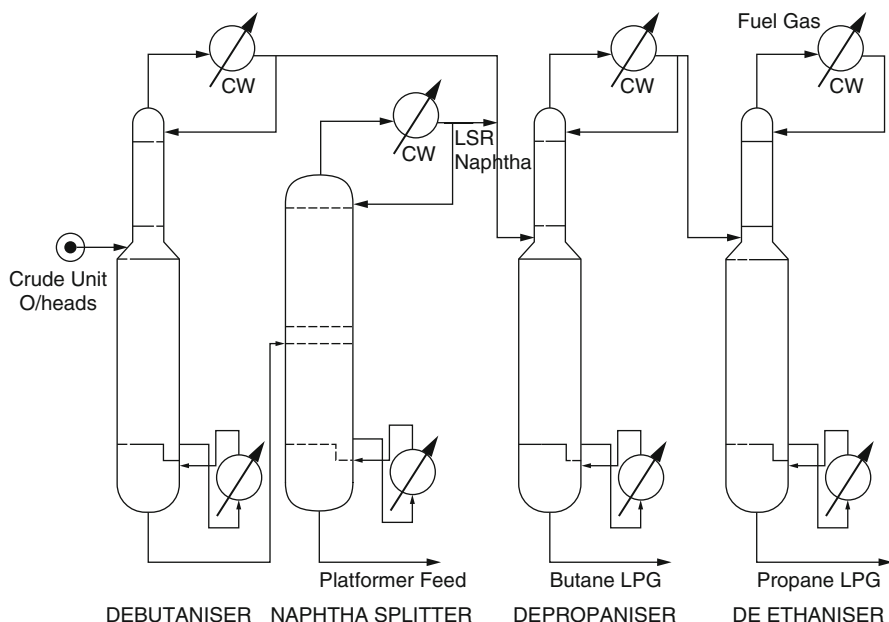


Fig. 1 A typical light ends unit configuration

In this configuration the total feed to the unit is debutanized in the first tower. The butanes and lighter hydrocarbons are totally condensed and collected in the column's overhead drum. Part of this condensate is returned to the tower top tray as reflux. The remainder is routed to a depropanizer column. The bottom product from the debutanizer is the full range naphtha product. This enters a naphtha splitter column where it is fractionated to give an overhead distillate of light naphtha and a bottom product of heavy naphtha.

The depropanizer separates the debutanizer overhead distillate to give a propane fraction as an overhead distillate stream and the butane fraction (butane LPG) as the bottom product. The overhead distillate is fractionated in a deethanizer column to produce a rich propane stream (propane LPG) as the bottom product. The overhead from this column is predominately hydrocarbons lighter than propane. This stream is only partially condensed to provide reflux for the tower. The uncondensed vapor is normally routed to the refinery's fuel gas system.

The products from the "light ends" unit are as follows:

Naphtha splitter – light naphtha (overhead distillate)

Heavy naphtha (bottom product-reformer feed)

Depropanizer – butane LPG (bottom product)

Deethanizer – propane LPG (bottom product)

Fuel gas (overhead vapor)

Developing the Material Balances for Light Ends Units

In light ends towers, the material balance is developed on a molal balance. This type of balance is determined by the degree of separation of the feed molal components that enter the distillate fraction and those that leave with the bottom product.

Effective separation by fractionation in light ends towers obeys the same laws as those in the crude distillation units, that is: the degree of separation is the result of the number of trays (or stages) and the reflux (or overflow) in the column.

In the crude unit, this separation was measured by the difference between the ASTM 95 % point of the lighter fraction and the 5 % ASTM point of the heavier fraction. This is the ASTM gap or overlap.

In light ends towers the degree of separation is a little more precise. This is determined by the distribution of key components in the two fractions to be separated. Key components may be real components (such as C₄s or C₅s) or pseudo-components defined by their mid-boiling points. Normally key components are adjacent components by boiling point in the feed composition. Any two key components may be selected – a light key and a heavy key. By definition, the light key has the lower boiling point. Both key components must, however, be present in the distillate and bottom products of the column. If a side stream exists, then these keys must also be present in the side stream product.

There are several correlations that describe the behavior of these key components in their distribution and relationship to one another. By far, the more common of these correlations is the Fenske equation which relates the distribution of key components at minimum trays with infinite reflux. The equation is relatively simple and does not require iterative calculation techniques to solve it. The Fenske equation is:

$$N_{m+1} = \frac{\text{Log} [((\text{LT key}/\text{HY key})_D \times (\text{HY key}/\text{LT key})_W)]}{\text{Log} \frac{K_{\text{LT key}}}{K_{\text{HY key}}}} \quad (1)$$

where:

N_m = minimum number of theoretical trays at total reflux. The +1 is the reboiler which is counted as a theoretical tray.

LT key = is the mole fraction of the selected light key.

HY key = is the mole fraction of the selected heavy key.

D = fractions in the distillate product.

W = fractions in the bottom product.

$K_{\text{LT key}}$ = the equilibrium constant of the light key at mean system condition of temperature and pressure.

$K_{\text{HY key}}$ = the equilibrium constant of the heavy key again at mean system conditions.

The ratio of the equilibrium constants is called the "relative volatility" of the keys. Setting fractionation requirements for light ends towers is usually done to meet a product specification. More often than not, this specification is the vapor pressure of the heavier fraction or the tolerable amount of a heavy key allowed in the lighter fraction.

Sometimes, however, a specification for the separation may not be given. Under this circumstance, some judgment must be made in determining the most reasonable separation that can be achieved with the equipment. This item of the handbook addresses calculation techniques that satisfy either premise.

It is common to use process simulation software (HYSYS, Pro/II, and others) to perform the design calculations for light ends units. This allows case studies to optimize the process design. The details of the computer simulations are beyond the scope of this handbook. The manual procedure for the calculations is described here. It is helpful to understand the manual approach to establish the background and limitations behind the computerized calculations.

Light Ends Specifications Cases

Case 1: Setting separation requirements to meet a specification

In this case it is required to determine the amount of butanes that can be retained by a light naphtha cut to meet a RVP specification. The steps are as follows:

Step 1. Calculate the properties of the C_5 + naphtha from a component breakdown. These properties should give weight and mole rates per hour.

Step 2. Carry out a bubble point calculation of the C_5 + fraction and inserting butane as the unknown quantity x .

Step 3. The equilibrium constant used (K) in the calculation will be at the temperature and pressure conditions of the RVP (i.e., normally at 100 °F which is the test temperature).

Step 4. Either the equilibrium constants given in the charts in Appendix 3 of the chapter entitled "Process equipment for petroleum processing" in this handbook or the relationship of vapor pressure divided by the total system pressure may be used.

Step 5. By definition, the total moles of liquid given is equal to the total moles of vapor (calculated) in equilibrium at the bubble point. Thus equate and solve for x as the quantity of butanes tolerable to meet RVP.

