# UNIVERSITY OF CAPE TOWN

# DEPARTMENT OF CHEMICAL ENGINEERING

## SEPARATION PROCESSES
## CHE343S

## TERM PROJECT

# THE DESIGN OF A DISTILLATION COLUMN

Date:        18 September 1998

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**Note:** References to material in the bibliography are indicated in square brackets: [XX, p yyy], where XX is the reference number in the bibliography and yyy is the page number in the reference.

# 1. INTRODUCTION

It is required to design a distillation column to separate a benzene–toluene mixture. The design specifications are shown in Table 1 below. The design approach was to specify the column operating conditions and physical column attributes (plate design and column length and diameter) that would be an optimum in terms of equipment and operating costs.

| Feed rate, F | 10 000 kg/hr feed ≡ 116.69 kmol/hr |
|---|---|
| Feed composition | Benzene 45 mol% <br> Toluene 55 mol% |
| Feed temperature | Bubble Point (93.6°C – see Appendix A) |
| Column operating pressure | Atmospheric |
| Column condenser | Total condenser |
| Column reboiler | Partial reboiler |
| Distillate composition, $x_D$ | 0.95 mol% benzene |
| Bottoms composition, $x_W$ | 0.95 mol% toluene |

*Table 1*: Specified column parameters

# 2. DESIGN CALCULATIONS AND ASSUMPTIONS

## 2.1 Design Procedure

The approach taken to design the column optimally was based on the methodology of [3, p 371 to 376] where various reflux ratios are selected and the appropriate calculations are performed at these ratios. A table of results is then drawn up and the column that appears to satisfy the costs most optimally is selected. The calculation method followed the different reflux ratios design procedure followed is outlined below and is adapted from [4, p 520]:

The design calculations are detailed in Appendix C.

1. Obtain the Antoine equation data to determine the Vapour–Liquid Equilibrium curve – see Appendix A.

2. Obtain the physical data of benzene and toluene required for the design – see Appendix B.

3. Calculate the flowrates of various streams inside and outside the column

4. Calculate the minimum reflux ratio and the minimum number of trays required

5. Using the physical data and flowrates calculate the reboiler and condenser duties

6. Calculate the maximum and minimum liquid and vapour flowrates

7. To start the iteration, select a reasonable plate spacing and using the trial plate spacing calculate the column diameter

8. Determine the liquid flow pattern (reverse, single or multiple pass)

9. Select a trial plate layout, that is, select downcomer area, active area hole area and size, weir height and length

10. From this data check that the weeping rate is satisfactory

11. Calculate the plate pressure drop

12. Check that the downcomer area backup is acceptable

13. If at any stage some of the values are too high or low select new trial values and repeat the iterations above

14. Draw up a table of results including the equipment and operating costs

## 2.2 *Assumptions made during the design*

Since we used the McCabe–Thiele method of column design [6; 7, p 402 to 419], the following assumptions are inherent in our calculations:

- The vapour and liquid flowrate from tray to tray do not change in a given section of the tower

- The latent heat of evaporation is approximately constant with composition and also does not vary much as we proceed from tray to tray

- The system is non–foaming and non corrosive, and thus we can use carbon steel rather than stainless steel as our material of construction.

## 3. RESULTS OF CALCULATIONS

The results from the calculations appear in Appendix C. All the designs satisfy the requirements for satisfactory column operation, and as such we will select the column which is most cost effective, that is the lowest reboiler and condenser duty. This is a column which has a reflux ratio $R = 1.3$ $R_{min} = 1.65$. The details of the column are:

### Column Parameters

| Parameter | Size |
|---|---|
| Column diameter | 1.25 m |
| Column Height | 10 m |
| Condenser duty | 1175.4 kW |
| Reboiler duty | 1858 kW |
| Number of actual trays *Theoretical* | 19  7  (14) |
| Feed Entry Tray | 10 |
| Pressure drop per plate | 1 020 Pa |
| Pressure drop over the column | 20 000 Pa |
| Percentage Flooding | 78.2% |

### Tray Parameters

| Parameter | Size |
|---|---|
| Material of construction | Carbon steel |
| Column area | 1.227 m$^2$ |
| Downcomer area (12% of column area) | 0.147 m$^2$ |
| Active area | 0.933 m$^2$ |
| Total hole area | 0.0933m$^2$ |
| Hole diameter | 5 mm |
| Unperforated zone around plate | 75 mm |
| Calming zone at each downcomer | 75 mm |
| Number of holes | 4 700 |
| Hole Pitch | Triangular |
| Plate Spacing | 0.5 m |
| Plate Thickness | 5 mm |

A view of the plate is as follows:
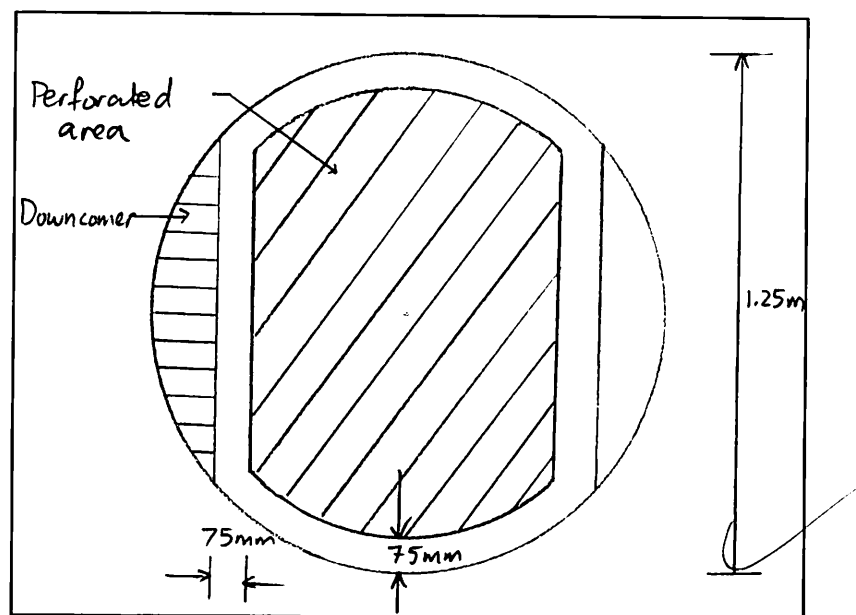


Figure 1:     A view of a typical column plate

## 4.    DISCUSSION

A design for the column that meets the specified criteria was designed and its characteristics appear in Section 3. The following should be noted in terms of the design:

