

Design of Debutanizer Column to Increase the Production of Polyisobutene

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Abstract: This work presents the design of a debutanizer column to increase the production of polyisobutene. The polyisobutylene is manufactured from LPG. A debutanizer is used to remove the butane and propane present in the feed, which affects the viscosity of the desired product polyisobutylene. Separation of polymer and removal of unreacted butane and propane is more effective in our process. So, we are designing a debutanizer column by the FUG (Fenske Underwood Gilliland) method. This method is familiar to multicomponent distillation. We are understood the manufacturing process of polyisobutene, the equipment's needed and the process involved. Polyisobutene is used in medicine plaster, food packing and lubricating oil.

Keywords: Design, Distillation, Debutanizer, Polyisobutene.

1. Introduction

A debutanizer column is commonly used in all petroleum refinery industries. This works on the distillation column principle. The distillation process is used to separate components in a feed mixture based upon their relative boiling points. A simple, continuous column can make the separation between two components into two product streams. In multi-component systems, the two main components to be separated are designated as the light and heavy keys. The light key is the more volatile component in greater purity in the top product stream, and the heavy key is the less volatile component in greater purity in the bottom product stream. The normal procedure for solving a typical multi-component distillation problem is to solve the MESH (Material balance, Equilibrium, Summation and Heat) balance equations stage-by-stage. Though computer programs are normally available for the rigorous solution of the MESH equations, short-cut methods are still useful in the preliminary design work, and as an aid in defining problems for computer simulation. This article describes a widely used short-cut distillation method commonly referred to as the Fenske-Underwood-Gilliland (FUG) method.

2. Literature review

As Hans P. Rath, Herwig Hoffmann, Peter Reuter (2005) said, a process for the preparation of polyisobutylene from isobutene and/or hydrocarbons containing isobutene in the presence of a complex of BF₃ and an alcohol, wherein the free BF₃ is substantially removed from the complex solution and/or

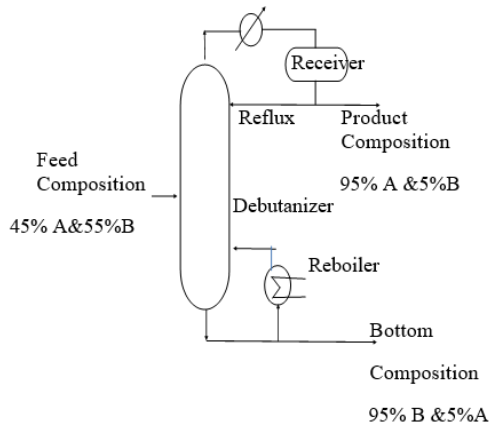
from the reaction chamber.

As L. Fortuna, S. Licitra, M. Sinatra, M. G. Xibilia said, this paper presents a neural approach to distillation column modelling is described. In particular, a debutanizer column is considered and a real-time estimate of the butane percentage (C₄) in the bottom draw (C₅) is obtained by a NARMAX model implemented with a Multi-Layer Perceptron. The analyser of the C₄ in C₅ percentage used at present, provides a measure after a great and unknown delay, and is therefore not suitable for closed loop control purposes. A neural-based model, acting as a virtual sensor, can therefore represent a suitable strategy in getting a real-time estimation of the C₄ in C₅ concentration. Neural networks are used both to evaluate the delay of the analyser and to provide the desired real-time estimate of the C₄ flow in the bottom draw of the debutanizer, overcoming the analyser's delay. To obtain more accurate results the model is built so that the measured output is used as an input of the model together with the predicted one, suitably delayed. The neural NARMAX model has been determined by using an appropriate set of measurements performed on a plant operating in Sicily (Italy) and is now working on the plant. A comparison between the estimated output and the analyser's measures confirms the validity of the proposed approach.

As V. R. Dholet and B. Linnhoff (1993) said, this paper relates to the design of distillation columns. In particular, it relates to the optimisation of a base case design. Column optimisation involves options such as different reflux ratios, pressures, side condensing/reboiling and feed preheating/cooling. We need to establish heat loads and temperature levels for such modifications and identify the best combined options. The paper presents a methodology based on a combination of thermodynamic and practical aspects of column modification. The methodology gives the engineer targets for design options ahead of design. In addition, it provides targets for the best combination of options. Alongside heat load and temperature targets, the methodology, clarifies the effect of design modifications on column capital cost, also ahead of design. The methodology is applicable to nonideal multicomponent systems and complex distillation configurations. There are downstream applications involving distillation column integration in the overall processes and column sequencing. The methodology has been tested in industrial applications involving ethylene,

LNG and refinery columns.