Case 2: Setting fractionation where no specification is given

Step 1. Determine the composition of the feed in terms of real components, pseudo-components, or both. Calculate this in moles/h and mole fractions.

Step 2. Select the key components, and decide the minimum distribution of one or other key. For example, in a debutanizer C_5 s allowed into LPG must not be more than 2 % of the total C_4 s and lighter. This is to protect the Butane LPG "Weathering" Test Specification.

Table 1 Rules of thumb for number of light ends column trays

Tower	Number of actual trays
Debutanizer	30–35
Depropanizer	35–40
Deethanizer	38–42
Naphtha Splitter	25–35

Step 3. Give values to the fraction of LT and HY keys in the distillate and bottoms. Use x as the unknown where appropriate.

Step 4. The Fenske equation will be used to calculate the distribution of keys. Determine the value of N_m by taking the actual number of trays and using an efficiency of 70–75 % to arrive at total theoretical trays. Divide this figure by 1.5 to arrive at the minimum theoretical trays N_m . Do not forget to add 1 for the reboiler to use in the equation.

Step 5. Estimate the mean tower conditions. This can be achieved by examining past plant logs and other sources. Determine K values for the keys at this mean tower condition. Use published data for real components and the ratio of vapor pressure divided by total systems pressure for pseudo-components.

Step 6. Solve for x in the Fenske correlation. This will be the split of the key components and the basis for the material balance.

Very often when setting up a design for a light ends tower, the actual number of trays is not known at the time when calculation number two is required. The rules of thumb in Table 1 may be used as a guide.

Example Case

An example for illustrating Case 1 above is given in the chapter entitled “► [Introduction to Crude Oil and Petroleum Processing](#)” of this handbook where the amount of butane LPG allowed to meet gasoline RVP is calculated. An example of Case 2 is given as follows:

The overhead distillate from an atmospheric crude distillation unit operating at 50,000 BPSD of Murban crude has the composition noted in Table 2.

The key components for the debutanizer will be nC_4 and iC_5 as LT and HY keys, respectively. In the Fenske equation let x be the moles/h of nC_4 in the distillate. To satisfy the weathering test for butane LPG, the maximum amount of C_5 s allowed in the debutanizer distillate is 2.0 mol% of the total C_4 s and lighter. A reasonable number of actual trays in a debutanizer is 30. Allowing an efficiency of 70 %, the number of theoretical trays will be 21. It is reasonable to predict that the minimum theoretical trays will be the actual theoretical number divided by 1.5. Then N_m will be $21/1.5 = 14$ adding one for the reboiler giving $N_{m+1} = 15$.

From past data, average operating conditions of temperature and pressure for a debutanizer are 210 °F and 110 psig. At these conditions, the equilibrium constants for both keys are read from curves given in the GPSA Engineering Data as:

Table 2 Full range naphtha composition

Comp	BPSD	lb/gal	Gals/h	lb/h	MW	Moles/h
C2	20	3.42	35	120	30	4
C3	240	4.23	420	1,777	44	40.38
iC4	275	4.70	481	2,262	58	39.0
NC4	735	4.87	1,286	6,264	58	108.0
iC5	915	5.21	1,601	8,343	72	115.87
nC5	1,216	5.26	2,126	11,184	72	155.33
C6	2,170	5.56	3,798	21,114	84	251.36
C7	2,930	5.71	5,128	29,278	100	292.78
Mbpt 224	1,075	6.12	1,881	11,513	106	108.62
239	1,525	6.15	2,669	16,413	109	150.58
260	1,345	6.22	2,354	14,640	115	127.31
276	895	6.26	1,566	9,805	120	81.71
304	895	6.35	1,566	9,946	130	76.51
Total	14,235	5.73	24,911	142,659		1,551.45

$$nC_4 = 1.48$$

$$iC_5 = 0.94$$

Then the relative volatility $\phi = 1.48/0.94 = 1.57$.

Using the Fenske equation:

$$N_{m+1} = \frac{\text{Log}((\text{LTkey}/\text{HYkey})_D \times (\text{HYKey}/\text{LTkey})_W)}{\text{Log}\phi}$$

$$15 = \frac{\text{Log}((x/3.8 \times 112.1/108 - x))}{\text{Log} 1.57}$$

$$15 \times 0.196 = \text{Log}((112.1x)/(410.4 - 3.8x))$$

$$871 = 112.1x/(410.4 - 3.8x)$$

$$x = 104.48 \text{ moles/h.}$$

Associated with nC_4 in the bottom product will be an equilibrium amount of iC_4 . Although small, this will have an effect on the bottom product bubble point and therefore the tower bottom temperature. The amount of this iC_4 in the bottom product can be calculated using a similar method as that for the split between nC_4 and iC_5 ; thus the keys for this calculation will be iC_4 and nC_4 . Let x be the moles/h of iC_4 in the bottom product. Then:

	Feed	Dist	Bottoms
iC4	39	x	$39 - x$
nC4	108	104.48	3.52

Using the Fenske equation again:

$$\begin{aligned}
 15 &= \frac{\log((x/104.48) \times (3.52/39 - x))}{\log(KiC_4/KnC_4)} \\
 &= \frac{\log((x/104.48) \times (3.52/39 - x))}{\log 1.3} \\
 1.709 &= \log((x/104.48) \times (3.52/39 - x))
 \end{aligned}$$

Then:

$x = 38.99$ mol/h. This gives zero iC_4 in the bottom cut.

The material balance for the debutanizer can now be written and is given in Table 3.

The material balances over the naphtha splitter, depropanizer, and the deethanizer follow the same technique in identifying key components and utilizing the Fenske equation.

Calculating the Operating Conditions in Light Ends Towers

Light ends units follow two calculation procedures for setting the conditions of temperature and pressure in the fractionating towers. The first procedure relates to towers in which the overhead product and reflux are totally condensed. The second procedure relates to those towers where the overhead product is not totally condensed.

Calculating the Tower Top Pressure and Temperature for Totally Condensed Distillate Product

This procedure commences with setting a realistic reflux drum temperature. This is fixed by the cooling medium temperature, such as ambient air temperature (for air coolers) or cooling water. As total condensation is required, then the pressure of the reflux drum must be the bubble point of the distillate product (and reflux) at the selected drum temperature. Once the drum pressure has been calculated, the tower top pressure can be determined by taking into account estimated or manufacturer's specified pressure drops for equipment and piping between the drum and the tower top. As a rough estimate, condensers and/or heat exchangers in the system have between 3 and 5 psi pressure drop. Allow also about a 2 psi pressure drop for piping.

The tower top temperature is calculated as the dew point of the distillate product at the total overhead pressure. There are usually no steam or inert gases present in the light ends tower overheads, so total pressure may be used. The following example uses the debutanizer overhead from the material balance given in section "[Developing the Material Balance for Light Ends Units](#)" above.

