



# **Distillation**

## **M111**

### **Distillation and Process control**

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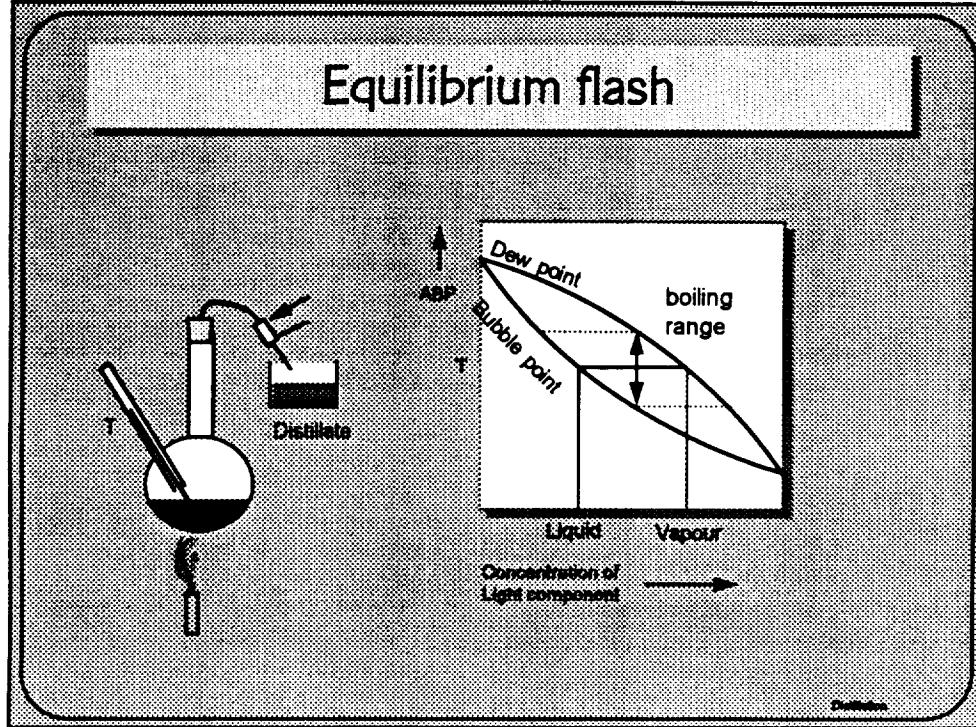
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**Binary separation**

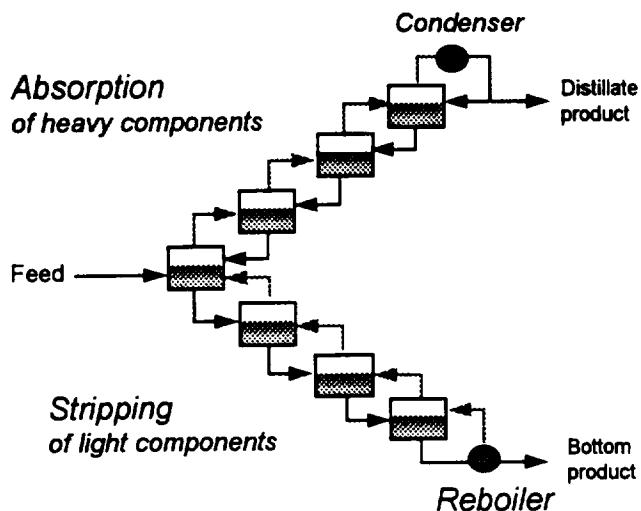
## Equilibrium flash



When the temperature of a liquid is raised above its bubble point, vapour with a high concentration of volatile components boil off.

With a further temperature rise will the amount of boil-off vapours increases. Both the remaining liquid as well as the accumulated vapour becomes more heavy; the concentration of volatile components decreases.

## Distillation as a series of flashes



The liquid left from the initial feed flash is stripped with vapour in the subsequent flashes to remove the light components.

More flash stages and/or more vapour will promote the stripping effect.

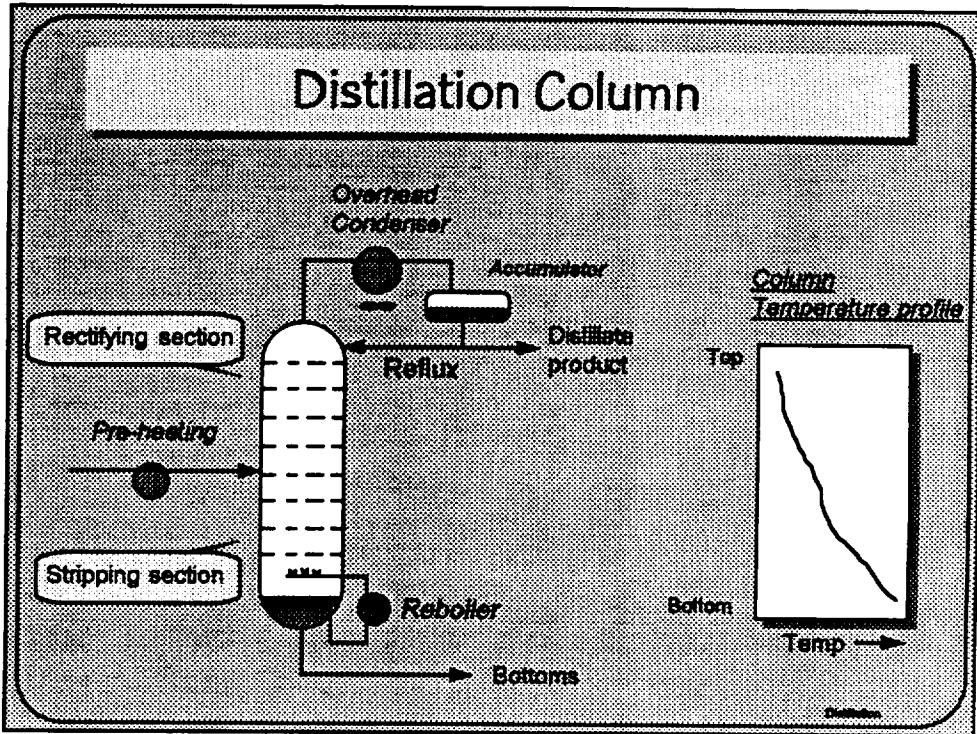
Vapour may be generated with a reboiler by boiling-off the volatile components from the bottom product. An option is to use steam as a stripping medium.

Either reboiler vapour or stripping steam, for the stripping effect it does not make any difference.

The vapour flashed off at the initial feed flash is rectified with liquid (called reflux) to absorb the most heavy components from the vapour in the subsequent flashes.

Increasing the number of flash stages and/or more liquid will improve the purity of the vapour.

Either condensed distillate product or a liquid from an external source may be used as reflux liquid. For the rectification of the vapour it does not make any difference.



A distillation column can be seen as a series of equilibrium flashes.

The column section above the feed inlet is the rectifying section.

Reflux liquid is used to absorb the relative heavy components from the internal vapour stream.

With the removal of these components the distillate product is purified or also called rectified.

The column section below the feed inlet is the stripping section.

Vapour is used to stripping-off the relative light components from the internal liquid stream.

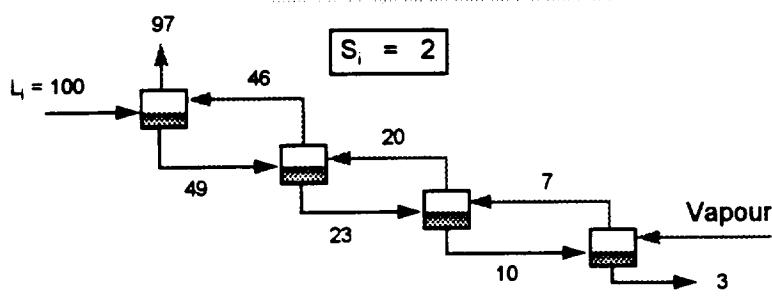
The purpose of this column section is to purify the bottom product.

An uniform temperature profile over the column indicates an effective distillation process; all distillation stages are efficiently used.

## Stripping section

$$\text{Stripping factor} = \frac{V_i}{L_i} = \frac{y_i V}{x_i L}$$

$$S_i = k_i \frac{V}{L}$$



The stripping effect to purify the bottom product depends on three variables:

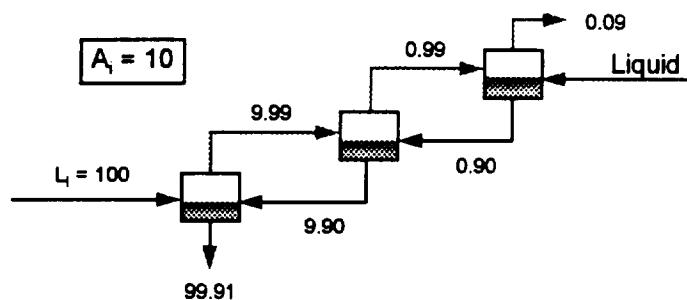
- 1 - The  $k$ -value of the component that is stripped-off; its volatility.
- 2 - The total molar vapour flow in relation to the liquid in the column section.
- 3 - The number of equilibrium stages in the stripping section of the column.

The first two variables are included in the stripping factor of the component while the number of stages is fixed by the hardware.

## Rectifying section

$$\text{Absorption factor} = \frac{L_i}{V_i} = \frac{x_i L}{y_i V}$$

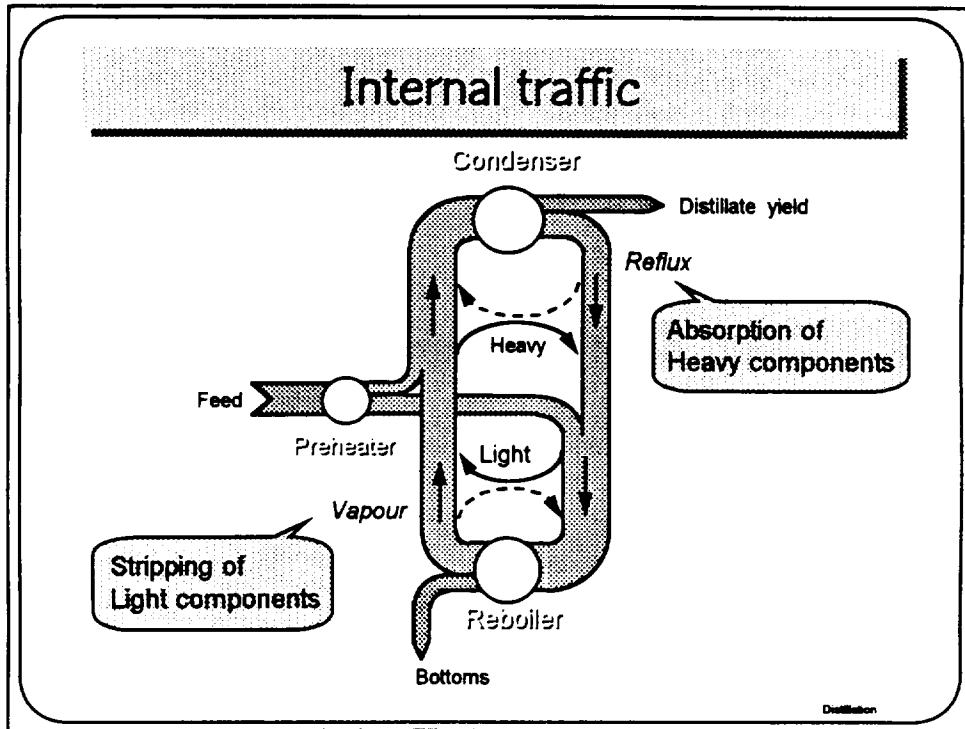
$$A_i = \frac{1}{k_i} \cdot \frac{L}{V}$$



The absorption factor is the reciprocal value of the stripping factor.

The rectification of the distillate depends on the amount of the reflux liquid in relation to the vapour flow.

Note that the internal reflux flow being at bubble point temperature should be considered. In some cases the external reflux may be subcooled (below its bubble point) and additional internal reflux will be generated by condensing vapour in the top of the column.



Separation performance of a distillation column can be increased with a higher internal traffic of vapour and liquid.

However for a constant distillate yield, more reflux requires also more vapour either from the reboiler or from the feed preheater. This additional overhead vapour is condensed in the condenser so more heat has to be discharged at a low temperature level. This will increase the operating cost of the column.

Vapour from the reboiler is more efficiently used than vapour from the feed preheater since the reboiler vapour is also utilized in the stripping section of the column.

Reboiler heat is more expensive because the reboiler operates at a higher temperature level than the feed preheater.

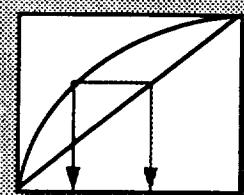
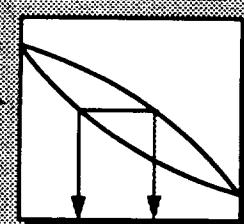
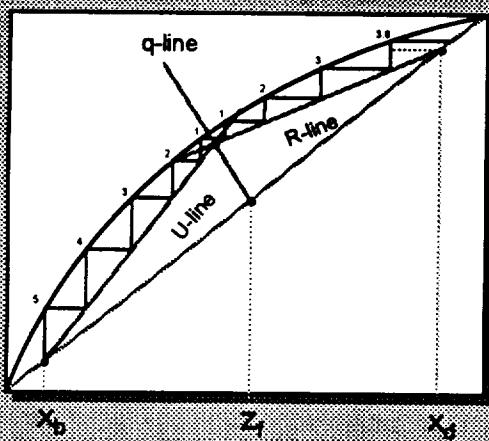
The optimal operation in terms of energy and column loading is when the amount of vapour from the feed preheater is about equal to the distillate yield of the column. This operation is called 'Top part vaporized'.

## Summary

- Fractionation is a series of flash stages
- Rectifying section is to recover the heavy components from the vapour
- Stripping section is to remove the light components from the liquid
- Internal traffic of vapour and liquid effects the Stripping factors, Absorption factors
- Reboiler is an additional single stage flash
- Condenser is an additional single stage flash only for a partial condenser with vapour top product

Binary separation

## McCabe-Thiele diagram

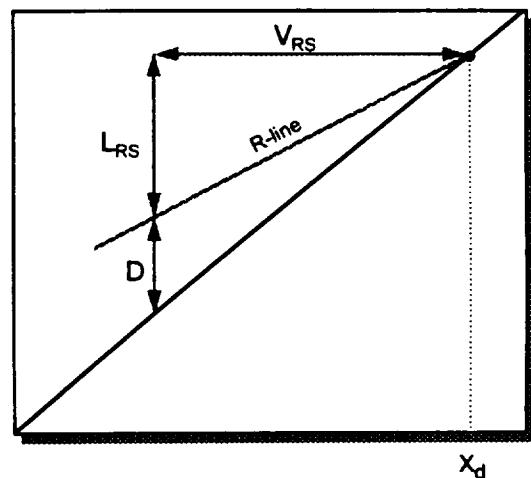


Equilibrium diagram.

## Rectifying section

Reflux ratio

$$R = \frac{L_{RS}}{D}$$

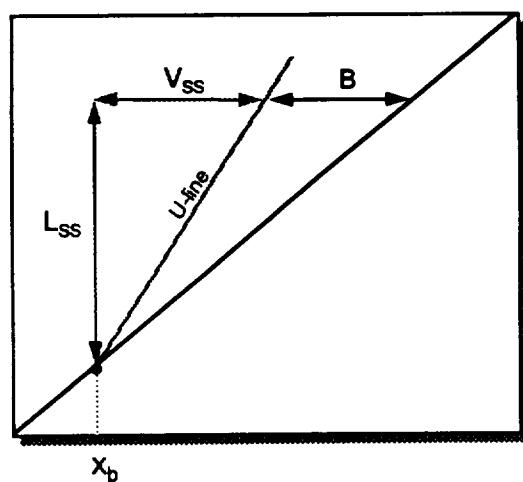


Binary separation

## Stripping section

Boil-up ratio

$$U = \frac{V_{SS}}{B}$$

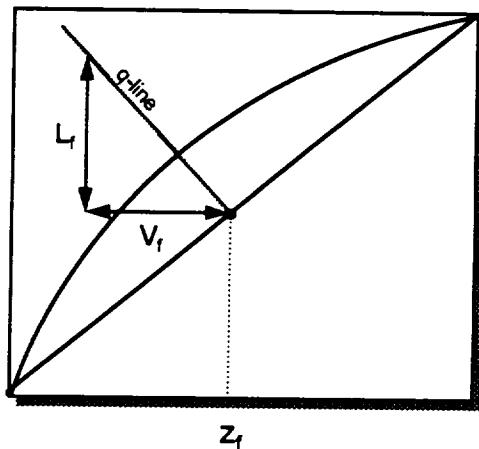


Binary separation

## Feed preheat

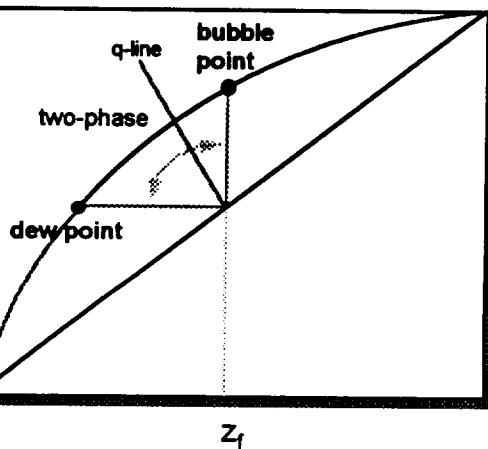
Feed flash

$$q = \frac{L_f}{F}$$



Binary separation

## feed inlet condition

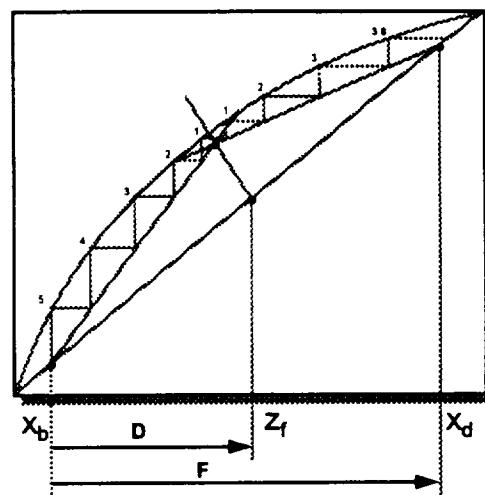


Binary separation

## Distillate yield

Distillate yield

$$\frac{D}{F} = \frac{Z_f - X_b}{X_d - X_b}$$



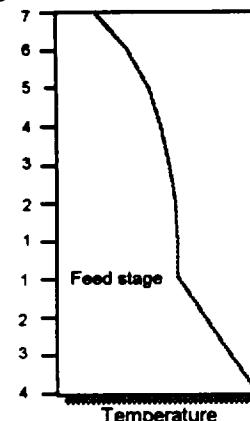
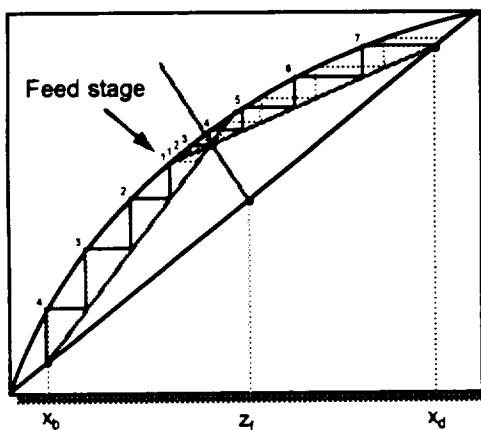
Binary separation

## Feed stage location

Feed stage

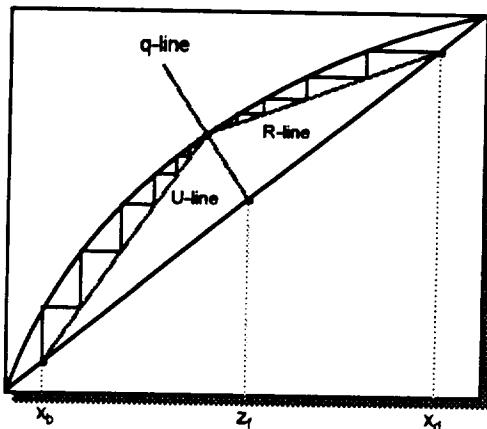
Column  
Temperature profile

Stages



Binary separation

## The minimum reflux case



Binary separation

## Summary

- Distillate yield is ONLY related to feed composition and product qualities
- The McCabe-Thiele diagram is a graphical presentation of a binary distillation process
- Slopes of the operating lines depends on Vapour and Liquid traffic in the column sections
- Required number of stages can't be specified, it results from the graph

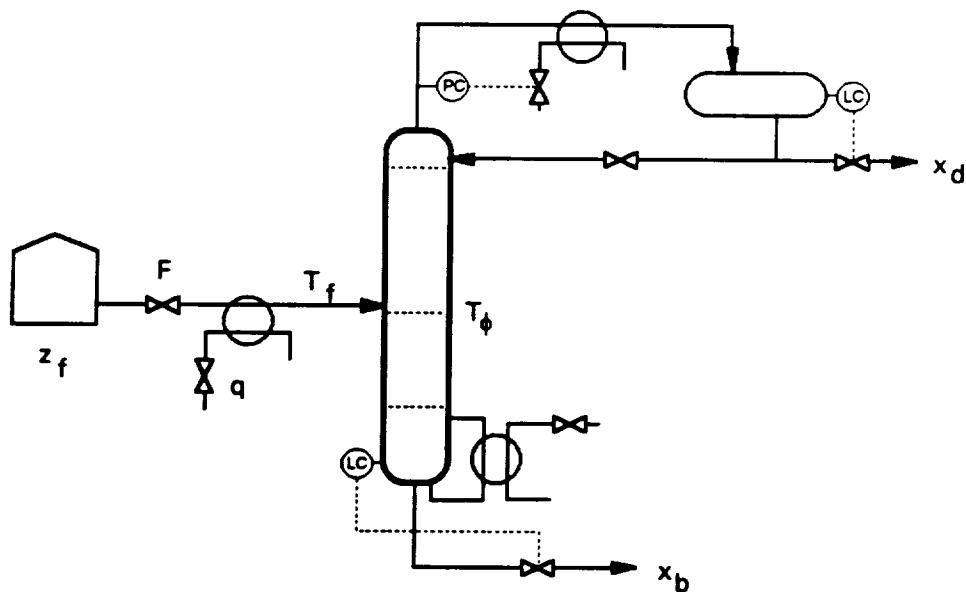
Binary separation

## Operational Cases

From experience we know that distillation columns are seldom operated at the original design conditions. Although they have been designed to process a certain feedstock, very often the actual feed to the column may be different in composition. This will have an effect on the distillate product quality or yield. To restore the required product quality the operation of the column need to be changed. In general there may be various options to shift the column operation. They can be demonstrated with the McCabe-Thiele diagram.

As an example we will consider a benzene/toluene splitter column shown in Figure 1. It has a feed preheater, a once-through reboiler and a total overhead condenser; no vapour is withdrawn from the accumulator. The column is essentially manually controlled, only the column pressure and liquid levels are controlled automatically.

Figure 1, benzene/toluene splitter



The benzene/toluene splitter column was originally designed for the following operation:

$$\begin{array}{ll}
 F = 100 \text{ mole/s} & x_d = 0.95 \\
 z_f = 0.50 & x_b = 0.10 \\
 q = 0.60 & \alpha = 2.40 \\
 T_\phi = T_f & R = 1.86
 \end{array}$$

This case is shown by the McCabe-Thiele diagram in Figure 2.

Now try to answer the following questions.

Binary separation

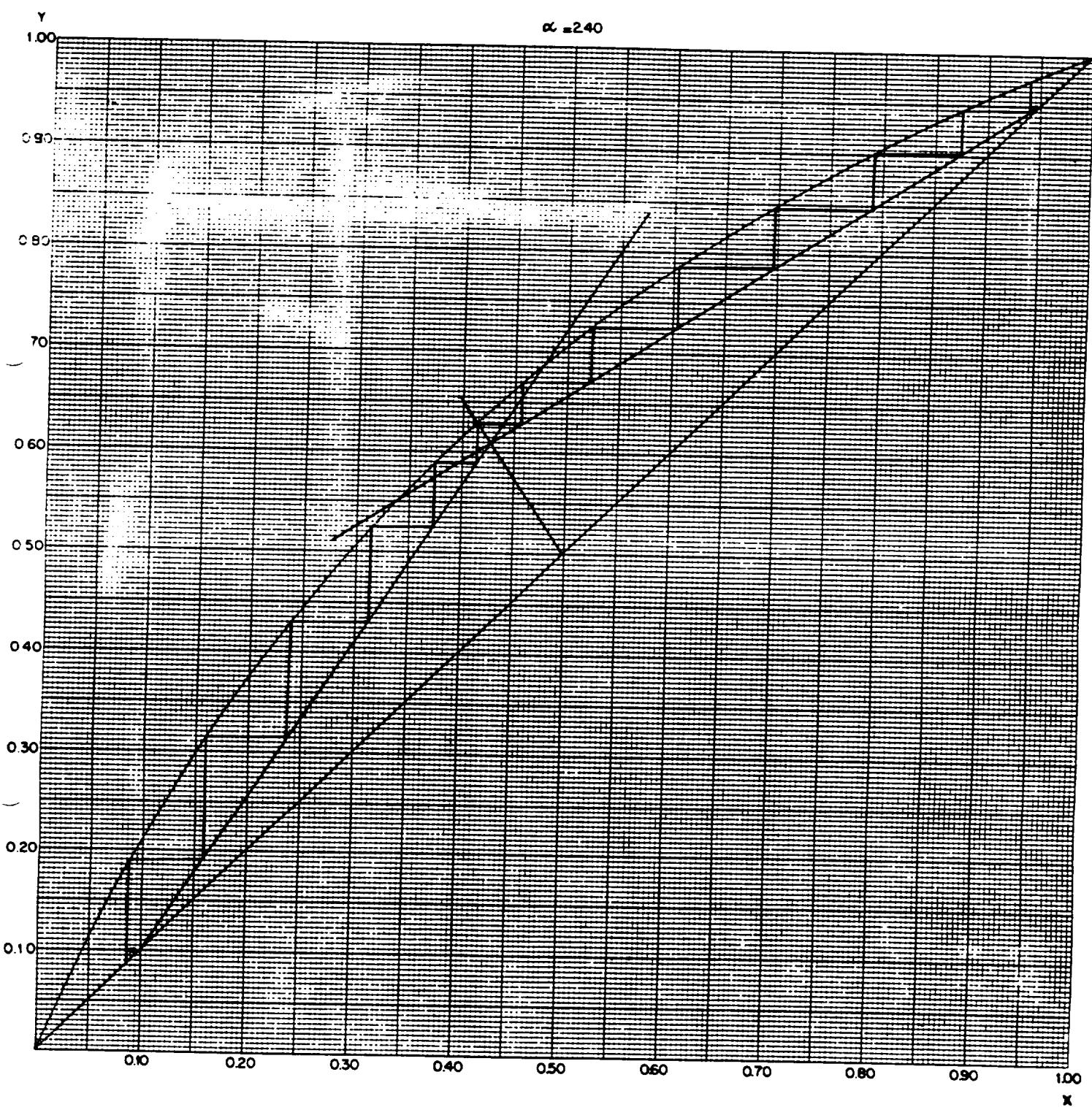


Figure 2, Design case

**Exercise 1**

- a. How many stages are required for the column in the rectifying section and in the stripping section ?

$$N_{RS} = \text{stages}$$

$$N_{SS} = \text{stages}$$

- b. What is the distillate yield for the design case ?

$$D/F = \text{mole/mole}$$

- c. Calculate the internal traffic in the column sections and complete the first column (design case) of Table 1.

**Table 1 Operational cases**

	Flow rate (mole/s)	
	Design case	Feed change
D		
$x_d$	0.95	0.95
$V_{RS}$		
$L_{RS}$		
$z_f$	0.50	0.40
$V_f$		
$L_f$		
$V_{SS}$		
$L_{SS}$		
B		
$x_b$	0.10	0.10

**Exercise 2**

After a tank switch the feed composition has changed to  $z_F = 0.40$ . Suppose that the column control remains as it is; the manual valves are left in their original positions.

- What happens to the distillate yield D/F and the internal traffic of the column ?
- What effect does the feed change have on the product purities  $x_D$  and  $x_B$  ?
- Show this case in a McCabe-Thiele diagram.
- Has the feed tray temperature  $T_\phi$  become higher or lower than the feed inlet temperature  $T_F$  .

**Exercise 3**

Given the new feed composition  $z_F = 0.40$  we want to restore the original product purities:

$$\begin{aligned}x_D &= 0.95 \\x_B &= 0.10\end{aligned}$$

- What will the distillate yield be ?

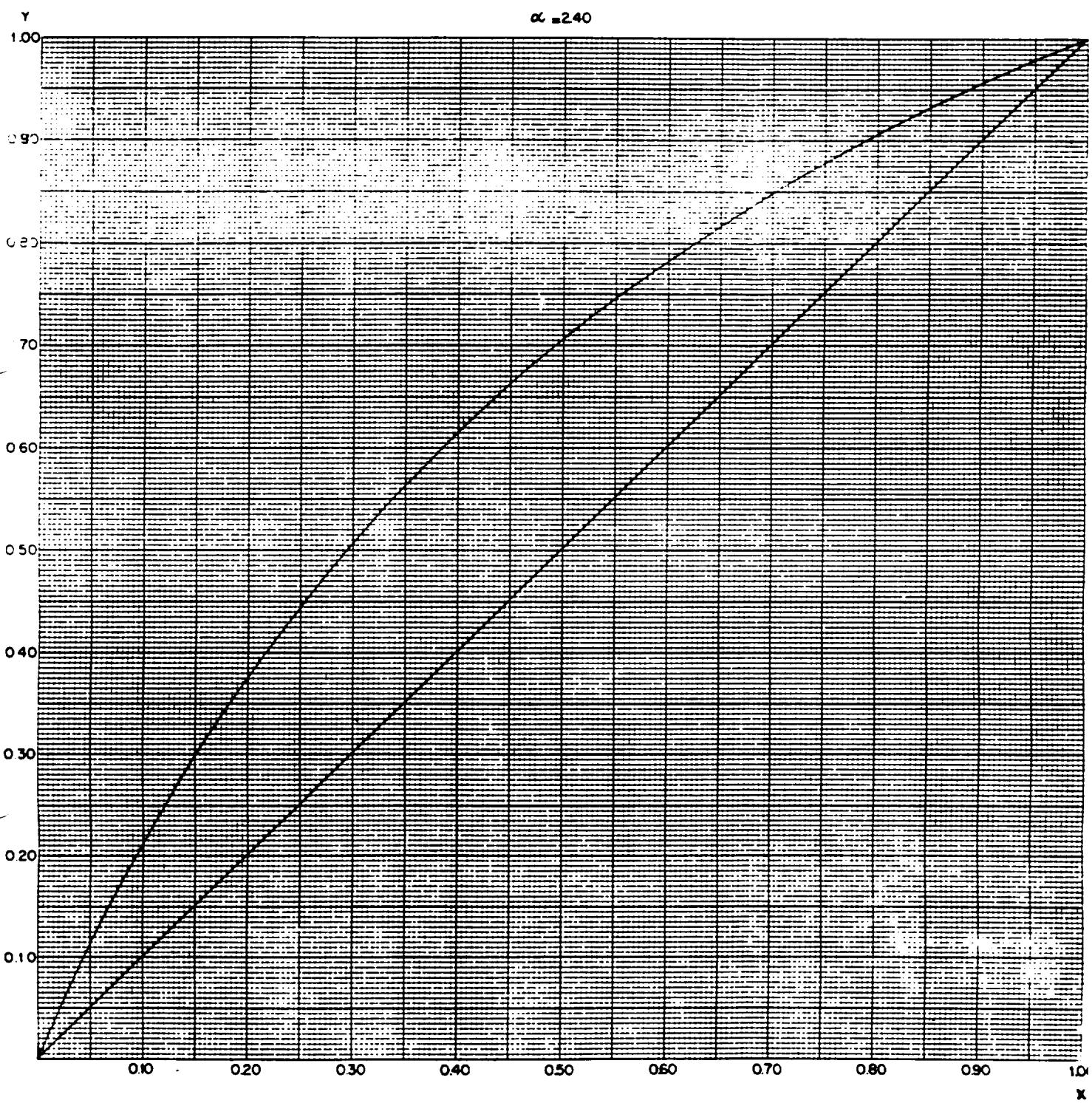
$$D/F = \text{mole/mole}$$

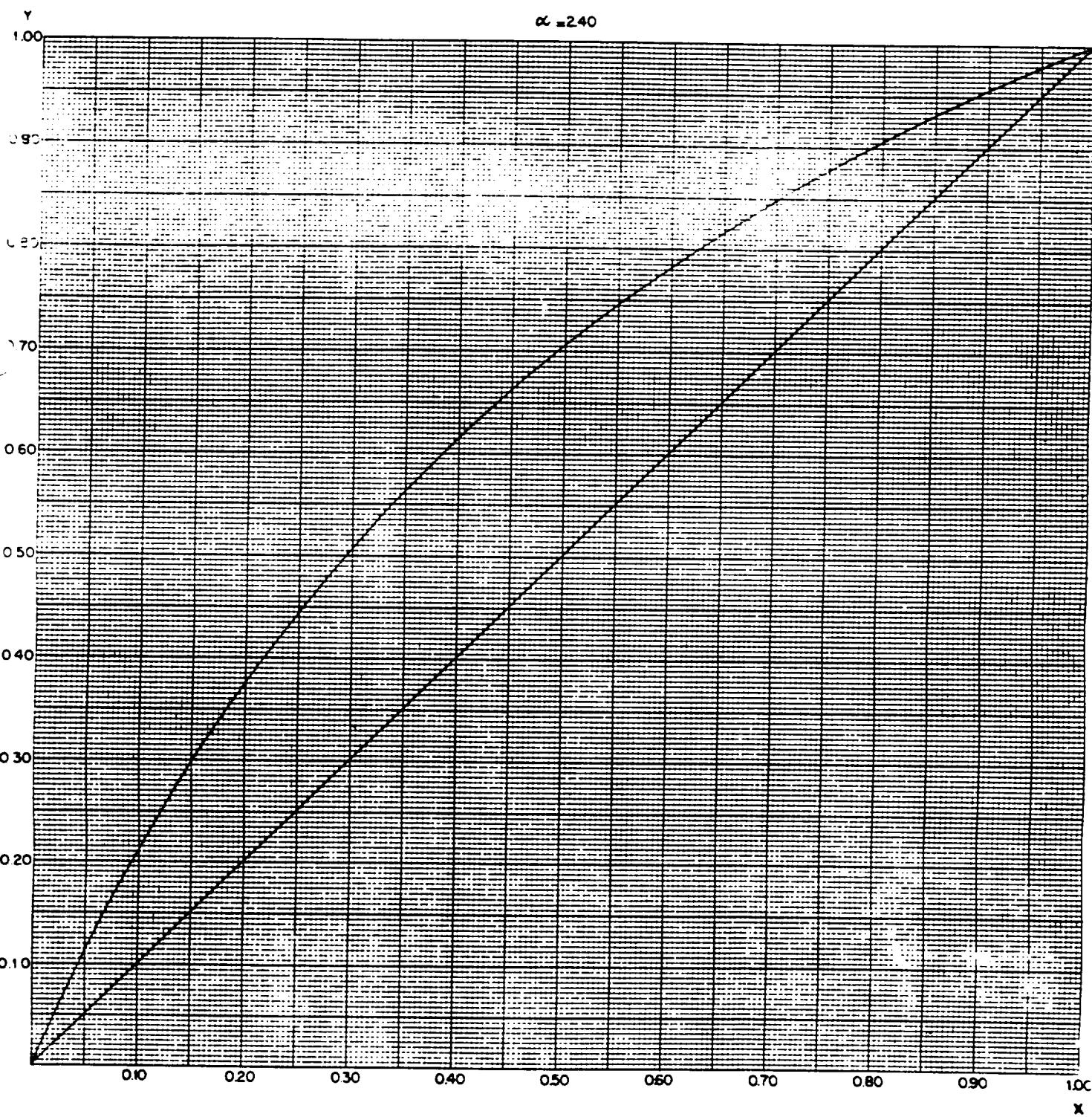
- Suppose we keep the reflux flow constant.  
Draw the new R-line and U-line in the McCabe-Thiele diagram.
- What is the required percentage change on reboiler duty if we ignore second-order effects?