The equipment and installation costs were calculated to be:

| Item | Cost (1998) | |
|---|---|---|
| Column Pipe | R | 155 000 |
| Plates | | 215 000 |
| Boiler | | 266 000 |
| Cooling Unit | | 70 000 |
| Piping and Instrumentation | | 70 000 |
| 10% Contingency | | 77 000 |
| **TOTAL** | **R** | **855 000** |

The operating costs for the reboiler and the condenser were estimated using [8]:

| Item | Cost per year |
|---|---|
| Steam | R     400 000 |
| Cooling Water | 2 050 000 |
| 5% Contingency | 122 500 |
| **TOTAL** | **R     2 573 000** |

Labor, maintenance, etc

The costing procedure is detailed in Appendix D.

Calculations for weeping, flooding and entrainment were made in Appendix C, and as can be seen the column operates at 78% away from flooding. The velocity through the holes exceeds the minimum velocity required to prevent weeping.

There is minimal entrainment (0.043 kg vapour/kg liquid flowrate) and the overall column efficiency was estimated to be 73%.

There is a pressure drop of 1020 Pa per plate and 20 kPa over the entire column.

The downcomers have a liquid backup (211 mm) that is less than half the plate spacing (500mm) and weir height (50mm) added, which will also ensure that flooding does not occur.

There is a residence time of 4.8 s in the downcomers which is sufficient time to allow the vapour and liquid to disengage.

Conclusions?

# 5. REFERENCES

1. Henley, EJ & Seader, JD "Equilibrium–Stage Separation Operations in Chemical Engineering", John Wiley, Chapter 13.

2. Himmelblau, DM, "Basic Principles and Calculations in Chemical Engineering", 6$^{th}$ Edition, Prentice Hall Series, CD ROM included in Textbook, File:\\PhysProp\Data.Exe, 1996.

3. Peters, MS & Timmerhaus KD, "Plant Design and Economics for Chemical Engineers", 4$^{th}$ Edition, McGraw–Hill, 1991, p 371 to 376.

4. Sinnott, RK, "Coulson and Richardon's Chemical Engineering" Revised 2$^{nd}$ Edition, Butterworth Heinemann, 1997, Chapter 6 and 11.

5. Smith, BD, "Design of Equilibrium Stage Processes", McGraw–Hill, 1963, Chapters 14 and 15.

6. Swartz, CLE "Lecture Notes for CHE343S" University of Cape Tow, 1998.

7. Treybal, RE, "Mass–Transfer Operations", 3$^{rd}$ Edition, McGraw–Hill, 1981, p 342 to 425.

8. Ulrich, GD, "How to Calculate Utility Costs" from "Calculations and Shortcut Deskbook", selected articles from *Chemical Engineering*, McGraw–Hill, p 130–133.

9. Ward, TJ, "A New Correlation for Sieve Trays" from "Calculation and Shortcut Deskbook", selected articles from *Chemical Engineering*, McGraw–Hill, p 57–58.

# APPENDIX A: ANTOINE DATA AND VAPOUR LIQUID EQUILIBRIUM CALCULATIONS

The Antoine data was obtained from [2] and are for the equation of the form:

$$\log(P^o_{vap}) = A - \frac{B}{T+C} \quad \text{with T in °C and } P^o_{vap} \text{ in mmHg} \tag{1}$$

The constants for this equation are:

| Compound | A | B | C |
|----------|-----|-----|-----|
| Benzene | 6.90565 | 1211.033 | 220.79 |
| Toluene | 6.95334 | 1343.943 | 219.377 |

Using this data the boiling points of the two components and the bubble point of the feed were determined to be:

$$T^{bp}_{Benzene} = 80.100 \text{ °C}; \quad T^{bp}_{Toluene} = 110.626 \text{ °C} \quad \text{and} \quad T^{bp}_{Feed} = 93.60 \text{ °C}$$

The following table of values are some of the data used for the vapour liquid equilibrium curve:

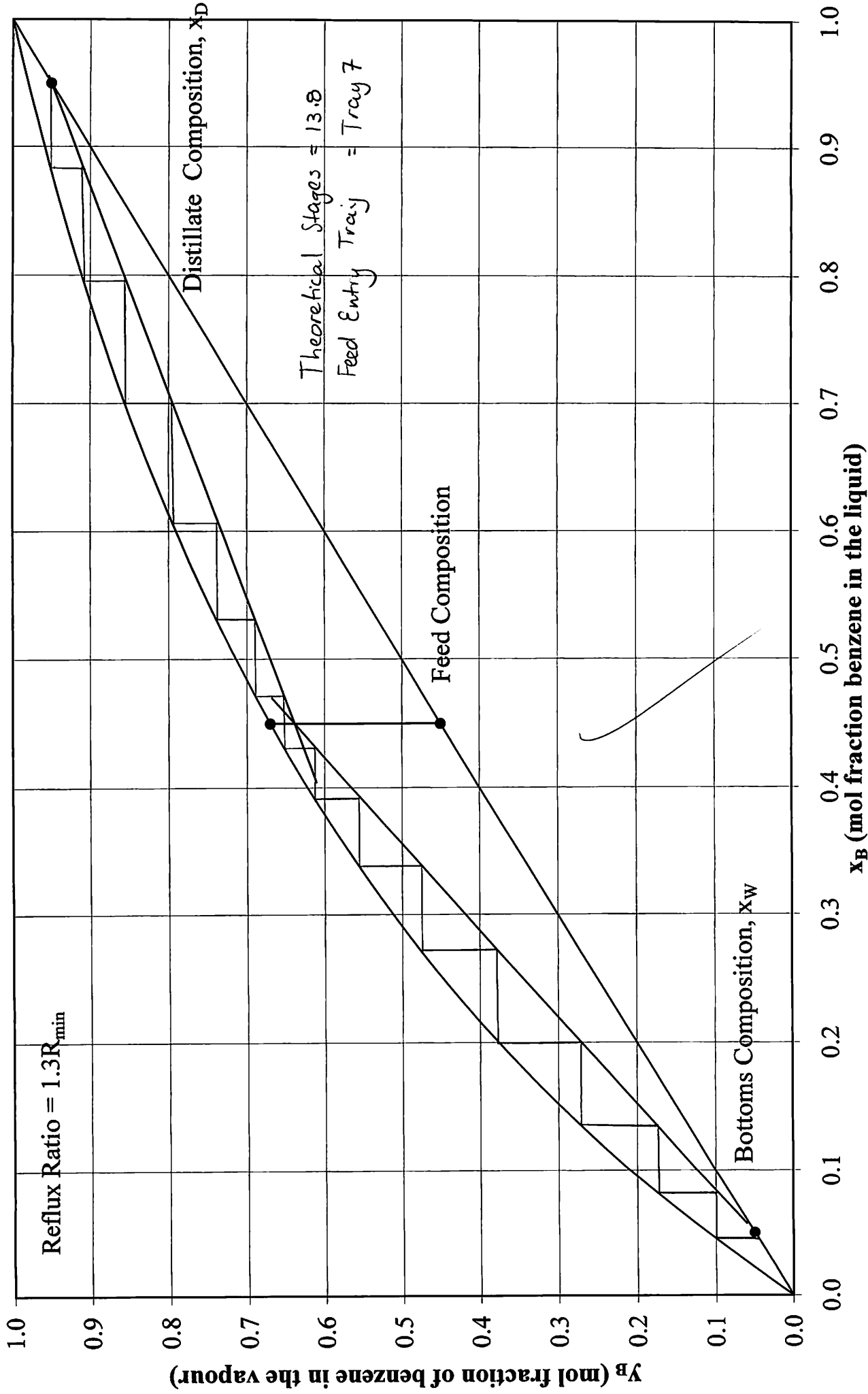| Temperature (°C) | $P^o_B$ | $P^o_T$ | $x_B$ | $y_B$ |
|----------|-----|-----|-----|-----|
| 80.1 | 760.0 | 292.2 | 1.000 | 1.000 |
| 82 | 805.5 | 311.9 | 0.908 | 0.962 |
| 84 | 855.7 | 333.7 | 0.817 | 0.919 |
| 86 | 908.3 | 356.8 | 0.731 | 0.874 |
| 88 | 963.3 | 381.1 | 0.651 | 0.825 |
| 90 | 1021.0 | 406.7 | 0.575 | 0.773 |
| 92 | 1081.3 | 433.7 | 0.504 | 0.717 |
| 94 | 1144.3 | 462.1 | 0.437 | 0.657 |
| 96 | 1210.1 | 492.0 | 0.373 | 0.594 |
| 98 | 1278.8 | 523.4 | 0.313 | 0.527 |
| 100 | 1350.5 | 556.3 | 0.256 | 0.456 |
| 102 | 1425.2 | 590.9 | 0.203 | 0.380 |
| 104 | 1503.1 | 627.2 | 0.152 | 0.300 |
| 106 | 1584.2 | 665.2 | 0.103 | 0.215 |
| 108 | 1668.6 | 704.9 | 0.057 | 0.125 |
| 110 | 1756.4 | 746.6 | 0.013 | 0.031 |
| 110.6255 | 1784.5 | 760.0 | 0.000 | 0.000 |

*Table 1*: Vapour–Liquid Equilibrium Data from the Antoine Equation

From this data it was also determined that the average relative volatility, $\alpha_{ave}$ = 2.467.

The data in Table 1 is sketched on the following page.

# Vapour Liquid Equilibrium Data for the Benzene/Toluene System



Distillate Composition, $x_D$

Theoretical Stages = 13.8
Feed Entry Tray = Tray 7

Feed Composition

Reflux Ratio = 1.3$R_{min}$

Bottoms Composition, $x_W$

$x_B$ (mol fraction benzene in the liquid)

$y_B$ (mol fraction of benzene in the vapour)

# APPENDIX B:    PHYSICAL DATA FOR BENZENE AND TOLUENE

The physical data obtained is from various sources as noted:

## Molar Masses

for Benzene:        78.114 kg/kmol ≈ 78 kg/kmol

for Toluene:        92.141 kg/kmol ≈ 92 kg/kmol

*Viscosity*:    $\log(1000\mu) = \text{VisA}\left[\dfrac{1}{T} - \dfrac{1}{\text{VisB}}\right]$, with T in K and $\mu$ in SI Units (Pa.s)

for Benzene:  VisA = 545.64        VisB = 265.34        $\mu$ave = 0.26 cP

for Toluene:  VisA = 467.33        VisB = 255.24        $\mu$ave = 0.27 cP

data from [4, p 870]

*Density in Liquids*:    $\rho = \text{A.B}^{-\left(1-\frac{T}{T_c}\right)^{2/7}}$    where T in K; $T_C$ is the critical temperature and $\rho$ in g/m$\ell$

for Benzene:  A = 0.3041        B = 0.269        $T_C$ = 562 K

for Toluene:  A = 0.2906        B = 0.265        $T_C$ = 593.1 K

From this data we can calculate that the density does not vary much over the entire column (between 790 and 810 kg/m$^3$), thus:   $\rho_{ave}$ = 0.8 g/m$\ell$ ≡ 800 kg/m$^3$

data from [2]

*Density in the Vapour:*        In the absence of data we can use the ideal gas law:

At the top of the column:     $\rho_{V,top}$ = 2.716 kg/m$^3$

At the bottom of the column: $\rho_{V,bot}$ = 2.905 kg/m$^3$

## Enthalpies:

for Benzene:  $C_{P,Top,Gas}$ = 99.04 J/mol.K        $C_{P,Bottom,Liquid}$ = 158.84 J/mol.K

for Toluene:  $C_{P,Top,Gas}$ = 124.58 J/mol.K        $C_{P,Bottom,Liquid}$ = 175.19 J/mol.K

data from [2]

### Heats of Vapourisation:

for Benzene:  $\Delta H_{vap,81°C} = 30.72$ kJ/kmol    $\Delta H_{vap,110°C} = 29.02$ kJ/kmol

for Toluene:  $\Delta H_{vap,81°C} = 34.90$ kJ/kmol    $\Delta H_{vap,110°C} = 33.22$ kJ/kmol

data from [2]