Table 3 Debutanizer material balance

Comp	Feed			Dist			Bottoms		
	BPSD	lb/h	mol/h	BPSD	lb/h	mol/h	BPSD	lb/h	mol/h
C2	20	120	4	20	120	4.00			
C3	240	1,777	40.38	240	1,777	40.38			
iC4	275	2,262	39.00	275	2,262	39.00			
nC4	735	6,264	108.00	711	6,061	104.48	24	203	3.52
iC5	915	8,343	115.87	30	274	3.80	885	8,069	112.07
nC5	1,216	11,184	155.33				1,216	11,184	155.33
C6	2,170	21,114	251.36				2,170	21,114	251.36
C7	2,930	29,278	292.78				2,930	29,278	292.78
MBP+224	1,075	11,513	108.62				1,075	11,513	108.62
239	1,525	16,413	150.58				1,525	16,413	150.58
260	1,345	14,640	127.31				1,345	14,640	127.31
276	895	9,805	81.71				895	9,805	81.71
304	895	9,946	76.51				895	9,805	76.51
Total	14,236	142,659	1,551.45	1,276	10,494	191.66	12,960	132,165	1,359.79

Table 4 Debutanizer reflux drum pressure. Distillate bubble point calculation at 100 °F (Trial 1)

Comp	Mol frac x	K at 125 psia	$Y = X \cdot K$
C2	0.0209	4.8	0.3906
C3	0.2107	1.48	0.4054
iC4	0.2035	0.68	0.0830
nC4	0.5451	0.5	0.1254
iC5	0.0198	0.23	0.0009
Total	1.0000		1.0053

Reflux drum pressure = 125 psia and 100 °F

Table 5 Tower top temperature

Comp	Mole fract (y)	K at 152 °F	Mole fract (x)
C2	0.021	6.2	0.003
C3	0.211	2.1	0.100
iC4	0.203	1.1	0.185
nC4	0.545	0.83	0.657
iC5	0.02	0.41	0.048
Total	1.000		0.993

The ambient air temperature for the site is 60 °F, and the operating temperature for the reflux drum will be set at 100 °F. The bubble pressure at this temperature is calculated and given in Table 4.

Allowing 3 psi pressure drop over the overhead condenser and 2 psi for the associated overhead piping, the tower top pressure becomes 130 psia (115 psig). The temperature of the tower top is the dew point of the distillate product at the top pressure. This is given in Table 5.

There is reasonable agreement that total $y =$ total x ; therefore the dew point and the tower top temperature at 115 psig is 152 °F.

Calculating the Overhead Conditions for Partially Condensed Distillate and Product

There are two circumstances where the overhead stream from a light ends tower may not be totally condensed. The most common of these is in the case of the deethanizer. Here, usually, only sufficient overhead stream is condensed to provide the overhead reflux stream. The reason for this is that at a normal condensing temperature, the pressure required at the reflux drum would be unacceptably high. It would be so high that the tower bottom pressure required would be higher than the bottom product's critical pressure, and thus fractionation would not be possible. In this case the tower pressure is set again by the reflux drum pressure. Unlike the case of total condensation, the reflux drum pressure for this partial condensation is found as the dew point at the condensing temperature of the distillate product. In fact the reflux drum becomes a theoretical fractionation stage. In certain cases refrigeration is used to totally condense the overhead stream at an acceptable pressure, but this selection would be as a result of a study of the operation's economics.

In certain processes where a high tower pressure (and consequently high temperature) may cause deterioration of one or more of the products from the tower, the tower pressure is reduced by condensing only a fraction of the overhead distillate product. For this purpose, a vaporization curve for the distillate product is constructed at a selected reflux drum temperature. This curve is developed by calculating a series of equilibrium compositions of the distillate product at the reflux drum temperature but over a pressure range. A reflux drum pressure can then be selected from the curve that satisfies an acceptable tower pressure profile. A final equilibrium calculation is then made at this pressure to provide the component and quantity composition of the liquid and vapor streams leaving the reflux drum.

An example of deethanizer overhead operating conditions is given below:

Assume that the overhead vapor from the reflux drum, which is the product in this case, has the following molar composition:

Mole	Fraction
C ₁	0.063
C ₂	0.443
C ₃	0.494

A reasonable temperature for operating the reflux drum is 100 °F (based on local ambient conditions). The reflux drum pressure is calculated from the dew point of the vapor at the reflux drum pressure, thus:

	Vapor mole fract y	K at 100 °F and 350 psia	Liquid mole fract x
C ₁	0.063	7.0	0.009
C ₂	0.443	2.0	0.222
C ₃	0.494	0.64	0.772
Total	1.000		1.003

This was the second trial starting with the pressure at 450 psia. The total x value is close enough to y to allow the pressure to be 350 psia.

Now the tower top conditions will be at the dew point of the reflux plus the product leaving the tower. The pressure will be the reflux drum pressure plus, say, 7 psi for the condenser and piping pressure drop. In this case this pressure will be 357 psia. Set the reflux ratio (moles reflux/mol product) to be 2.0:1 where the moles/h rate of the product is 7.9. The composition of the reflux stream is the x value of the product vapor. The dew point calculation to establish the tower top temperature therefore will be as shown in Table 6.

Note this dew point calculation may change if subsequent calculations show that the reflux ratio required will be significantly different to the one assumed here.

It is not proposed to show the case of the partially condensed product here. This calculation would be similar to the deethanizer and in this case the reflux composition will be the liquid phase from the equilibrium flash of the product.

Table 6 Calculating the tower top temperature for a deethanizer

Component	Mole/h prod	Mole/h reflux	Total moles/h	Mole fraction y	K at 357 psia and 133 °F	Mole fract liquid x
C1	0.5	0.14	0.64	0.027	0.42	0.064
C2	3.5	3.49	6.99	0.295	2.00	0.148
C3	3.9	12.17	16.07	0.678	0.86	0.788
Total	7.9	15.80	23.70	1.000		1.000

Table 7 Debutanizer bottom temperature (bubble point)

Components	Moles/h	Mol fract x	K at 138 psia and 358 °F	$Y = x/K$
nC4	3.5	0.003	3.7	0.010
iC5	112.1	0.082	2.4	0.198
nC5	155.33	0.114	2.0	0.228
C6	251.35	0.185	1.4	0.259
C7	292.78	0.215	0.77	0.166
Mpt 224 ° F	108.62	0.080	0.58	0.046
239 °F	150.58	0.111	0.49	0.054
260 °F	127.31	0.094	0.38	0.036
276 °F	81.71	0.060	0.29	0.017
304 °F	76.51	0.056	0.21	0.012
Total	1,359.79	1.000		1.026

The tower bottom condition is therefore 138 psia and 358 °F

Calculating the Tower Bottom Conditions for Light Ends Towers

The calculation to establish the tower bottom conditions for all light ends towers is the same. Only the values for pressure and composition change. The temperature of the product leaving the bottom of the tower will be at its bubble point at the tower bottom pressure. Consider the debutanizer column whose material balance was given in Table 2.