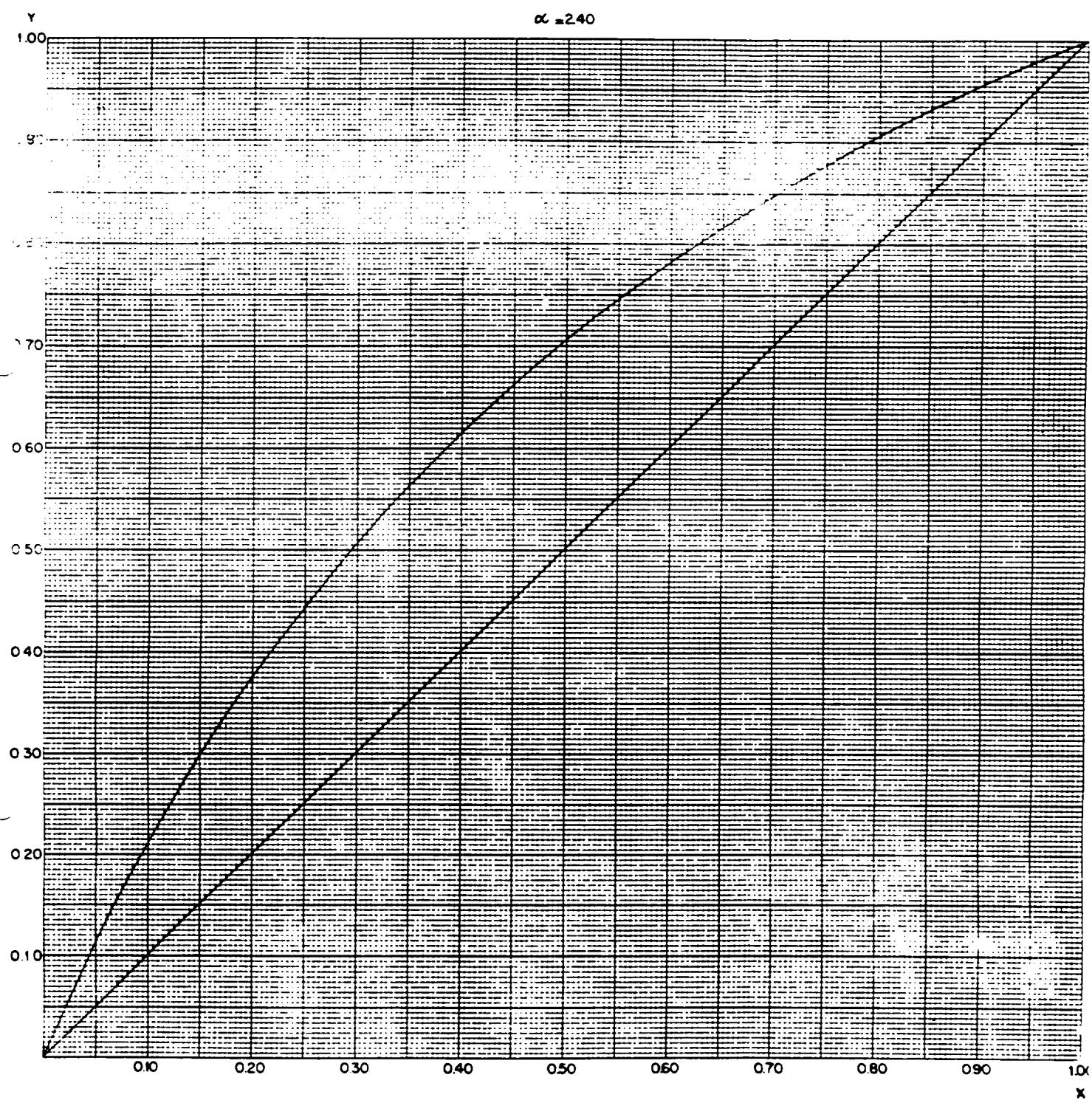
**Exercise 4**

Instead of changing the reboiler duty what else could you do to restore the product compositions.

- Show this in the McCabe-Thiele diagram.
- Calculate the internal traffic of the column and complete the last column of Table 1.







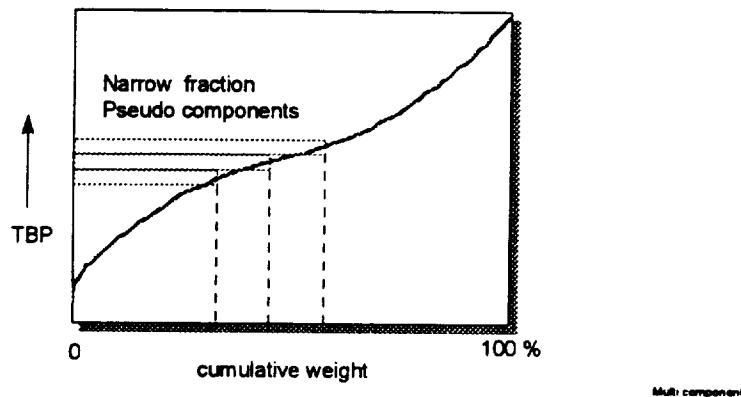


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## **Multi component separation**

## TBP analysis

- Test column (sharp fractionation)
- Gas-Liquid Chromatograph (TBP-GLC)



A TBP curve gives the cumulative component distribution against the true boiling point (TBP) of the components.

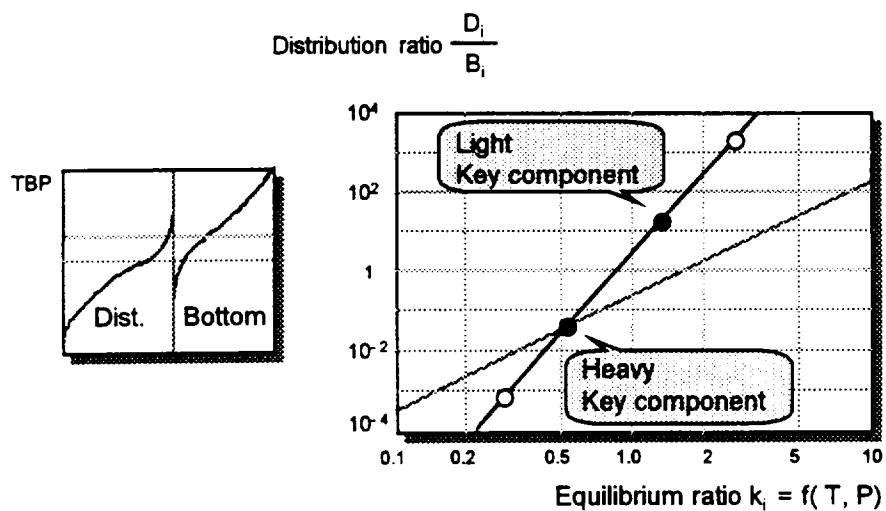
A TBP curve can be obtained in the laboratory either by a test column or with a Gas Chromatograph.

To limit the number of data points the components which do have a similar boiling point are lumped. Together they are considered as being one single component; a pseudo component.

A pseudo component represent all components boiling within a narrow range of say 5C or 10C ; narrow fraction pseudo component.

Properties of pseudo components like d 15/4, UOP-K and V50 are the average properties of its constituents.

## The 'vanDijck' line



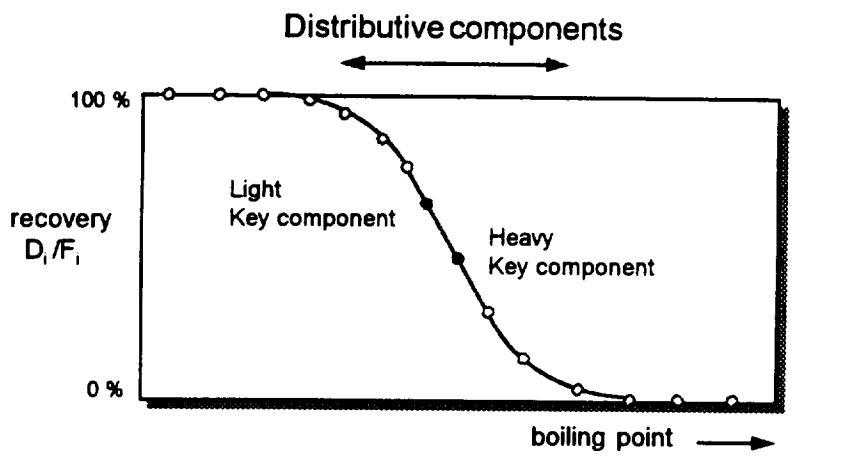
All components in a two-cut multi component separation fall on a single straight line in a distribution ratio versus equilibrium ratio plot with logarithmic scales.

If you know the position of two (key) components you can draw the line; the 'vanDijck' line.

Given the equilibrium ratio, you can read the distribution ratio of any component.

The slope of the 'vanDijck' line is equivalent to the sharpness of separation between the distillate and bottom product. A higher slope of the line indicates a better separation performance.

## Distribution curve



'vanDijck' lines are usually presented as recovery of distillate plotted against atmospheric boiling point. This gives the typical double bent distribution curve to show the separation.

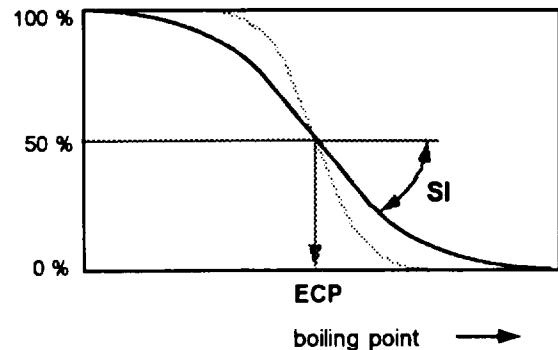
The area on the left of the distribution curve is the yield of distillate product. The area on the right is the amount of bottom product.

The components which are partly present in the distillate and in the bottom product are called 'distributive components'. Key components are always distributive components.

Non-distributive components appear only in the distillate product or in the bottom product. The most volatile components on the 100% recovery line and the heavy components on the 0% line.

## ECP and SI

recovery  $D_i/F_i$



### Separation Index

- 0.3 Equil. Flash
- 1.2 Steam stripping
- 2.0 No gap / no overlap
- >4 Sharp fractionation

Multi component

The boiling point of the component which is equally (50%) distributed over the distillate and bottom product is called the effective cutpoint (ECP).

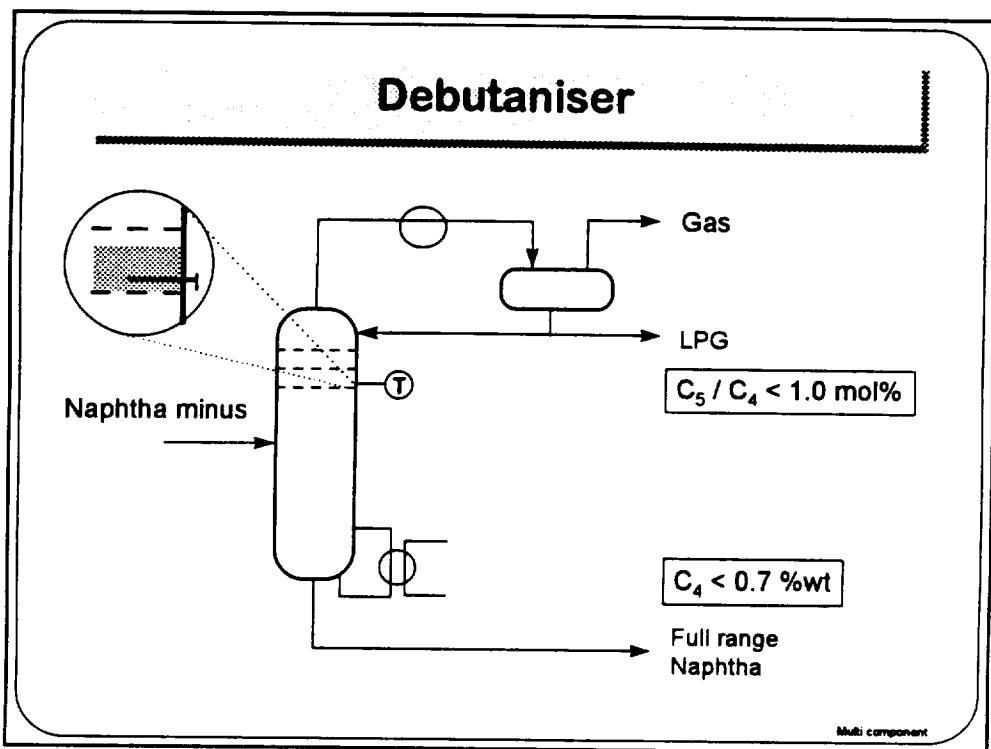
A shift in the ECP will change the distillate yield.

The separation index (SI) is the slope of the distribution curve at the ECP in (%/F). A higher SI means a better separation performance.

The SI is not related to the distillate yield.

For a Fractionator column with side strippers the distribution curve may not be symmetric. In that case there are two SI's; one for the front end of the heavy fraction and one for the tail end of the light fraction.

With an asymmetric separation the TBP cutpoint will be different to the ECP and the SI's may have an effect on the yield.



Multi component distillation resembles binary separation by considering the distribution of light and heavy key components only, assuming that the other components in the feed are non-distributive. For a Debutaniser column key components are typically C4 and C5.

In the operation of distillation columns temperature can be used as an indication of composition at various points in the column. The justification is the Gibbs Phase Rule:

*For homogeneous systems with two phases (vapour and liquid) the number of degrees of freedom equals the number of components;  $F = C$ .*

The composition is given by  $C-1$  component variables. This leaves one free variable. If the pressure is fixed the temperature is related to the composition of the mixture.

- **Dew Point Condenser**

Only the reflux is condensed. The top product is left in the vapour phase; at its dew point. The accumulator acts as an additional theoretical stage.

- **Total Condenser**

All overhead vapours are condensed. The column pressure depends on the condenser temperature. The liquid can even be subcooled.

- **Partial Condenser**

There are two top product streams; heavy components as liquid, light components as vapour.

## Effect of Non-distributive components

- Light components in feed
  - Absorption factor  $A = 1/k * (L / V)$
  - Vapour loading of trays in Rectifying section
  - Equilibrium temperature Condenser
- Heavy components in feed
  - Stripping factor  $S = k * (V / L)$
  - Liquid loading of trays in Stripping section
  - Reboiler temperature

Multi component

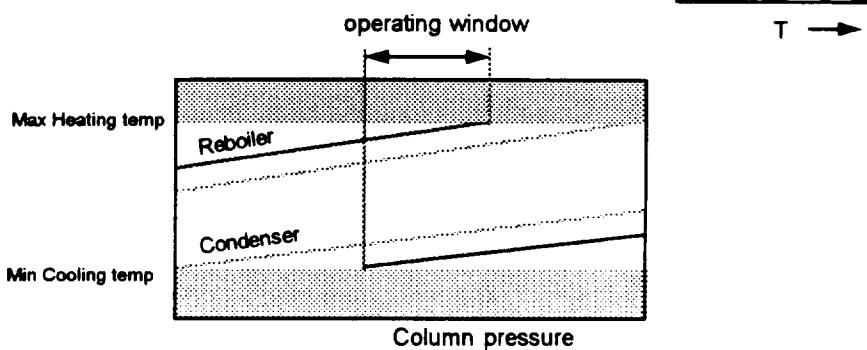
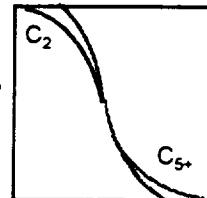
The amount of non-distributive components has direct and also an indirect effect on the tray loading. The reduction of Absorption and Stripping factors needs to be compensated by an increased traffic; more reflux and more boil-up vapour.

It has also a negative effect on the column pressure operating window as shown in the next slide.

## Column operating window

- Condenser temperature
- Reboiler temperature

Column  
Temperature  
profile



The relative amount of non-distributive components has an effect on the equilibrium temperature in the top and bottom of the column.

With a fixed temperature level of the heating and cooling media the column pressure operating window will change.

To relax on the condenser equilibrium temperature part of the distillate yield can be shifted from liquid product to the vapour phase.

## Free variables

	<i>Operation</i>	<i>Flowsheeting</i>	<i>Design</i>
			<i>McCabe-Thiele</i>
• feedstock	- Flow rate - composition	component breakdown	$F \quad z_F$
• system	- Nr of trays - Feed tray	$n_{ss} \quad n_{rs}$	$T_\phi \quad \alpha$
• separation performance	- Reflux flow - Reboiler duty	ECP    SI	$x_d \quad x_b$
• operating condition	- Pressure - Feed temp	- Condenser temp - Feed flash	$R \quad U$

Multi component

The number of 'degrees of freedom' for a two-cut distillation column is:

$$F = 6 + C$$

where C is the number of variables required to define the feed inlet:  
C-1 variables for the composition and 1 variable for the feed inlet flow.

This leave 6 independent variables to define the distillation process. In general they can be grouped in three pairs.

- **System**

Two variables are required to define the system. Typical example is the number of stages in the two column sections.

- **Separation performance**

Two variables are related to the product separation. They can be either Top and Bottom qualities, ECP / SI or any other combination which sets the separation performance.

- **Operating condition**

Two variables can be assigned to the operating of the column. In general it are the free variables which can be used to optimise the process.

## Summary

- Component breakdown
  - pure components
  - pseudo components with a narrow boiling range
- Distributive components
  - Key components on an S-shaped curve
  - Separation performance given by two variables ECP and SI
- Non-distributive components
  - Effects to the Stripping and Absorption factors
  - Contribute to the tray loading of the column
  - Sets the top and bottom temperature
  - Limits the pressure operating window

Multi component

## Depropaniser operation

The Depropaniser column separates LPG into Propane and Butane.

The composition of the LPG feedstock is given:

LPG	t/d
C2	1.0
C3	60
i-C4	40
n-C4	100
i-C5	1.0
Total	202.0

Propane is a final product. The commercial specifications is :

### Propane product spec

Vapour pressure @ 100° F	200 psig maximum (14.8 bara @ 38° C)
Total C4+ content	< 2.5 vol.%

With the present column operation the recovery in the Propane product is:

- C3 = 90 %wof and
- n-C4 = 0.5 %wof.

### Exercise

- 1 What is the total C4+ content (vol %) of the Propane product.
- 2 Calculate the vapour pressure of the Propane product at 38 C.
- 3 What recovery is required for on spec Propane product.

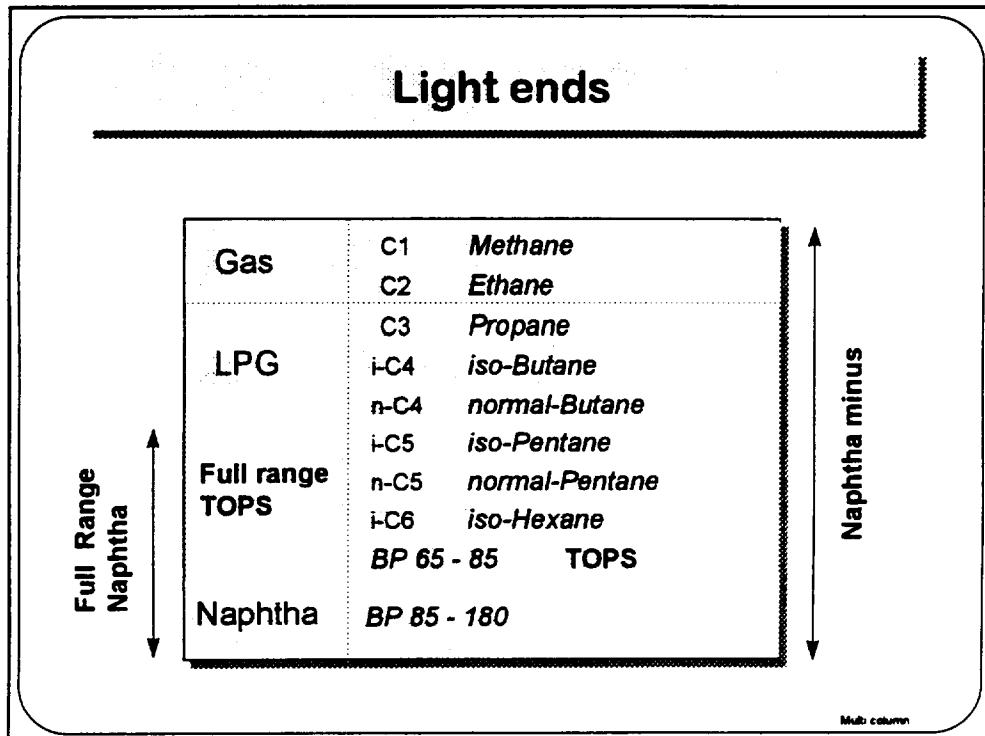




**3**

**Multi column systems**

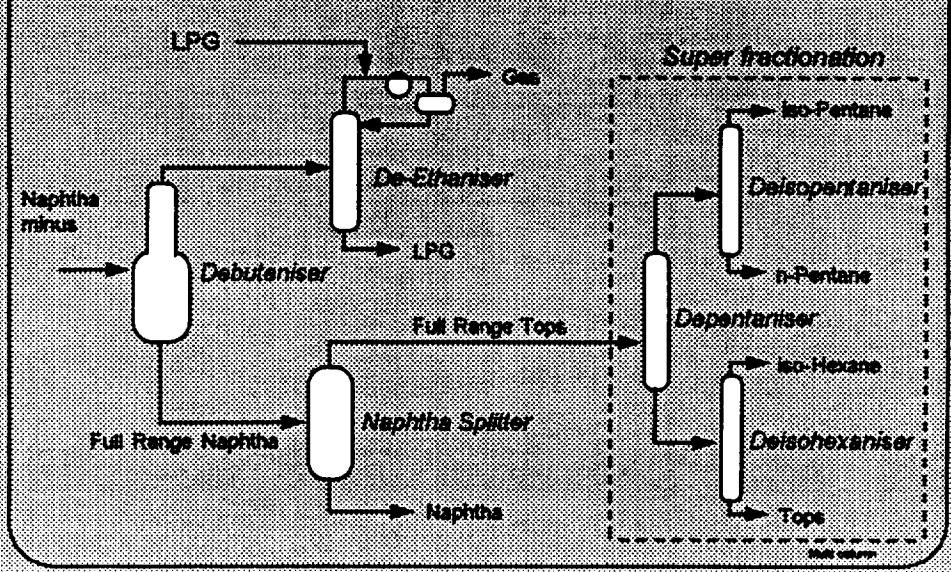
## Light ends



Naphtha minus fraction is a mixture of low boiling point hydrocarbons:

- Methane and Ethane is used as refinery fuel gas
- LPG is a final product
- Full Range TOPS is used either as a Mogas blending component or can be sold as a chemical feedstock. An option is a Super Fractionation Section to separate the iso-Pentane and iso-Hexane fractions
- Straight Run Naphtha is the routed to the Platformer unit to improve the octane number.

## Naphtha minus separation

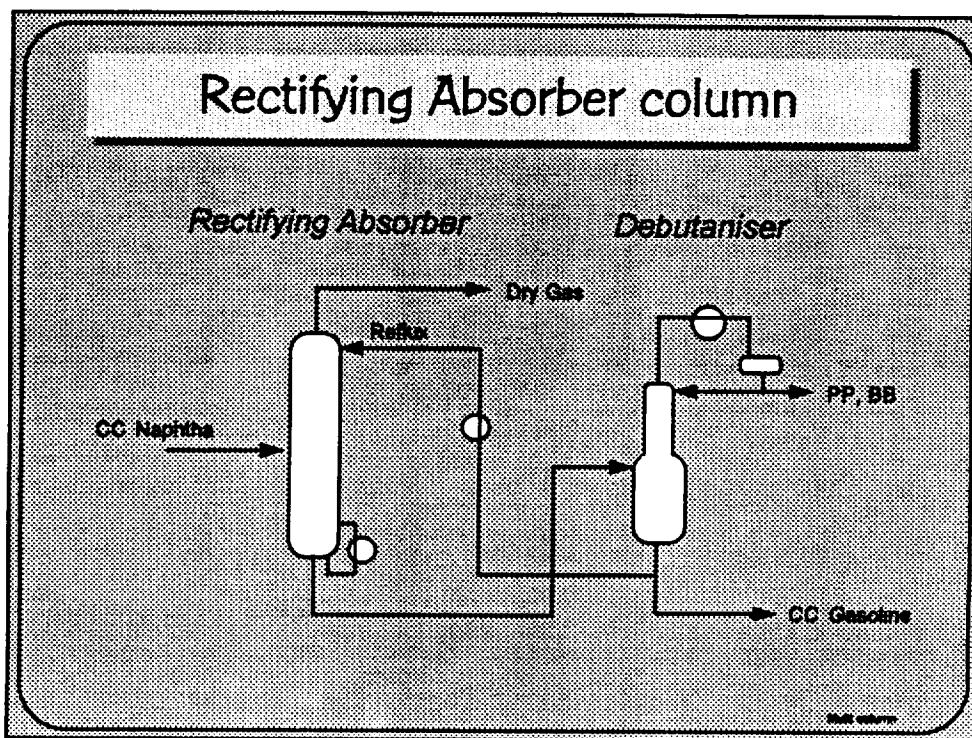


The light ends section of the CDU is a typical example of a topping multi-column system. First the LPG is separated (or topped off) and in the following step the next lighter fraction, Full Range Tops, is topped off from the FR Naphtha.

The advantage is that the second column operates at a much lower pressure. Low pressure facilitates the separation between TOPS and Naphtha.

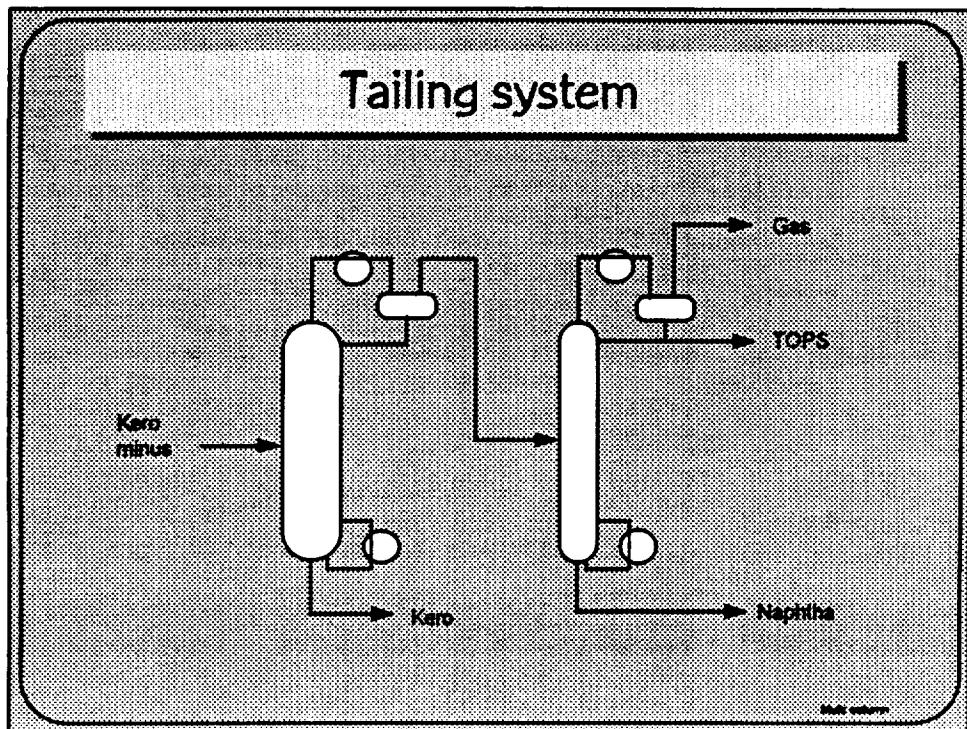
For the De-ethaniser column it is in general not possible to condense all the reflux liquid from the overhead vapour. Although the column pressure is high (~25 bar) and the condenser temperature low (~20C) most of the overhead vapour will not condense and remain as a vapour top product. To have sufficient reflux in the rectifying section of the column liquid is imported from an external source, eg. LPG liquid from the platformer stabiliser.

## Rectifying Absorber column



A typical example of reflux coming from an external source is a Rectifying Absorber column. The overhead vapours are too volatile to condense to liquid. The column has no condenser and accumulator. All reflux to the column has to be imported from a external source.

In the Cat Cracker Gas Plant a Rectifying Absorber is used to remove a relative large amount of Methane and Ethane. Down stream of the Rectifying Absorber is a Debutaniser column. The bottom product of the Debutaniser can be used as reflux for the Rectifying Absorber.

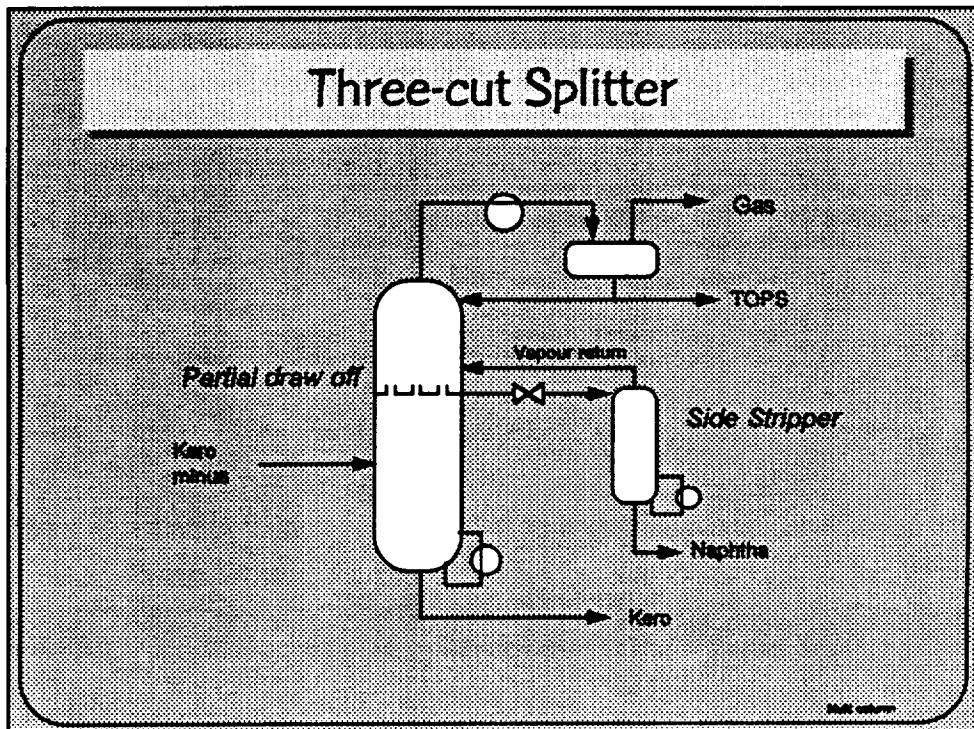


In a Tailing system the most heavy fraction is separated, or tailed-off first.

An example is the overhead product from a Kero minus CDU where a mixture of Kero, Naphtha, Tops and lighter is produced as overhead product.

The first column has a partial condenser. Only the reflux is condensed. The top product is left in the vapour phase to avoid revaporation in the second column.

For all practical purposes this type of configuration is only used in the form of a Three-cut splitter.



A three-cut splitter is a combination of two two-cut splitter columns in a tailing set up.

The column sections of the second column are detached. The rectifying section is lined up on top of the first column and the stripping section is used as side stripper.

The overhead condenser/accumulator systems of the two columns are combined.

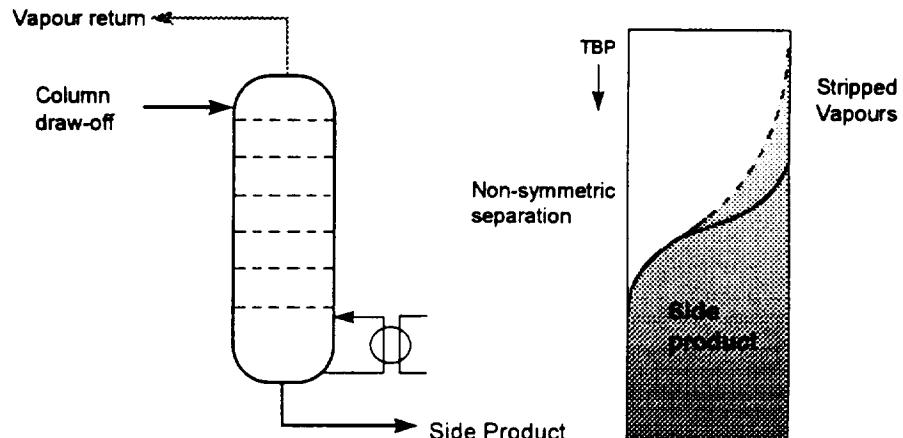
The (internal) reflux of the first column is drawn via a partial draw-off tray from the column section above. The side cut draw-off flow to the side stripper has two functions:

a- It sets the ECP between Naphtha and Kero and because of this it effects the yield on Naphtha.

b- It sets indirectly the internal reflux to the first column and thus the SI of the separation between Naphtha and Kero.

Because of the dual function of the draw off valve it is difficult to keep the Naphtha FBP constant.