### Surface Tensions:

for Benzene and Toluene the surface tensions are approximately = 20 dyne/cm = 20 mJ/m$^2$
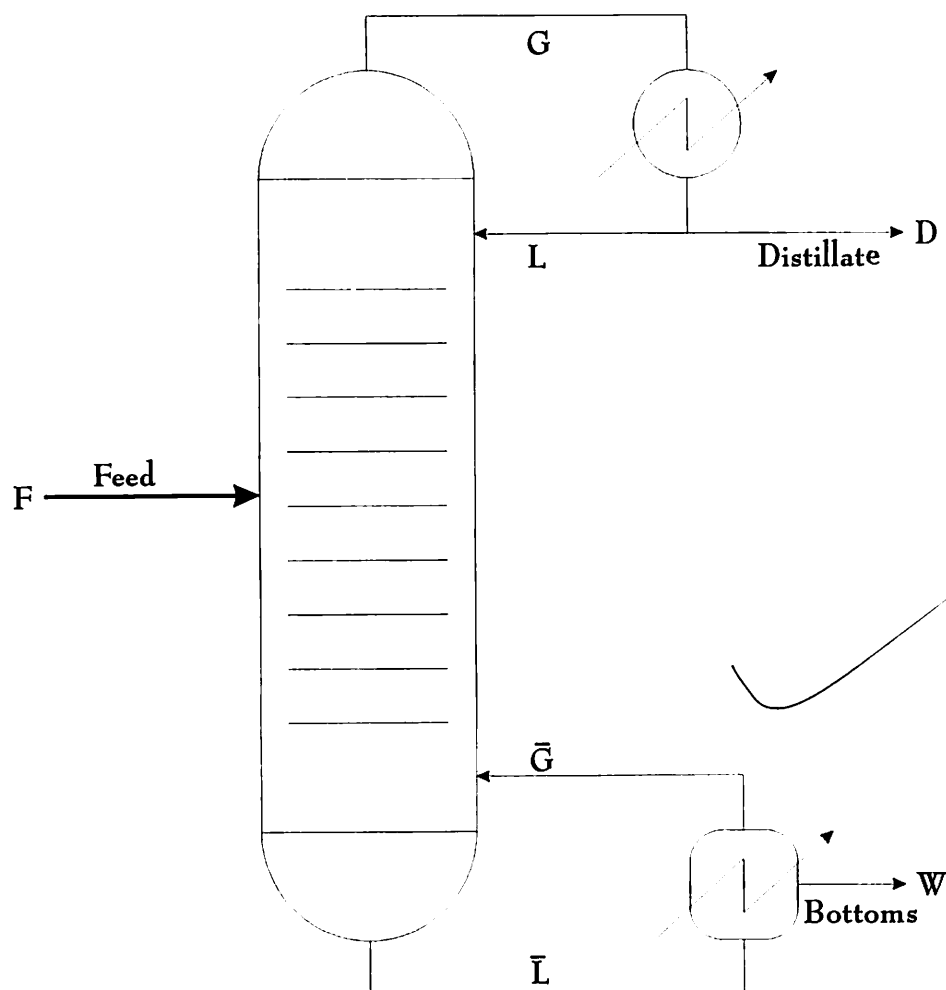
data from [5]

### Heat Capacity of Water

$C_{p,water} = 4.190$ kJ/kg.K

data from [2].

The method followed is detailed in the main report, section 2. Some of the notation used in this Appendix is illustrated on the diagram below:



*Flowrates of Various Streams Inside and Outside the Column*

The following flowrates were given:

Feed rate, $F = 116.686$ kmol/hr

Distillate rate, $D = 51.861$ kmol/hr

Bottoms rate, $W = 64.826$ kmol/hr

These flowrates are fixed, whereas the flowrates inside the column are dependant on the reflux ratio, R, and may be calculated as shown on the following page.

$\bar{L}$ = liquid flow rate in the bottom section $= RD + F$     kmol/hr         (1)

$\bar{G}$ = G = vapour flow rate in the column    $= (R + 1)D$    kmol/hr         (2)

L = liquid flow rate in the top section       $= RD$           kmol/hr         (3)

### Minimum Reflux Ratio, $R_{min}$

The q–line on the McCabe–Thiele diagram in Appendix A is used to calculate our minimum reflux ratio. The q–line is vertical since the feed enters at its bubble point. The minimum reflux ratio is calculated using the 'pinch-point' condition on the McCabe-Thiele diagram. The value obtained was $R_{min} = 1.26$.

### Minimum number of stages

The minimum number of stages required ($N_m$) is calculated from the Fenske equation [6] below. This is for a system with a partial reboiler and a total condenser.

$$N_m + 1 = \frac{\log\left[\dfrac{x_D}{1-x_D} \cdot \dfrac{1-x_w}{x_w}\right]}{\log \alpha_{ave}} \tag{4}$$

From Appendix A, $\alpha_{ave}$ = average relative volatility = 2.467, we obtain from (4) that $N_m$ = 5.5 stages.

The minimum number of stages can also be read off the McCabe-Thiele diagram if we have infinite recycle. It is found that this case gives $N_m$ = 5.4 stages.

### Feed Stage Location

The Feed point location can be determined by the Kirkbridge Equation [4, p 480]

$$\log\left(\frac{N_R}{N_S}\right) = 0.206 \log\left[\left(\frac{D}{W}\right)\left(\frac{x_{f,HK}}{x_{f,LK}}\right)\left(\frac{x_{b,LK}}{x_{d,HK}}\right)^2\right] = 0.206 \log\left[\left(\frac{51.861}{64.826}\right)\left(\frac{0.55}{0.45}\right)\left(\frac{0.05}{0.05}\right)^2\right] \tag{5}$$

From this we obtain that $N_R / N_S = 1.091$, thus our feed needs to enter in the middle tray of the column.

### Condenser and Reboiler Duties

From the Antoine data the temperatures at various points in the column may be determined. In particular we have that: $T_{top}$ = 81 °C and $T_{bottom}$ = 110 °C.

The condenser at the top of the column has to only remove the heat of condensation of the gaseous stream, G. Thus we have that:

$$\text{Condenser duty} = G \times \Delta H_{vap, ave, 81°C} \tag{6}$$

where $\Delta H_{vap, ave, 81°C}$ = molar weighted heat of vaporisation at 81 °C – see Appendix B for the values.