The number of actual trays (estimated) to accomplish the separation between the C_{4s} and the C_{5s} is 30. The pressure drop for fully loaded trays can be taken as being between 0.15 and 0.2 psi. Assume then a pressure drop of 0.17 psi per tray. The tower top pressure calculated earlier is 125 psia, and then the bottom pressure is $125 + (0.17 \times 30) = 130$ psia (Table 7).

Calculating the Number of Trays in Light Ends Towers

For definitive design work, one of the many excellent simulation packages should be used. However most simulation packages require good quality input data. This often means a fairly accurate estimate of the number of theoretical trays and the

liquid/vapor traffic in the tower. An acceptably accurate "short-cut" method to arrive at the number of theoretical trays is given here. The estimate of the liquid/vapor traffic is discussed later in this chapter.

The Short-Cut Method for Predicting Number of Theoretical Trays

Three calculations or relationships are used to determine the number of theoretical trays in this method. These are:

The Fenske calculation to determine the minimum number of trays at total reflux
The Underwood calculation to determine the minimum reflux at infinite number of trays

The Gilliland correlation which uses the result of the two calculations to give the theoretical number of trays

The Fenske Equation This has been discussed earlier under the section dealing with the "Material Balance for Light Ends Towers." The equation is as follows:

$$N_{m+1} = \text{Log} \left[\frac{(\text{LT key}/\text{HY key})_D (\text{HY key}/\text{LT key})_W}{K_{\text{LT key}}/K_{\text{HY key}}} \right] \div \text{Log} (K_{\text{LT key}}/K_{\text{HY key}}) \quad (2)$$

where:

N_m = minimum number of theoretical trays at total reflux. The +1 is the reboiler which is counted as a theoretical tray.

LT key = is the mole fraction of the selected light key.

HY key = is the mole fraction of the selected heavy key.

D = fractions in the distillate product.

W = fractions in the bottom product.

$K_{\text{LT key}}$ = the equilibrium constant of the light key at mean system conditions of temperature and pressure.

$K_{\text{HY key}}$ = the equilibrium constant of the heavy key again at mean system conditions.

The Underwood Equation and Calculation The Underwood equation is more complex than the Fenske and requires a trial and error calculation to solve it. The equation itself is in two parts: the first looks at the vapor volatilities (ratio of the K s) of each component to one of the keys and then by trial and error arriving at an expression for a factor B that forces the equation to zero. This first equation is written as follows:

$$\sum ((\phi_i) \cdot (x_i F) \div (x_i F) - B) = 0 \quad (3)$$

where:

ϕ_i = the relative volatility of component i

x_iF = the mole fraction of component i in the feed

B = the factor that forces the expression to zero

The second part of the equation is expressed as follows:

$$R_{(m+1)} = \sum ((\phi_i) \cdot (x_iD)) \div ((x_iD) - B) \tag{4}$$

where:

R_m = minimum reflux at infinite number of trays

x_iD = the mole fraction of i in the distillate

The relationship between the Fenske equation and the Underwood is given by the Gilliland correlation shown in Fig. 2.

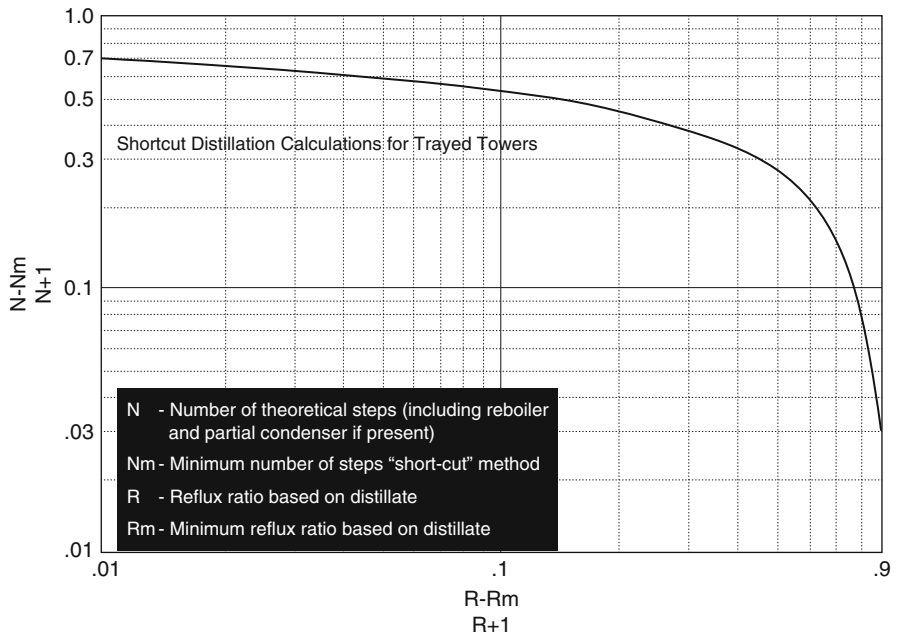


Fig. 2 The Gilliland correlation for calculating theoretical trays

Table 8 Underwood equation, Part 1, calculation example

Comp	Mol fract x_f	K at Ave ^a cond	Rel vol Φ	$x_f \times \Phi$	Trial 3(1) $B = 0.815$
C2	0.003	9.0	4.5	0.0135	0.0037
C3	0.026	4.2	2.1	0.0545	0.0424
iC4	0.025	2.4	1.2	0.0300	0.0776
nC4 (Key)	0.070	2.0	1.00	0.0700	0.3753
iC5	0.075	1.2	0.62	0.0450	-0.2108
nC5	0.100	1.0	0.52	0.0500	-0.1595
C6	0.162	0.49	0.245	0.0397	-0.0698
C7	0.188	0.25	0.125	0.0235	-0.0341
M Bpt 224 °F	0.070	0.16	0.08	0.0056	-0.0076
239 °F	0.097	0.13	0.065	0.0040	-0.0084
260 °F	0.082	0.097	0.0485	0.0039	-0.0052
276 °F	0.053	0.069	0.0345	0.0018	-0.0023
304 °F	0.049	0.042	0.021	0.0010	-0.0013
Total	1.000				0

^aAve conditions are 134 psia and 255 °F

$$\Sigma((\phi_i) \cdot (x_iF) \div (x_iF - B)) = 0(1)$$

Table 9 Underwood equation, Part 2, calculation example

Comp	Mole fract x_D	Rel vol Φ	$(x_D)(\Phi)$	$(x_D)(\Phi)/(B - \Phi)$
C2	0.022	4.5	0.0945	0.0257
C3	0.217	2.1	0.4431	0.3444
iC4	0.207	1.2	0.2436	0.6303
nC4 (Key)	0.544	1.00	0.5450	2.9223
iC5	0.010	0.60	0.012	-0.0562
Total	1.000			3.8665

$$R_{m+1} = 3.8665 \quad R_m = 2.87 \quad \text{and} \quad R = 2.87 \times 1.5 = 4.3$$

An Example of the Underwood Equation Calculation

Consider the material balance of a debutanizer developed in Table 2.