## Side Stripper



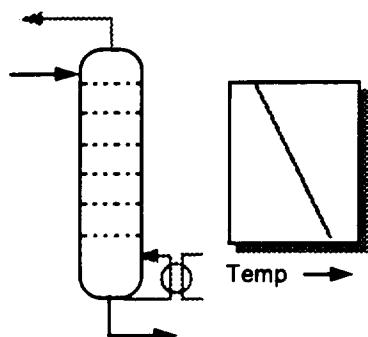
The function of the side stripper is to correct the product on flash point.

Separation of the adjacent lighter fraction in the fractionation section of the main column is typically non-symmetric. The drawn off side cut has a long front end.

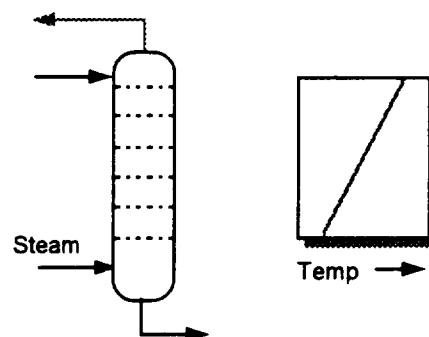
In the side stripper the light ends are stripped off and the vapour is returned to the main column.

## Reboiled or Steam stripped

Stripped vapours  
is proportional to  
Reboiler Duty



Stripped vapours  
is proportional to  
Temperature Difference



Multi column

Strippers can have a reboiler or are operated with steam in the bottom.

With a reboiler, vapour is generated from the bottom product by raising the temperature.

The maximum amount of vapour is limited by the capacity of the reboiler.

With steam stripping there is no external heat source to provide the heat for evaporation of the stripped vapours. The heat is drawn from the liquid.

The more vapours are stripped-off the lower the temperature in the bottom of the stripper will be. When there is no difference in temperature between the feed inlet and the bottom run down for sure no vapours has been stripped-off in the side stripper, no matter how much steam was used.

Because of the drop in temperature the additional stripped-off vapour will be less each time the strip steam is increased. The maximum strip steam is 3% on stripper bottom product.

Note that the steam temperature or pressure has no effect on the stripping.

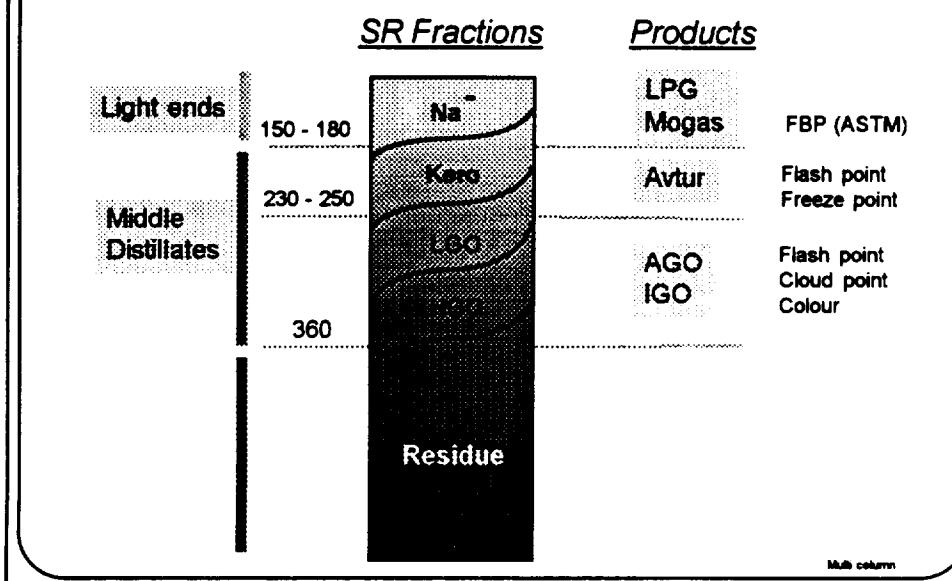
Wet steam will damage the trays.

## **Steam Stripping**

- **advantages**
  - easy to operate
  - hardly any maintenance
  - lower temperature level of bottom product
- **however . . .**
  - reduced stripping effect, maximum 3 % steam
  - stripped product is wet
  - possible steam explosion will damage trays

Multi column

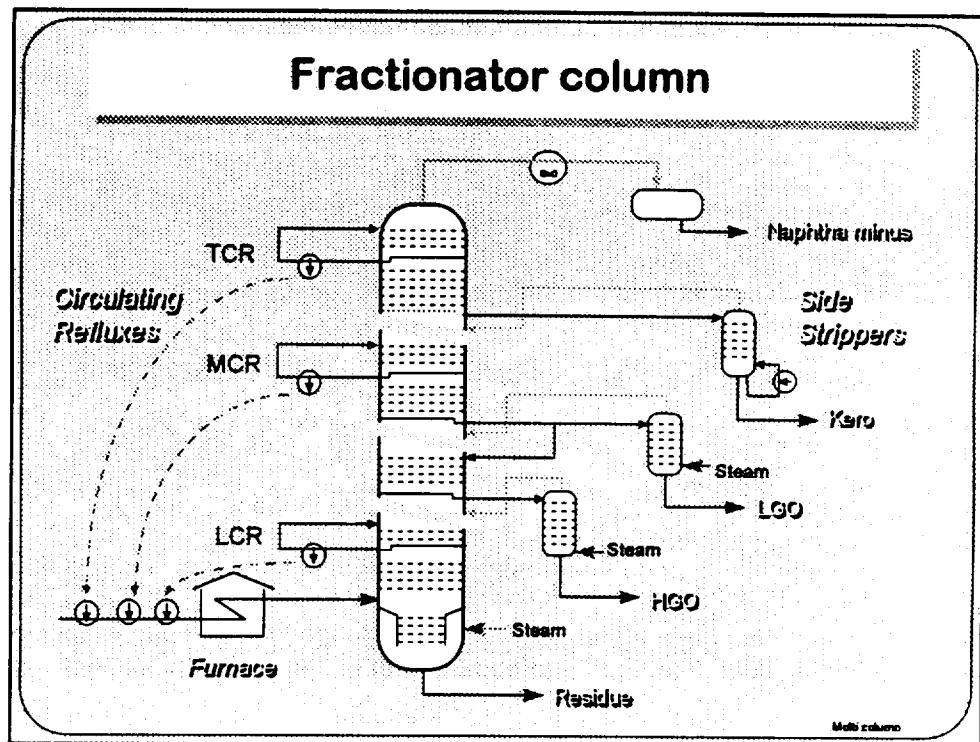
## Fractions and Products



Yields on Naphtha, Kero and GO fractions are interchangeable by adapting the cutpoints.

The cutpoints between Na minus, Kero and LGO are limited by the product specifications and/or plant constraints.

A high cutpoint between HGO and LR increases the distillate yield. It is constrained to the maximum furnace outlet temperatures to avoid unwanted cracking in the furnace.



A Fractionator column is a combination of a number of distillation columns. The rectifying sections of the columns are stacked on top of each other. The side strippers are the stripping sections.

The feed inlet to the columns is in the vapour phase and comes from the section below.

Note that a Fractionator has no external reflux. The refluxes are generated internally in the circulating reflux sections or may be drawn from the column section above like a three-cut splitter.

The separation between the fractions is made in the fractionation sections of the main column. In the side strippers only the volatile components are stripped-off to correct the front end of the product on flashpoint.

## Circulating Reflux

- **Purpose**
  - *Partial* condenser to generate internal reflux
- **advantage**
  - heat recovery at higher temperature level
  - independent operation of lower column sections
  - reduces the amount of OVHD vapours
- **however . . .**
  - less vapour/liquid traffic in higher column sections
  - large amount of liquid to be pumped around

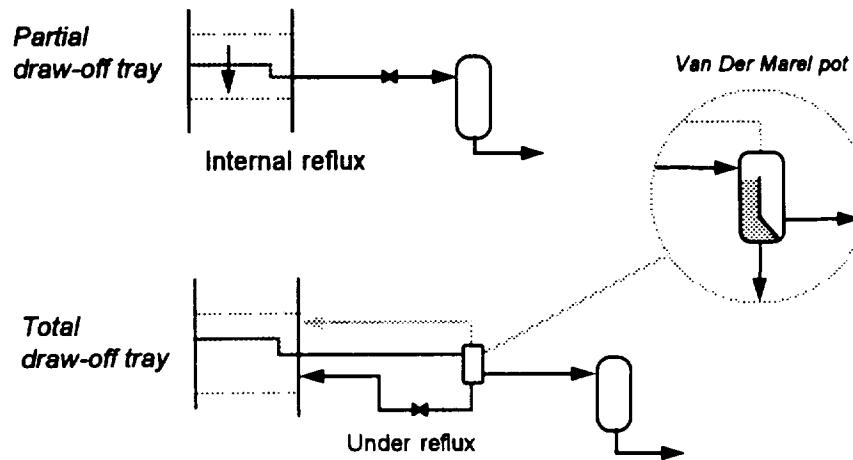
Multi column

## Column loading

- The highest loaded trays of a Fractionator are in general:
  - Top tray of rectifying section
  - Bottom tray circulation reflux section
  - Top tray **steam** stripping section
  - Bottom tray **reboiled** stripping section

Multi column

## Side cut draw-off



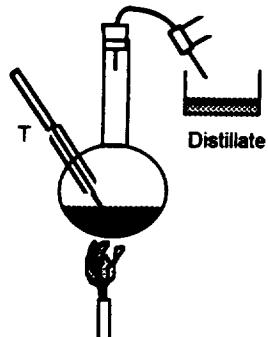
When the internal reflux is drawn from the column section above it can be routed via a partial draw-off tray or a total draw-off tray.

With a partial draw-off the flow to the side stripper is controlled with a draw-off valve. The excess of liquid collected on the draw-off tray is passed to the section below; the (not controlled) internal reflux to this section.

A total draw-off tray is in general used in combination with a 'Van Der Marel pot'. Now the under reflux is controlled. The excess of liquid is the (not controlled) draw-off flow to the side stripper.

## ASTM analysis

### Boiling ranges



% vol	Naphtha	Kero	LGO
IBP		162	204
10		178	243
30		190	256
50	102	200	268
70	125	210	281
90	150	226	304
FBP	162	240	331

No gap / No overlap  
overlap = 36 C

Multi column

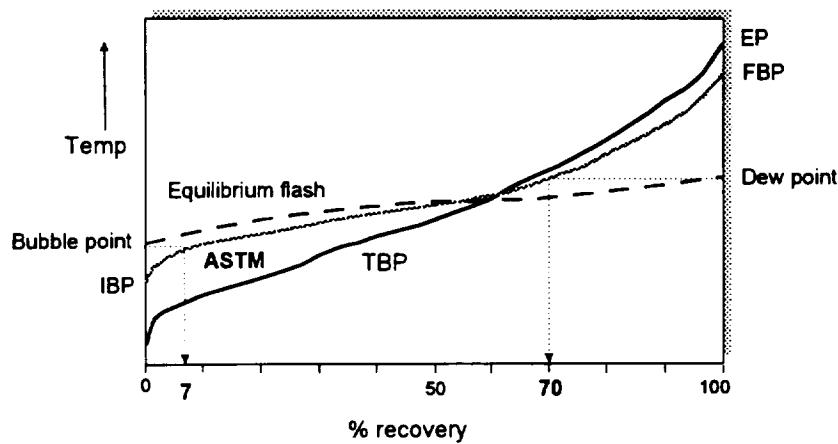
ASTM D86 test is used to observe the boiling ranges of the more heavy distillate products. The analysis is simple and has a good repeatability. For the light end fractions with narrow boiling ranges the ASTM is not very suitable.

ASTM data have no direct relation with the ECP and SI.

A typical separation between straight run Naphtha and Kero is characterised by 'no gap, no overlap'

The boiling ranges of the Middle Distillate fractions may have some overlap.

## Boiling curves



ASTM and TBP analysis are two distinct test methods. The boiling curves are completely different for the same product sample. TBP data is more accurate but the analysis requires also more time.

**ASTM recoveries are in % volume.**

IBP is the ASTM 0 % temperature

FBP is the ASTM 100 % temperature

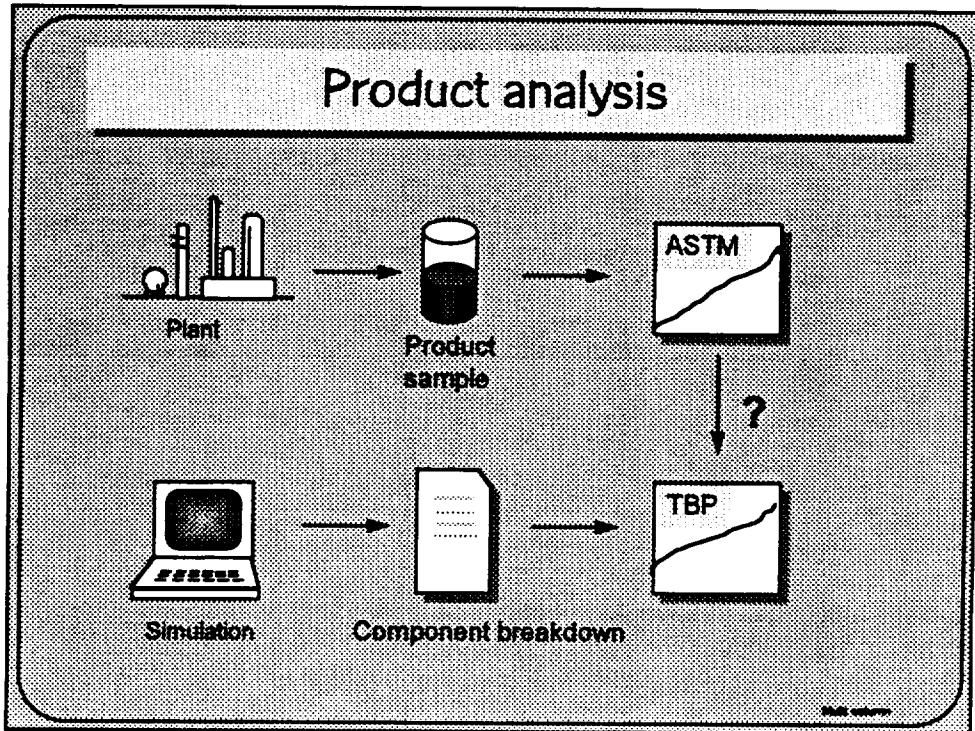
Boiling range is given by the IBP and FBP temperatures

**TBP recoveries are in % weight.**

EP is the TBP 100 % temperature

The atmospheric bubble point of a fraction is approximately equal to the 7 % ASTM point.

The atmospheric dew point is approximately equal to the 70 % ASTM point.



ASTM data is extensively used to monitor the distillation process and trim the plant operation if required.

TBP curves are suitable for analysing the composition of process streams and are used in process simulation (flow sheeting).

TBP data can not be derived from ASTM test results.



**4**

## **Distillation trays**

## 1. Introduction

The Calming Section tray is a vapour/liquid contacting device with liquid downcomers of a special box-like construction known as '*Calming Sections*' (Appendix 1). A number of these Calming Sections can be used on a tray instead of one downcomer which is typical for a conventional tray design. The Calming Sections are evenly distributed over the tray in a prescribed manner, trays with higher liquid loads being designed with more Calming Sections.

When the area occupied by the Calming Sections exceed 12-14%, the layout and construction are different, and use is made of extended 'HiFi' Calming Sections.

Calming Section trays have been successfully applied in a wide range of services including distillation, absorption and stripping. They can be designed for conditions varying from high vacuum to high pressure. Unlike conventional side-downcomer trays, Calming Section trays are easy to scale up to large diameters. Also the capacity of Calming Section trays is greater than that of conventional downcomer trays, because the typical problems in large size conventional trays are avoided, e.g. severe backmixing and operational instability due to excessive weir loading.

In addition there are certain advantages in the simple mechanical construction of Calming Section trays. The tray is supported on a continuous ring welded to the column wall, and in some cases, depending on the column diameter, by one or two support beams.

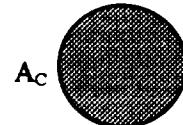
Calming Section downcomers are not fixed to the wall, so that the bolting bars and lugs necessary with conventional downcomer trays are avoided.

On Calming Section trays there are no weirs or downcomers extending from wall to wall, so that the construction is less rigid and can more easily accommodate out-of-roundness of the column than conventional downcomer trays.

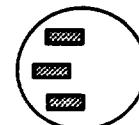
For column diameters smaller than 1.5 m the above advantages of Calming Section trays are less pronounced and it is recommended to apply Conventional Downcomer trays.

The following definitions of tray areas are used.

- Area of column cross section ( $m^2$ )

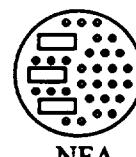


- Calming Section Area (% of  $A_c$ )



CSA

- Net Free Area (% of  $A_c$ )



NFA

- Basic Free Area (% of  $A_c$ )

For non-perforated areas see Appendix 2.



BFA

- Bubbling Area ( $m^2$ )

A<sub>BA</sub>

## 2. Basic tray performance considerations

### 2.1. Maximum operating limits

A tray is performing well if there is an intimate contacting between the vapour and liquid followed by good separation of the two phases. To establish this, the vapour and liquid streams have to be maintained within certain limits.

The *flooding capacity* is the maximum operating limit when the tray has lost its function. If flooding crops up it normally propagates throughout the column section and the separation performance is completely lost. It can be observed by a rapidly increase of pressure drop across the column section and sometimes by a sudden drop of liquid level in the bottom of the column.

Flooding can be caused by a number of phenomena which will be briefly described below. For a visual presentation of the operating limits and regimes on a distillation tray, please refer to Appendix 3.

- **Flooding by priming**

Flooding by priming occurs when the froth bed on the tray expands and reaches the tray above. This is usually the case when the vapour flow is increased beyond the maximum capacity of the tray.

- **Flooding by entrainment.**

Flooding by entrainment occurs in the so-called spray regime when vapour velocities become so high that excessive quantities of liquid are blown to the tray above. It is usually encountered at vacuum conditions where liquid loads are relatively very low.

- **Flooding by downcomer backup limitation**

Flooding by downcomer backup limitation occurs when the downcomer boxes are not able to handle the liquid load. Above the flooding point, the downcomer boxes get completely filled with liquid and overflow. The result is an inter tray space which is entirely filled with froth.

- **Flooding by downcomer choking**

Downcomer choking occurs when the liquid flow into the downcomers is obstructed by the engaged vapour. The insufficient discharging of liquid through the downcomer leads to accumulation of froth on the tray. Choking is most likely to occur when the difference between vapour and liquid density is small in combination with a too high liquid velocity at the downcomer for an efficient settling.

It will usually not be easy to determine which of the above phenomena is the cause of column flooding. To find this out exactly, one often has to do tray hydraulics calculations and/or apply diagnostic measuring techniques such as gamma ray scanning.

The operator will simply notice an increase in column pressure drop, loss of level in the bottom compartment, accompanied by a sharp drop in efficiency. For low flow-parameter applications, the sharp drop in efficiency may start at vapour loads up to 20% lower than the hydraulic limitation.

Because a tray can not be operated at its flooding point, reference is often made to the "maximum capacity", defined as the highest load at which the tray may be safely operated. This definition is somewhat arbitrary, but it generally means a loading of about 95% of its flooding capacity.

## 2.2. Minimum operating limits

The minimum operating limits of a tray are determined by weeping and insufficient downcomer sealing.

- **Weeping**

Weeping occurs when the vapour flow rate through the bubbling area is too low to prevent liquid to 'weep' through the perforations of the bubbling area. In most cases, at the minimum load, about 10-30% of weeping can be tolerated without noticing a significant reduction of the separation efficiency. For design purposes therefore, the minimum load in view of weeping is specified as that load at which 10% of weeping occurs.

- **Sealing**

Sealing of the downcomer by the descending liquid is under all circumstances desired. When too little liquid is flowing out of the downcomer, vapour may flow through the downcomers instead of through the bubbling area (loss of downcomer sealing). The bypassing of the bubbling area by the vapour will result in a lower separation efficiency. To ensure sufficient liquid flow through the downcomers a minimum liquid flow is specified for design purposes and a check is performed to establish whether the vapour flow is sufficiently high to prevent liquid flowing through the bubbling area instead of into the downcomer.

## 2.3. Column capacity

The maximum capacity of a column results from the capacity of the most heavy loaded tray to handle the internal traffic of vapour and liquid.

### Tray loading

The tray capacity is based on the vapour and liquid traffic on the tray.

We take vapour *to the tray*:

$Q_v$	= actual volumetric flow rate	(m <sup>3</sup> /s)
$M_v$	= mass flow rate	(kg/s)
$\rho_v$	= density of vapour	(kg/m <sup>3</sup> )

and liquid *from the tray*:

$Q_l$	= actual volumetric flow rate	(m <sup>3</sup> /s)
$M_l$	= mass flow rate of liquid	(kg/s)
$\rho_l$	= density of liquid	(kg/m <sup>3</sup> )

The flow parameter is defined by:

$$\varphi = \frac{M_l}{M_v} \sqrt{\frac{\rho_v}{\rho_l}} = \frac{Q_l}{Q_v} \sqrt{\frac{\rho_l}{\rho_v}}$$

The limiting capacity  $\lambda_{max}$  of Calming Section trays as a function of the flow parameter  $\varphi$  is shown in Figure 2.1. This correlation is based on the concept that for every flow parameter  $\varphi$  there is an optimal tray layout with a and an acceptable tray efficiency.

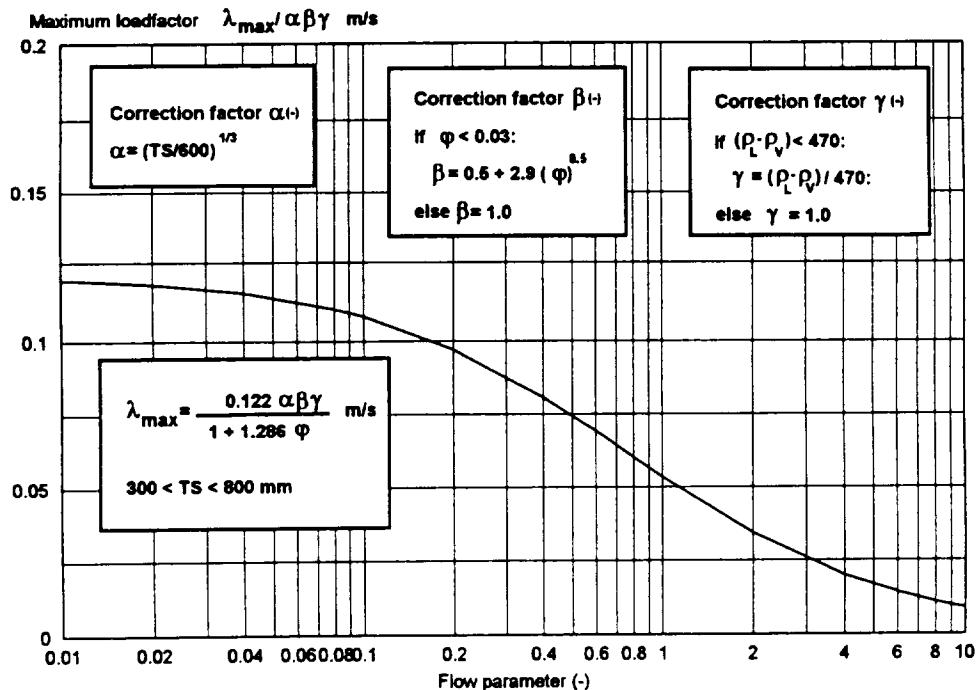


Figure 2.1: Maximum load factor of Calming Section trays.

The maximum capacity curve shown in the Figure is given by the equation:

$$\lambda_{\max} = \frac{0.122 \alpha \beta \gamma}{1 + 1.286 \varphi} \text{ (m/s)}$$

The correction factors  $\alpha$ ,  $\beta$  and  $\gamma$  are given as follows:

- **Tray spacing correction factor**

$$\alpha = \left( \frac{TS}{600} \right)^{\frac{1}{3}}$$

where TS is the tray spacing ranging from 300 to 800 mm.

- **Entrainment correction factor**

$$\beta = 0.5 + 2.9\sqrt{\varphi} \quad \text{with } \beta \leq 1.0 \text{ (for } \varphi < 0.03\text{)}$$

It takes into account the capacity reduction due to entrainment and is based on an entrainment rate of about 20% of the total liquid rate at maximum load.

- **Density difference correction factor**

$$\gamma = \frac{(\rho_L - \rho_v)}{470} \quad \text{with } \gamma \leq 1.0$$

The density difference correction factor  $\gamma$  should be applied only when the density difference is less than  $470 \text{ kg/m}^3$ .

### Foaming systems

Trays operating with foaming systems are found to flood at a *lower load factor* than with non-foaming systems. The main causes of foaming are surface properties which can not be predicted in most cases. For this reason, the foaming tendency is best judged on the basis of experience with similar systems. Examples of foaming liquids are some crude oils, heavy residues, absorption and extraction solvents.

For foaming systems the calculated maximum load factor should be multiplied by a correction factor F (F less than 1.0) which is related to the severity of the expected foaming. The maximum load factor multiplied with F then represents the limiting capacity of Calming Section trays with the particular system considered. A list of typical foam factors is given in table 2.1.

**Table 2.1: Typical foam factors for foaming systems**

System	Foam factor, F
Deethaniser	1.0 *)
Depropaniser	1.0 *)
Rectifying absorber	0.65
Sponge oil section	0.85
Sulfinol absorber	0.85
Sulfinol regenerator	0.85
EO absorber	0.85
EO stripper	0.85
EO dehydrator	0.85
Glycol regenerator	0.85
Butadiene, butylenes extractive distillation with acetonitrile/water mixture	0.73
CO <sub>2</sub> strippers using steam (EO plants)	0.73
Amine regenerator	0.71
SCOT absorber	0.71
Amine absorber	0.71
Acetylene absorber (dimethyl formamide)	0.70
Long residue stripper	0.65
Short residue stripper	0.65
CO <sub>2</sub> absorber using K <sub>2</sub> CO <sub>3</sub> solution (EO plants)	0.60
Condensate stripper	0.65

#### Notes:

\*) If free water is present: F = 0.85

\*\*) The foam factor F applies to the column cross sectional area. The increase in downcomer area is larger than that of the column cross sectional area (See Equations A3.2.6, A3.3.1, A3.3.2 and A3.3.3)

### Column loading

The actual loading of a column section is expressed as a percentage of ( $F \cdot \lambda_{\max}$ ); the flooding condition for the most heavily loaded tray in that section.

The column load factor is defined as:

$$\lambda_c = \frac{Q_v}{A_c} \sqrt{\frac{\rho_v}{\rho_l - \rho_v}} \quad (\text{m/s})$$

or as  $\lambda_c = \frac{Q^*}{A_c}$

where  $Q^*$  is the volumetric load factor defined by:

$$Q^* = Q_v \sqrt{\frac{\rho_v}{\rho_l - \rho_v}} \quad (\text{m}^3/\text{s})$$

and  $A_c$  the cross-sectional area of the column in  $\text{m}^2$ .

The column loading as percentage flood is:

$$P_f = \frac{\lambda_c}{\lambda_{\max}} \cdot \frac{1}{F} \cdot 100\%$$

The choice of the maximum allowable percentage of flood is matter of judgement. Normally it is within 75% and 85% for new designs. For actual operating conditions values up to  $P_f < 95\%$  could well be acceptable.

### 2.4 Calming section area

Calming section tray layouts are standardised. This means that for each combination of column diameter and calming section area the number, size and distribution of calming sections has been determined.

The required calming section area results from the maximum allowable superficial liquid velocity in the tops of the calming sections. For trays with a tray spacing of 600 mm the maximum allowable downcomer entrance velocity is 0.18 m/s.

For systems where the disengagement of vapour is more difficult, such as foaming systems or systems with a low density difference, lower velocities should be applied. For tray spacing lower than 600 mm less height is available for vapour/liquid disengagement and therefore lower velocities should be applied.

The maximum allowable *liquid velocity* in the top of the Calming Sections is:

$$\text{For } 0.01 < \varphi < 0.28 \quad U_{DC} = \frac{12.2 \varphi}{(4 + 35.7 \varphi)} \frac{F^2}{(1 + 1.286 \varphi)} \left( \frac{\rho_l - \rho_v}{\rho_l} \right)^{\frac{1}{2}} \left( \frac{TS}{600} \right)^{\frac{1}{3}}$$

At very low flow parameters ( $\varphi < 0.01$ ) the liquid is highly aerated and therefore lower velocities should be applied.

$$\text{For } \varphi > 0.28 \quad U_{DC} = 0.18 F^2 \left( \frac{\rho_l - \rho_v}{\rho_l} \right)^{\frac{1}{2}} * \left( \frac{TS}{600} \right)^{\frac{1}{3}} \quad (\text{m/s})$$

For higher flow parameters ( $\varphi > 0.1$ ) and liquid densities below 600 kg/m<sup>3</sup> an additional check for downcomer choking is required. When choking is found, the calming section area should be increased.

For existing trays higher velocities (maximum 30%) can be accepted at the cost of a proportional decrease in vapour handling capacity.

Except for the cases where the flow parameter  $\varphi < 0.01$ , the CSA can be calculated using the above equations for the maximum allowable liquid velocity and:

$$A_{DCT} = Q_l \frac{100}{P_F} \frac{1}{U_{DC}} \quad (\text{m}^2)$$

where  $Q_l$  = volumetric liquid flow rate at design, m<sup>3</sup>/s.

The Calming Section Area (CSA) is given by:

$$CSA = \frac{A_{DCT}}{A_c} 100 \quad (\%)$$

The nearest standard layout whose CSA is not less than required should be chosen from the list of standard layouts. These standard layouts have been developed to optimally use the total area available, provide good heat and mass transfer etc.

Note that the minimum calming section area is 4%.

## 2.5 Net Free Area

The net free area is defined as the area of the perforations expressed as a percentage of the column area. The NFA should be chosen such that excessive weeping will not occur at the lowest anticipated vapour load, whilst at the same time the tray pressure at maximum load should not be too great. Sieve trays are preferred due to lower cost, valve trays are used when the turndown required cannot be met using sieve perforations.

The NFA for sieve trays is usually between 6 and 14% and for valve trays between 12 and 18%.

### Weeping and sealing

The *weep point* is defined as that vapour load - or rather lambda hole - at which 10% of the net liquid flow rate leaks through the bubbling panel perforations.

The *seal point* is defined as the point at which 100% of the liquid is weeping through the bubbling area perforations. It should be noted that vapour bypassing can already start above this point if the liquid slot velocity is below 0.10-0.15 m/s (e.g. with large slot areas). In these cases, the seal point could be reached before the weep point. Hence to ensure proper sealing of the downcomers separate check has to be carried out. If the downcomers are not properly sealed, the separation efficiency will drastically drop and stable operation of the tray will hardly be possible.

In normal cases the minimum allowable vapour load will usually be in the range of 20-50% of the design load and is verified with the weep and seal point correlation's.

If  $\lambda_{HW} > \lambda_{HS}$ , the weeping hole load factor  $\lambda_{HW}$  should be taken as the minimum load. In exceptional cases, where high turndown is required,  $\lambda_{HS}$  can be taken as the minimum load, but this will result in some loss of efficiency for vapour loads below  $\lambda_{HW}$ .

For  $\lambda_{HW} < \lambda_{HS}$ , the sealing hole load factor  $\lambda_{HS}$  should be taken as the minimum load.

Note that in most cases the sealing criteria will be limiting at high flow parameter applications.

In all cases, the hole load factor at minimum load should not be less than 0.2 m/s to ensure a proper tray performance at minimum load.

### Sieve trays

In the absence of other constraints the NFA should be chosen such that the hole load factor at design load is about 0.8 m/s, which gives:

$$NFA = 125 \lambda_c \quad \% \quad (1)$$

The minimum allowable vapour load is determined either by the weep point or by the seal point of the lowest loaded tray in a section and can be calculated as follows:

- **Weep point**

The weep point for Calming Section Sieve trays is calculated with:

$$\lambda_{HW} = 0.28 \left( \frac{H_w}{50} \right)^{\frac{1}{4}} \varphi^{-0.11} \quad \text{m/s}$$

For HiFi Calming Section Sieve trays, in case the area underneath the discharge slots is perforated, weeping starts at higher hole load factors due to impingement of the liquid from the slots on the bubbling area and the turndown capacity of the tray will be less and the weep point is to be taken equal to:

$$\lambda_{HW} = 0.41 \quad \text{m/s}$$

- **Seal point:**

$$\lambda_{HS} = \frac{0.39}{(1 + 4.2 \cdot \varphi)} \left( \frac{H_w}{50} \right)^{\frac{1}{2}} \left( 1 + \frac{A_s}{A_h} \right) \quad \text{m/s}$$

for downcomers with seal pans  $A_s = 0$  and  $\varphi \leq 1.0$

For CSS and HiFi trays the preferred sieve hole diameter is 12 mm. For applications with very low BFA, sieve holes with a diameter of 6 mm may be considered. For fouling services, the hole diameter may be increased up to 20 mm, but it should be noted that the turndown capacity reduces with larger hole diameters.

Sieve perforations should be square-edged, punched so as to be burr-side up. For a given pressure drop the weep rate of trays with square edged is similar to that of trays with edge rounded sieve holes and hence the cheaper square edged holes are preferred.

It should be noted that the weeping - and sealing correlation's mentioned in this manual are only valid for trays with square edged sieve holes.

### Valve trays

Valve trays are normally equipped with the Shell snap-in valves. The dry tray pressure drop of this valve is shown in Appendix 7.

In the absence of other constraints the NFA should be chosen so that at the design load all valves are just fully open. This occurs at the hole load factor at which the horizontal part of the characteristic in Figure intersects the steeply rising part.

This gives:

$$\lambda_{Hdes} = \frac{12.3}{\sqrt{\rho_1}} \quad \text{m/s}$$

$$A_H = \frac{Q^*}{\lambda_{Hdes}} \quad \text{m}^2$$

$$NFA = \frac{A_H}{A_c} \cdot 100 \quad \%$$

$$N_V = \frac{A_H}{0.001257}$$

where  $N_V$  is the number of valves with holes of 40 mm.

- Weep point

$$\lambda_{HW} = 0.056 \left( \frac{H_w}{50} \right)^{\frac{1}{2}} \varphi^{-0.3} \quad \text{m/s}$$

- Seal point

$$\lambda_{HS} = \frac{0.17}{(1 + 4.2 \varphi)} \left( \frac{H_w}{50} \right)^{\frac{1}{2}} \left( 1 + \frac{3 A_s}{A_H} \right) \quad \text{m/s}$$

for downcomers with seal pans  $A_s = 0$  and  $\varphi \leq 1.0$

Apart from ensuring that the trays are operated above their weep - or seal point, it should also be ensured that the trays are *operated outside the unstable region* where valve oscillations are possible (Appendix 7). If the weep point or seal point calculated according to the above described methods are within the unstable region, the minimum hole load factor is to be taken equal to:

$$\lambda_{HSstab} = \frac{60}{\rho_1} + 0.1 \quad \text{m/s}$$

## 2.6. Pressure drop and downcomer backup

Due to the pressure drop over the tray, the liquid through the downcomer flows into a zone with a slightly higher pressure. To balance this pressure difference, a certain liquid head is required in the downcomer; called 'downcomer backup'. If the amount of froth in the downcomer exceeds the downcomer height, the tray will flood.