3. Design calculation of debutanizer column



A-High volatile mixture of butene and butane
 B-Less volatile mixture of high polymer and polybutene

Data:

mass flow rate of feed, $F=1963.13 \text{ kg/h}$
 Avg molar mass of feed $M_{avg}=(0.45*57)+(0.55*350)=218.15$

Molecular weight of butane and butene= 57 kg/mole

Molecular weight of light polymer and polybutene= 350 kg/mole

Molar flow rate of feed= $1963/218.15=8.99 \text{ kmol/h}$

mol% of A in feed $= (45/57) / \{ (45/57) + (55/350) \}$
 $X_f=0.83$

mol% of B in feed= 0.17

Distillate (top) contains 10% of A and 10% of B
 $= (95/57) / \{ (95/57) + (5/350) \}$
 $X_D=0.99$

Bottom contains 10% of A and 90% of B
 $X_B=0.01$

$D = F(X_f - X_B) / (X_D - X_B)$
 $= 8.99 \{ (0.83 - 0.01) / (0.99 - 0.01) \}$

$D = 6.05 \text{ kmol/h}$

$F = B + D$

$B = F - D$

$F X_f = B X_B + D X_D$

$B = F - D = 8.99 - 6.05$

$= 2.94 \text{ kmol/h}$

Relative volatility

$\alpha_{AB} = k_A / k_B = (P_A / P_B) (P_B / P_i)$

where,

P_A, P_B are vapour pressure of pure A and B

P_{in} -total pressure

vapour pressure is calculated from the poling correlation

For polybutene	for butene
At $210^\circ\text{C} = 9445.51 \text{ Pa}$	25184 Pa
At $100^\circ\text{C} = 406 \text{ Pa}$	6124.9 Pa
At $60^\circ\text{C} = 179 \text{ Pa}$	6638.2 Pa

Total pressure

At $210^\circ\text{C} = 11016 \text{ Pa}$

At $100^\circ\text{C} = 406 \text{ Pa}$

At $60^\circ\text{C} = 179 \text{ Pa}$

At 210°C

$\alpha_{AB} = 25184 / 9445 = 2.66$

Relative volatility $\alpha_{AB} =$

$R_{\text{min}} = 1 / (\alpha - 1) * ((X_D / X_F) - \alpha * (1 - X_F) / (1 - X_F))$

$R_{\text{min}} = 1 / 1.66 * ((0.99 / 0.01) - 2.66 * (0.01 / 0.17))$

$R_{\text{min}} = 0.624$

Minimum no. stages by fensky equation

$N_m = \log(X_D / (1 - X_D)) (1 - X_w / X_w) / \log \alpha_{av}$

where

$\alpha_{av} =$ average relative volatility for light key component

$\alpha_{av} = (\alpha_{\text{top}} + \alpha_{\text{bottom}})^{1/2} = (0.02 + 2.66)^{1/2} = 1.6$

$N_m = \log((0.99 / (1 - 0.99)) * (1 - 0.01 / 0.01)) / \log(1.6)$

Minimum no of stages (or) trays, $N_m = 19.55 \sim 20$

Number of theoretical stages by Gilland's correlation developed by Malokonav

$f(N) = N - N_m / N$

$f(N) = 1 - \exp$

$\{ (1 + 54.4 \Psi) / (11 + 117.2 \Psi) (\Psi - 1) / (\Psi^{0.05}) \}$

where

$\Psi = R - R_m / R + 1$

From the above equation,

$R = 3.5$

$\psi = (3.5 - 0.624) / 4.5$

$\Psi = 0.639$

$N - N_m / N = 1 - \exp \{ (1 + 54.4 * 0.639) / (11 + 117.2 * 0.639) ((0.639 - 1) / (0.639^{0.05})) \}$