Part 1 of the Underwood equation is calculated as shown in Table 8.

Part 2 of the Underwood equation is calculated as shown in Table 9.

Using the Gilliland Curve: $\frac{(R-R_m)}{R+1} = 0.27$

And from the curve: $\left(\frac{N-N_m}{N+1}\right) = 0.4$.

Then number of theoretical trays N will be $N - N_m = 0.4 N + 0.4$.

N_m calculated from the Fenske equation (see section on "[Developing the Material Balance for Light Ends Units](#)") is 14.

Then $N = 24$.

Assume an average tray efficiency of 70 % then total actual trays = 34.

Condenser and Reboiler Duties

Both the condenser and reboiler duties are the result of the light ends tower heat balance that meets the degree of separation required in the products. It may be said that the condenser duty and operation determine the amount of wash liquid flow in the tower to meet the degree of rectification required for the lighter product. The reboiler in turn generates the vapor flow in the tower to satisfy the degree of stripping required for the particular separation and the quality of the heavy product. At a constant feed temperature and pressure, changes to either the condenser duty or the reboiler duty will affect the duty of the other. In other words the tower must always be in heat balance.

Calculating the Condenser Duty

The duty of the overhead condenser is determined by a heat balance over the tower top (above the top tray) and reflux drum, with the condenser duty being the unknown quantity. Using the data already determined for the debutanizer in the previous sections, the following is a calculation to determine the condenser duty for this unit.

Consider the following heat balance sketch (Fig. 3).

In the following heat balance, the liquid overflow from the top tray is based on the internal reflux ratio obtained from the Fenske, Underwood equations, and the Gilliland correlation. This ratio is in moles and its molecular weight is derived from the molal composition of the dew point calculation to establish the tower top temperature. It is the mole weight of the liquid in equilibrium with the overhead distillate vapor. Thus the total moles of the overflow liquid are $4.3 \times 191.66 = 824.14$ mol/h. Its mole weight is 56.79. Therefore overflow liquid is 46,803 lb/h. The heat balance now follows (Table 10).

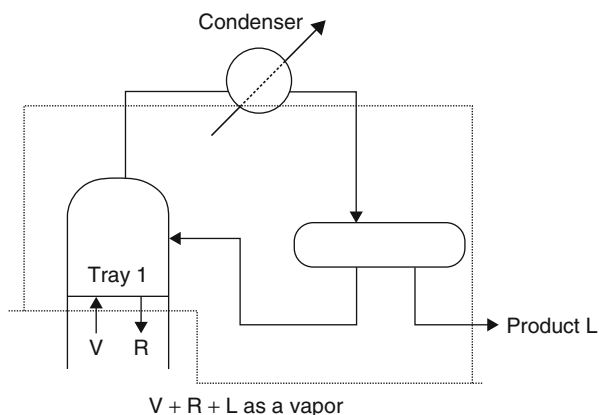


Fig. 3 The overhead heat balance figure

Table 10 Tower top heat balance example

Stream	L/V	Mole wt	°F	lb/h	Btu/lb	MMBtu/h
In						
Ref o/flow	V	56.8	160	46,803	305	14.275
Distillate	V	55	160	10,494	306	3.211
Total in				57,297		17.486
Out						
Dist Prod	L	55	100	10,494	150	1.574
Ref Liquid	L	56.8	155	46,803	190	8.893
Condenser				By difference		7.019
Total out				57,297		17.486

The top tray temperature is taken as 3° F above the tower top temperature. The vapor to the top tray is taken to be about 5° F above the top tray.

Condenser duty is 7,020,000 Btu/h.

Calculating the Reboiler Duty

It is important that the reboiler duty achieve a balance between generating an effective rate of vapor for stripping the bottom product while maintaining an economic vapor load on the stripping trays. As a rule of thumb, this can be achieved with a stripping rate as a percent of bottom product of between 70 and 90 mol %. The tower bottom temperature has already been calculated as 358 °F at 138 psia. The composition of the vapor in equilibrium at these conditions has also been calculated. Thus the moles/h of bottom product are 1,359.79; then the moles of strip-out required will be $1,359.79 \times 0.9 = 1,223.8$ mol/h. The molecular weight of the strip-out is 88 (from the equilibrium composition). The heat balance to determine the reboiler duty now follows. In this calculation the draw-off tray to the reboiler is about 10° F lower than the tower bottom temperature, that is, 348 °F (Table 11).

Calculating the Overall Tower Heat Balance

Knowing the reboiler and condenser duties, an overall heat balance over the tower can be calculated and will provide the preheat required in the feed and consequently its temperature. A calculation to determine the bubble point of the feed at 134 psia (mean inlet pressure) was made. This temperature was 305 °F. Should the calculated temperature be above this, then an enthalpy curve would need to be developed for the feed to determine its actual enthalpy and temperature. The overall heat balance now follows (Table 12).

Enthalpy of the feed is $\frac{21,713,000}{142,659} = 152.2$ Btu/lb; from the enthalpy tables, the temperature is found to be 276 °F. The feed tray is based on a preliminary tower

Table 11 Bottom tray heat balance

Stream	L or V	Temp °F	lb/h	Btu/lb	MMBtu/h
In					
Liquid ex tray 34	L	348	239,860	198	47.492
Reboiler			By difference		13.403
Total in					
Out					
Bot product	L	358	132,165	200	26.433
Strip out	V	358	107,695	320	34.462
Total out			239,860		60.895

Reboiler duty is 13,403,000 Btu/h

Table 12 Overall tower heat balance

Stream	V or L	Temp °F	lb/h	Btu/lb	MMBtu/h
In					
Feed	L	T	142,659	By diff	21.713
Reboiler					13.403
Total in					35.116
Out					
Distillate	L	100	10,494	150	1.574
Bot prod	L	358	132,165	200	26.433
Condenser					7.109
Total out			142,659		35.116

temperature profile with a 75 % efficiency for rectifying trays and a 65 % efficiency for the stripping trays (see Fig. 4). Both the temperature profile and the tray efficiencies will be finalized by a suitable computer simulation package or a rigorous tray-to-tray calculation. Note: The calculation techniques and examples given here are good input to computer simulation packages; wherever possible, these packages should be used for final process design.