Calculation of the downcomer backup is hence an important check on the workability of the tray design. The pressure drop and downcomer backup check should be carried out for maximum (flooding) flow rates to ensure that there is no premature downcomer limitation. To calculate the downcomer backup a pressure balance around the tray as shown in figure 2.6.1 below is set up. The contributions to the backup are given below.

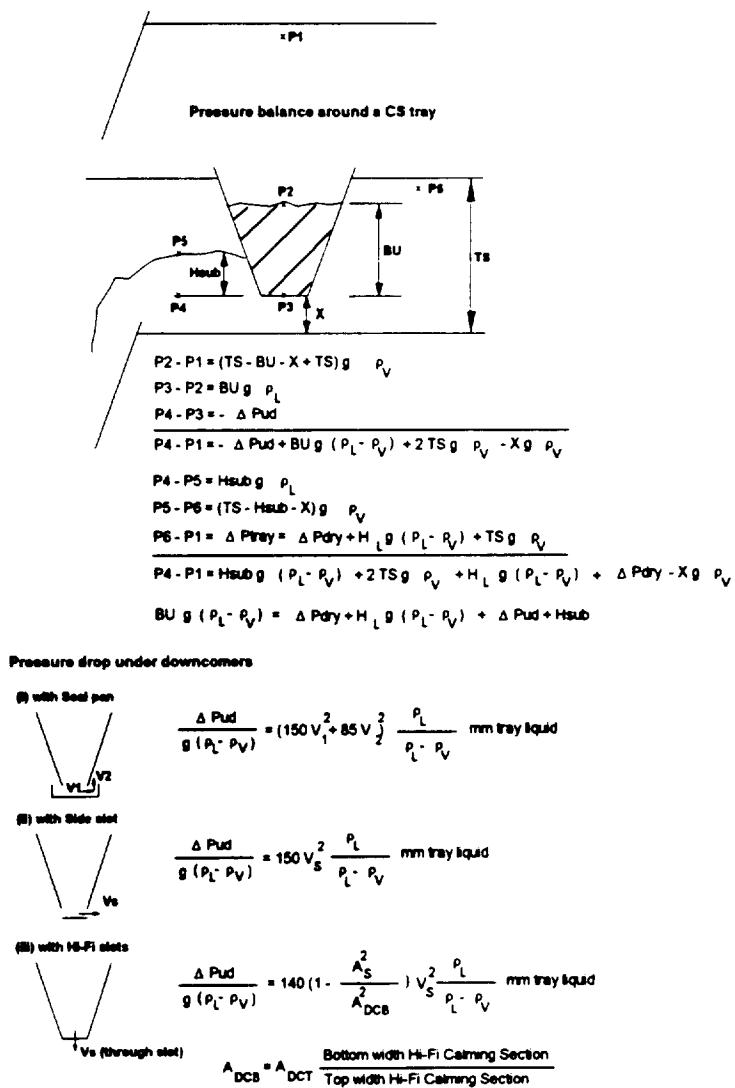


Fig.2.6.1 Pressure balance around a tray and downcomer outlet pressure drop.

### Tray pressure drop

The pressure drop and downcomer back-up should be checked at maximum (flooding) flow rates to ensure that there is no premature downcomer limitation. Use the values

$$Q_{L_{\max}} = Q_l \frac{100}{P_{Fl}} \quad \text{m}^3/\text{s}$$

$$Q_{V_{\max}} = Q_v \frac{100}{P_{Fl}} \quad \text{m}^3/\text{s}$$

The tray pressure drop is the sum of the dry tray pressure drop, given in mm tray liquid  $\Delta P_{dry}/(g \cdot \Delta \rho)$ , and the clear liquid height on the tray ( $H_l$ ).

For valve trays equipped with Shell snap-in valves,  $\Delta P_{dry}/(g \cdot \Delta \rho)$  can be read direct from Figure 8a

For sieve trays:

$$\frac{\Delta P_{dry}}{g (\rho_l - \rho_v)} = C \lambda_H^2 \quad \text{mm tray liquid}$$

where the coefficient C is given in Figure 8b as a function of BFA and hole diameter/tray thickness ratio. Calming Section trays are commonly made of stainless steel 2 mm thick and sieve holes of 8-15 mm are usually applied.

The liquid head on the tray ( $H_l$ ) is the sum of the liquid content of the dispersion below the top of the weir and the crest over the weir:

$$H_l = \varepsilon_l H_w + H_{ow} \quad \text{mm tray liquid}$$

where  $\varepsilon_l$  is the liquid fraction in the dispersion, as given in Figures 8a and 8b.

In these figures  $\lambda_{BA}$  is the load factor based on bubbling area ( $= Q^*/(A_c - A_{DCT})$ ). The crest over the weir  $H_{ow}$  can be calculated with:

$$H_{ow} = 830 \left( \frac{Q_{L_{\max}}}{L_w} \right)^{\frac{2}{3}} \left( \frac{\rho_l}{\rho_l - \rho_v} \right)^{\frac{1}{3}} \quad \text{mm tray liquid}$$

The factor 830 in this equation takes into account the aeration of the liquid flowing over the weir. For Calming Sections the weir length is given by the sum of the long sides plus one top width. For the semi-segmental HiFi Calming Section downcomers the length of the curved side is not included.

The total tray pressure drop is then given by:

$$\frac{\Delta P_{tray}}{g (\rho_l - \rho_v)} = \frac{\Delta P_{dry}}{g (\rho_l - \rho_v)} + H_l + TS \frac{\rho_v}{(\rho_l - \rho_v)}$$

The minimum allowable weir height is 25 mm. Weir heights higher than  $TS/6$  are not recommended since it will result in a reduction of capacity.

### Downcomer backup

The pressure balance in Figure 2.6.1 shows that the downcomer back-up increases with increasing tray pressure drop, pressure drop under downcomer ( $\Delta P_{ud} / (g \Delta \rho)$ ) and downcomer submergence ( $H_{sub}$ ). We have:

$$BU = \frac{\Delta P_{dry}}{g (\rho_l - \rho_v)} + H_1 + \frac{\Delta P_{ud}}{g (\rho_l - \rho_v)} + H_{sub} \text{ mm tray liquid}$$

The calculation of  $\Delta P_{ud} / (g \Delta \rho)$  for Calming Sections with side slots, seal pans or HiFi Calming Section slots is indicated in Figure 2.6.1 above.

A conservative estimate of the back-up due to the submergence of the Calming Sections in the froth bed at maximum loading is:

$$H_{sub} = (CSH - H_w) \frac{H_1}{TS} \text{ mm tray liquid}$$

### Allowable backup

The liquid back-up calculated from equation above (BU) can be converted into a height of froth with the aid of Appendix 11. Note that the back-up in CSS and CSV trays, whether in mm froth or mm liquid, is always measured from the level of the top of the slots or the top of the seal pan. The Calming Section height is measured from the same level.

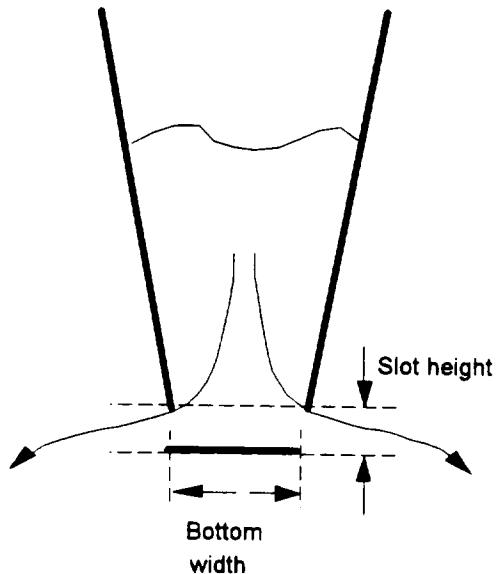
If we denote the calculated froth height in the downcomer as  $BU_{fr}$ , then the requirement is that at flood:

$$\frac{BU_{fr}}{\sqrt{F}} < 0.9 CSH$$

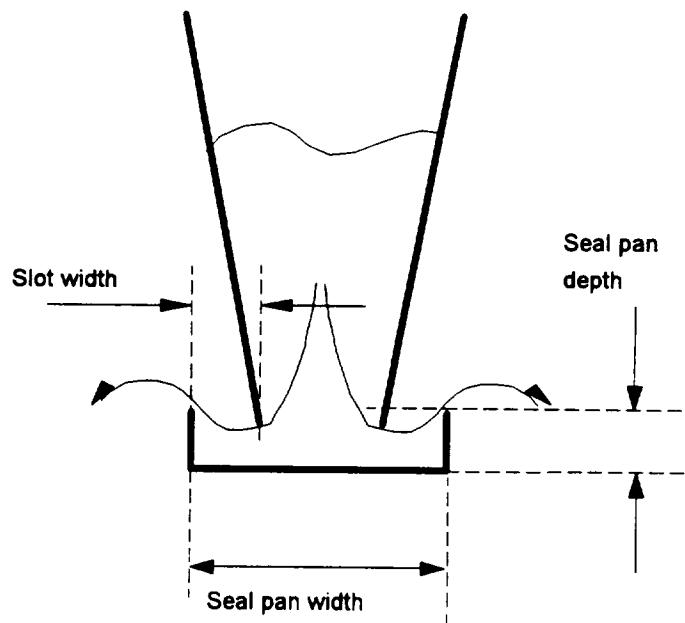
where F is the foam factor (table 2.1).

Calming Sections may have slot heights between 15 and 40 mm. The height of the slot is usually chosen such that the slot velocity is about 0.6 m/s for Calming Section trays and 0.8 m/s for HiFi trays. To ensure proper downcomer sealing, the minimum slot velocity should always be above 0.10 m/s for Calming Section trays and above 0.15 m/s for HiFi trays.

In case a higher turndown is required, seal pans can be installed. A seal pan should have twice the bottom width of a Calming section and the depth should be half the bottom width.



Side view Calming Section box  
with open slot



Side view Calming Section box  
with seal pan

## 2.7. Entrainment

Severe entrainment of liquid from a tray to the tray above may be one of the phenomena which can limit tray capacity. However, when the entrainment rate is not excessively high, flooding may not be observed and 'only' a reduction in separation efficiency may be caused (entrainment rates > 20%). In general entrainment may become important at low flow parameters ( $\phi < 0.03$ ), at high vapour loads ( $\lambda > 0.07 \text{ m/s}$ ), low tray spacing and with systems with low surface tension ( $\sigma < 10 \text{ mN/m}$ ). A reliable correlation to determine the entrainment rate does not exist. For vacuum services Appendix 6 can be used to estimate the entrainment rate at a given flow rate.

## 2.8. Tray efficiency

Process simulations of distillation columns are usually carried out in terms of 'equilibrium' or 'theoretical' stages. The number of real or 'actual' trays required for the given separation is then calculated using an overall efficiency figure  $E_o$ , defined as the ratio between the number of theoretical stages required and the number of actual trays.

The tray efficiency is mainly determined by the system involved, but is also dependent on the tray design. In general acceptable tray efficiencies can be expected when the liquid height on a tray is above 40 mm and the flow path length more than 200 mm.

System	Overall Tray Efficiency $E_o$
Demethaniser	0.90
Deethaniser	0.90
Wesseling Stripper	0.80
PP splitter	0.85
Depropaniser	0.85
C4-splitter	0.85
BB splitter	0.85
Deisopentaniser	0.80
Deisohexaniser	0.75
Absorber	0.75
Sponge oil absorber	0.60
Debutaniser	0.80
Gasoline splitter	0.75
CDU/CCU main fractionation rectifying sections Naptha/Kero, Naptha/ LCO	0.70
Kero/LGO, LCO/HCO	0.60
LGO/HGO, HCO/Slurry	0.50
HGO/Residue	0.45
Vacuum distillates	0.40
Stripping sections HT Naptha, Reboiled	0.75
HDS Kero, Reboiled	0.65
HDS Gasoil, Reboiled	0.55
Kero/LCO, Reboiled	0.65
Kero/LCO, Steam	0.45
Gasoil/HCO, Reboiled	0.55
Gasoil/HCO, Steam	0.40
Vacuum Gasoil/Luboil distillates	0.35
Long Residue	0.25
Short Residue	0.20
Water stripping	0.15
H2S stripping	0.15

### 3. Summary

$Q_v$	=	actual volumetric flow rate of vapour to the tray, $\text{m}^3/\text{s}$
$M_v$	=	mass flow rate of vapour to the tray, $\text{kg/s}$
$Q_l$	=	actual volumetric flow rate of liquid from the tray, $\text{m}^3/\text{s}$
$M_l$	=	mass flow rate of liquid from the tray, $\text{kg/s}$
$\rho_v$	=	density of vapour flowing to the tray, $\text{kg/m}^3$
$\rho_l$	=	density of liquid flowing from the tray, $\text{kg/m}^3$

The flow parameter

$$\varphi = \frac{M_l}{M_v} \sqrt{\frac{\rho_v}{\rho_l}} = \frac{Q_l}{Q_v} \sqrt{\frac{\rho_l}{\rho_v}}$$

The volumetric load factor

$$Q^* = Q_v \sqrt{\frac{\rho_v}{\rho_l - \rho_v}} \quad \text{m}^3/\text{s}$$

Tray (flooding) capacity

$$\lambda_{\max} = \frac{0.122 \alpha \beta \gamma}{1 + 1.286 \varphi} \quad \text{m/s}$$

Tray spacing correction factor

$$\alpha = \left( \frac{TS}{600} \right)^{\frac{1}{3}}, \quad \text{TS is the tray spacing in mm.}$$

Entrainment correction factor

$$\beta = 0.5 + 2.9 \sqrt{\varphi} \quad \text{for } \varphi < 0.03$$

Density difference correction factor

$$\gamma = \frac{(\rho_l - \rho_v)}{470} \quad \text{for } \Delta\rho < 470 \text{ kg/m}^3.$$

Foam correction factor: F ( $0 < F < 1.0$ ), see table 2.1.

Column load factor

$$\lambda_c = \frac{Q^*}{A_c} = \frac{Q_v}{A_c} \sqrt{\frac{\rho_v}{\rho_l - \rho_v}} \quad \text{m/s}$$

where  $A_c$  is the cross-sectional area of the column in  $\text{m}^2$ .

$$\lambda_c = \frac{Q^*}{A_c} = \frac{P_{fl}}{100} F \lambda_{\max} \quad \text{m/s}$$

Column cross sectional area

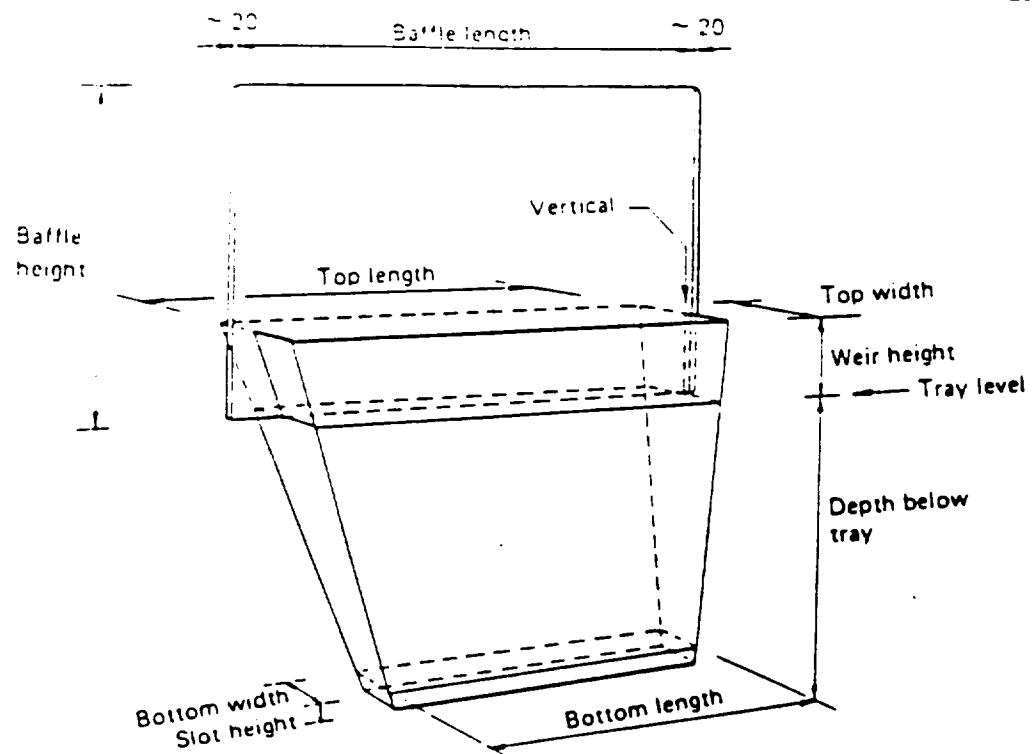
$$A_c = Q^* \frac{100}{P_{fl}} \frac{1}{F} \frac{(1 + 1.286 \varphi)}{0.122 \alpha \beta \gamma} \quad \text{m}^2$$

## SYMBOLS LIST

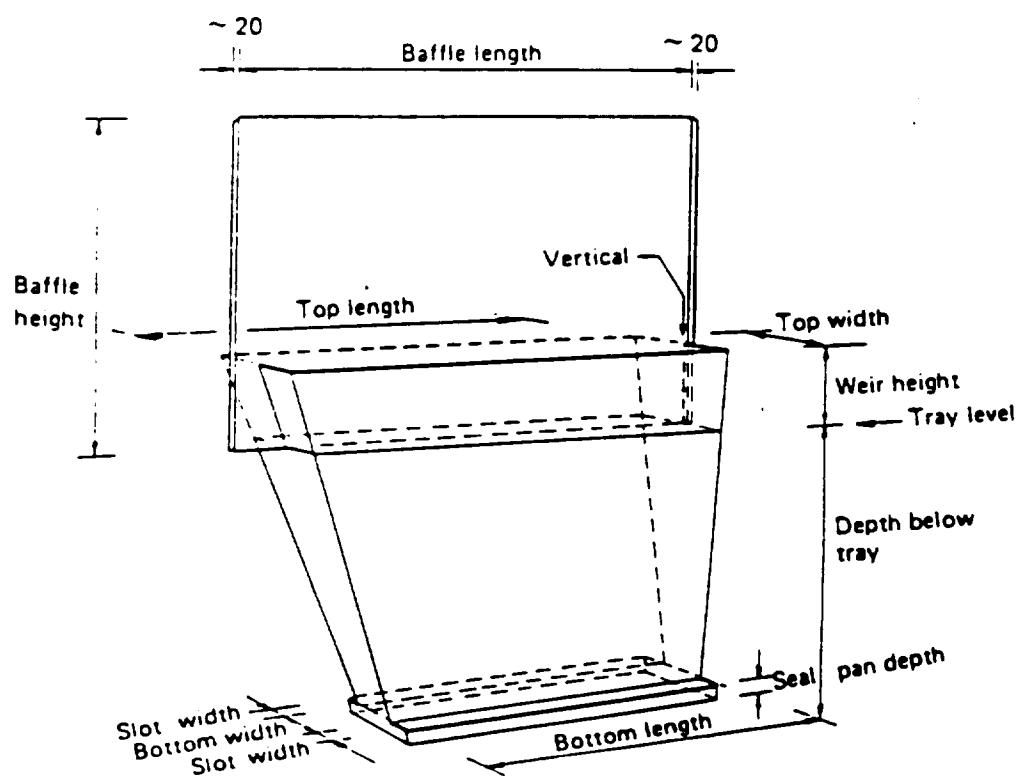
$A_c$	Cross-sectional area of column ( $m^2$ )
$A_{DCB}$	Area of downcomer bottom ( $m^2$ )
$A_{DCT}$	Area of downcomer top ( $m^2$ )
$A_H$	Hole area ( $m^2$ )
$A_{NPA}$	Non perforated area underneath HiFi Calming Section slots ( $m^2$ )
$A_s$	Slot area ( $m^2$ )
$B$	Downcomer bottom width (mm)
$BFA$	Basic free area (%)
$BU$	Dowcomer back-up (mm tray liquid)
$BU_{fr}$	Froth height in dowcomer (mm)
$C$	Dry tray pressure drop coefficient ( $mm/(m/s)^2$ )
$C_f$	Capacity factor (-)
$CSA$	Calming Section area (%)
$CSH$	Calming Section height (mm)
$CSH_{max}$	Maximum Calming Section height (mm)
$CSS$	Calming Section sieve tray
$CSV$	Calming Section valve tray
$C_{ssp}$	Calming Section spacing
$D_c$	Column diameter (m)
$E_o$	Overall efficiency (%)
$F$	Foam factor (-)
$g$	Gravitational constant ( $m/s^2$ )
$H_L$	Clear liquid height on tray (mm tray liquid)
$H_{ow}$	Liquid crest over Calming Section weir (mm tray liquid)
$H_{sub}$	Liquid back-up in the dowcomer due to submergence (mm tray liquid)
$H_{wn}$	Height of the overflow weir (mm)
$H_w$	Outlet weir height of tray (mm)
$HIFIS$	HiFi Calming Section Sieve tray
$HIFIV$	HiFi Calming Section Valve tray
$L_w$	Total weir length on the tray (m)
$L_{FP}$	Flow path length (mm)
$M_v$	Mass flow rate of vapour flow to the tray at design (kg/s)
$M_l$	Mass flow rate of liquid flow from the tray at design (kg/s)
$NFA$	Net free area (%)
$N_{act}$	Actual number of trays for given separation (-)
$N_{th}$	Number of theoretical stages for given separation (-)
$N_v$	Number of valves (-)
$P_{Fl}$	Percentage of flood (%)
$PW$	Panel width (mm)
$Q^*$	Volumetric load factor at design ( $m^3/s$ )
$Q_v$	Volumetric flow rate of vapour to the tray at design ( $m^3/s$ )
$Q_{Vmax}$	Maximum value of $Q_v$ ( $m^3/s$ )
$Q_l$	Volumetric flow of liquid from the tray at design ( $m^3/s$ )
$Q_{Lmax}$	Maximum of $Q_l$ ( $m^3/s$ )
$Q_{Mmax}$	Maximum volumetric flow rate of vapour and liquid phase ( $m^3/s$ )
$TS$	Tray spacing (mm)
$TL$	Top length of a Calming Section (mm)

$U_{dcMax}$	Max. superficial liquid velocity in the top of the Calming Sections (m/s)
$V_1, V_2$	Slot velocities in seal pan (m/s)
$V_s$	Liquid velocity in slots (m/s)
$W_r$	Width of tray support ring (m)
$X$	Clearance underneath Calming Section downcomer (mm)
$\alpha$	Tray spacing correction factor (-)
$\beta$	Entrainment correction factor (-)
$\gamma$	Density difference correction factor (-)
$\Delta P_{dry}$	Dry tray pressure drop (mPa)
$\Delta P_{ud}$	Pressure drop under downcomer (mPa)
$\Delta P_{tr}$	Tray pressure drop (mPa)
$\Delta \rho$	Density difference of liquid and vapour phase (kg/m <sup>3</sup> )
$\varepsilon_l$	Liquid fraction in dispersion on tray (-)
$\lambda_c$	Load factor based on column area (m/s)
$\lambda_{BA}$	Load factor based on bubbling area (m/s)
$\lambda_H$	Hole load factor (m/s)
$\lambda_{Hdes}$	Hole load factor at design load (m/s)
$\lambda_{Hstab}$	Hole load factor at which unstable tray behaviour may occur (m/s)
$\lambda_{HS}$	Hole load factor at seal point (m/s)
$\lambda_{HW}$	Hole load factor at weep point (m/s)
$\lambda_{max}$	Maximum value of $\lambda$ (m/s)
$\lambda_{maxCh}$	Maximum value of $\lambda$ limited by choking (m/s)
$\rho_v$	Vapour density (kg/m <sup>3</sup> )
$\rho_l$	Liquid density (kg/m <sup>3</sup> )
$\varphi$	Flow parameter (-)
$\sigma$	Surface tension (mN/m)

APPENDIX 1.



CALMING SECTION WITH OPEN SLOT



CALMING SECTION WITH SEAL PAN

## Appendix 2

**Non-perforated areas of the tray**

- The area of the Calming Sections plus a narrow zone around them. For estimating purposes take this area as  $1.6 A_{DCT} m^2$ .
- The area occupied by major support beams, which may be assumed to be 100 mm wide. This area is therefore  $0.1 L_B m^2$ , where  $L_B$  is the total length of major beams in metres.
- Surface of the tray support ring (40 mm wide below 2.0 m diameter, 50 mm up to 6.0 m and 60 mm above 6.0 m). This area is therefore  $(\pi D_c W_r) m^2$ , where  $W_r$  is the width of the ring in metres.
- The non-perforated area underneath the liquid discharge slots of HiFi Calming Section sieve trays. The non-perforated area underneath the liquid discharge slots of HiFi Calming Section sieve trays  $A_{NPA}$  can be estimated as the HiFi Calming Section bottom area  $A_{DCB} m^2$ . For all other tray types or in case perforations are located underneath the HiFi Calming Section downcomers,  $A_{NPA}$  is equal to zero.

The BFA is then given by:

$$BFA = NFA \frac{A_c}{(A_c - 1.6 A_{DCT} - 0.1 L_B - \pi D_c W_r - A_{NPA})} \%$$

Appendix 3.

Operating limits and regimes of a distillation tray

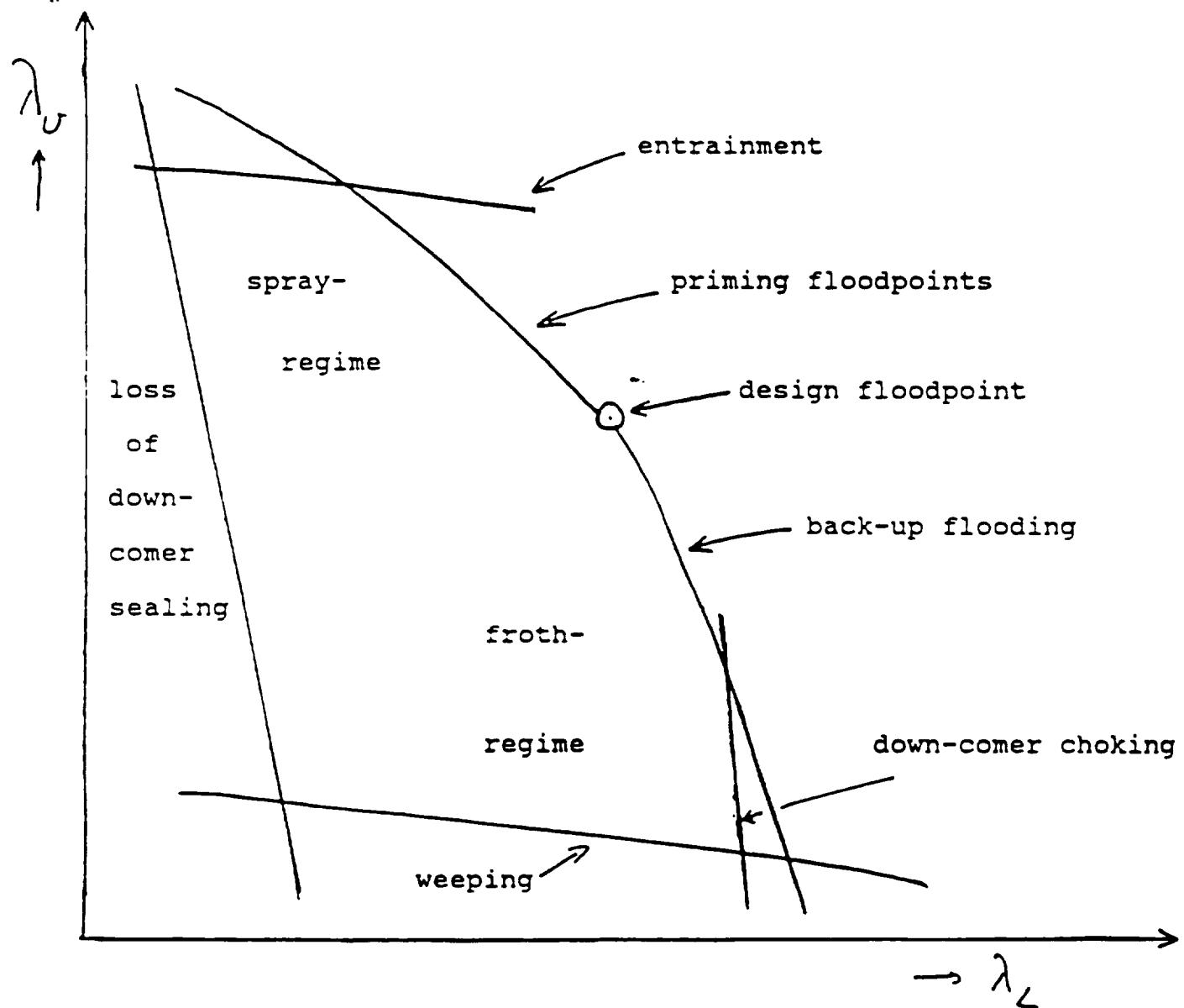
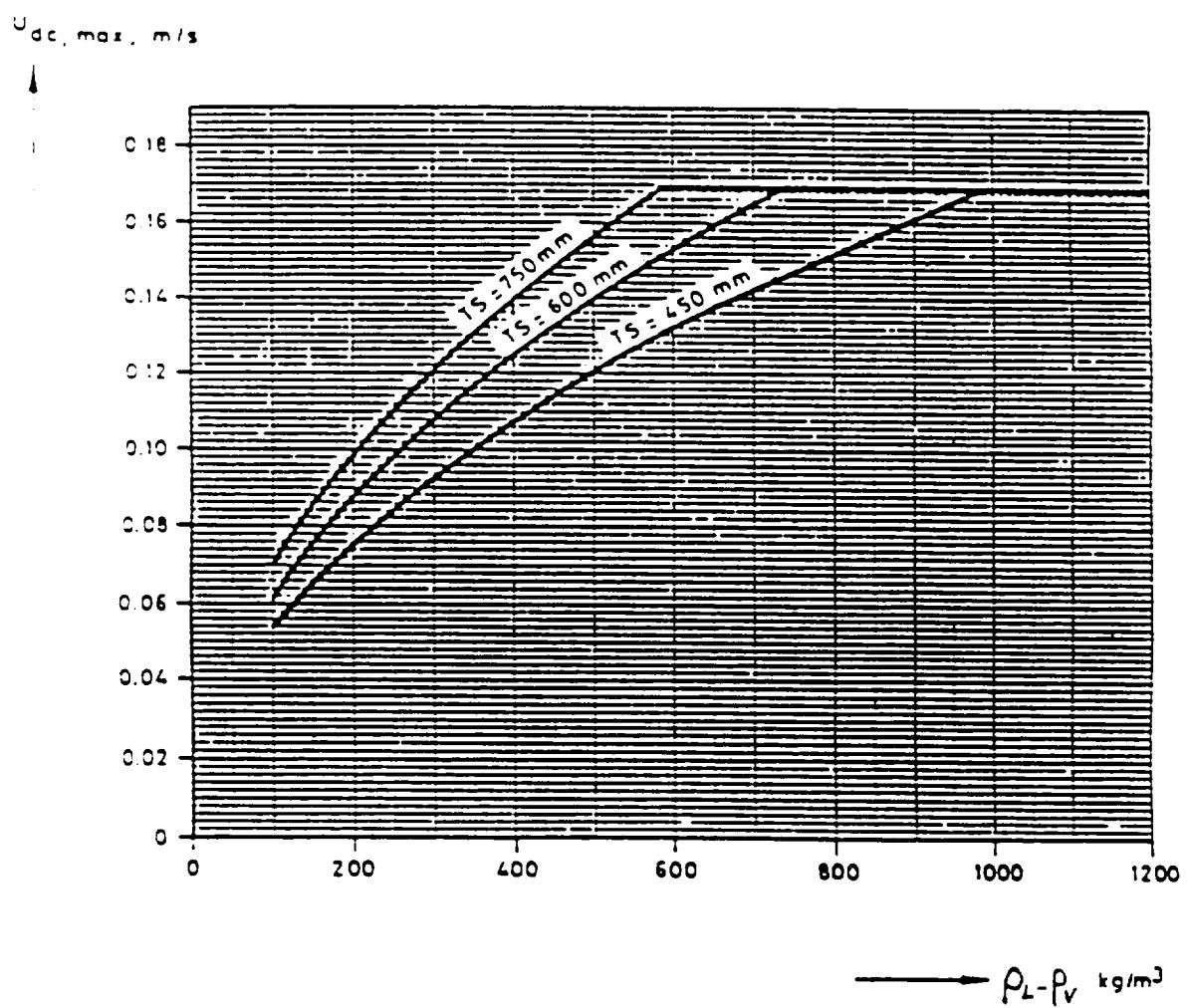


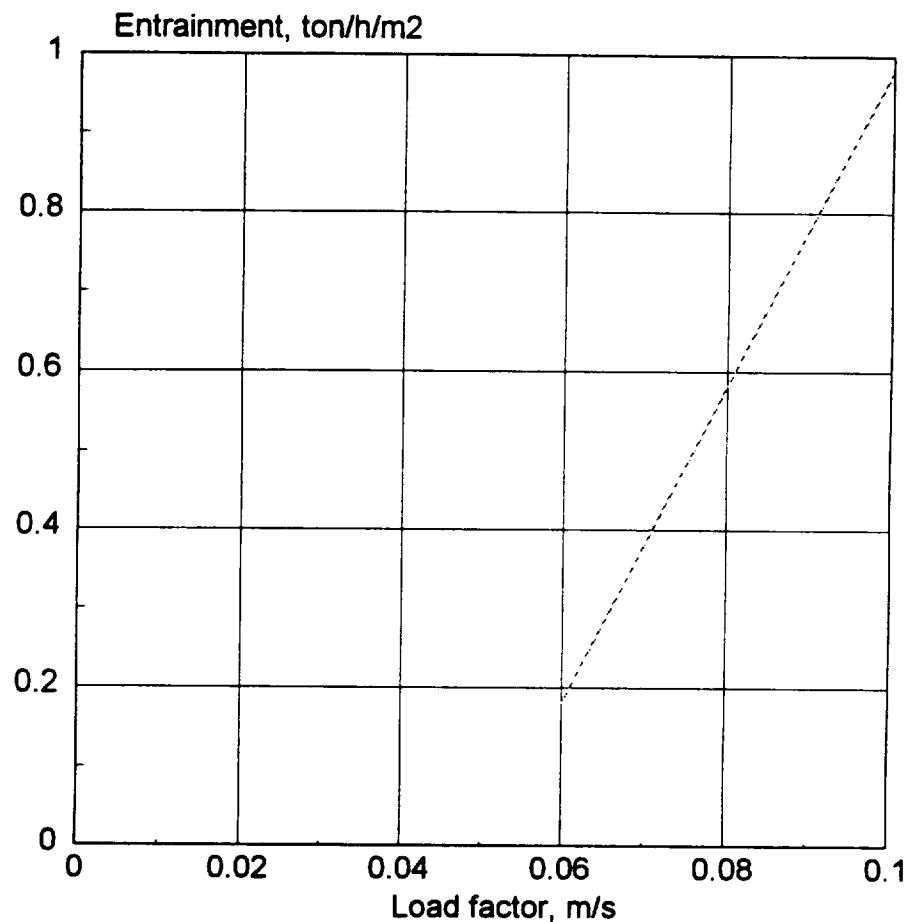
figure 3.