$N = 24.38 \sim 24$

No. of theoretical stages $N = 24$

Molar flow rates of vapour and liquid the enriching section

$L = RD = 3.5 * 6.05 = 21.17 \text{ kmol/h}$

$V = (R + 1)D = (3.5 + 1) * (6.05)$

$= 27.22 \text{ kmol/h}$

Molar flow rates of vapour and liquid the stripping section

$L_s = L + Fq = 21.17 + (8.99 * 1.2)$

$= 31.95 \text{ kmol/h}$

$$V_s = F(q-1) + V = 8.99(1.2-1) + 27.22 = 29.01 \text{ kmol/h}$$

Using mass balance

$$L_s X_1 = V_s Y_w + B X_w$$

$$X_B = 0.01$$

$$Y_w = \alpha X_w / (1 + (\alpha - 1) X_w)$$

$$Y_w = (2.66 * 0.01) / (1 + (1.66 * 0.01))$$

$$Y_w = 0.0261$$

Tray 1:

$$X_1 = [(29.01 * 0.0261) + (2.94 * 0.01)] / 3$$

$$X_1 = 0.0246$$

$$Y_1 = (2.66 * 0.0246) / (1 + (1.66 * 0.0246))$$

$$Y_1 = 0.0624$$

TRAY 2:

$$X_2 = V_s Y_1 + B X_1 / L_s$$

$$X_2 = [(29.01 * 0.0627) + (2.94 * 0.01)] / 31.95$$

$$X_2 = 0.0591$$

$$Y_2 = (2.66 * 0.0591) / (1 + (1.66 * 0.0591))$$

$$Y_2 = 0.1431$$

$$\text{Efficiency } \eta = \frac{Y_n - Y_{n+1}}{Y_{a^*} - Y_{a+1}} * 100$$

$$\eta = (0.1431 - 0.0627) / (0.1681 - 0.0246)$$

$$\eta = 0.56 * 100$$

$$\eta = 56\%$$

Feed tray location by Kirkbride equation

$$\log \{N_r / N_s\} = 0.206 \log [W/D (X_r / 1 - X_r) (X_b / X_b)^2]$$

where,

N_r = number of stages above the feed

N_s = number of stages below the feed

$$\log \{N_r / N_s\} = 0.206 \log [2.94 / 6.05 (0.55 / 0.45) (0.05 / 0.05)]$$

$$= -0.0466$$

$$N_r = 0.8982 N_s$$

$$N_r + N_s = 24$$

$$0.8982 N_s + N_s = 24$$

$$N_s = 12$$

$$N_r = 12$$

Actual no. of trays = no. of theoretical trays / tray efficiency

$$= 24 / 0.56$$

Actual no. of tray = 42

Tower diameter at enriched section

Pressure at top = 5992.28 Pa

Temperature = 60°C

$V = 27.22 \text{ kmol/h}$

$L = 21.17 \text{ kmol/h}$

Density of vapour, $\rho_v = P M_{avg} / RT$

$$M_{avg} = (0.99 * 57) + (0.01 * 350)$$

$$= 59.93$$

$$R = 8.314 \text{ kPa m}^3 / \text{kmol K}$$

$$\text{Density of vapour, } \rho_v = (5.99228 * 59.93) / (8.314 * 333)$$

$$= 0.1297 \text{ kg/m}^3$$

Density of liquid, $\rho_L = 1 / (X/\rho)$

$$= 1 / (0.99 / 273.76) + (0.01 / 714)$$

$$= 275.45 \text{ kg/m}^3$$

Liquid vapour flow factor at top,

$$F_{LV} = (L/V) (\rho_v / \rho_L)^{0.5}$$

$$= (21.17 / 27.22) (0.129 / 275.45)^{0.5}$$

$$F_{LV} = 0.0168$$

Flooding velocity,

$$V_f = K_1 ((\rho_L - \rho_v) / \rho_v)^{1/2}$$

K_1 = Coefficient obtained from K_1 vs F_{LV}

Assuming the plate spacing = 0.5m

$$K_1 = 0.09$$

$$V_f = 0.09 ((275.45 - 0.129) / 0.129)^{1/2}$$

$$V_f = 4.15$$

Actual velocity, $v = 0.85 V_f$

$$= 0.85 * 4.15$$

$$v = 3.52 \text{ m/s}$$

volumetric flow rate of vapour at top,

$$Q = V M_{avg} / \rho_v$$

$$= 27.22 * 59.93 / 0.129$$

$$= 12645.69 / 3600$$

$$Q = 3.51 \text{ m}^3/\text{s}$$

Net area required at top,

$$A_n = Q / v$$

$$= 3.51 / 3.52$$

$$A_n = 0.997 \text{ m}^2$$

$$A_d = 0.12 A_c$$

$$A_n = A_c - A_d$$

$$0.997 = A_c - 0.12 A_c$$

Cross sectional area, $A_c = 1.13 \text{ m}^2$

Down corner area, $A_d = 0.12 * 1.13$

$$= 0.1356 \text{ m}^2$$

Active area, $A_a = 2 * A_d$

$$= 2 * 0.1356$$

$$A_a = 0.2712 \text{ m}^2$$

Inside diameter of column at top,

$$D_i = [(4 * A_c) / \pi]^{1/2}$$

$$= [(4 * 1.13) / \pi]^{1/2}$$

$$D_i = 1.19 \text{ m}$$

Weir length = $0.77 * 1.19 = 0.9163 \text{ m}$

Weir height = 50 mm = 0.05 m

Tower Diameter At Stripping Section:

Pressure at top = 11016 Pa

Temperature = 210°C

$V_s = 29.01 \text{ kmol/h}$

$L_s = 31.95 \text{ kmol/h}$

Density of vapour, $\rho_v = P M_{avg} / RT$

$$M_{avg} = (0.99 * 350) + (0.01 * 57)$$

$$= 347.07$$

$$R = 8.314 \text{ kPa m}^3 / \text{kmol K}$$

$$\text{Density of vapour, } \rho_v = (11.016 * 347.07) / (8.314 * 483)$$

$$= 0.952 \text{ kg/m}^3$$

Density of liquid, $\rho_L = 1/(X/\rho)$
 $= 1/(0.99/714) + (0.01/273.76)$
 $= 702.69 \text{ kg/m}^3$

Liquid vapour flow factor at bottom,
 $F_{LV} = (L_s/V_s)(\rho_v/\rho_L)^0$
 $= (31.95/29.01)(0.952/702.69)^{0.5}$
 $F_{LV} = 0.04053$

Flooding velocity,
 $V_f = K_1((\rho_L - \rho_v)/\rho_v)^{1/2}$
 $K_1 = \text{Coefficient obtained from } K_1 \text{ vs } F_{LV}$
 Assuming the plate spacing = 0.5m
 $K_1 = 0.06$
 $V_f = 0.06((702.69 - 0.952/0.952)^{1/2})$
 $V_f = 1.62$

Actual velocity, $v = 0.85V_f$
 $= 0.85 * 1.62$
 $v = 1.37 \text{ m/s}$

volumetric flow rate of vapour at top,
 $Q = VM_{avg}/\rho_v$
 $= 29.01 * 347.07/0.952$
 $= 10576.15/3600$
 $Q = 2.93 \text{ m}^3/\text{s}$

Net area required at bottom,
 $A_n = Q/v$
 $= 2.93/1.37$
 $A_n = 2.14 \text{ m}^2$
 $A_d = 0.12A_c$
 $A_n = A_c - A_d$
 $2.14 = A_c - 0.12A_c$

Cross sectional area, $A_c = 2.43 \text{ m}^2$
 Down corner area, $A_d = 0.12 * 2.43$
 $= 0.29 \text{ m}^2$
 Active area, $A_a = 2 * A_d$
 $= 2 * 0.29$
 $A_a = 0.58 \text{ m}^2$

Inside diameter of column at bottom,
 $D_i = [(4 * A_c)/\pi]^{1/2}$
 $= [(4 * 2.43)/\pi]^{1/2}$
 $D_i = 1.19 \text{ m}$
 Weir length = $0.77 * 1.75 = 1.35 \text{ m}$
 Weir height = $50 \text{ mm} = 0.05 \text{ m}$
 Height of column = (no. of trays - 1) * (plate spacing + top spacing + bottom spacing)
 $= (42 - 1) * (0.45 + 0.5 + 0.5)$
 $= 59.45 \text{ m}$