The reboiler has to vapourize the bottoms liquid flow as well as heat it slightly. The amount of heating required can be found from the McCabe–Thiele diagram, since the last stage on the diagram is the equilibrium reboiler. One reads off the mol fraction from the diagram and the consults the data from which the diagram was drawn (Appendix A) to obtain the temperature difference. Thus the equation for the reboiler duty is:

$$\text{Reboiler duty} = \overline{L} \times \Delta H_{vap, ave, 110°C} + (\overline{L})(C_{p,bottom,ave,110°C})(\Delta T) \tag{7}$$

where the physical data referred to is the molar weighted data of the mixture.

For the calculation of the operating costs it is not assumed that the reboiler duty is this exact figure. Heat exchange equipment and thermosyphons are often used as reboilers and are not 100% efficient; there is also heat loss to the surroundings from the column and associated piping. For the costing calculations and efficiency of 70% is assumed.

Both G and $\overline{L}$ are dependent on the reflux ratio. The condenser and reboiler duties increase with increasing R, which can be seen from the table of results at the end of this Appendix.

*Liquid and Vapour Flowrate through the Column*

The liquid and vapour flowrates will be different at the two ends of the column. Although one of the assumptions in the design is that the flowrates from tray to tray do not vary, this is a very good approximation in the McCabe–Thiele diagram. The compositions vary and as such the average vapour and liquid densities vary. In our column design we consider the two ends of the column, since they are where the maximum liquid (bottom) and vapour (top) volumetric flows occur. We design our column for the largest of these and then the other trays will have sufficient capacity.

$$V_W = (G)(MW_{ave}) = \text{vapour mass flowrate, top (kg/s)} \tag{8}$$

$$L_W = (\overline{L})(MW_{ave}) = \text{liquid mass flowrate, bottom (kg/s)} \tag{9}$$

$$V_V = (V_W) \div (\rho V) = \text{vapour volumetric flowrate, top (m}^3\text{/s)} \tag{10}$$

$$L_V = (L_W) \div (\rho L) = \text{liquid volumetric flowrate, bottom (kg/s)} \tag{11}$$

## Column Diameter

The column diameter must be selected so that flooding does not occur, however at the same time one needs vapour velocities that are high for greater plate efficiencies. In our calculations we will design for operation at 85% flooding velocity. This velocity may be determined from a

correlation by Fair [4, p 521]: $u_f = K_1 \sqrt{\dfrac{\rho_L - \rho_V}{\rho_V}}$ (12)

where $K_1$ = constant evaluated from Figure 11.27 [4, p 522]. Alternatively this figure has been

curve fitted by Ward in the article [8], where $K_1 = \dfrac{0.26t - 0.029t^2}{\sqrt{1 + 6F_{LV}^2 t^{0.7498}}}$ (13)

where $K_1$ here is given in ft/s, and t = tray spacing (ft). The parameter $F_{LV} = \dfrac{L_w}{V_w} \sqrt{\dfrac{\rho_V}{\rho_L}}$ (14)

is dimensionless.

This equation assumes one knows the plate spacing. In our initial calculation we have t = 0.5 m.

Every plate has a downcomer to lead to the following plate, and a downcomer area from the previous plate. Assuming for all our calculations that one downcomer area = 12%, we have an active area, $A_A = (1-2\times0.12)A_C$. The column area, $A_C$ will be different at the top and bottom of the column, as such we calculate:

$$A_{C, top} = \frac{\text{volumeric flow}}{\text{velocity}} = \frac{V_v}{0.85u_f} \quad \& \quad A_{C, bottom} = \frac{L_v}{0.85u_f} \qquad (15)$$

Although in all the calculations these two values are approximately the same, we take the larger of the two for the rest of the design. The column diameter can easily be obtained, knowing the area.

## Liquid Flow Pattern

The choice of plate type (reverse, single pass or multiple pass) depends on the liquid flowrate and the column diameter. The selection of the liquid flow arrangement can be made with the aid of figure 11.28 [4, p523]. The chart requires the maximum liquid flowrate. This occurs in the bottom section of the column, $\overline{L}$. It is found for this design that a single pass plate is suitable.

## Plate Layout

The design method followed requires that a certain plate layout is guessed initially. Calculations for key performance factors such as entrainment and downcomer liquid backup and then carried out. If the results are not acceptable the plate layout needs to be modified. Our initial layout is as follows:

$d_C$:     Column diameter

$A_C$:     Column area

$A_N$:     Net area = $0.88A_{column}$          (12% downcomer area)

$A_A$:     Active area = $(1-2\times0.12)A_C$

$A_H$:     Hole area = 0.10AA          (10% of the active area are hole)

The dimensions for the weir also need to be guessed. [4, p 527] recommend the following initial guesses:

$l_W$:     Weir length = 0.77(Column Diameter)

$h_W$:     Weir height = 50 mm

Since the weir height determines the volume of the liquid on the plate it is an important factor in determining the plate efficiency. A high weir height will increase the plate efficiency but at the expense of a higher plate pressure drop. It has been found from practical experience that for columns operating above atmospheric pressure the weir heights will normally be between 40mm and 90mm with 40 – 50 mm being recommended.

Other recommended initial guesses are:

$d_h$:     Hole diameter = 5 mm, with equilateral triangle pitch

$T_p$:     Plate thickness = 5 mm

## Check Weep Point

The lower limit of the operating range occurs when liquid leakage through the plate holes becomes excessive. This is known as the weep point. The vapour velocity at the weep point is the minimum value for stable operation. The hole area must be chosen so that at the lowest operating rate the vapour flow velocity is still well above the weep point.