Appendix 5.



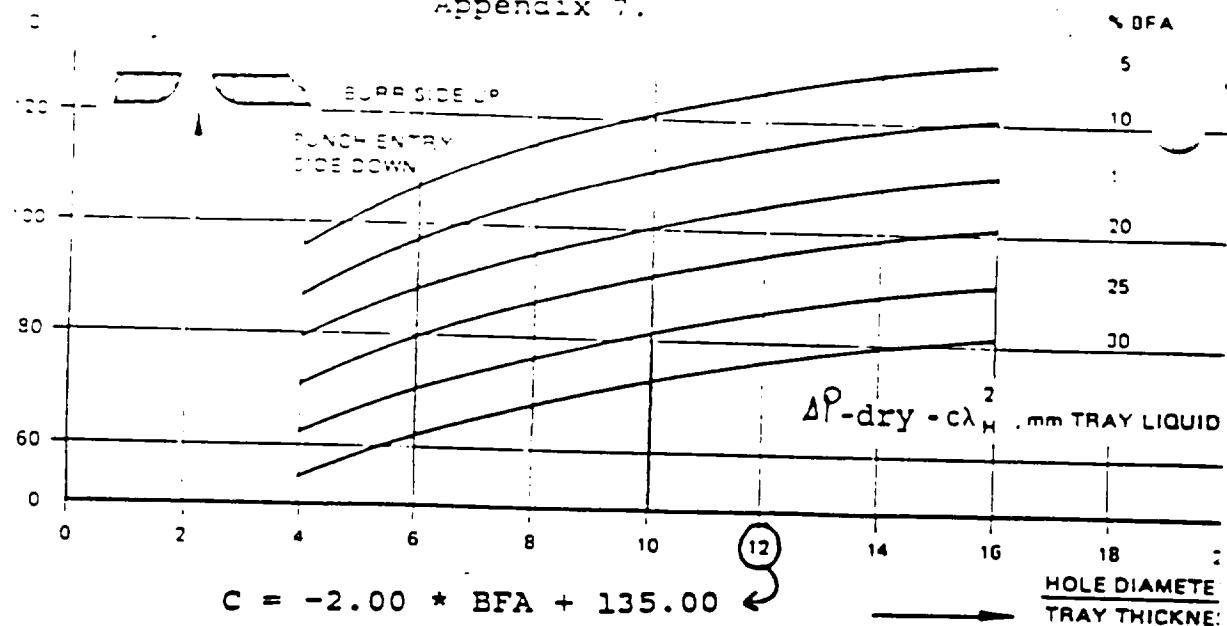
(DRAWN FOR FOAM FACTOR,  $F=1.0$ )

## Appendix 6



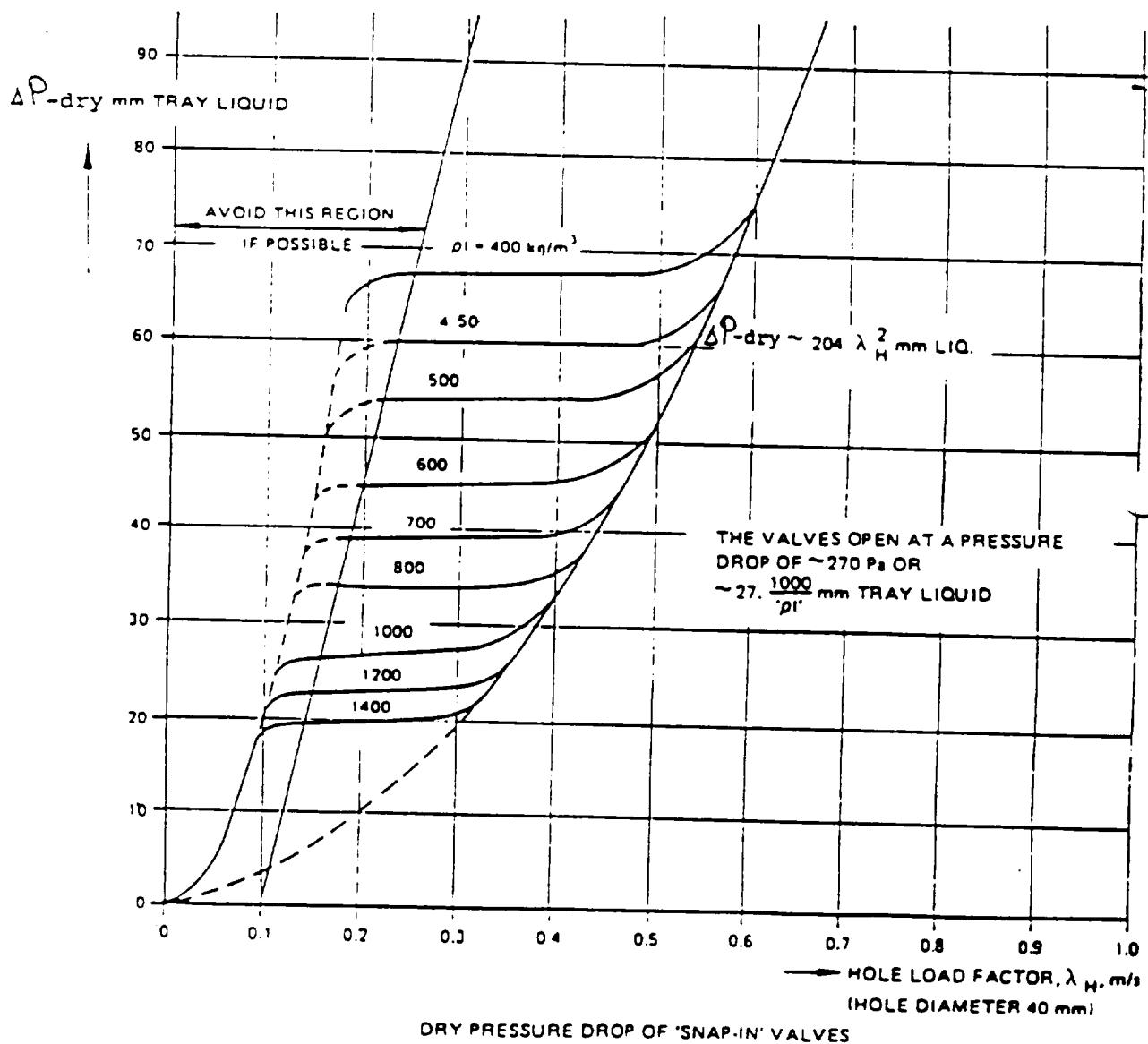
**Figure A3.8.1:** Entrainment of trays under vacuum conditions valid for liquid loads of 4 ton/h/m<sup>2</sup> and lower

Appendix 7.



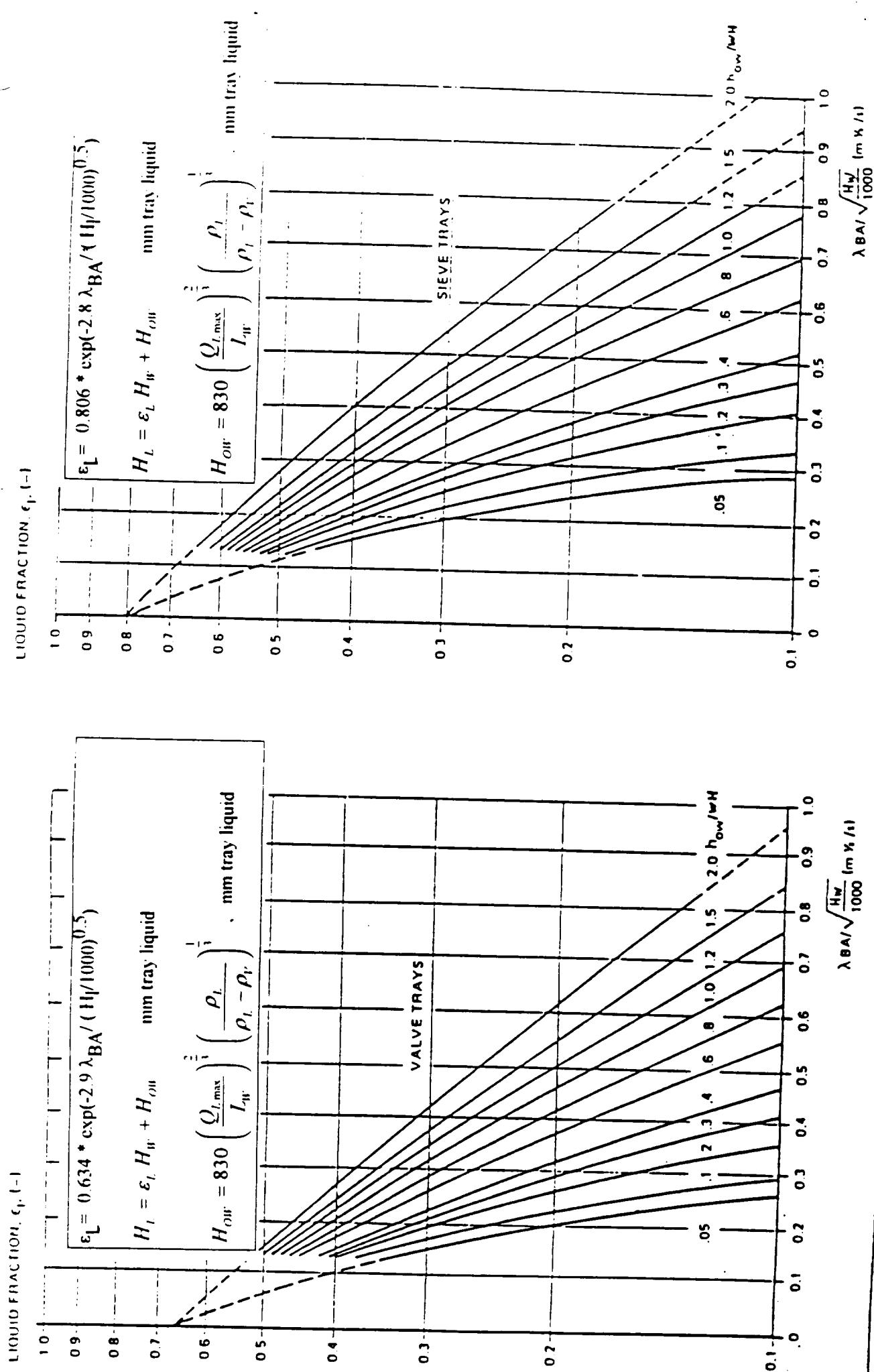
DRY TRAY PRESSURE DROP COEFFICIENT OF COMMERCIAL SIEVE PLATE

note 1: standard tray thickness is 2 mm.  
note 2: no burrs are any more applied,  
sieve-holes should be square edged.



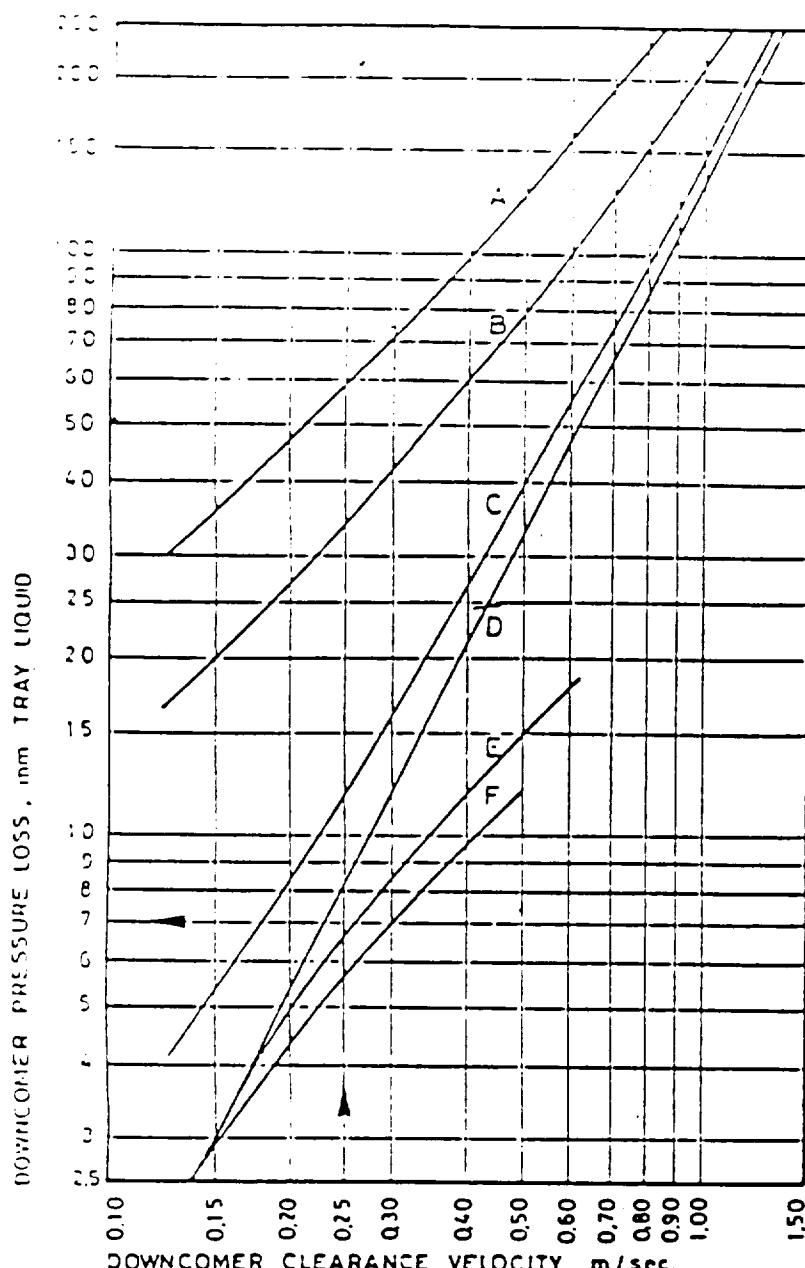
DRY PRESSURE DROP

figure 7.

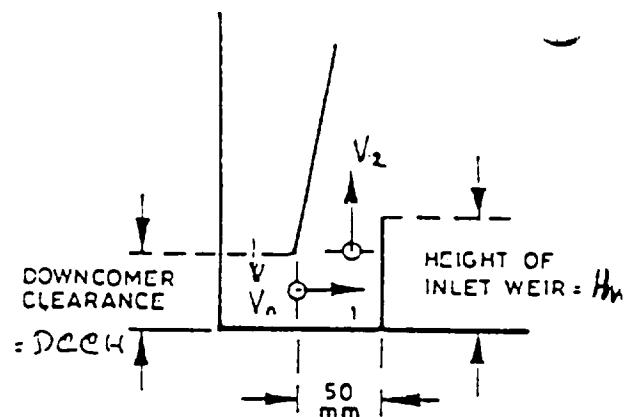


LIQUID FRACTION IN DISPERSIONS ON VALVE TRAY

FIG. 8a. LIQUID FRACTION IN DISPERSIONS ON SIEVE TRAYS FIG. 8b.



(actually this should be:  $V \sqrt{\frac{S}{\rho_f g}}$ )



	DOWNCOMER CLEARANCE mm	HEIGHT OF INLET WEIR mm
A	38	115
B	38	50
C	38	20
D	38	-
E	75	-
F	115	-

#### NOTE :

IN CASE OF A DOWNCOMER CONSIDERABLY DIFFERENT FROM THE ONES SHOWN HERE OVER THE DOWNCOMER PRESSURE LOSS CAN BE ESTIMATED AS FOLLOWS:-

1. DOWNCOMER FRICTION LOSS AND UNDERFLOW LOSSES =  $153 (V_1)^2$  mm TRAY LIQUID.
2. WHEN AN INLET WEIR IS INSTALLED ADD:-

- c. FRICTION LOSS IN UPFLOW AREA =  $82 (V_2)^2$  mm TRAY LIQUID
- d. STATIC HEIGHT OF FROTH IN UPFLOW AREA = WEIR HEIGHT  $\times$  FROTH DENSITY
- e. CREST OVER INLET WEIR.

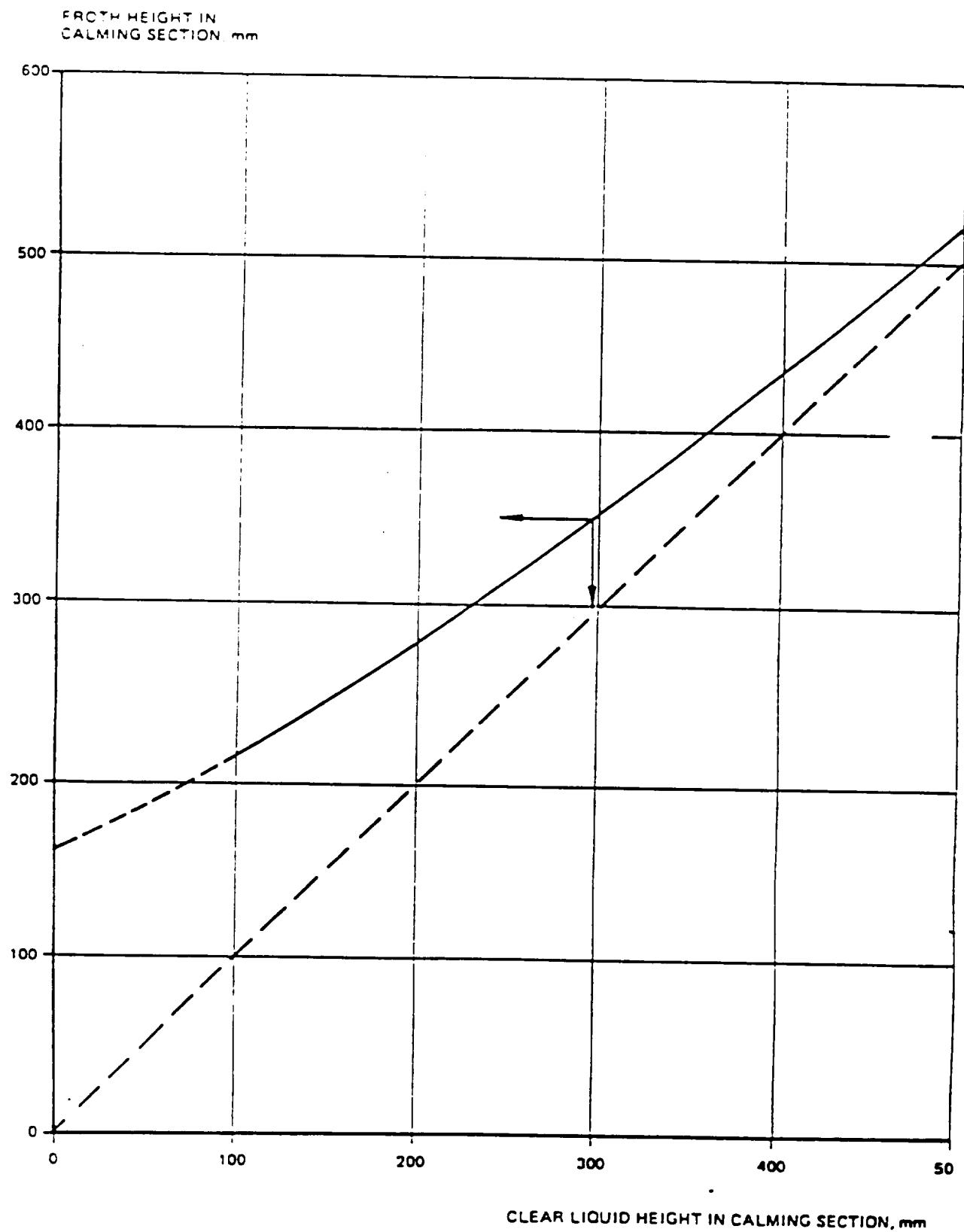
THE DOWNCOMER PRESSURE LOSS IS THE SUM OF 1 AND 2

$V_1$  = CLEAR LIQUID VELOCITY IN THE CLEARANCE AREA , m/sec.

$V_2$  = CLEAR LIQUID VELOCITY IN THE UPFLOW AREA , m/sec.

FOR DOWNCOMER FROTH DENSITY SEE FIG. 3 - 57

Appendix 11.



CONFIDENTIAL

Appendix 13.

CS DOWNCOMER DIMENSIONS

Top width, mm	Top length, T1, mm	Bottom width, B mm	Bottom length, mm
140	< 400	60	T1-100
140	400 ≥ T1 < 600	60	T1-150
140	600	60	T1-200
-----			
160	< 450	70	T1-75
160	≥ 450	70	T1-100
-----			
180	all lengths	80	T1-50

Hi-Fi downcomer dimensions

top width	top length	bottom width	bottom
140	all	equals half	as
160	lengths	the top width	top length
180	from	plus 10	
200	beam		
220	to wall		

slot area SA in Hi-Fi bottom plat consists of transverse slots of 15 mm width at 15 mm spacing in groups of about 8 slots and a spacing between the groups of 60 mm. Total slot area should always be below 40 % of bottom area.

Note on Flow Path Length

Recent work on tray efficiency has resulted in new guidelines on layout of downcommers for Hi-Fi trays.

Layouts with short flow path length ( $L_{FP}$ ) should be avoided. The maximum CS top width is now 300 mm instead of 220 mm.

$L_{FP} < 150$  mm is not acceptable

$L_{FP} > 150 - 200$  should be avoided

## **Exercise**

### **Tray trouble shooting**

A kero minus Debutaniser column is giving problems with regard to the separation between top and bottom products. It is suspected that the trays are not performing properly.

A performance test run of the Debutaniser was carried out. All relevant process data is given in Figure 1.

The tray data/requisition sheets from the design book are attached.

#### **Exercise**

- Identify the most probable cause of the poor separation.
- What can be done immediately to improve column operation.
- What modifications do you suggest for a longer term solution.

Be creative. As usual there are more possibilities to solve the problem.

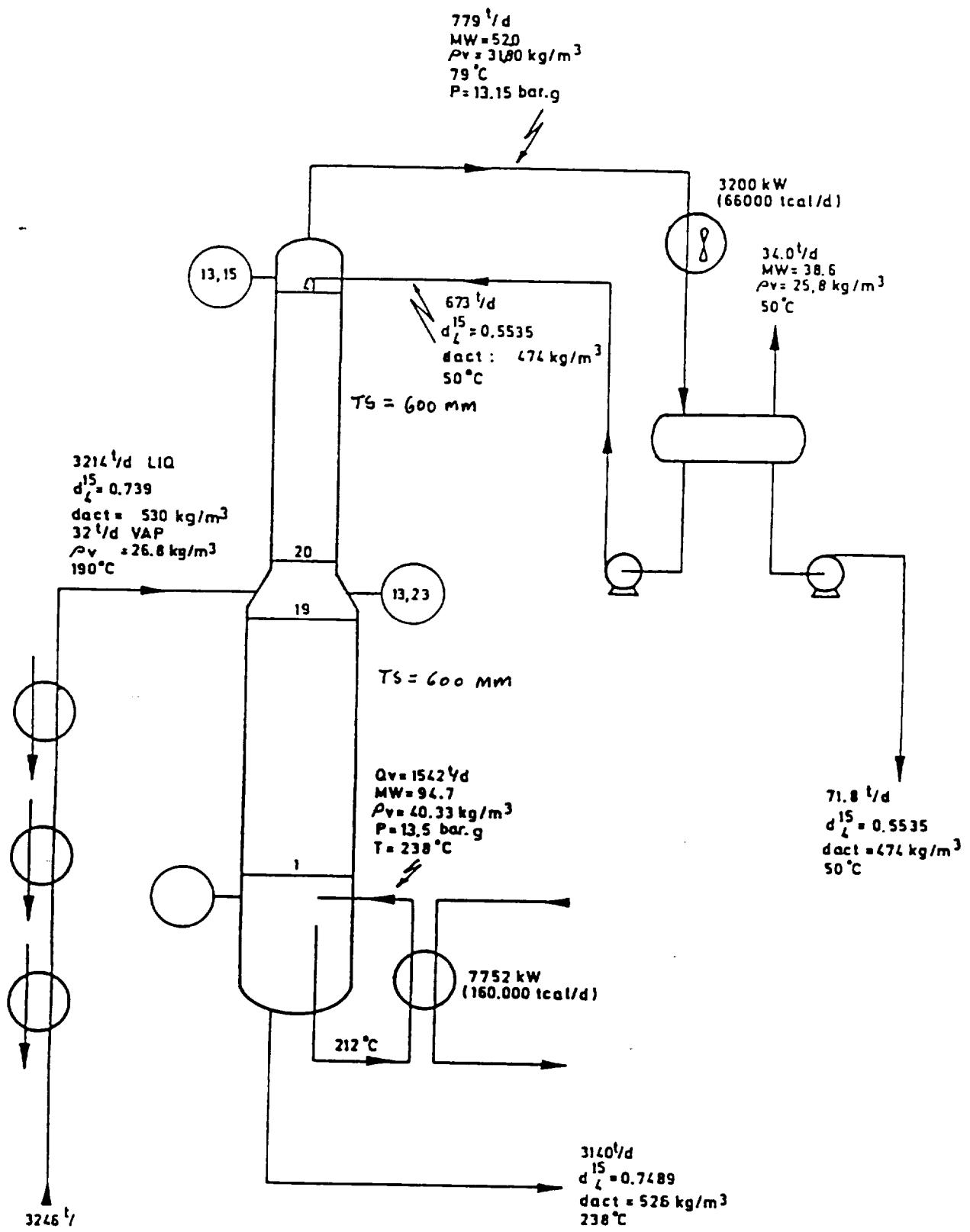
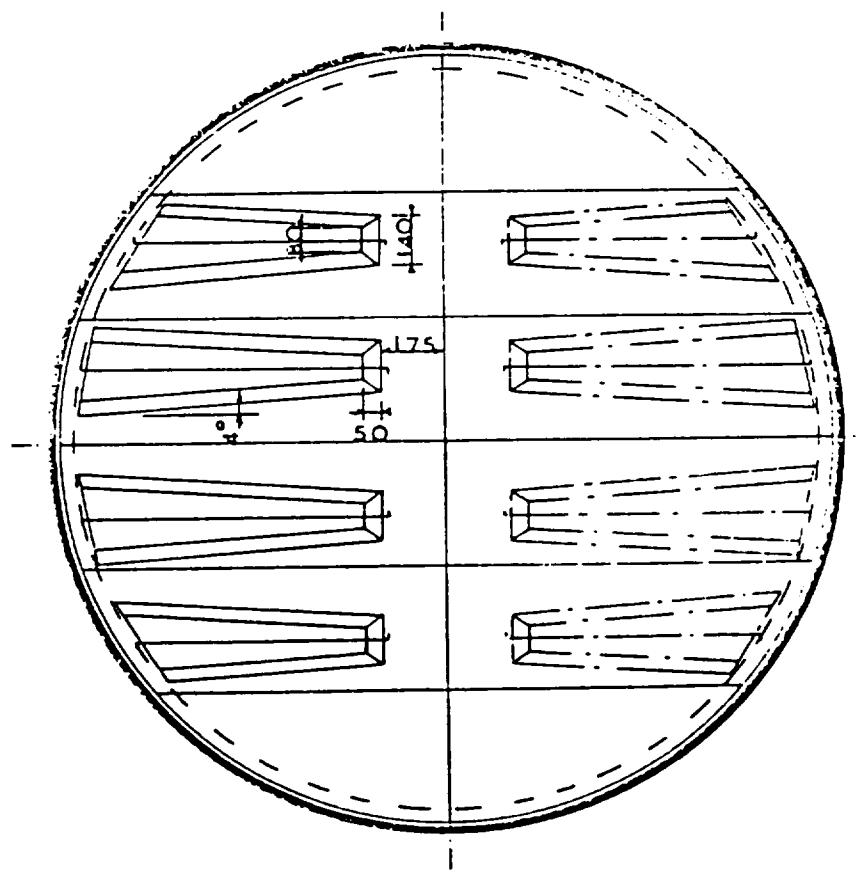


FIG-1

Data / requisition sheet for TRAYS					Design book No.:		
					Contr. Job No.:		
					MESC No.:		
For equipment No.: C - X			Specification		DEP 31.20.20.31 - Gen.		
No. of tray(s)	1-5	6-15	16-40		MATERIAL SPECIFICATION		
OD/ID <sup>a</sup> of column/tube	2240	2240	2240	mm	Parts	ASTM No.	
Space above tray				mm	Tray panels, valves	AISI - 310 S or 405	
SIEVE TRAY - BUBBLING AREA					Support beams, hang rods	~	
Required net free area	9.0	1.5	5.5	%	Calming Sections	~	
Basic free area (estimated)				%	Weir, downcomers	~	
Diameter of holes <sup>b</sup> )	12	12	12	mm	Wedges, clamps	~	
Pitch in rows (rectangular)				mm	Bolts and nuts	~	
Pitch between rows (rectang.)				mm			
VALVE TRAY - BUBBLING AREA						GENERAL	
Required net free area				%	Approved welding electrodes, etc.		
Number of valves per tray				%	DEP 40.10.64.10 - Gen., List		
Type of valve					Corrosion allowance		
Diameter of holes				mm	Reference drawings	T	
Pitch in rows (rectangular)				mm			
Pitch between rows (rectang.)				mm			
GRID TRAY - BUBBLING AREA						TYPE OF SEALING OF CALMING SECTION	
Type of bar	-	-	-		A: SEAL PAN	B: OPEN SLOTS	C: SLOTS WITH FLAPS
Slot width				mm			
Maximum free area				%			
To be blanked off to				%			
CONVENTIONAL DOWNCOMERS						NOTES	
Type of downcomer					1) The manufacturer is responsible that design and construction are in accordance with the specifications referred to on the requisition and/or drawings. Furthermore, the manufacturer is responsible for ensuring that the mechanical design is satisfactory for the design conditions indicated on the requisition and/or drawings. Calculations and thicknesses of material supplied to the manufacturer are for information and tendering purposes only. The manufacturer shall make his own calculations for which he is fully responsible		
Downcomer area					2) Edge rounding  : yes/no		
Height of inlet weir					3) For A, width of seal pan is twice bottom width of Calming Section		
Height of outlet weir							
Downcomer clearance							
Depth of seal pans							
CALMING SECTIONS							
Number of Calm. Sect per tray	4	4	4				
Length at top of weir	SEE SHEET 2			mm			
Width at top of weir	SEE SHEET 2			mm			
Height of weir above tray	10	10	10	mm			
Height of baffle above tray	330	330	330	mm			
Depth below tray, excl. slot	470	470	470	mm			
Length at bottom	SEE SHEET 2			mm			
Width at bottom	SEE SHEET 2			mm			
Slot width = slot height	40	40	40	mm			
Type of sealing (A, B or C) <sup>b</sup> )	B	B	B				
Lay-out No.	2240-10						
INFORMATION TO BE SUBMITTED WITH THE TENDER							
43							
44							
45							
46							
47							
48							
49							
50	REMARKS ON REVISIONS						
51							
52							
53							
54							
55	* Delete what is not applicable						
Manager by <i>[Signature]</i>	Date	EQUIPMENT : TRAYS 1-19			Rev. letter		
Checked by	Date	PLANT : DEBUTANIZER C-X			Date		
Ascr. by	Date	CONSIGNEE :			Sign.		
					Sheet No. 1 over'd on sheet No. 2		
					Equipm. No.	C - X	
					Req. No.		

Data/requisition sheet for  
TRAYS - cont. sheet

Design book No.:  
Contractor Job No.:  
MESC No.:



TRAYS 1-19  
DEBUTANIZER C-X

LAYOUT NR. 2200-10

4 C.S.

Sheet No. 2 cont'd on sheet No. 3

Eng. by:  
Principal:

Reg.  
No.

Data / requisition sheet for  
TRAYS

Design book No.:

Contr. Job No.:

MESC No.:

For equipment No.: C - X

Specification : DEP 31.20.20.31 - Gen.