Tray pressure drop for enriching section
 $h_d = 51(v_h/c_o)^2 * (\rho_v/\rho_L)$
 $v_h = Q/A_h$
 $= 3.51/0.0271$
 $v_h = 129.52 \text{ m/s}$
 $h_d = 51(129.52/0.842)^2 * (0.1297/275.45)$
 $= 568.2 \text{ mmLC}$

$h_w = 50 \text{ mmLC (to be considered)}$
 $h_{ow} = 750(L_w/\rho_w * l_w)^{2/3}$
 $L_w = 0.7 * L * M_{avg}$
 $= 0.7 * 21.17 * 59.93$
 $L_w = 0.24 \text{ kg/s}$
 $h_{ow} = 750(0.24/275.45 * 0.9163)^{2/3}$
 $= 7.25 \text{ mmLC}$
 Residual pressure drop,
 $h_r = (12.5 * 10^3)/\rho_L$
 $= 12.5 * 10^3/275.45$
 $= 45.38 \text{ mmLC}$
 Total pressure drop,
 $h_t = h_d + h_{ow} + h_w + h_r$
 $= 568.2 + 7.25 + 50 + 45.38$
 $h_t = 670.83 \text{ mmLC}$

Tray pressure drop for stripping section
 $h_d = 51(v_h/c_o)^2 * (\rho_v/\rho_L)$
 $v_h = Q/A_h$
 $= 2.93/0.058$
 $v_h = 50.51 \text{ m/s}$
 $h_d = 51(50.51/0.842)^2 * (0.952/702.69)$
 $= 248.64 \text{ mmLC}$

$h_w = 50 \text{ mmLC (to be considered)}$
 $h_{ow} = 750(L_w/\rho_w * l_w)^{2/3}$
 $L_w = 0.7 * L * M_{avg}$
 $= 0.7 * 31.95 * 347$
 $L_w = 2.15 \text{ kg/s}$
 $h_{ow} = 750(2.15/702.69 * 1.35)^{2/3}$
 $= 12.94 \text{ mmLC}$
 Residual pressure drop,
 $h_r = (12.5 * 10^3)/\rho_L$
 $= 12.5 * 10^3/702.69$
 $= 17.78 \text{ mmLC}$
 Total pressure drop,
 $h_t = h_d + h_{ow} + h_w + h_r$
 $= 248.64 + 12.94 + 50 + 17.78$
 $h_t = 329.36 \text{ mmLC}$

Column pressure drop
 Pressure drop at top most tray
 $= 670.83 \text{ mmLC}$
 Pressure drop at bottom most tray = 329.36 mmLC
 Assume tray pressure drop for entire column = 500.09 mmLC
 $\Delta P_T = \text{Number of trays} * \text{pressure drop}$
 $= 42 * 500.09$
 $= 21003.78 \text{ mmLC}$
 $\Delta P_T = 21003.78 * 10^{-3} * 9.81$
 $= 206.04 \text{ N/m}^2$
 $\Delta P_T = 206.04 \text{ Pa}$

4. Design summary

STRIPPING SECTION
Plate spacing=0.45m
Down corner area, $A_d=0.29m^2$
Cross sectional area, $A_c=2.43m^2$
Active area=2.14m ²
Weir height=0.05m
Weir length=1.13m
Net area, $A_n=2.14m^2$
Inside diameter, $D_i=1.75m$

ENRICHING SECTION
Plate spacing=0.45m
Down corner area, $A_d=0.1356 m^2$
Cross sectional area, $A_c=1.13 m^2$
Active area=0.2712 m ²
Weir height=0.05m
Weir length=0.9163m
Net area, $A_n=0.997 m^2$
Inside diameter, $D_i=1.19m$
Minimum no. of trays =20
No. of theoretical trays=24
No. of actual trays =42
Feed tray location =12
Efficiency =56%
Total pressure drop =206.04Pa

5. Conclusion

Although debutanizer columns are not rare in industry, shortcut methods were available for their design. Whenever such columns were designed by hand, completely arbitrary values were assigned to the number of trays required, which could be far from the correct or optimal values. The equations we presented here are simple to use and provide the design engineer with accurate estimations of the vapor rate, the number of trays, provided that the mixture exhibits a nearly ideal behavior. This design method will also allow debutanizer columns in the synthesis of liquid separation systems and to compare designs with traditional sequences of simple columns.

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