Tower Loading and Sizing

Light ends towers all follow the following principles of cross-sectional area sizing and tower height. These dimensions are interrelated by the height required between trays to ensure proper separation of clear liquid from the frothy mixture of the tray inlet fluid. There are many procedures and correlations to determine these dimensions. The following method is just one which will provide a good estimate for tower design. For definitive design, however it is essential that tray manufacturers be consulted and their methodology be used. After all they, the manufacturers, will be required to guarantee the performance of the unit in terms of flooding capacity and tray efficiency. In all cases this has to be a significant consideration, but in case

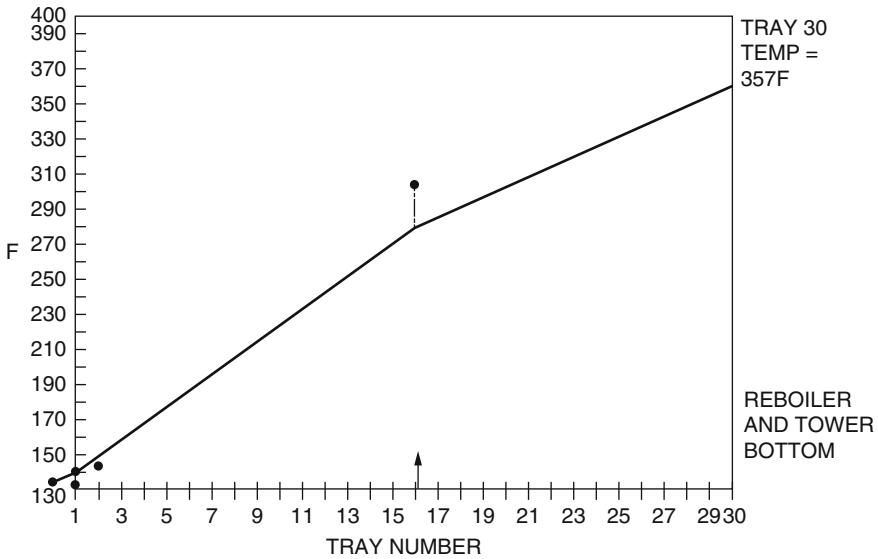


Fig. 4 Tower temperature profile

of super fractionating units with over a hundred trays in many cases, this has to be a primary consideration.

There are many light ends tower computer simulation packages in the market, and most of these calculate vapor and liquid traffic in the tower on a tray-to-tray basis. The input for these programs however does require a fairly accurate estimate of these values for easy convergence and subsequent use of the program. The following procedure with the equations used can provide details of tower loading and tray criteria at critical trays in the tower. A linear correlation between these points will then provide a reasonable loading profile over the tower sufficient for good computer input.

Tower Loading and Overall Tower Diameter

The diameter of a fractionating tower is usually based on the vapor loading on two critical trays. These are the tower top tray which will set the diameter for the rectifying section (i.e., the trays above the feed tray) and the tower bottom tray. This lower tray sets the tower diameter for the stripping section of the tower (i.e., the feed tray and the trays below the feed tray). These two sections may have different diameters.

The vapor loading is based on the rate of vapor passing through the tray and the density of the vapor and the liquid on the tray through which the vapor bubbles. This relationship is given by the Brown and Souders equation. There are several

forms of this equation which can be used with the appropriate physical tray constants. One of the forms used here is as follows:

$$G_f = K\sqrt{(\rho_v \times (\rho_l - \rho_v))} \quad (5)$$

where:

G_f = mass of vapor per sq foot of tray at flood (lb/h · ft²)

K = a constant based on tray spacing at flood (see chapter “► [Atmospheric and Vacuum Crude Distillation Units in Petroleum Refineries](#)”)

ρ_v = density of vapor at tray conditions of temperature and pressure in lb/cuft

ρ_l = density of liquid at tray conditions in lb/cuft

The area thus determined is the “bubble” area of the tray. Normally trays are designed at 85–90 % of flood. Therefore for good design, the G_f is divided by this percentage to give the actual or design area of the tray. The whole tray is made up of two other areas: that for the downcomers and a waste area which is allocated to calming the liquid leaving the bubble area before entering the downcomer. The relationship of these areas to one another is given in Table 13.

Using the criteria in Table 13 and the value of the bubble area based on the vapor loading, the total tray area and therefore the tower diameter can be determined. This relationship is summed up by the following expression:

$$A_s = A_B + A_{dc} + A_w \quad (6)$$

where:

A_s = total tray area

A_B = bubble area

A_{dc} = downcomer area (inlet + outlet)

A_w = waste or calming zone area (usually 15 % of A_s)

Tray Spacing

The tray spacing used in the initial determination of flood loading needs to be checked. If necessary, the spacing and the calculation will be revised to meet the correct spacing criteria. Usually this first guess at tray spacing is taken as 24". The following equations are then applied to determine whether this spacing is satisfactory. These equations calculate the pressure drop across the tray in terms of the clear liquid holdup (or height) in the downcomer. This height of liquid must be less than 50 % of the tray spacing for most applications. In the case of a high foaming process, this height must be less than 40 % of the tray spacing. These pressure drop criteria concerning the tray hydraulics and their associated equations now follow:

Table 13 Valve and sieve tray characterization

Design feature	Suggested value	Alternate values	Comment
1. Valve size and layout			
(a) Valve diameter	–	–	Valve diameter is fixed by the vendor
(b) Percent hole area A_o/A_b	12	8–15	Open area should be set by the designer. In general, the lower the open area, the higher the efficiency and flexibility, and the lower the capacity (due to increased pressure drop). At values of open area toward the upper end of the range (say 15 %), the flexibility and efficiency are approaching sieve tray values. At the lower end of the range, capacity and downcomer filling become limiting
(c) Valve pitch diam ratio	–	–	Valve pitch is normally triangular. However, this variable is usually fixed by the vendor
(d) Valve distribution	–	–	On trays with flow path length $\geq 5'$, and for liquid rates $> 5,000$ GPH/ft. (diameter) on trays with flow path length $< 5'$, provide 10 % more valves on the inlet half of the tray than on the outlet half
(e) Bubble area, A_b , or A_B	–	–	Bubble area should be maximized
(f) Tray efficiency	–	–	Valve tray efficiency will be about equal to sieve tray efficiency provided there is not a blowing or flooding limitation
(g) Valve blanking	–	–	This should not generally be necessary unless tower is being sized for future service at much higher rates. Blanking strips can then be used. Blank within bubble area, not around periphery to maintain best efficiency
2. Tray spacing	–	12–36	Generally economic to use min. values which are usually set by maintenance requirements. Other considerations are downcomer filling and flexibility. The use of variable spacing to accommodate loading changes from section to section should be considered

(continued)

Table 13 (continued)