No. of tray(s)	20-41		
ODTID <sup>o</sup> of column/tray <sup>o</sup>	1905		mm
Space above tray			mm

MATERIAL SPECIFICATION

SIEVE TRAY - BUBBLING AREA	
Required net free area	12.0
Basic free area (estimated)	%
Diameter of holes ?)	12
Pitch in rows (rectangular)	mm
Pitch between rows (rectang.)	mm

Tray panels, valves	ASTM No. A151 4105 or 405
Support beams, hang rods	
Calming Sections	
Weirs, downcomers	
Wedges, clamps	
Bolts and nuts	

VALVE TRAY - BUBBLING AREA	
Required net free area	%
Number of valves per tray	± .....
Type of valve	
Diameter of holes	mm
Pitch in rows (rectangular)	mm
Pitch between rows (rectang.)	mm

GENERAL	
Approved welding electrodes, etc.	
DEP 40.10.64.10 - Gen., List	
Corrosion allowance	
Reference drawings	T

GRID TRAY - BUBBLING AREA

Type of bar			
Slot width			mm
Maximum free area			%
To be blanked off to			%

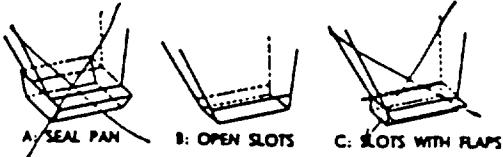
CONVENTIONAL DOWNCOMERS

Type of downcomer			
Downcomer area			%
Height of inlet weir			mm
Height of outlet weir			mm
Downcomer clearance			mm
Depth of seal pens			mm

CALMING SECTIONS

Number of Calm. Sect. per tray	4		
Length at top of weir	SEE SHEET 4		mm
Width at top of weir	SEE SHEET 4		mm
Height of weir above tray	40		mm
Height of baffle above tray	330		mm
Depth below tray, excl. slot	460		mm
Length at bottom	SEE SHEET 4		mm
Width at bottom	SEE SHEET 4		mm
Slot width = slot height	30		mm
Type of sealing (A, B or C) <sup>3)</sup>	B		
Lay-out No.	1000 - 10		

TYPE OF SEALING OF CALMING SECTION



NOTES

- 1) The manufacturer is responsible that design and construction are in accordance with the specifications referred to on the requisition and/or drawings. Furthermore, the manufacturer is responsible for ensuring that the mechanical design is satisfactory for the design conditions indicated on the requisition and/or drawings. Calculations and thicknesses of material supplied to the manufacturer are for information and tendering purposes only. The manufacturer shall make his own calculations for which he is fully responsible
- 2) Edge rounding ~~mm~~: yes/no
- 3) For A, width of seal pen is twice bottom width of Calming Section

INFORMATION TO BE SUBMITTED WITH THE TENDER

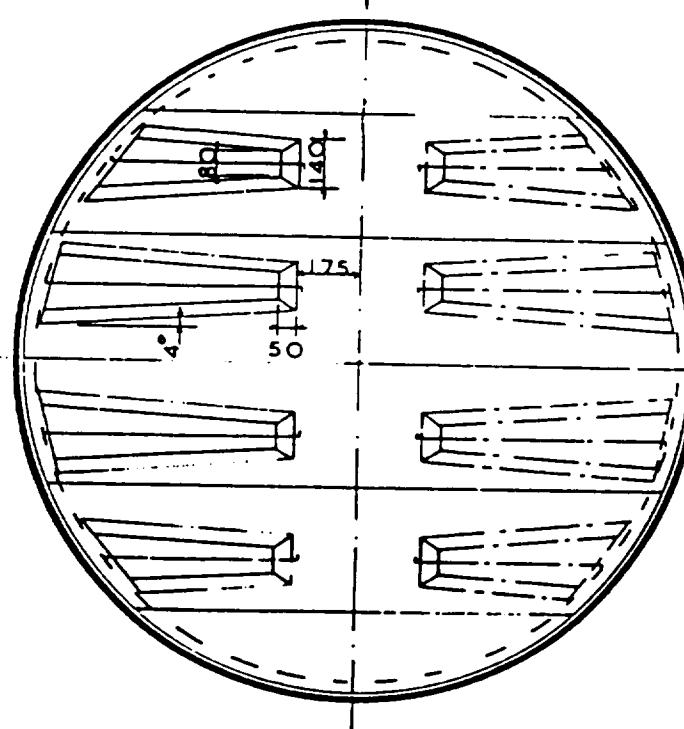
43	
44	
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50	REMARKS ON REVISIONS
51	
52	
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54	

\* Delete what is not applicable

Planned by <i>ME</i>	Date	EQUIPMENT : TRAYS 20-41	Rev. letter			
Checked by	Date	PLANT : DEBUTANIZER C-X	Date			
Appr. by	Date	CONSIGNEE:	Sign.			
Eng. by :						
Principal:						
Eng. No. : Principal : Sheet No. 3 cont'd on sheet No. 4				Equipm. No. : C - X		
				Rev. No.		

Data/requisition sheet for  
TRAYS - cont. sheet

Design book No.:  
Contractor Job No.:  
MESC No.:



TRAYS 20-41

LAYOUT NR. 1900-10

DEBUTANIZER C-X

4 C.S.

Sheet No. 4 cont'd on sheet No. -

Eng. by:  
Principal:

Req.  
No.



**5**

**Auxiliary column internals**

# Auxiliary Column Internals

## 1 Introduction

Distillation columns have all sorts of auxiliary internals such as feed inlet devices, draw-off trays, etc. This equipment is equally vital for the performance of the distillation process as the distillation trays. Any improperly designed column internal may limit the capacity of the distillation column.

Most of the auxiliary internals as discussed here were specially developed in conjunction with the Shell calming section trays.

## 2. Column Inlets

The main requirement for a column inlet device is to prevent any disturbance of the froth bed on the tray by the incoming steam. In addition the inlet stream should not provoke entrainment of liquid to the column section above. Both effects may cause premature flooding of the column.

Depending on the condition of the incoming stream, different types of inlet devices are used.

### 2.1 Liquid streams

A liquid inlet requires an internal pipe to distribute the liquid on the tray. Although the stream may also contain some vapour, the average velocity should be limited to maximal 2.5 m/s to avoid erosion enhanced corrosion of the inlet device. The shape of the internal pipe is related to the type of tray.

#### 2.1.1 Conventional downcomers trays

Reflux liquid that enters at the top of a column is simply fed to the tray via an elbow pipe that is bent downwards as shown in Figures 1a and 1b. It is good practice to use a separate inlet section(s) on the tray to submerge the pipe end. In the case of low liquid loads this inlet weir may be provided with a few overflow notches in order to improve liquid distribution.

If the liquid is fed to a tray that also receives liquid from the tray, the best technique is a half open pipe with a closed end, as shown in Figure 2a. This type is used for small columns with a diameter of 2 metres or less. The pipe should preferably be parallel to the outlet weir(s) of the tray.

Columns with a larger diameter have a perforated inlet pipe as shown in Figure 2b. The pipe extends over the full diameter of the column. The 10 mm holes splash the liquid at an angle into the froth bed. For an equal liquid distribution over the full tray area, the velocity in the holes should be more than twice the pipe velocity.

#### 2.1.2 Calming section trays

The liquid inlet is shown in Figure 3a. The perforated pipe extends over the full column diameter. The pipe is provided with rows of holes of approximately 10 mm in diameter, through which the feed jets into the froth bed with a velocity of maximum 5 m/s. The position of the holes is such that the liquid is evenly distributed over the froth bed and is not squirted directly into the calming section boxes.

For smaller diameter columns simple elbow pipes may be used (Figures 3b and 3c). The pipe outlets with dollar plate are submerged into the froth bed.

### 2.1.3 Dual-flow trays

Grid trays and dual-flow sieve trays are prone to channelling. A uniform liquid distribution is essential for the performances of these types of trays.

The liquid is sprayed on the tray with a spider as shown in Figure 4. It consists of a large-diameter central pipe with parallel side branches. The pipes are provided with rows of small holes, through which the liquid squirts into the froth bed. The hole diameter is about 8 mm and the density must be at least four holes per m<sup>2</sup> tray area. Spiders with smaller diameter holes require filters in the feed line to protect the spider against plugging with scale.

## 2.2 Partially vaporised streams

When the vapour fraction of the two-phase inlet stream exceeds 15% volume, an inlet pipe with holes or slots in the sides is used. A typical example is reboiler outlets. The inlet device is a T-shaped shown in Figure 5. To reduce the velocities, the pipe ends have normally a larger diameter than the column inlet nozzle. They are perforated throughout with 25 mm holes. The perforation should be ample such that in the holes or slots

$$\rho_{(v+l)} U^2_{(v+l)} \leq 1000 \text{ N/m}^2.$$

The orientation of the distributor with regard to calming sections or downcomers is not critical. However, prevent that hot vapour blows against downcomers or recessed parts of calming sections. It might cause downcomer flooding by boiling liquid.

No inlet device is required when at the inlet nozzle  $\rho_{(v+l)} U^2_{(v+l)} \leq 1000 \text{ N/m}^2$ .

## 2.3 Highly vaporised streams

For larger diameter columns when the inlet stream has more than 85 vol.% vapour a Schoepentoeter is applied. This Shell-patented vane type inlet device is shown in Figures 6, 7 and 8. It combines a high capacity with an excellent performance to reduce the velocity, much better than the T-shaped inlet nozzle.

The Schoepentoeter vane type inlet device is specifically designed for operation in the mist flow region. The mist droplets are dispersed in a vapour flow or a very thin liquid layer flows along the pipe wall (annular mist flow). The purpose is to obtain separation of entrained liquid from the vapour and good distribution of liquid and vapour. This is obtained by slicing up the inlet stream into a series of thin flat jets. These jets dissipate a large part of their kinetic energy by a volumetric expansion at the vanes.

The Schoepentoeter can be inserted between two sections of trays. The orientation of the Schoepentoeter is independent of the trays and requires much less column height for the feed section. It permits an inlet nozzle perpendicular to the column wall which is much cheaper for larger column diameters than a tangential inlet.

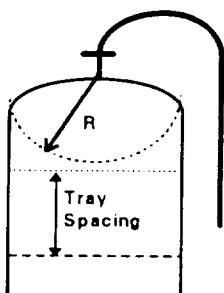
For HVU columns also downwards blowing Schoepentoeters have been used in combination with M-cap de-entrainment decks as shown in Figure 8. The M-cap deck is built directly on top of the vane type inlet device.

## 2.4 Completely vaporised streams

No inlet device is required for small diameter columns ( $D \leq 1.5$  m) when at the inlet  $\rho_v U_v^2 \leq 1000$  N/m<sup>2</sup>. Otherwise a T-shaped perforated pipe is recommended, similar to the device for partly vaporised feeds (Figure 5). However, to prevent that the vapour blows directly into the froth bed the bottom of the pipe should not be perforated. For stripping steam inlets sometimes a ladder-type spider is used as shown in Figure 10. The shape is identical to the liquid spider for dual-flow trays.

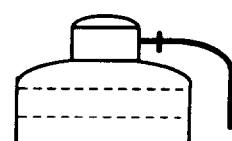
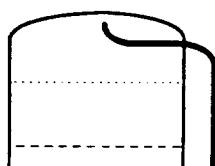
## 3. Vapour Outlets

For smaller diameter columns, the overhead vapour line is normally connected to a nozzle at the top of the column. There should be a minimum distance left above the froth bed of the top tray to prevent that liquid is surge into the overhead line by the increased vapour velocity towards the outlet nozzle. This distance is determined by the radius  $R$  of a virtual sphere. When the surface area of the lower half of the sphere is set equal to column cross sectional area then there is a constant of vapour velocity outside the sphere.



$$\begin{aligned}\text{Surface area half sphere} &= 2\pi R^2 \\ \text{Column cross section area} &= \pi/4 D_{\text{col}}^2 \\ \text{Clear distance} &\quad R \approx 0.4 D_{\text{col}}\end{aligned}$$

To simplify the construction of the overhead vapour line the outlet nozzle can also be placed at the side of the column above the top tray. In that case the outlet nozzle is provided with an internal pipe with an upward bend right to the top of the column. Larger diameter columns may have a top dome with a side nozzle.



## 4. Draw-off trays

A Fractionator column has a number of circulating reflux sections and fractionating sections. Draw-off trays are required to collect the liquid and withdraw it from the column either as a side-cut product or as a circulating reflux stream.

Liquid draw-off with conventional downcomer trays require special arrangements to avoid vapour entrainment. Because these types of trays are seldom applied in Fractionator columns they will not be discussed further.

For CS trays and packing the liquid is drawn-off with a separate tray. These draw-off trays draw either the total liquid from the column (TDO trays) or only a part of the liquid (PDO trays).

#### **4.1 TDO tray**

Three types of draw-off trays are in use:

1. Flat gutter type draw-off tray with rectangular risers between the gutters. The tray has one or two side sumps to collect the liquid (Figures 11a, b, and c). For larger diameter columns it has a central trough (Figure 11d).
2. The tray consists of a non-perforated horizontal plate with chimney vapour risers (Figure 11e). This type is generally used for column diameters below 1 m.
3. The M-cap deck (Figures 8 and 12). This draw-off tray is exclusively used in combination with a downwards-blowing vane type inlet device. The construction is expensive. It has been applied in High Vacuum units.

To prevent that liquid is splashing through the tray, vapour risers have caps. The drip ring collects the liquid that runs along the column wall.

#### **4.2 PDO tray**

PDO trays are identical to flat gutter TDO trays, but have a liquid build-up in the gutters. The stagnant liquid level may be that high that the bottom side of the calming sections of the tray above may be partially submerged in clear liquid. The remaining liquid that is not drawn from the column passes via the overflow boxes and pipes to the preferred spots on the tray below (see Figure 11a).

### **5. Bottom compartment of column**

Figures 13a and b show typical bottom compartments for columns equipped with thermo-siphon type of reboilers. The basic difference between the two arrangements is the baffle that guides the liquid from the bottom tray to the reboiler compartment. This ensures that the bottom product has gone at least once through the reboiler.

To give the thermo-siphon sufficient driving force the upper tube sheet of the reboiler is at the same elevation as the liquid level in the reboiler compartment. The clear liquid velocity in the reboiler compartment should not be too high otherwise the segregation of vapour bubbles becomes rather difficult. Entrainment of vapour to the reboiler decreases the liquid density and thus reduces driving force. For a stable circulation of the thermo-siphon the amount of vapour generated in the reboiler should be not exceed 25% wt of the liquid inlet.

Hot vapour from the reboiler should not blow against the tray downcomers. Boiling liquid in the calming sections causes flooding by back-up limitation.

The cover plate of the bottom compartment has a chimney to vent some vapour.

### **6. De-entrainment devices**

Wire-mesh demister mats are used to diminish heavy liquid entrainment. The entrained liquid, in the form of fine droplets, impinges on the wire-mesh and coalesces to large droplets that fall down. The de-entrainment efficiency is high for a wide vapour load region (see Figure 14). At lower vapour loads the entrained liquid follows the vapour streams without impinging on the wires. At higher vapour loads the impinged droplets are re-entrained, causing flooding of the mat.

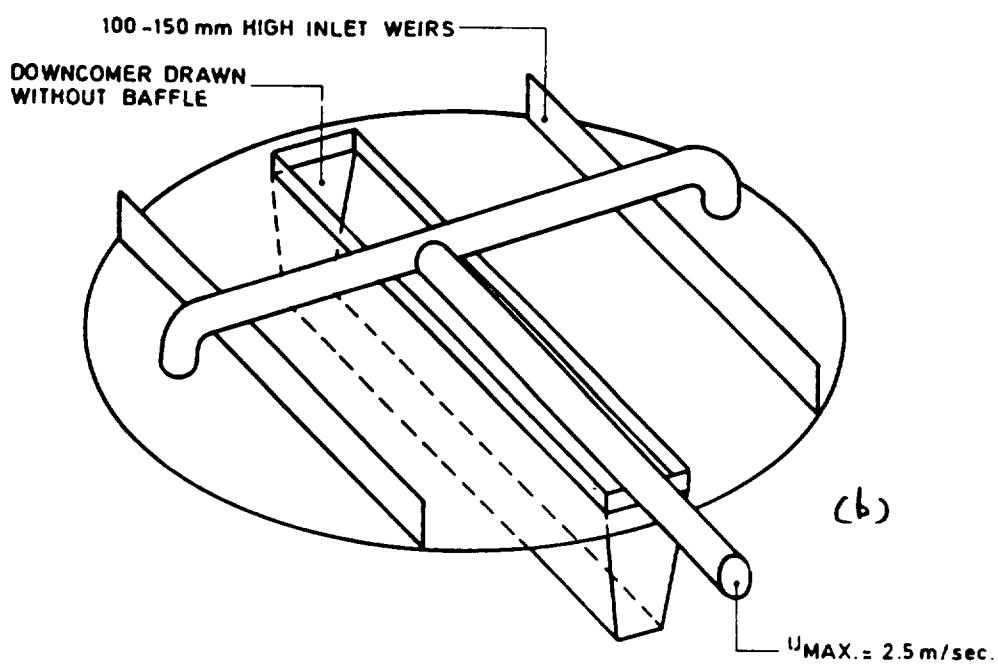
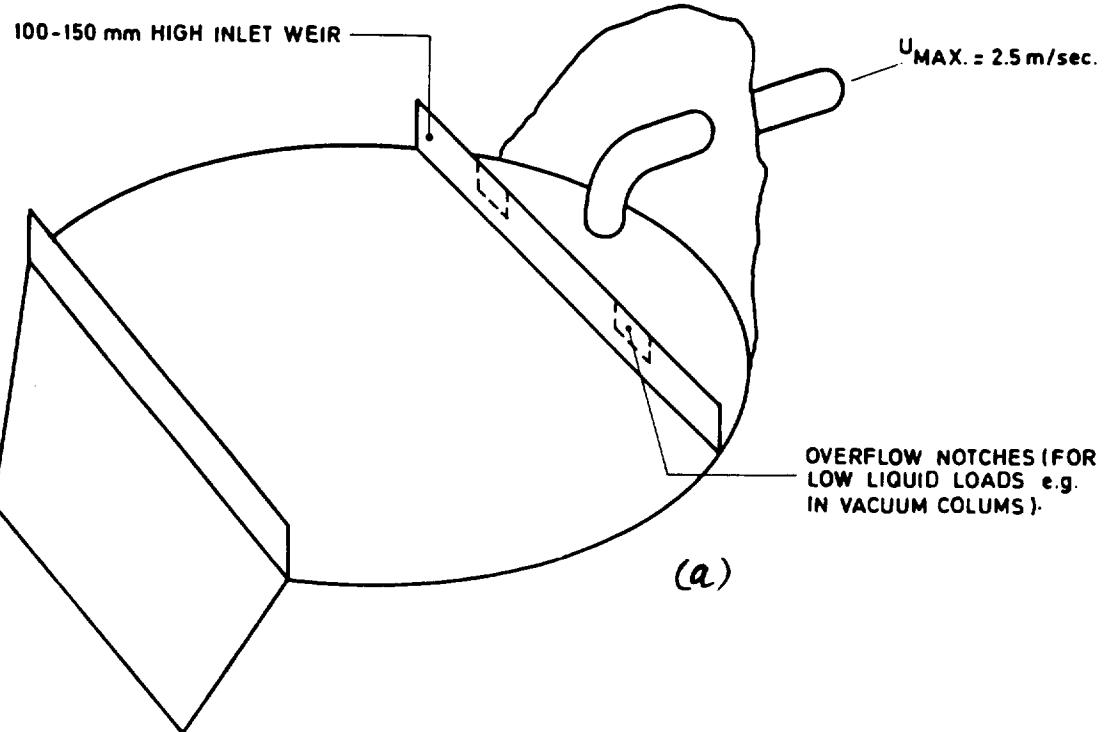
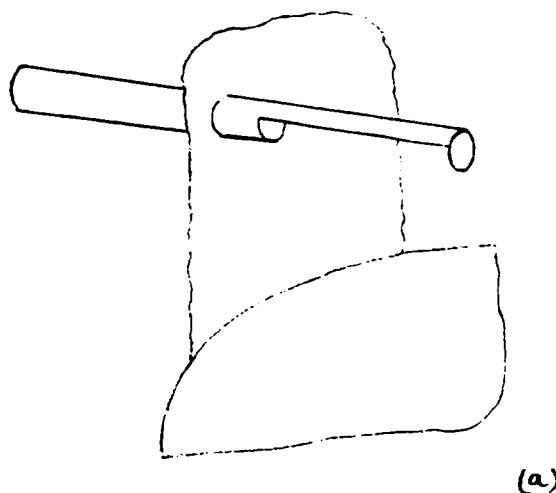
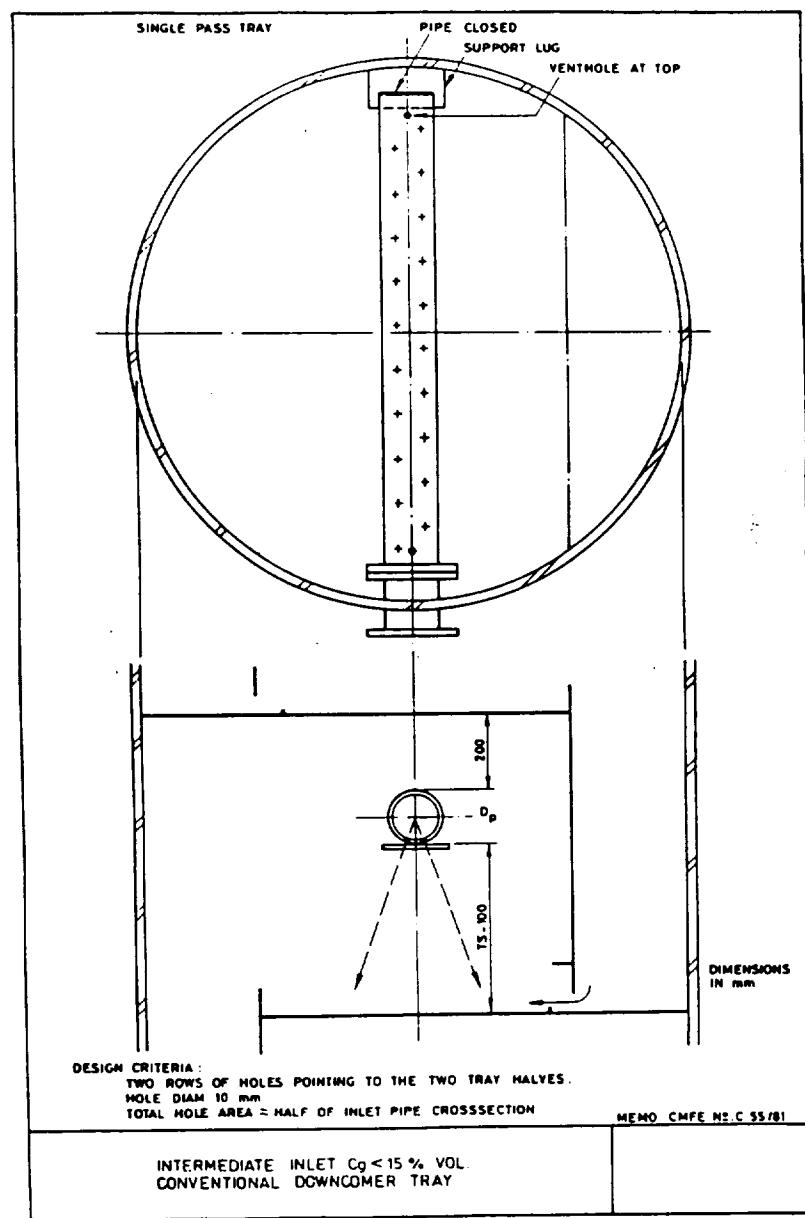


FIGURE 1      REFLUX INLETS FOR TRAYS WITH CONVENTIONAL DOWNCOMERS.

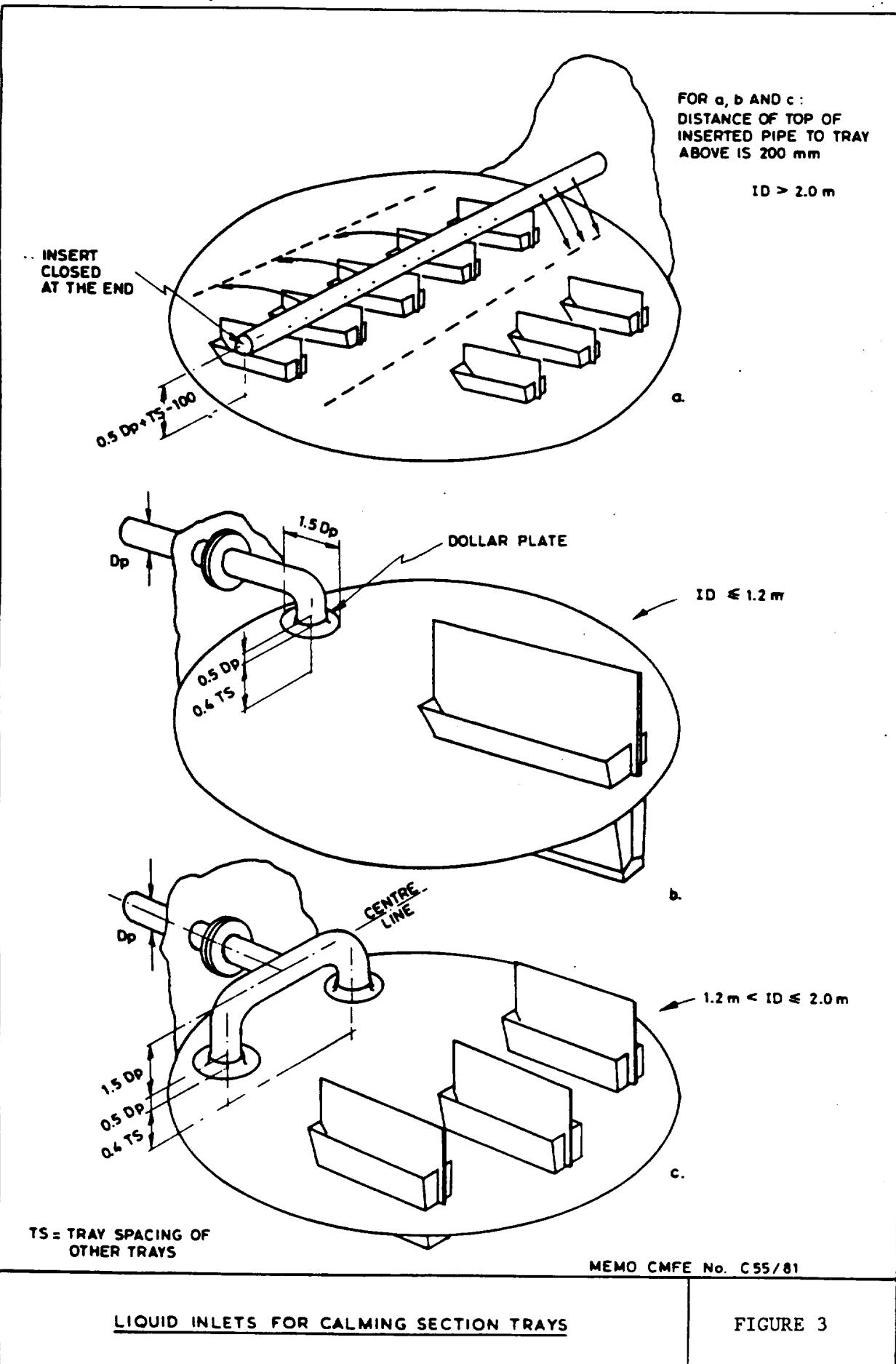
FIGURE 2

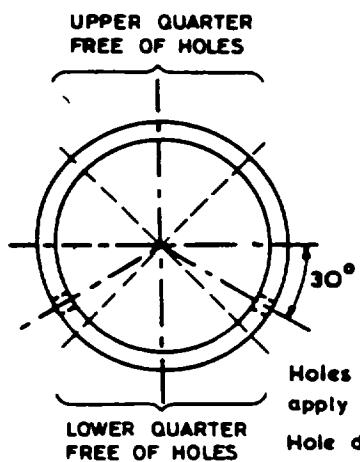
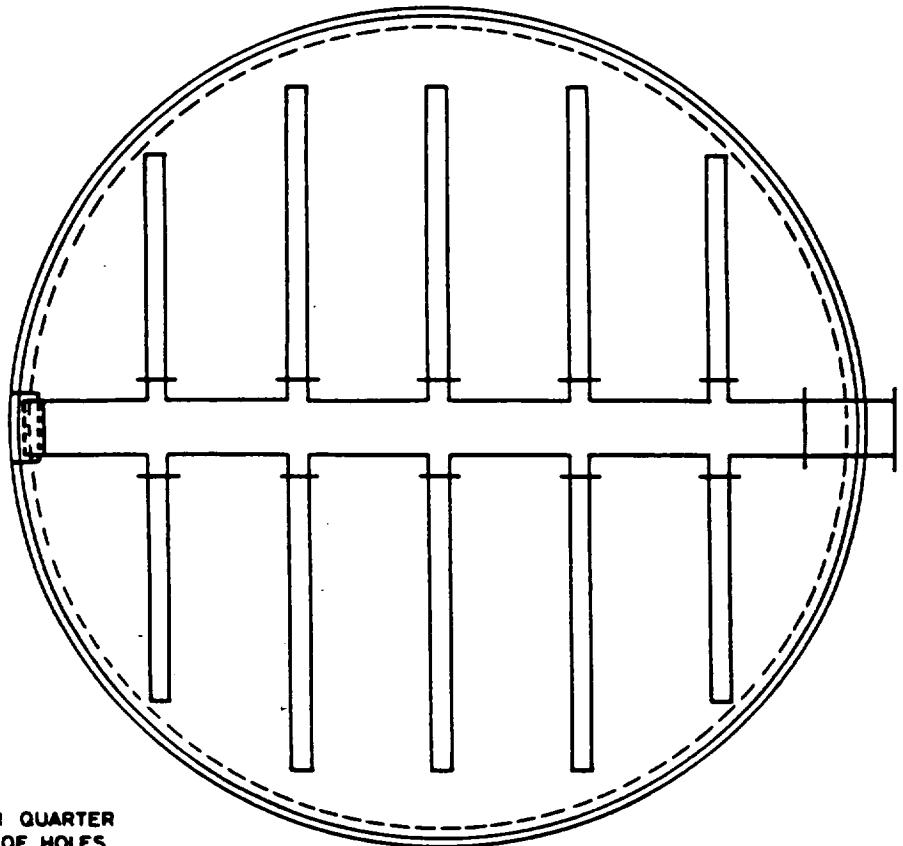


(a)



(b)





Holes alternately drilled in right and left-hand side of tube  
apply at least 4 holes/m<sup>2</sup> tray area.

Hole diameters : min.3 mm.

max.12mm.

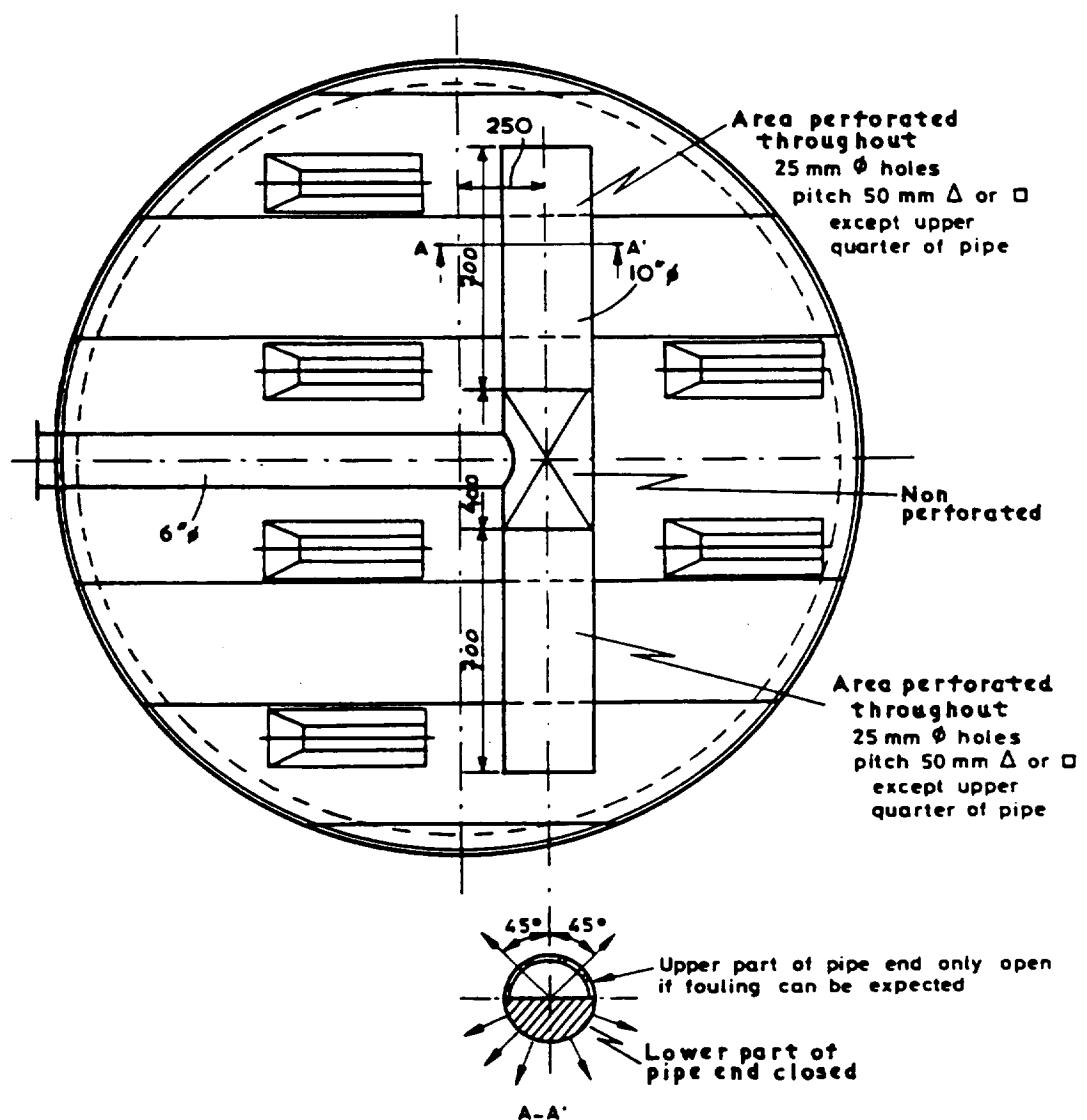
preferably 10mm.

Note : Use strainers, non-rusting and non-scaling pipe if hole dia. < 10mm

MEMO CMFE NO. C55/81

TYPICAL LIQUID SPIDER FOR DUAL-FLOW TRAYS

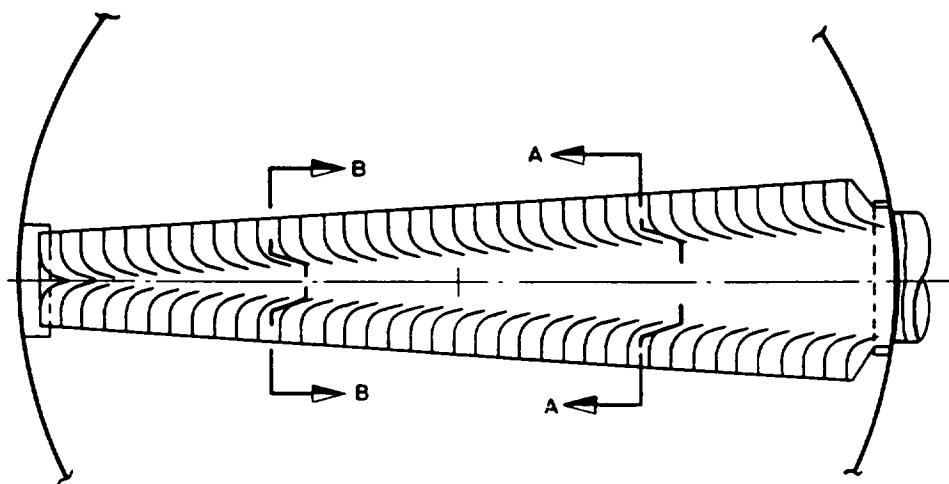
FIGURE 4



MEMO CMFE N°. C 55/81

TYPICAL EXAMPLE OF FEED INLET FOR PARTIALLY VAPORIZED  
FEED

FIGURE 5



NOTE: THERE IS NO PREFERENCE FOR ORIENTATION ANGLE OF FEED INLET WITH REGARD TO EITHER TRAYS ABOVE OR BELOW.