The following correlation is used for predicting the vapour velocity at the weep point.

$$u_h = \frac{[K_2 - 0.90(25.4 - d_h)]}{(\rho_v)^{\frac{1}{2}}}$$  (16)

where

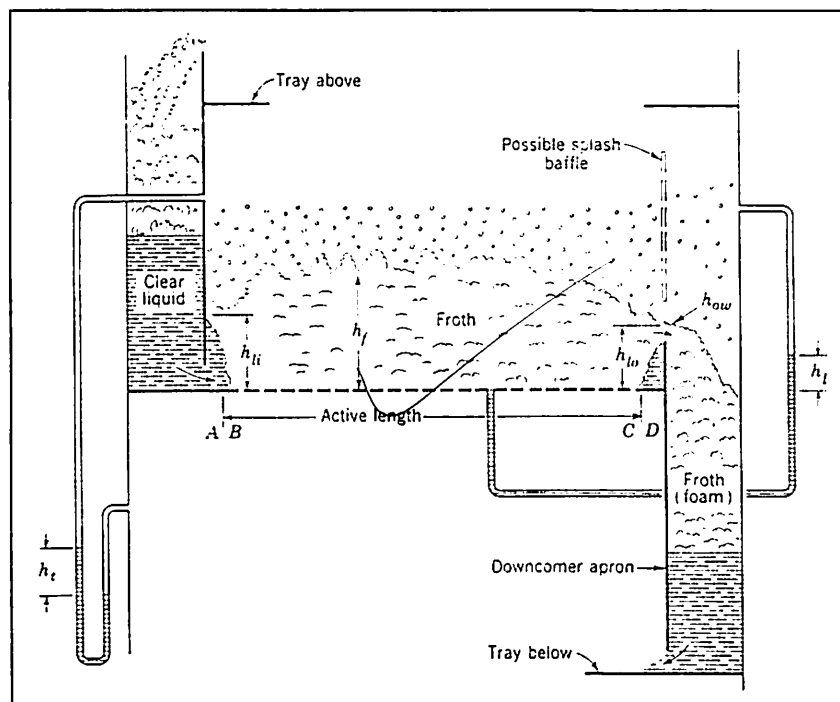$u_h$ = minimum vapour velocity through the holes (based on the hole area) [m/s]

$d_h$ = hole diameter [mm]

$K_2$ = constant, obtained from Figure 11.30 [4, p 525].

This figure requires that we need to know the height of liquid flowing over the weir crest, which

may be obtained from the Francis weir formula $h_{ow}$ (mm) $= 750 \left[ \dfrac{L_w}{\rho_L \cdot l_w} \right]^{\frac{2}{3}}$  (17)

See the diagram below for an illustration of where these heights occur.



*from* [5, p 542]

Since we have calculated the minimum vapour velocity through the holes, we need to determine our actual velocity through the holes. This is found from our minimum vapour rate per unit hole

area: $u_{actual} = \dfrac{0.7 \times V_{max, base}}{A_h}$ (m/s)  (18)

where minimum vapour rate is taken at 70% turndown i.e. $L_{min} = 0.7 \bar{L}$

This actual velocity from (18) should be well above the weep point.

## Check the plate pressure drop

The pressure drop over the plates is an important design consideration. A simple additive model is normally used to predict the total pressure drop. The total drop is taken as the sum of the pressure drop for the vapour flow vapour through the dry plate ($h_d$); the head of clear liquid on the plate ($h_{ow} + h_w$); and a term to account for other, minor, sources of pressure loss, the so-called residual loss $h_r$.

## Dry plate drop

For flows through orifices we have the equation:
$$h_d \text{ (m)} = 51 \left[\frac{u_h}{C_o}\right]^2 \frac{\rho_v}{\rho_L} \tag{19}$$

where: $C_o$ = orifice coefficient , is obtained from Figure 11.34 [4, p 530]

$u_h$ = velocity through the holes = $V_{max,base} / A_h$

## Residual head

The value for this pressure drop is found from:
$$h_r \text{ (m)} = \frac{12.5 \times 10^3}{\rho_L} \tag{20}$$

## Total pressure drop per plate

The total pressure drop per plate is then cumulatively found from:
$$h_t \text{ (m)} = h_d + (h_w + h_{ow}) + h_r \tag{21}$$

## Total pressure drop over the column

The pressure drop over the entire column may be then calculated from the formula:
$$\Delta P_{total} = (9.81 \times 10^{-3})(h_t)(\rho_L) \tag{22}$$

## Downcomer area liquid backup

The downcomer area and plate spacing must be such that the level of the liquid and froth in the downcomer is well below the top of the outlet weir on the plate above. The back-up of liquid in the downcomer is caused by the pressure drop over the plate and the resistance to flow in the downcomer itself. In terms of clear liquid the downcomer back-up is given by:
$$h_b \text{ (mm from plate surface)} = h_t + (h_w + h_{ow}) + h_{dc} \tag{23}$$
where all previous terms except $h_{dc}$ have already been calculated.

$h_{dc}$ may be evaluated from $\quad h_{dc}\,(mm) = 166\left[\dfrac{L_{wd}}{\rho_L \cdot A_m}\right]^2 \qquad\qquad (24)$

$L_{wd}$ = liquid flow rate in the downcomer [kg/s]

$A_m$ = either the downcomer area, or the clearance area under the downcomer, whichever

     is smaller [$m^2$]

The clearance under the downcomer, $A_{ap}$ may be calculated from:

$$A_{ap} = (h_W\,(mm) - 10)(l_W) \qquad\qquad (25)$$

For safe design to avoid flooding, we should have that the liquid back-up does not exceed half the

plate spacing, that is $\qquad\qquad h_b \leq \frac{1}{2}(t + h_W) \qquad\qquad (26)$

If this condition is met, then our plate spacing is acceptable.

### Downcomer Residence Time

Sufficient residence time must be allowed in the downcomer for the entrained vapour to disengage

from the liquid stream. A time of at least 3 seconds or more is recommended. The residence time

can be found from the following equation: $\tau = \dfrac{A_d \cdot h_{bc} \cdot \rho_L}{L_W}$ ($\tau$ is seconds) $\qquad (27)$

### Entrainment

The entrainment of liquid droplets in the vapour phase moving upwards affects the plate

efficiency. The entrainment may be estimated from Figure 11.29 [4, p 524]. The percentage

flooding parameter on the chart is given by %flood = $\dfrac{u_{actual}\ (\text{from equation }18)}{u_f} \qquad (28)$

An entrainment less than 0.1 kg/kg liquid flow has very little effect on the plate efficiency.

*Compare this with what you got*

### Overall Column Efficiency

The overall column efficiency, $E_O$, may be calculated from the O'Connell correlation for overall

tray efficiencies. It requires the value of $\alpha_{ave}$ and $\mu$ (cP) of the mixture. A correction term is then

added on for the length of the liquid flow path.