Design feature	Suggested value	Alternate values	Comment
3. Number of liquid passes	1	1–2	Multi-passing improves liquid handling capacity at the expense of vapor capacity for a given diameter column and tray spacing. Cost is apparently no greater – at least, for tower diameters <8 ft.
4. Downcomers and weirs			
(a) Allowable downcomer inlet vel ft/s of clear liq		0.3–0.4	Lower value recommended for absorbers or other systems of known high frothiness
(b) Type of downcomer	Chord	Chord, arc	Min. chord length should be 65 % of tray diameter for good liquid distribution. Sloped downcomers can be used for high liquid rates – with maximum outlet velocity = 0.6 ft/s. Arc downcomers may be used alternatively to give more bubble area (and higher capacity) but are somewhat more expensive. Min. width should be 6 in. for the latter
(c) Inboard DC width (inlet and outlet)		Min. 8 in.	Use of a 14–16" "jump baffle" suspended lengthwise in the center of the inboard downcomer and extending the length of the downcomer is suggested to prevent possible bridging over by froth entering the downcomer from opposite sides. Elevation of base of jump baffle should be level with outlet weirs. Internal access way must be provided to allow passage from one side to another during inspection
(d) Outlet weir height	2"	1–4"	Weir height can be varied with liquid rate to give a total liquid head on the tray (hc) in the range of 2.5–4" whenever possible. Lower values suggested for vacuum towers, higher ones for long residence time applications
(e) Clearance under DC	1.5"	1" min	Set clearance to give head loss of approximately 1 in. Higher values can be used if necessary to assure sealing of downcomer

(continued)

Table 13 (continued)

Design feature	Suggested value	Alternate values	Comment
(f) DC seal (inlet or outlet weir height minus DC clearance)	Use outlet weir to give min. 1/2" seal in tray liquid	Inlet weir or recessed inlet box	In most cases tray liquid level can be made high enough to seal the downcomer through use of outlet weir only. Inlet weirs add to downcomer buildup; in some cases they may be desirable for 2-pass trays to ensure equal liquid distribution. Recessed inlets are more expensive but may be necessary in cases where an operating seal would require an excessively high outlet weir
(g) DC filling (% of tray spacing)		40-50	Use the lower value for high pressure towers, absorbers, vacuum towers, known foaming systems, and also for tray spacing of 18" or lower

Clear Liquid Height, h_{cl}

$$h_{cl} = 0.5 \times [V_L \div (N_p \times l_o)]^{2/3} \tag{7}$$

where:

- h_{cl} = height of clear liquid on tray in inches of hot liquid
- V_L = liquid loading in gallons per minute of hot liquid
- N_p = number of passes on tray
- l_o = length of outlet weir in inches

Effective Dry Tray Pressure Drop, h_{cd} The effective dry tray ΔP shall be the greater of the following two expressions:

(a)
$$\Delta P_{po} = 1.35 t_m \cdot \rho_m / \rho_1 + K \cdot (V_o^2) \cdot \rho_v / \rho_1 \tag{8}$$

(b)
$$\Delta P_{fo} = K_2 (V_o^2) \rho_v / \rho_1 \tag{9}$$

where:

- ΔP_{po} = dry tray ΔP , valve partially open. In inches of hot liquid
- ΔP_{fo} = dry tray ΔP , valve fully open. In inches of hot liquid
- t_m = valve thickness in inches (see Table 14)
- ρ_m = valve metal density in lb/cuft (see Table 15)
- V_o = vapor velocity through valves in ft/s (= cuft/s/ A_o)

Table 14 Valve thickness in inches

Gage	t_m
20	0.037
18	0.050
16	0.060
14	0.074
12	0.104
10	0.134
8	0.250

Table 15 Valve metal densities (in lb/cuft)

Metal	Density, ρ_m
C.S.	480
S.S.	510
Nickel	553
Monel	550
Titanium	283
Hastelloy	560
Aluminum	168
Copper	560
Lead	708

Table 16 Values of K_1 and K_2

Type of unit	K_1	K_2			
Deck thickness ins		0.074	0.104	0.134	0.25
Normal valves	0.2	1.05	0.92	0.82	0.58
Vacuum valves	0.1	0.50	0.39	0.38	–

Constants K_1 and K_2 are given in Table 16.

In calculating V_o , assume ratio hole to bubble area (A_o/A_B) is 12 %.

Total Tray, $\Delta P \cdot ht$

$$\text{Total tray } \Delta P = h_{cl} + h_{ed} \quad (10)$$

where:

h_{cl} = height of clear liquid on tray in inches of hot liquid

h_{ed} = effective dry pressure drop in inches of hot liquid

Head Loss Under Downcomer, h_{ud}

$$h_{ud} = 0.06 [\text{GPM} \div (cL_i - N_p)]^2 \quad (11)$$

where:

h_{ud} = head loss under downcomer in inches of hot liquid

c = constant = 1.5

L_i = length of inlet weir in inches

Inlet Head in Inches of Hot Liquid, h_i When there is an inlet weir, use:

$$h_i = 0.06[GPM \div (N_p \cdot L_i)]^{2/3} + h_{wi} \quad (12)$$

where:

h_i = tray inlet head in inches of hot liquid

L_i = length of inlet weir in inches

h_{wi} = height of inlet weir in inches

Where there is no inlet weir, then $h_i = h_{cl}$.

Downcomer Filling in Inches of Hot Liquid

$$L_d = h_i + (h_t + h_{ud}) \cdot [\rho_l \div (\rho_l - \rho_v)] + 1.0 \quad (13)$$

where:

L_d = downcomer filling in inches of hot liquid

Example Tower Sizing Calculations

The following is an example of tower sizing and hydraulic analysis. Tray loading data that is used for this example are those calculated for the debutanizer column earlier in this chapter. Thus, all data are at tray conditions of temperature and pressure (Table 17):

Bubble area on top tray at flood:

$$G_f = K\sqrt{(\rho_v \times (\rho_l - \rho_v))} \quad (14)$$

where:

G_f = load at Flood in lb/h ft²

$K = 1,110$

Table 17 Example tower sizing basis

Vapor	Top tray	Liquid	Top tray
Moles/h	1,012	lb/h	46,803
lb/h	57,297	GPM	187
ACFS	14.3	CFS	0.147
ρ_v lb/cuft	1.11	ρ_l lb/cuft	31.18
Vapor	Bottom tray	Liquid	Bottom tray
Moles/h	1,223.9	lb/h	239,860
lb/h	107,695	GPM	768.4
ACFS	8.7	CFS	1.715
ρ_v lb/cuft	3.44	ρ_l lb/cuft	38.9

$$\begin{aligned}\rho_v &= 1.11 \\ \rho_l &= 33.37 \\ G_f &= 6,415.8 \text{ lb/h ft}^2\end{aligned}$$

Tower diameter will be designed to 80 % of flood. Then $G_A = 5,132.6 \text{ lb/h ft}^2$
Bubble section A_B of tray at 80 % flood will be $\frac{57,297}{5,132.6} = 11.16 \text{ ft}^2$.

Downcomer area (inlet and outlet) A_{dc} :

Downcomer velocity will be 0.4 ft/s (see Table 13).

Then area of one downcomer will be $0.417 \div 0.4 = 1.04 \text{ ft}^2$.

Total downcomer area $A_{dc} = 2.08 \text{ ft}^2$.

Waste area of tray A_w will be 15 % (see Table 13).

Total tray area $A_s = A_B + A_{dc} + A_w$.