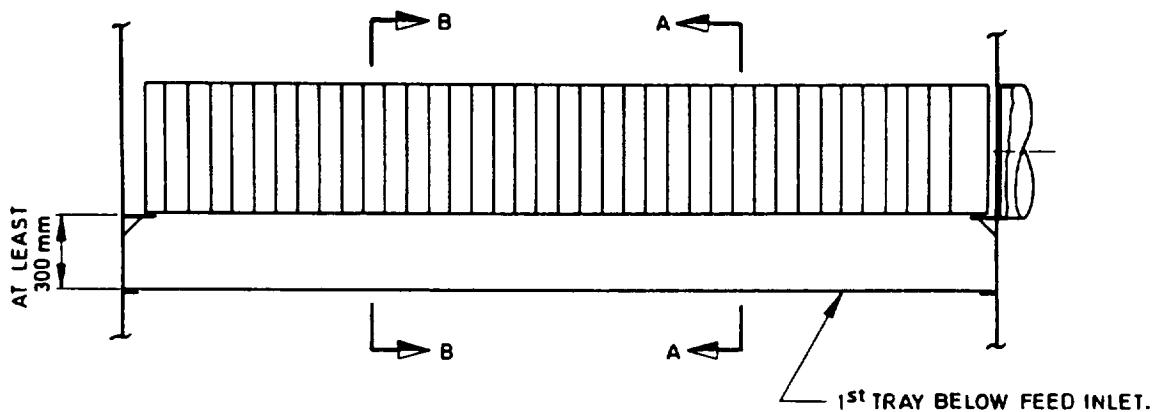
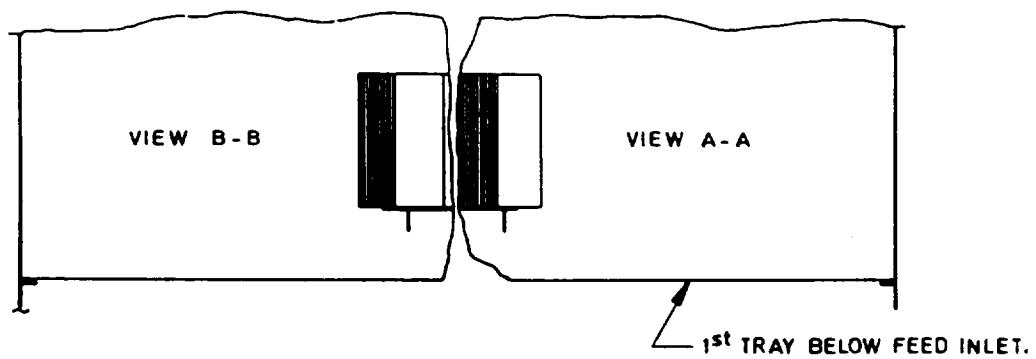


FIGURE 6

FEED INLET DEVICE  
(EXAMPLE OF HORIZONTAL BLOWING TYPE).

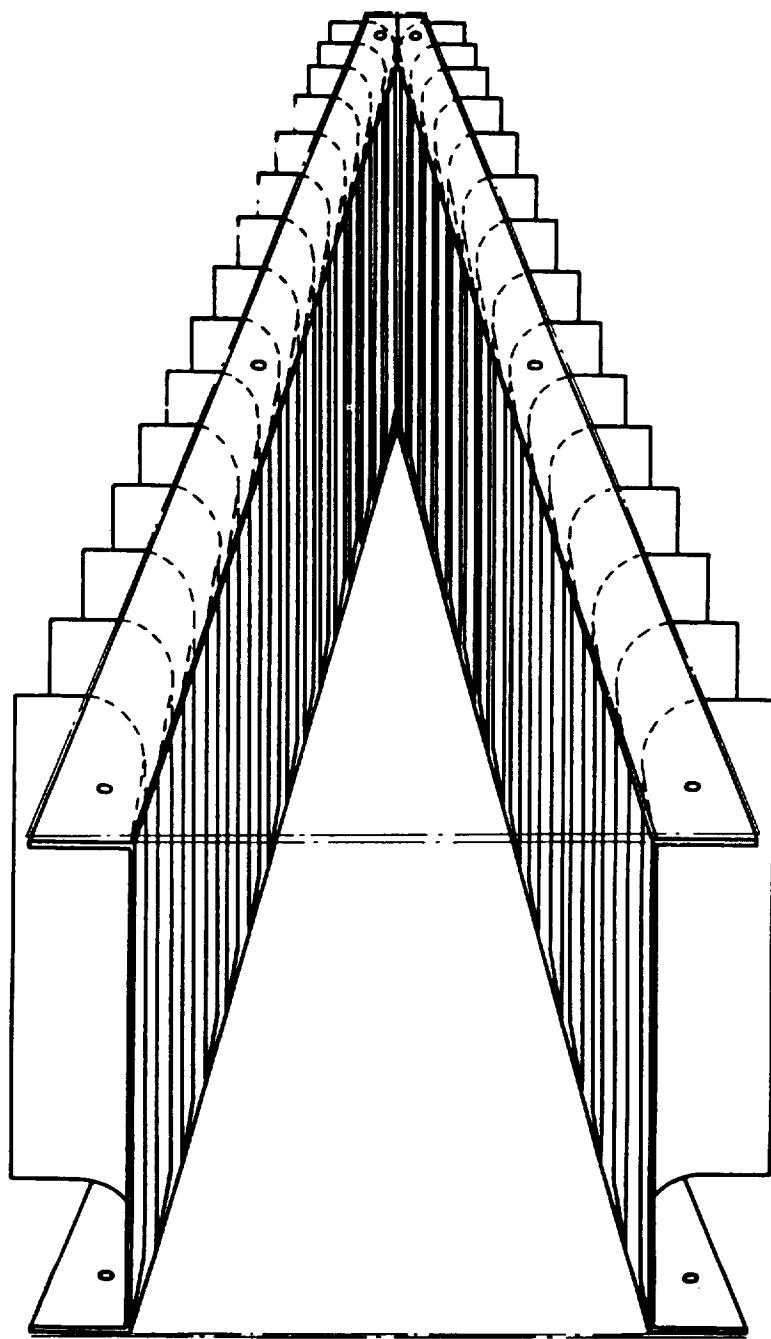
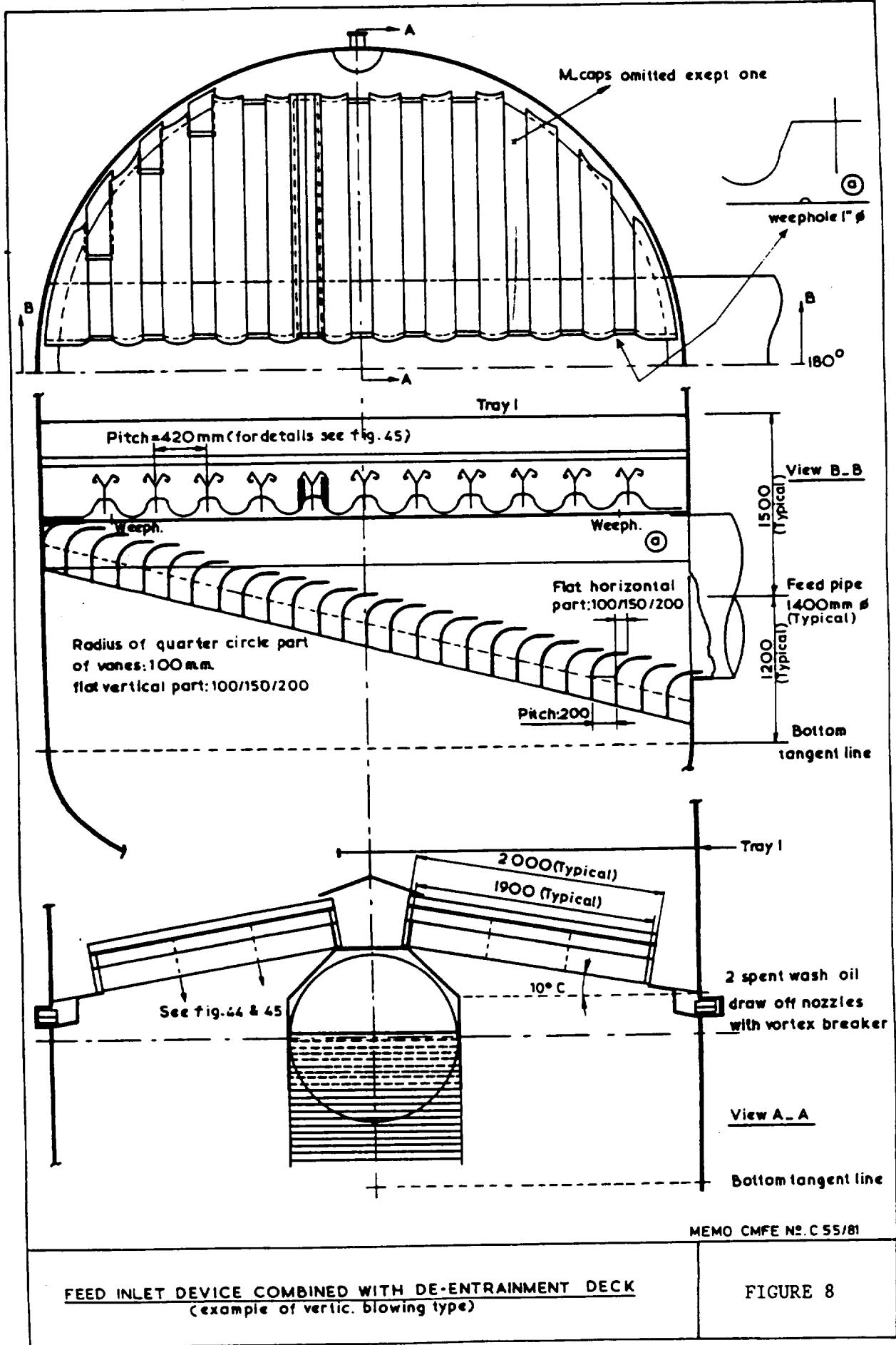


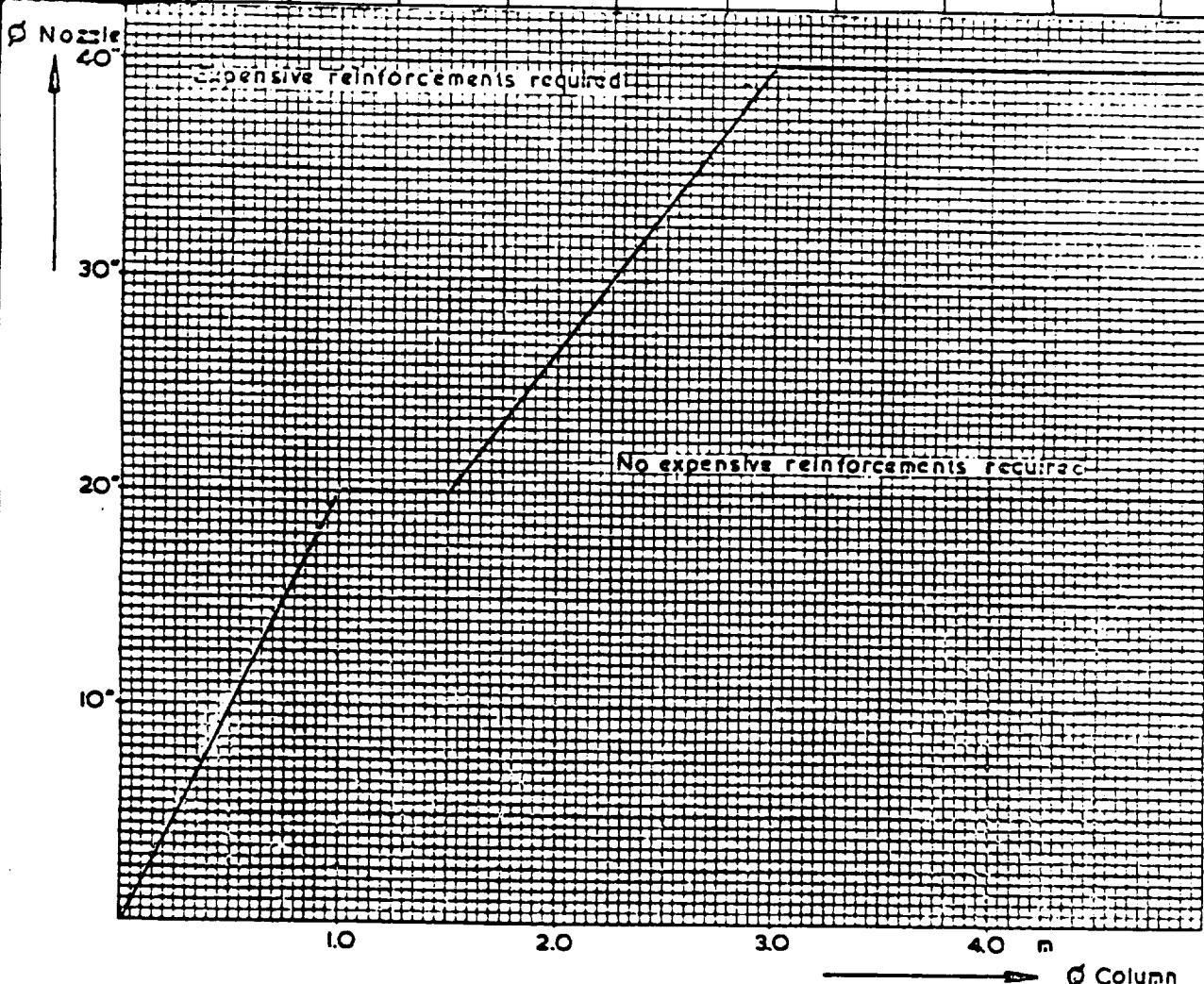
FIGURE 7      ASSEMBLY OF VANE TYPE FEED INLET DEVICE  
(HORIZONTAL BLOWING TYPE).



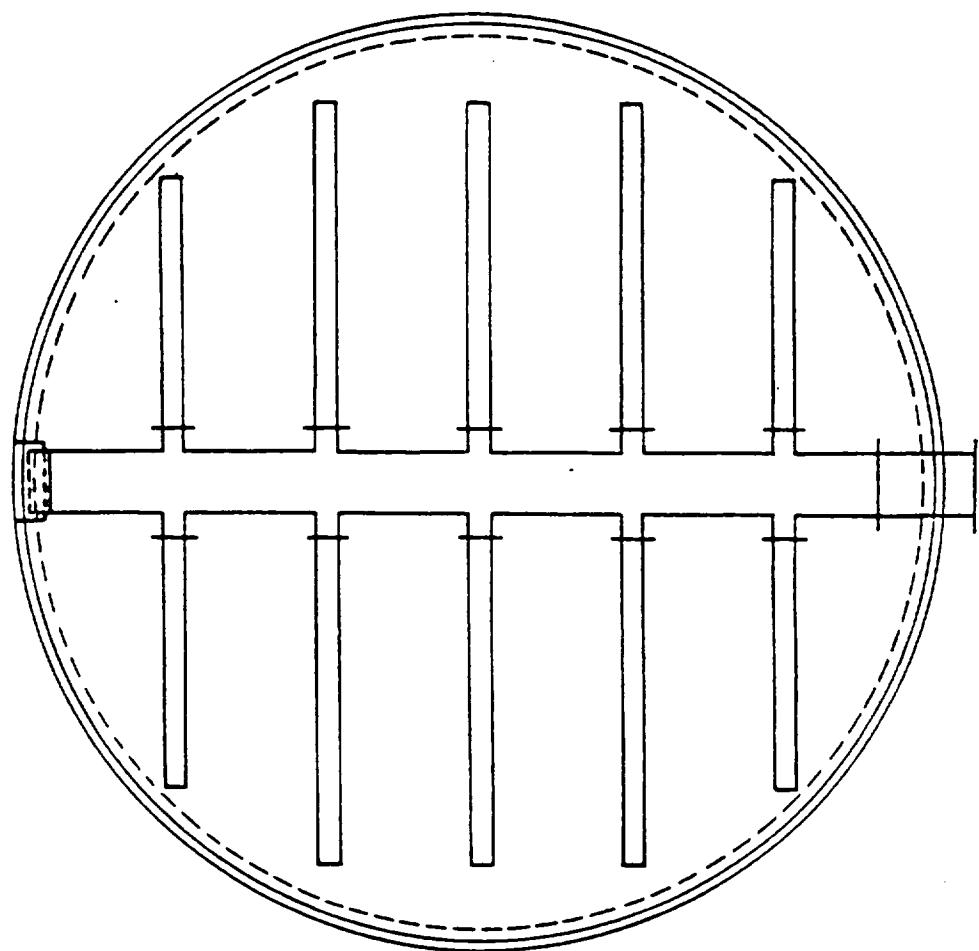
Nom. pipe diam.	Schedule 80					Schedule 40			
	7/8"	1/2"	3/4"	1"	1 1/2"	2"	3"	4"	6"
I.D. of pipe (mm)	10.74	13.07	13.8	24.3	38.1	52.5	77.9	102.2	154.0
$\square$ of pipe (m <sup>2</sup> )	0.000009	0.00015	0.00023	0.00048	0.00114	0.002166	0.00477	0.00821	0.01834

Nom. pipe diam.	Schedule 30								
	8"	10"	12"	14"	16"	18"	20"	24"	30"
I.D. of pipe (mm)	205.0	257.4	307.1	336.5	387.3	435.0	432.6	531.0	730.2
$\square$ of pipe (m <sup>2</sup> )	0.0330	0.0520	0.0741	0.0839	0.1178	0.1486	0.1829	0.2551	0.4182

Nom. pipe diam.	Schedule 20			Schedule 10			10 mm plast		
	20"	24"	30"	20"	24"	30"	36"	42"	48"
I.D. of pipe (mm)	489.0	590.5	736.6	495.3	596.9	746.1	894.4	1047.0	1199.0
$\square$ of pipe (m <sup>2</sup> )	0.1870	0.2732	0.4351	0.1929	0.2806	0.4373	0.6283	0.8502	1.1290



CMF/MFD Memo No 39/69



TYPICAL

SPIDER FOR

STRIPPING STEAM

FIG. 10

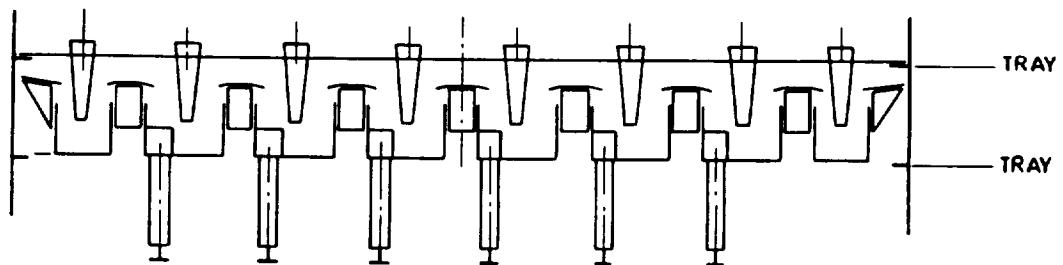
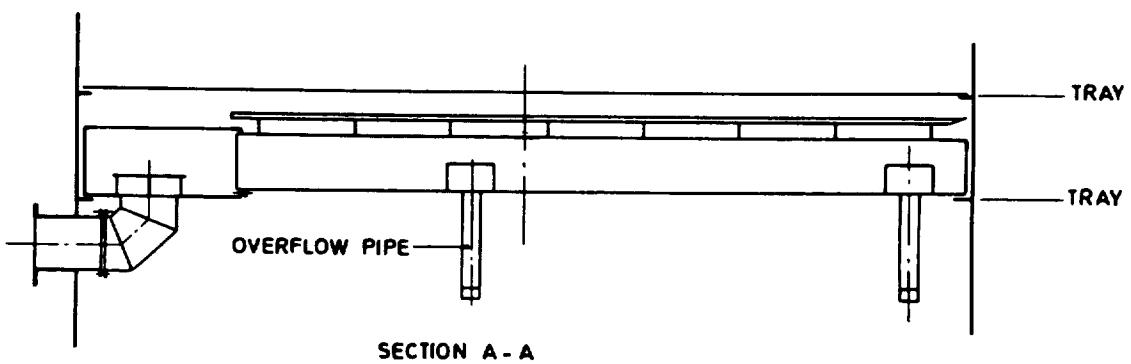
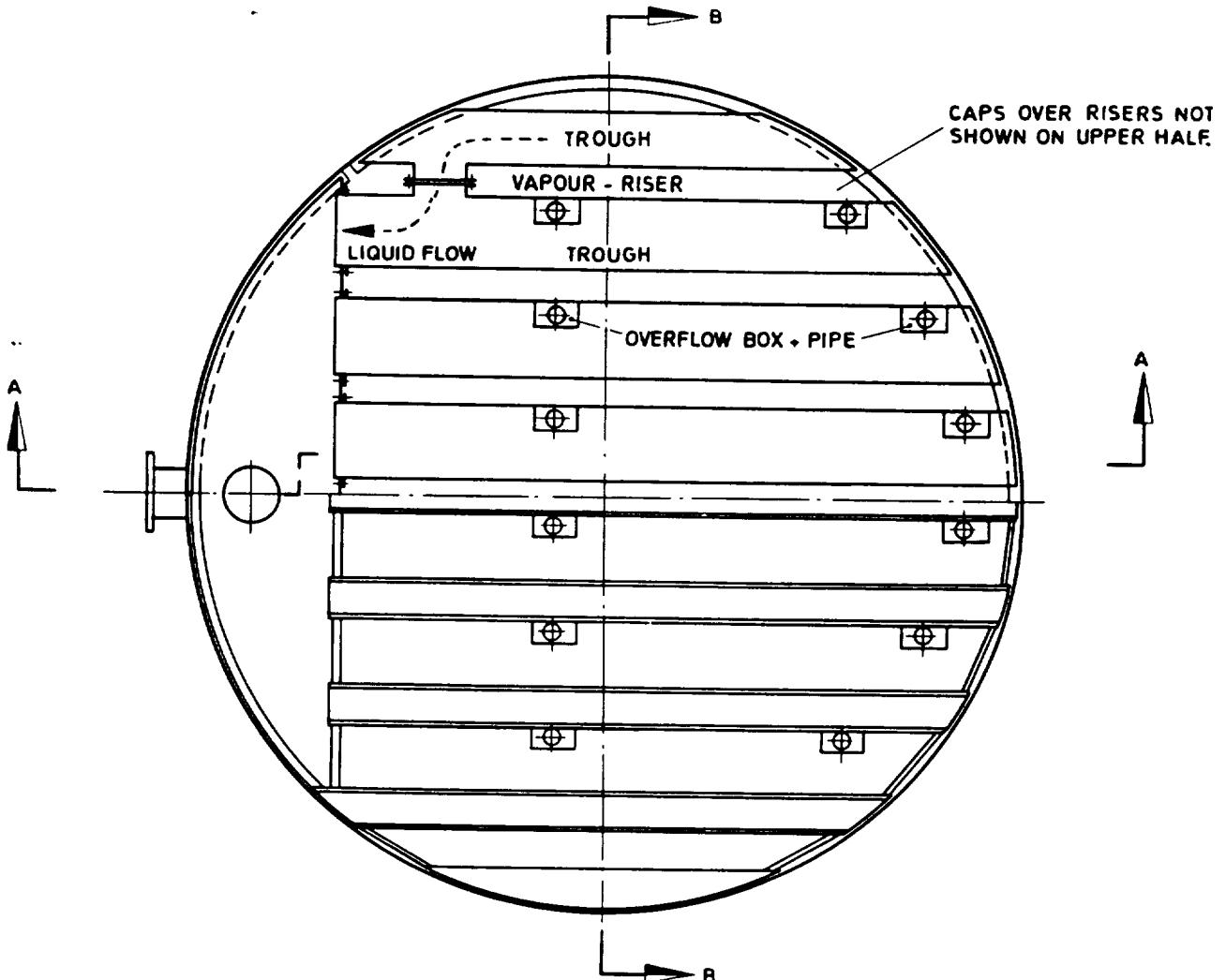


FIGURE 11a TYPICAL PARTIAL DRAW-OFF TRAY BELOW A CSS OR A CSV TRAY  
LOCATING OFF NOZZLE

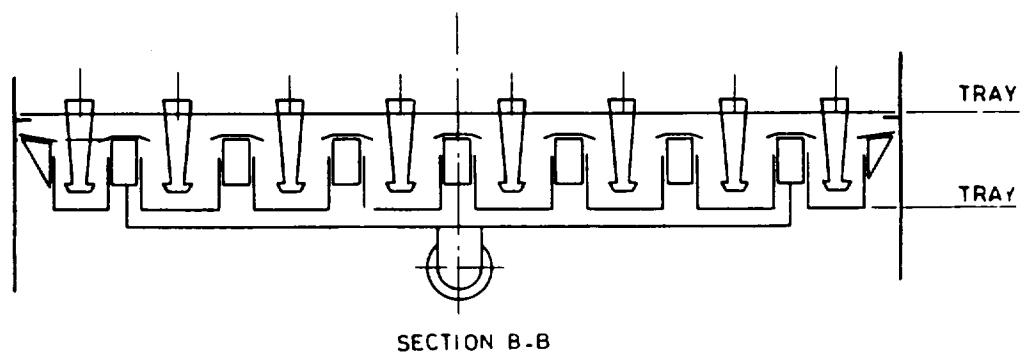
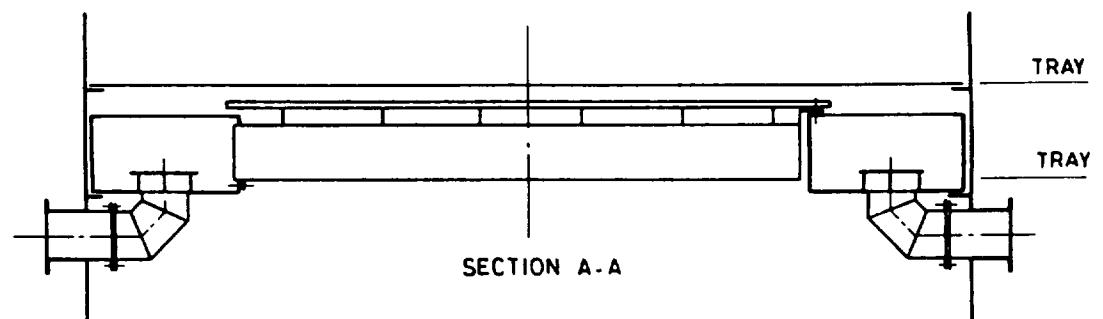
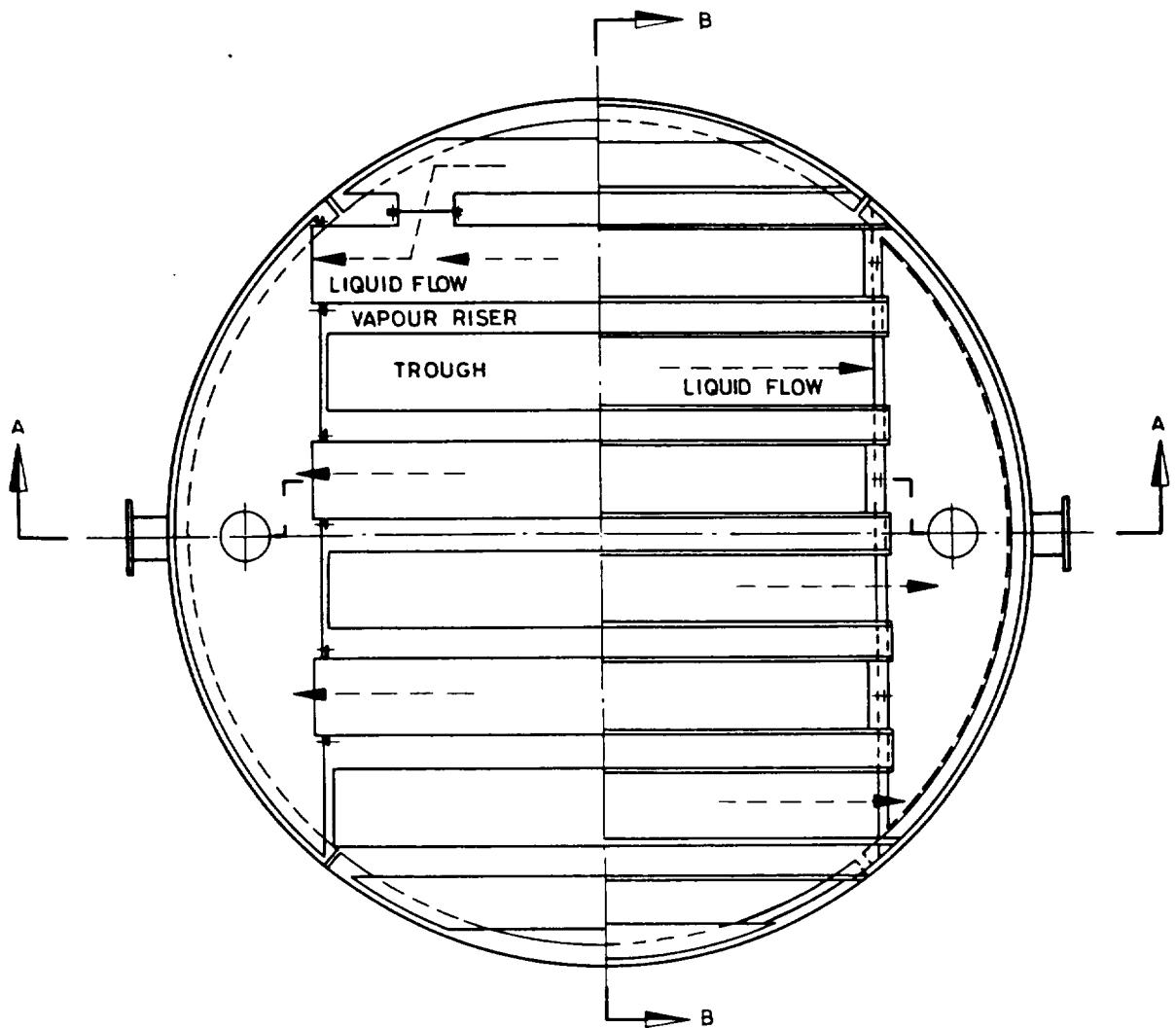
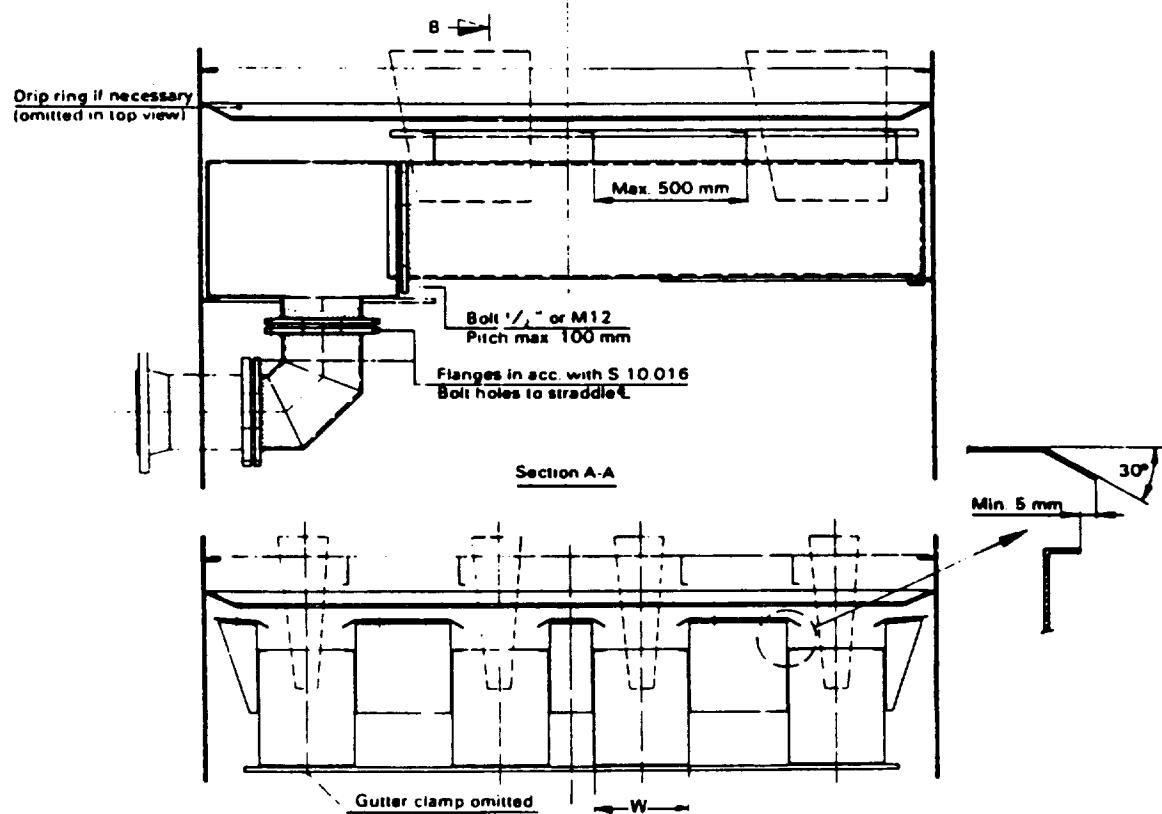
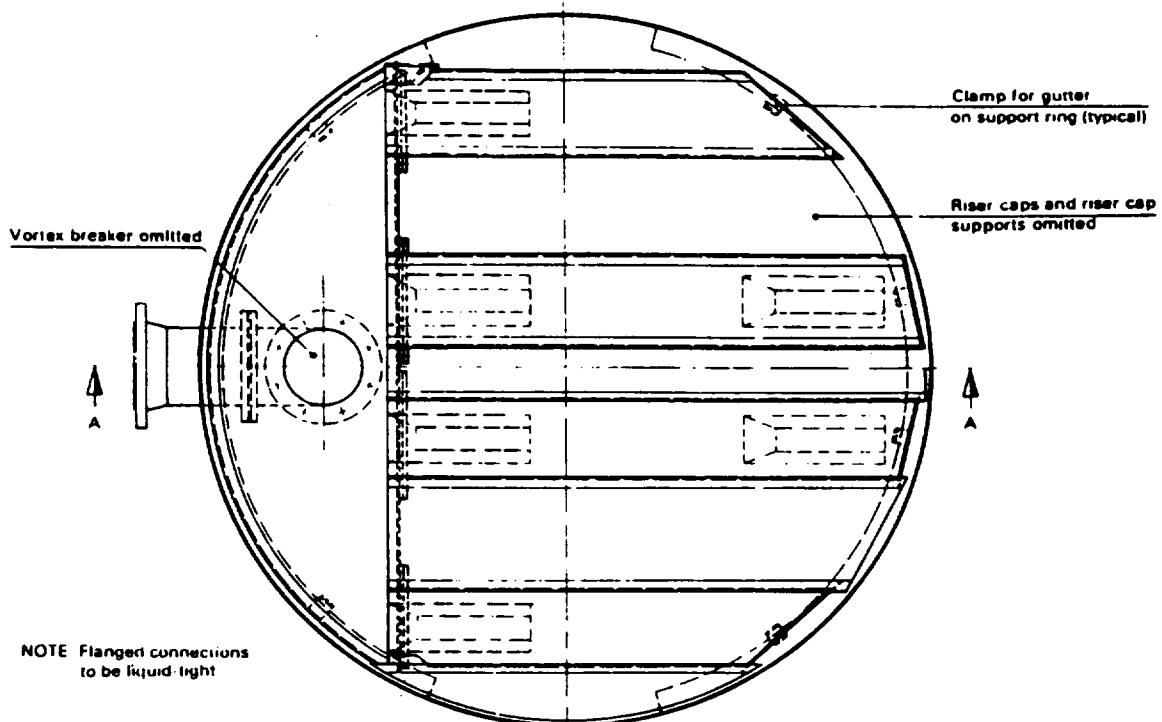


FIGURE 11 b TYPICAL TOTAL DRAW-OFF TRAY BELOW A CSS OR A CSV TRAY  
2 DRAW-OFF NOZZLES.