Then we have that $E_O = \dfrac{\text{number of theoretcal trays}}{\text{number of actual trays}} \qquad\qquad (29)$

### Summary of the above

In the above calculations we have given methods to estimate various column parameters. In our design once we have selected our tray layout we calculate these parameters then and check that the following hold:

| | | | |
|---|---|---|---|
| Flooding: | $u_{actual} < u_f$ | | – see equations 12 and 18 |
| Weeping: | $u_{actual} > u_h$ | | – see equations 16 and 18 |
| Pressure drop per plate: | Check that it is not excessive | | – see equation 21 |
| Back–up in downcomer: | Check $h_b \leq \frac{1}{2}(t + h_w)$ | | – see equation 26 |
| Residence time: | Check $\tau > 3$ seconds | | – see equation 27 |
| Entrainment: | Entrainment $< 0.1$ kg/kg liquid flow | | – see equation 28 |

If any of the inequalities do not hold new values should be selected and the calculations repeated.

### Calculations for various reflux ratios

The following reflux ratios were used to calculate the above:

$R = 1.3 R_{min}$

$R = 1.5 R_{min}$

$R = 2 R_{min}$

$R = 3$

All of the calculations had the following common values:

Downcomer area = 12%

Hole area = 10% of active area

Hole diameter of 5 mm

Tray spacing = 0.5 m

Weir height = 50 mm

Overall column efficiency = 73 %

The results for all of the calculations appear on the following page:

|  | Unit | 1.3 $R_{min}$ | 1.5 $R_{min}$ | 2 $R_{min}$ | R = 3 |
|---|---|---|---|---|---|
| L | kmol/hr | 84.95 | 98.02 | 130.69 | 155.58 |
| $G = \overline{G}$ | kmol/hr | 136.81 | 149.88 | 182.55 | 207.44 |
| $\overline{L}$ | kmol/hr | 116.69 | 116.69 | 116.69 | 116.69 |
| Condenser Duty | kW | 1175.4 | 1287.7 | 1568.3 | 1782.2 |
| Reboiler Duty | kW | 1858.0 | 1979.1 | 2297.8 | 2529.0 |
| Theoretical n° of trays |  | 13.8 | 12 | 10.5 | 9.4 |
| Actual number of trays |  | 19 | 17 | 15 | 13 |
| Feed Entry Tray |  | 10 | 9 | 8 | 7 |
| Column Diameter | m | 1.24 | 1.30 | 1.41 | 1.53 |
| Flooding velocity (SGV) | m/s | 1.56 | 1.54 | 1.52 | 1.46 |
| Actual velocity (SGV) | m/s | 1.04 | 1.03 | 1.07 | 1.03 |
| Percentage flooding | (%) | 78.2 | 78.8 | 82.5 | 83.4 |
| Weeping velocity (VTH) | m/s | 7.18 | 7.18 | 7.24 | 7.12 |
| Actual velocity (VTH) | m/s | 8.40 | 8.37 | 8.67 | 8.36 |
| Pressure drop per plate | mm liquid | 130.0 | 130.1 | 133.9 | 133.2 |
| Total column pressure drop | Pa | 20 000 | 17 400 | 15 300 | 13 300 |
| Downcomer liquid height | mm | 211 | 212 | 217 | 217 |
| Tray Residence Time | s | 4.79 | 4.96 | 5.19 | 5.55 |
| Fractional Entrainment | kg/kg | 0.043 | 0.045 | 0.05 | 0.045 |

SVG = Superficial gas velocity

VTH = velocity through the holes

All these designs meet the criterion in the summary, and as such we will choose the design which will have the minimum cost, that is R = 1.3Rmin that is R = 1.65.

# APPENDIX D:    COSTING CALCULATIONS

The costs for the distillation column are made up of the initial equipment and installation costs (fixed cost) and then operating or working cost.

## FIXED COST

A quick order–of–magnitude cost can be arrived at using the Bridgewater

step–counting equation [4, p 218]:    $C = 128N\left(\dfrac{Q}{s}\right)^{0.675}$    (1)

where:    C = capital cost in GB£
N = number of units (1 in our case)
Q = capacity (tonnes per year)
$s = \text{conversion} = \dfrac{\text{mass of desired product}}{\text{mass into the reactor}} = 1$ (in our case)

For the distillation column design Q = 87600 tonnes/year, and thus

C = GB£ 277 600 ≡ R 2.94 million ≈ **R 3 million** (at the current exchange rate)

Breaking down the fixed costs as follows we can arrive at a better estimate:

Fixed Costs =  Column pipe (carbon steel)
+    Plates
+    Boiler  (for steam generation)
+    Cooling unit
+    Piping
+    Instrumentation

Individual costs may be found from various sources

**Column Pipe:**    from [4, p 223]

Pressure factor = 1;    Material = carbon steel;    Diameter = 1.25 m;

Height = 10 m;    Cost = GB£ 13 000 (1992)

Cost = R 155 000 (1998)

**Plates:** from [4, p 224]

Diameter = 1.25 m    Cost = GB£ 18 000 (1992)

Cost = R 215 000 (1998)


**Boiler:** from [4, p 225]

A preliminary estimate of a boiler cost can be made from the following equation:

$$C = BS^n \qquad (2)$$

We will assume a boiler at 500 kPa, producing saturated steam at 525K, which has the following properties, $\Delta H_{lg} = 2108$ kJ/kg.

The boiler requires 1858 kW, but taking a 70% efficiency due to heat losses we require 2655 kW. Thus we have steam from 1.25 kg/s at 500 kPa or 4.5 ton/hr.


From Table 6.2 [4, p 225] our constant B = 30, n = 0.8 and S = 4500 kg/hr. We then obtain: $C_{boiler}$ = GB£ 25 000 (1992)

$C_{boiler}$ = R 266 000 (1998)


**Cooling Unit:** Costs for a cooling unit could not be found, and thus we assumed a cost which is 25% of the Boiler = R 70 000 (1998)


**Piping and Instrumentation:** The costs for this component are assumed to be 10% of the fixed costs so far:    R 70 000 (1998)


| Item | Cost (1998) | |
|------|------|------|
| Column Pipe | R | 155 000 |
| Plates | | 215 000 |
| Boiler | | 266 000 |
| Cooling Unit | | 70 000 |
| Piping and Instrumentation | | 70 000 |
| 10% Contingency | | 77 000 |
| **TOTAL** | **R** | **855 000** |

Table of Estimated Fixed Costs

## OPERATION COSTS

Operation costs include:     Steam

                            Water for cooling

A quick estimation as to the operating costs can be provided by a correlation using the current fuel price and Chemical Engineering Plat Cost Index, which appears in the journal *Chemical Engineering* [8]

The data obtained was for 1990, which is then adjusted for 1998.