Then for the top section of the tower, the diameter will be:

$$\begin{aligned}A_s - 0.15A_s &= 2.08 + 11.6 \text{ ft}^2 \\ A_s &= 15.58 \text{ ft}^2 \text{ diameter} = 4.45 \text{ f. say } 4.5 \text{ ft}\end{aligned}$$

Similarly for the stripping side of the tower:

$$\begin{aligned}G_f &= 1,110\sqrt{3.44} \times (38.9 - 3.44) \\ &= 12,259 \text{ lb/h ft}^2 \text{ and at } 80 \% \text{ flood } G_a = 9,807 \text{ lb/h ft}^2\end{aligned}$$

Downcomer area = $1.715/0.4 = 4.29 \text{ ft}^2$ and $A_{dc} = 8.58 \text{ ft}^2$.
 $A_s = 24 \text{ ft}^2$ and diam is 5.53 ft, say 6 ft.

Tower Hydraulics and Downcomer Filling

Using the pressure drop equations defined earlier, the percentage of downcomer filled by liquid is calculated. This calculation is based on the stripping section of the tower only. A similar one will be completed for the rectifying section. Thus:

Clear Liquid Height, h_{cl}

$$h_{cl} = 0.5 \times [V_L \div (N_p \times L_o)]^{2/3} \quad (7)$$

$V_L = 768.4 \text{ GPM}$.

$N_p = 2$ (liquid loading is relatively high so the option of a two pass tray is used).

$L_o = 58.8 \text{ in}$. Use the correlation given in Appendix 1 of the chapter entitled

“► [Process Equipment for Petroleum Processing](#)” in this handbook.

Then $h_{cl} = 1.74 \text{ in}$. of hot liquid.

Effective Dry Tray, ΔP

$$(a) \quad \Delta P_{po} = 1.35t_m \cdot \rho_m / \rho_l + K_1 \cdot (V_o^2) \cdot \rho_v / \rho_l \quad (8)$$

(b)

$$\Delta P_{fo} = K_2(V_o)\rho_v/\rho_l \quad (9)$$

Use $A_o/A_B = 12\%$ giving A_o as 1.32 ft^2 .

Valve thickness is 0.05 in. , and metal density is 480 lb/cuft.

$$K_1 = 0.2, K_2 = 0.92$$

$$V_o = 6.6 \text{ ft/s}$$

$$\Delta P_{po} = 1.603 \text{ in. of hot liquid}$$

$$\Delta P_{fo} = 3.54 \text{ in. of hot liquid}$$

$$\text{Total Tray, } \Delta P \cdot h_t \quad h_t = 1.74 + 3.54 = 5.28 \text{ in. of hot liquid}$$

Head Loss Under Downcomer, h_{ud}

$$h_{ud} = 0.06[\text{GPM} \div (cL_i \cdot N_p)]^2$$

$$c = 3.2 \text{ (estimated)}$$

$$L_i = 58.8 \text{ in.}$$

$$h_{ud} = 0.96 \text{ in. of hot liquid}$$

Inlet Weir Head, h_i There is no inlet weir therefore $h_i = h_{cl}$.

Total downcomer filling, L_D

$$L_D = 1.74 + (5.28 + 0.96)(38.9/35.46) + 1$$

$$= 9.54 \text{ inches of hot liquid.}$$

$$= 40\% \text{ of tray spacing, which is satisfactory.}$$

Checks for Light Ends Tower Operation and Performance

Most light ends towers are very stable in their operation. That is, once they are lined out for an operating requirement under normal unit control, they maintain their stability. When performance falls off, it can be attributed to one of a few reasons. This section looks at some of these reasons and how they can be evaluated and checked. Performance in this case is meant the ability of the unit to make product quality at the prescribed throughput.

Cold Feed

The condition of the feed entering the tower is very important to the tower operation. Ideally the feed should enter the tower at as close to a calculated feed

tray temperature as possible. If the feed is well below its bubble point on entering the tower, several trays below the feed tray are taken up for heat transfer before effective mass transfer can begin. This could prevent the specified product separation occurring and tray efficiency in this section of the tower falls off dramatically. Feed condition can be checked by bubble point calculation and a flash calculation.

Hot Feed

This situation is probably the more serious regarding feed condition. If the feed enters at a temperature far above its bubble point, its resulting enthalpy will be such as to reduce the reboiler duty. This will occur automatically as the tower must always be in heat balance. The tower controls will maintain the product quantity and split. However, if the reboiler duty is drastically reduced, insufficient stripper vapors will be available for the stripping function. Poor separation will result.

As a rule of thumb, the stripping vapor to the bottom tray must be at least 70 % mole of the bottom product made. In super fractionation, such as a deisopentanizer, this figure would be at least 80–100 % of bottoms made.

Heat balances as shown will quickly determine the stripping vapor status.

Ideal Feed Condition

Ideally the feed should enter the tower close to the feed tray temperature. Usually then at the inlet pressure, the feed will be in a mixed phase with the vapor portion very close in quantity to the distillate product. As the feed to these units is generally heated by the bottom product heat exchange, the approach temperatures are always a consideration. To maintain good feed conditions, however it is often beneficial to include a separate steam (or hot oil) feed preheater.

Entrainment

A common cause of poor plant performance at high throughput or high reflux rates is liquid entrainment or carryover from tray to tray. Very often in a high load and entrainment situation, the problem is further aggravated by increasing reflux to attempt separation improvement.

A well-designed light ends tower can operate up to about 120 % of allowable flood before substantial carryover occurs. Loading above this figure would result in some degree of entrainment.

Downcomer Backup and Flooding

If tower loadings are increased well above allowable flood point, there is a real danger that downcomers become unable to cope with the liquid load. They would fill, and the tower would be in a state of flood. This will be very apparent with very

high, abnormal pressure drops occurring across the tower. Separation by fractionation is not possible under these conditions. Heat input (and feed rate) to the tower must be reduced to bring the unit back to a normal pressure drop.

Low Tower Loading

Most towers have been designed with at least a 50 % turndown ratio for the trays. This means that the trays should operate satisfactorily at 50 % of their loading. Nevertheless tray performance does fall off at these low loadings. Below this turndown ratio, performance, particularly in sieve trays, is drastically reduced. This is almost certain to be due to "weeping" where liquid falls from tray to tray. If the low loads are to be for only a short time due to temporary reduced throughput, tray loading can be increased by increasing reflux. If the low throughput is to continue for an extended period of time, a tray blanking schedule should be considered to reduce the active tray area.

Operating Close to Critical Conditions

Deethanizers in particular operate close to critical pressure at the bottom of the tower. Careful attention should be paid to avoid any pressure surges in this unit. Feed to the unit and reflux streams should be on flow control.

No separation by fractionation can occur at pressures in excess of critical. Very often chilled water is used for overhead condensing to reduce reflux drum pressure but maintaining minimum C_3 loss in the case of deethanizers.