1990:     CE Plant Cost Index = 357.6

          SA Fuel Price $\approx$ R 1.00 per liter

          Exchange rate = R3 / US$

          Energy from fuel = 37 GJ/m$^3$

          Density of fuel = 700 kg/m$^3$                    – Values from [8]

The correlation is     $C_{utility} = a(CE) + b(FP)$

FP in 1990 is found to be FP = (333 US$/m$^3$)(1/37 m$^3$/GJ) = 9 US$/GJ

For *steam* production: $a = \dfrac{2.3 \times 10^{-5}}{m_s^{0.9}}$ and $b = 0.0034(p^{0.05})$

where   $m_s$ = steam flow in kg/s = 1.25 kg/s

and     p  = pressure in bar     = 5 bar

thus    $a$ = 1.882×10$^{-5}$

        $b$ = 3.685×10$^{-3}$

$C_{steam}$ (US$/kg)       = 1.882×10$^{-5}$(357.6) + 3.685×10$^{-3}$(9)

          $C_{steam}$  = 0.0399 US$/kg (1990)

          $C_{steam}$  = 0.260 R/kg (1998)

Our plant requires 4500 kg/hr $\equiv$ R 1170 / day $\approx$ R 400 000 / year

For *cooling water* from a new plant the constants are found as follows:

Our condenser duty is 1175 kW. Assuming we have cooling water entering at 20°C and leaving at 40°C, we then require a flowrate of 14 kg/s

$$a = 1.856 \times 10^{-3}$$

$$b = 0.0056$$

$$C_{cooling\ water}\ (US\$/m^3) = 0.1.856 \times 10^{-3}(357.6) + 0.0056(9)$$

$$= 0.714\ US\$/m^3\ (1990)$$

$$= R\ 4.64\ /m^3$$

Flowrate = 441 800 m³/year

$C_{cooling\ water}$ = R 2.05 million per year

Thus our operating costs are:

| Item | Cost per year |
|------|------|
| Steam | R      400 000 |
| Cooling Water | 2 050 000 |
| 5% Contingency | 122 500 |
| **TOTAL** | R      2 573 000 |

Table of Operating Costs

What about Labor, maintenance etc

# APPENDIX E: NOMENCLATURE

| Symbol | Description | Units |
|---|---|---|
| | | |
| $a, b$ | Constants used in Appendix D | Dimensionless |
| A, B, C | Constants in the Antoine Equation | Dimensionless |
| $A_A$ | Active area of column | $m^2$ |
| $A_{ap}$ | Clearance under the downcomer | $m^2$ |
| $A_C$ | Cross sectional area of entire column | $m^2$ |
| $A_H$ | Hole area | $m^2$ |
| $A_M$ | Downcomer area | $m^2$ |
| $A_N$ | Net column area | $m^2$ |
| C | Cost | Rands |
| CE | Chemical Engineering Plant Cost Index | Dimensionless |
| $C_o$ | Orifice coefficient | Dimensionless |
| $C_p$ | Heat Capacity | $J.kg^{-1}.K^{-1}$ |
| $d_C$ | Column diameter | m |
| $d_h$ | Hole diameter | mm |
| $E_O$ | Column efficiency | Dimensionless |
| $F, D, W, G, \overline{G}, L, \overline{L}$ | Molar flowrates in the column | $kmol.hr^{-1}$ |
| $F_{LV}$ | Liquid vapour flow parameter – Eqn (14) | Dimensionless |
| FP | Fuel Price Index | $US\$.GJ^{-1}$ |
| $h_b$ | Height of liquid backup | mm of liquid |
| $h_d$ | Dry plate pressure drop | mm of liquid |
| $h_{dc}$ | Height of downcomer backup | mm of liquid |
| $h_{OW}$ | Height of liquid over weir | m |
| $h_r$ | Residual head | mm of liquid |
| $h_t$ | Total pressure drop per plate | mm of liquid |
| $h_W$ | Weir height | mm |
| $K_1$ | Constant | Dimensionless |

| | | |
|---|---|---|
| $K_2$ | Constant | Dimensionless |
| $l_W$ | Weir length | m |
| $m_s$ | Mass flowrate of steam | $kg.s^{-1}$ |
| MW | Molecular weight | g/mol |
| N | Number of Trays | Dimensionless |
| $N_m$ | Minimum number of trays at infinite R | Dimensionless |
| $N_R$, $N_S$ | Number of stages in the rectification or stripping section | Dimensionless |
| P | Pressure | Pa |
| Q | Capacity | Ton/year |
| R | Reflux ratio | Dimensionless |
| $R_{min}$ | Minimum reflux ratio | Dimensionless |
| s | Conversion – defined in Appendix D (1) | Dimensionless |
| T | Temperature | °C or K as noted |
| t | Tray spacing | m or ft |
| $T_p$ | Plate thickness | mm |
| $u_{actual}$ | Actual velocity through the column | $m.s^{-1}$ |
| $u_f$ | Flooding velocity | $m.s^{-1}$ |
| $u_h$ | Weep point velocity | $m.s^{-1}$ |
| $V_V$ or $L_V$ | Vapour or liquid volumetric flowrate | $m_3.s_{-1}$ |
| $V_W$ or $L_W$ | Vapour or liquid mass flowrate | $kg.s^{-1}$ |
| $x_D$, $x_W$ | Mol fraction of stream in the distillate or bottoms | Dimensionless |
| $\Delta H_{vap}$ | Molar heat of vapourisation | kJ/mol |
| $\Delta P_{total}$ | Total pressure drop per plate | Pa |
| $\alpha_{ave}$ | Average relative volatility | Dimensionless |
| $\mu$ | Viscosity | Pa.s or cP |
| $\rho$ | Density | $kg.m^{-3}$ |
| $\tau$ | Downcomer residence time | s |