### TYPICAL TOTAL DRAW-OFF TRAY WITH SIDE SUMP

Recommended when tray above has no main beam  
(Column ID  $\leq 2500$  mm)



Section A-A

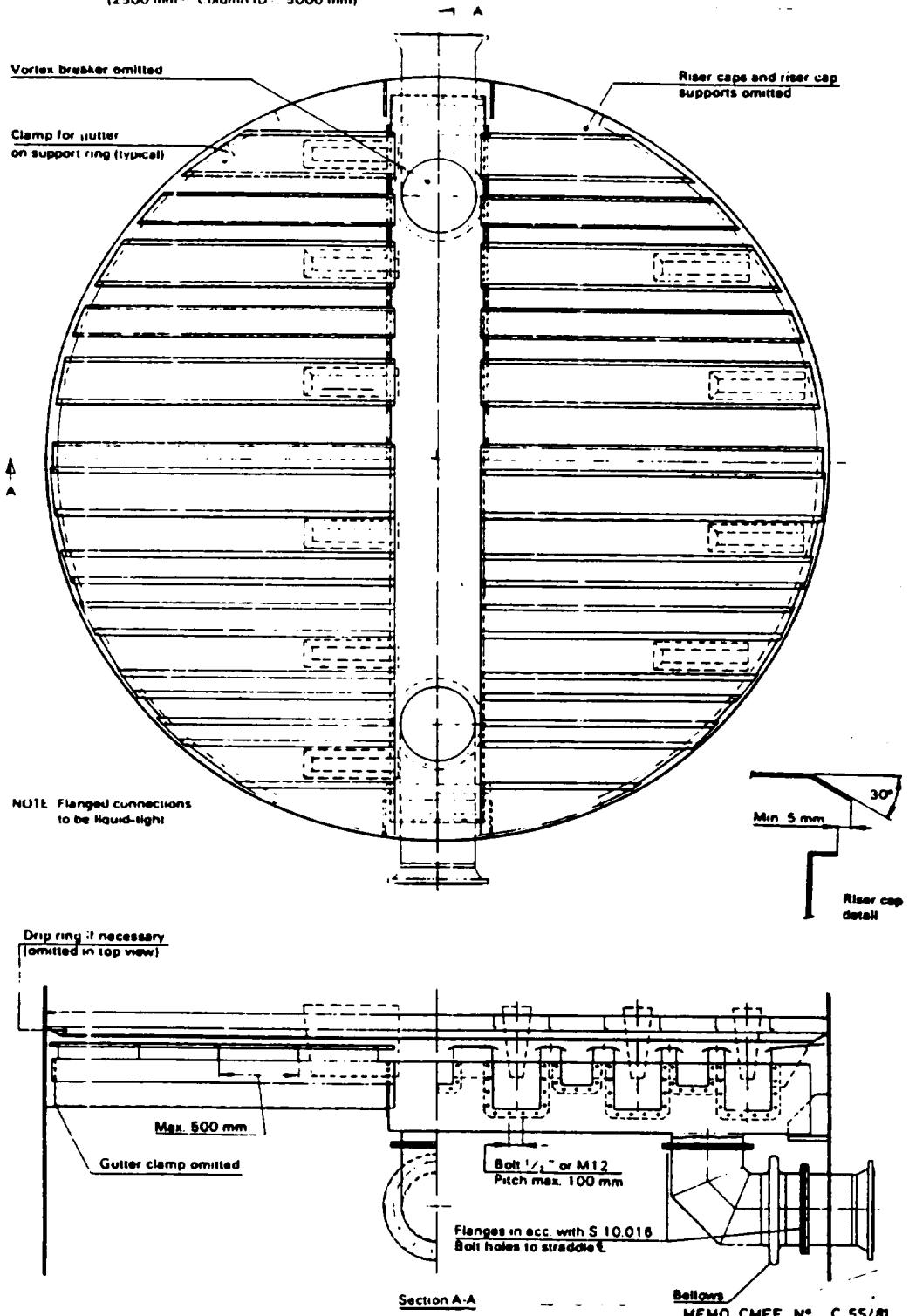
MEMO CMFE N°. C 55/81

TOTAL DRAW OFF-TRAY, TYPE I

FIGURE 11 c

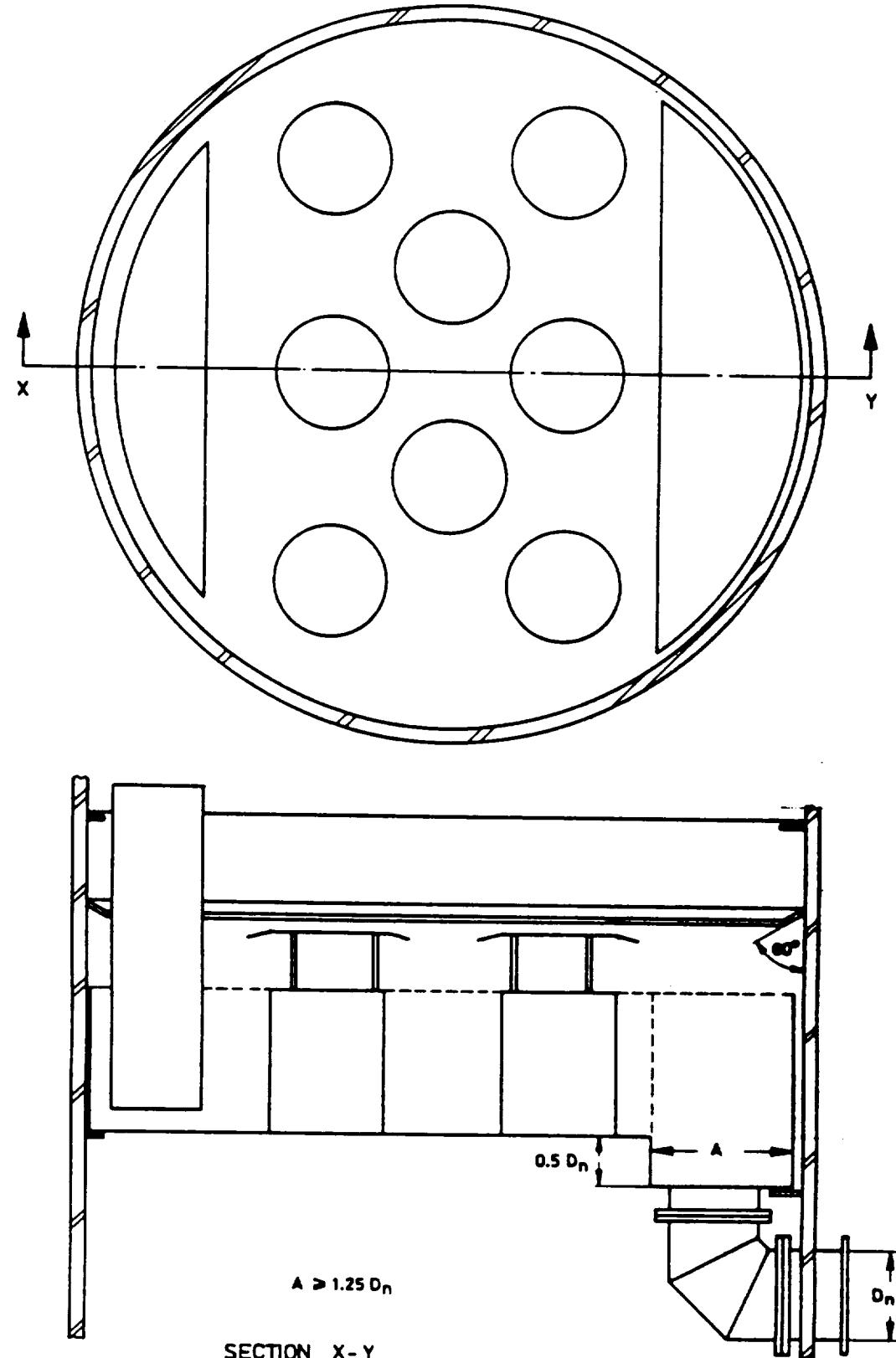
## TYPICAL TOTAL DRAW-OFF TRAY WITH CENTRAL TROUGH

Not recommended when tray above has one main beam  
(2500 mm) - Column ID 1. 5000 mm)



TOTAL DRAW-OFF TRAY, TYPE II

FIGURE 11d



MEMO CMFE No. C55/81

TOTAL DRAW-OFF TRAY WITH  
ROUND CHIMNEYS

FIGURE 11 e

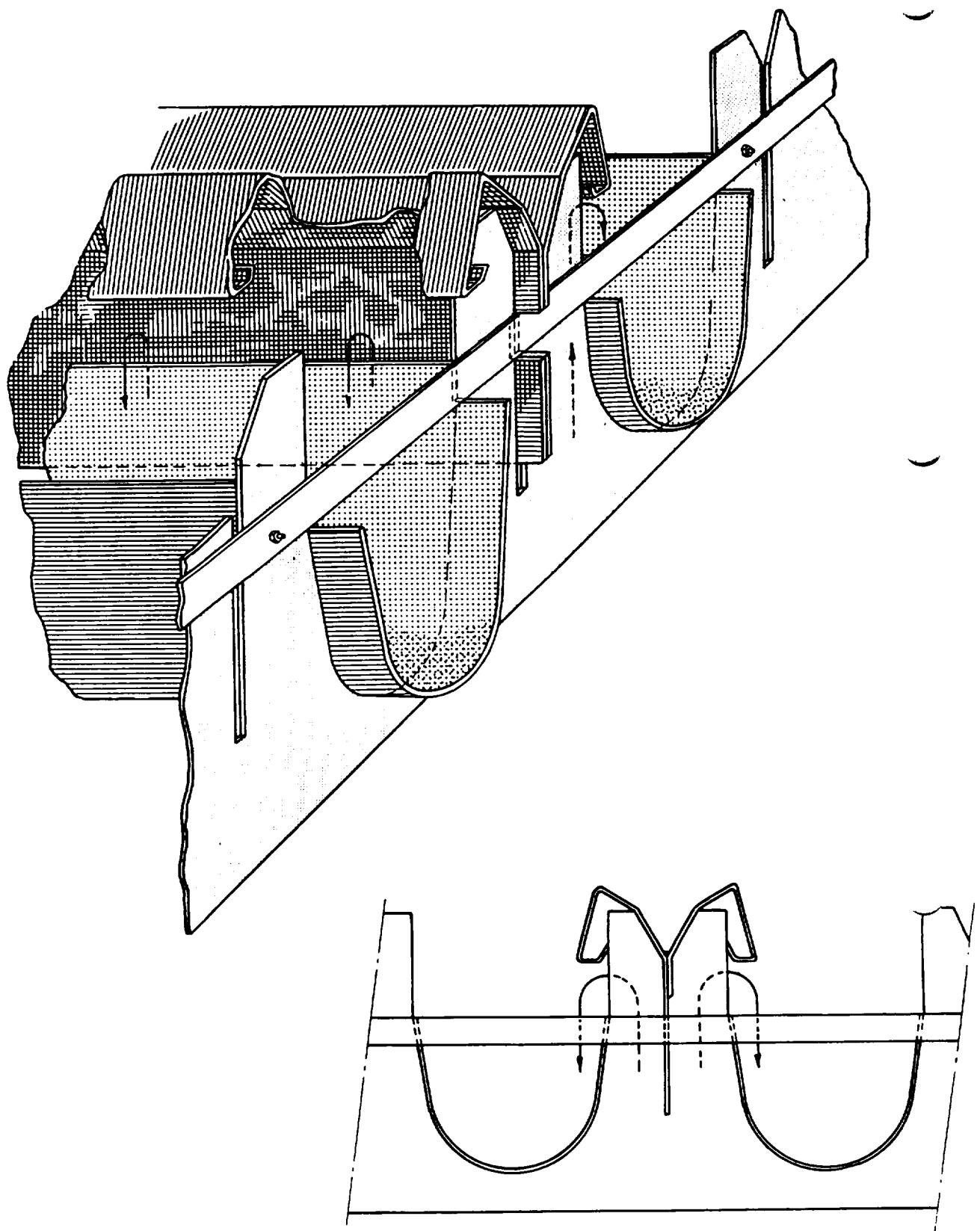
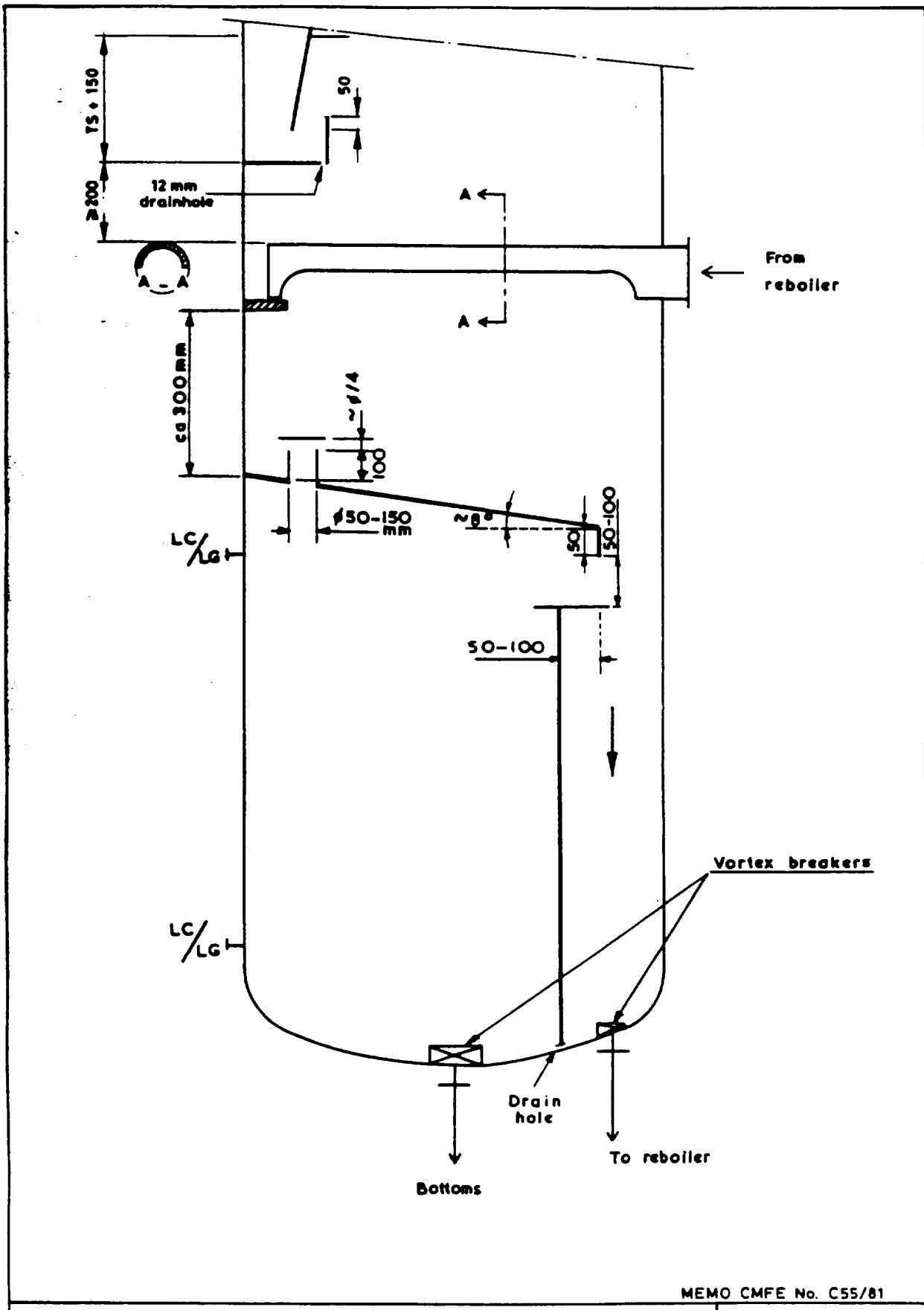
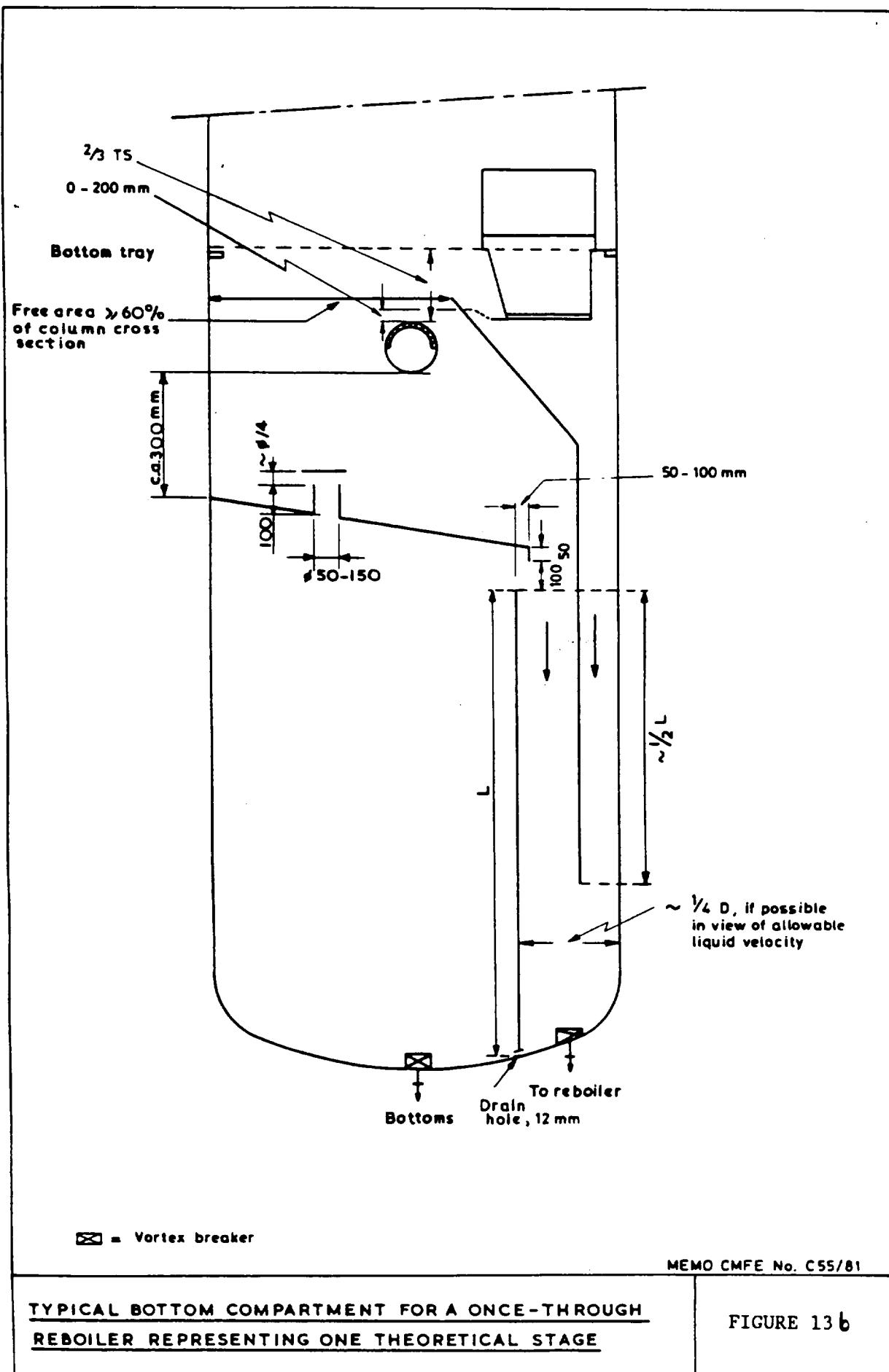


FIGURE 12 ASSEMBLY OF M CAP DECK.



TYPICAL BOTTOM COMPARTMENT FOR A CIRCULATING - TYPE  
REBOILER

FIGURE 13a



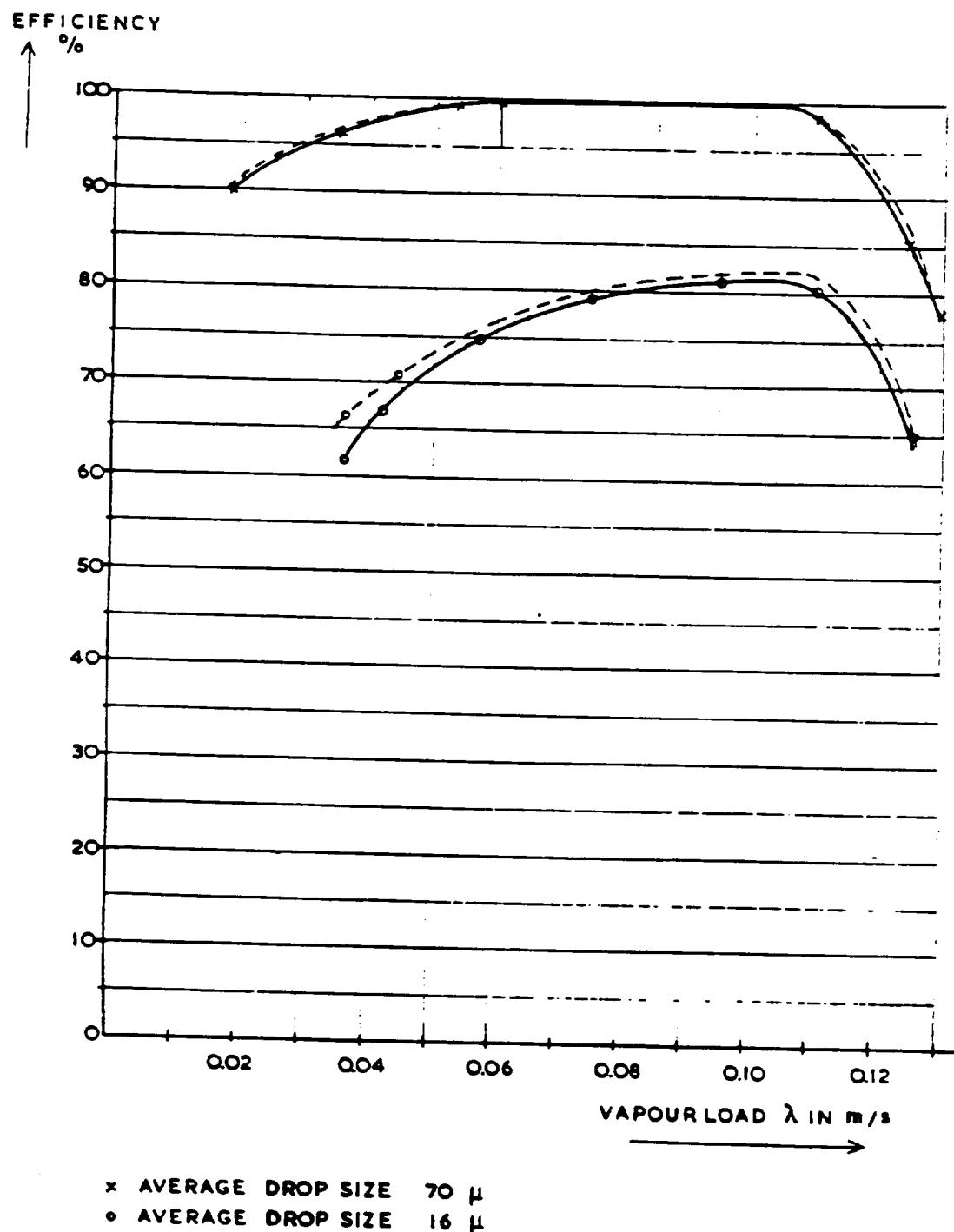


Figure 14 - Wire-mesh Demister Mat



**6**

**Basic data systems**

## **Basic data systems**

### **1. INTRODUCTION**

Reliable estimation of the phase behaviour and physical properties of process streams is essential for the analysis and simulation of oil, gas and chemical processes. The data, models and methods required for this purpose are collectively known as *basic data*. Various methods are available. They are all different and an appropriate choice has to be made. Even though designers use a number of personal guidelines, there is no general understanding of *when* to use *which* basic data system. As a result a basic data system can inadvertently be chosen, which is not the most appropriate for the specific application. This may lead in seemingly converged and accurate output, but the obtained results may contain *hidden errors*.

In the PRO/II manual and the PRO/II reference manual some 36 basic data systems are described. However, for the standard user it is not an easy task to see the wood through the trees. In Ref. 1 a very useful excerpt is presented of the various methods for use in EP. Since all methods boil down to only 4 different application areas (low pressure oil, high pressure oil/gas, extraction/chemicals and specials), and because many of the eligible systems are of only academic interest, we tried to clarify the applicability of the various basic data systems and to give some practical guidelines regarding the selection of the most suitable basic data system.

The general recommendations for some typical applications are summarised in Table I. A complete survey of the available basic data methods is given in Ref. 7.

### **2. K-VALUES AND ENTHALPIES**

Each separation step or heat calculation in PRO/II is assumed to be an equilibrium step : The vapour, liquid (and sometimes second liquid) are in thermodynamic equilibrium and the mass balance must be fulfilled. The necessary condition for equilibrium between vapour/liquid is the equality of temperature and pressure and the equality of the fugacity (or partial pressure) of each component in each phase :

$$f_j^V = f_j^L$$

The program estimates a starting value for the composition in each phase, the fugacity of each component in each phase is calculated, the equalities are checked and new estimates of the vapour/liquid composition are calculated. The program continues until the mass and enthalpy balances are fulfilled, the T,P and fugacities in each phase are the same and until there is no change in these properties between the previous and actual iteration cycle.

For each step the thermodynamic package must be called to calculate the enthalpies and the fugacity of each component in each phase. Since this determines the final result, the thermodynamic package is the heart of the flowsheet. Therefore lots of effort has been invested to get the K-values, enthalpies and properties right, and therefore there are many basic data systems specially tuned to a specific application.

## 2.1 Raoult's law

The most simple basic data system is Raoult's law (Ideal basic data) :

$$f_i^V = Y_i P \text{ and } f_i^L = x_i p_i^0 \quad \text{and} \quad K_i = Y_i/x_i = p_i^0 / P$$

where :  $Y_i$  = mole fraction of component i in the vapour phase

$x_i$  = mole fraction of component i in the liquid phase

$p_i^0$  = vapour pressure of component i

P = total pressure

Only the vapour pressure of the component is required. In the API data books the graphs of K-values of oil fractions were calculated from Raoult's law. These graphs were translated into equations (e.g. in the BK-10 system). This system should not be used at high pressures (> 15 bar) or when polar components are present (alcohol, glycol, ethers etc.). For water the decant option is used : from generalised solubility charts the water solubility is calculated. The rest of the water is decanted as a pure water phase. Also for sharp separations this system should not be used.

## 2.2 Equations of state

Most of the equations of state are extensions of the 'van der Waals' equation of state, e.g. SRK, PR., SRK:

$$P/RT = 1/(V-b) - a(T) / (V(V+b))$$

The fugacities  $f_i$  are calculated by integrating the pressure derivatives over the total volume :

$$RT \ln f_i^V = \int_{V^V}^{\infty} (\delta P / \delta n_i - RT/V) dV - RT \ln [V^V / (Y_i RT)]$$

where : P and V are Pressure and Volume

$n_i$  = number of moles of component i in the phase

a and b are pure component E.o.s. parameters

(available in the data banks or from oilfraction correlation methods)

This equation holds also for the liquid phase and a second liquid phase if present. The superscript V is than replaced by the superscripts L<sub>1</sub> and L<sub>2</sub>.

For the mixture the pure component a's and b's are summed with the mixing rule :

$$A_{\text{mix}} = \sum_i \sum_j x_i x_j (a_i a_j)^{0.5} (1 - \theta_{ij})$$

$$B_{\text{mix}} = \sum_i x_i b_i$$

where  $\theta_{ij}$  is the binary interaction parameter

The parameter  $\theta_{ij}$  is used to tune the predictions to the experimental data. Therefore, for accurate calculations, all binary interactions in the mixture must be available in the data base or must be generated from generalised methods. When the user selects PRINT INPUT=ALL, all binary interaction coefficients are printed and can be inspected. For missing important binaries the user should either generate them from binary VLE or LLE data (The REGRESS program can be used for this purpose) or ask the basic data experts to generate them for you and incorporate them into the Simsci or PEPPER data base.

The SRK or PR has the disadvantage that Liquid-liquid equilibrium calculations cannot be performed accurately. Therefore the original equation was extended to a larger temperature

range and the mixing rule was modified with a second interaction coefficient (See Refs. 2 and 3). This proprietary SMIRK equation of state was developed in 1980 :

$$\text{SMIRK: } P/RT = 1/(V-b(T)) - a(T) / (V(V+b(T)))$$

$$A_{\text{mix}} = \sum_i \sum_j x_i x_j (a_i a_j)^{0.5} (1 - \theta_{ij}^0 - \theta_{ij}^1 * (\sum_p x_p - \sum_{ap} x_{ap}))$$

Where the second term sums over the polar (water, glycol, alcohol etc.) and the a-polar components in the mixture. The availability of the binary interaction coefficients is again of paramount importance for the accuracy of the predicted results.

In ca. 1990 PRO-II incorporated SRKKD and SRKM, which has a similar mixing rule as SMIRK. The combination SRK with a slightly different formulation and with another mixing rule is the subject of research of many universities (SRKP, SRKS, SRKH etc.). It is advised to use only SRK or SRKM or SMIRK. For water solubility's the SRKKD is advised, since it was specially devised for that. It is less good for hydrocarbon solubility's in water.

### 2.3 Activity coefficient models

The activity coefficient models use a combination of a corrected Raoult's law and the E.o.s. This gives the user more flexibility to model e.g. the liquid-liquid separation more accurately :

$$f_i^V = Y_i P \phi_i^V \text{ and } f_i^L = x_i p_i^0 \gamma_i \phi_i^0 \exp(v_i^0 (P - p_i^0)/RT)$$

$$\text{and } K_i = Y_i/x_i = p_i^0 \gamma_i \phi_i^0 \exp(v_i^0 (P - p_i^0)/RT) / P \phi_i^V$$

where :  $v_i^0$  = liquid volume of component i

$\phi_i^V$  and  $\phi_i^0$  are vapour and liquid fugacity coefficients, which correct for the non-ideality due to the pressure and are calculated from the E.o.s.

The activity coefficient,  $\gamma_i$  is calculated from a local composition mixing rule, which is more sophisticated than the E.o.s. mixing rule and therefore more accurate for highly non-ideal mixtures.

E.g. NRTL :

$$\ln \gamma_i = \frac{\sum_j \tau_{ji} G_{ji} x_i}{\sum_k G_{ki} x_k} + \sum_j \frac{G_{ji} x_j}{\sum_k G_{kj} x_k} \left( \tau_{ij} - \frac{\sum_k \tau_{kj} G_{kj} x_k}{\sum_k G_{kj} x_k} \right)$$

where :  $G_{ji} = \exp(-\alpha_{ji} * \tau_{ji})$

$\tau_{ji}$  and  $\alpha_{ji}$  = NRTL binary interaction parameters (to be tuned on VLLE data)

Again, when important interaction parameters are lacking, the results may be worthless. Give PRINT INPUT=ALL, then all NRTL data are printed.

When data are missing, the user should provide them in the input. A very useful method to fill missing NRTL data is the UNIFILL method. From group contribution methods the missing NRTL data are calculated. This is better than nothing, but measured data are preferred. An alternative is to copy the data from similar binaries to the missing binary (e.g. take for 2 methyl pentane-Sulfolane those of 3 methyl pentane-Sulfolane).

### **3. PHYSICAL PROPERTIES**

To start the calculations, the minimum required properties for each component are the name, boiling point and density and preferably also the molecular weight. Such properties are extracted from the pure component data bases, or must be supplied by the user (for pure components via PETRO records or for streams in the ASSAY records).

All other properties can be generated from oil fraction correlation's like FRACHAR or C7plus (Refs. 4 and 5). Ref. 6 summarises the available physical properties in PRO/II and gives the formula's for some 60 properties. These properties are available in PROMOTE-PRO-II.

- In PRO/II the methods to be selected for the properties are spread all over the manual :

- Reid vapour pressure : In the General Data section (RVPbasis = e.g. D323)
- ASTM/TBP curves : In the General Data section : PRINT TBP.
- The method for ASTM calculation : In the Component data section : ASSAY Conversion=API63
- Oil fraction characterisation method : In the Component data section : ASSAY Characterize=SIMSCI. When STND, SMIRK or SNRTL (Shell methods) are chosen, automatically FRACHAR is selected.
- Transport properties (Viscosity, thermal conductivity, surface tension) are not automatically printed out. In the Thermodynamic data section : METHOD SYSTEM=xxx, TRANSPORT=PURE or PETRO
- Special properties : In the thermodynamic data section : METHOD SYSTEM=xxx, FLPOINT=NELSON, CETANE=ASTM. For other properties (Cloud point, pour point, sulphur content, kinematic viscosity and "own" properties (via SPROP, e.g. V50, Smoke point etc)) the user has to supply average stream properties in the Stream data section (for streams) and in the Thermodynamic data section for all library components.

### **4. BASIC DATA SYSTEMS**

The Shell basic data systems - the PEPPER data base, the oil fraction characterisation models C7plus and FRACHAR and the basic data (K value and enthalpy) systems SMIRK and SNRTL - are coupled to PRO/II via the SPPTS interface. These methods are only available in the special Shell/PROMOTE version of PRO/II.

When in PRO/II a general thermodynamic method is chosen (e.g. METHOD SET=SRK), automatically the pure compound data bank is attached and the most appropriate set of methods for K-values, enthalpies and property methods (e.g. density, entropy) is chosen. The user may override these choices by selecting other methods individually per property.

Application guidelines for the different basic data systems are given in the PRO/II keyword input manual chapter 21. They can be generalised in 4 parts:

#### **4.1 Oil and gas refinery applications at low pressures (< 15 bar)**

For example crude and vacuum columns, simple oil distillation. For such application simple methods based on e.g. API's K-value charts (BK10, GS) can be applied. When substantial light material is present, the extended or improved methods can be applied (GSE, IGS). GS gives in general more vapour than BK10. When in high vacuum furnaces cracking occurs, GS will be somewhat better than BK10. In such cases it is of course better to take the cracking into account by replacing a part of the LR feedstock by a cracked gas. Also an other oil fraction characterisation method can produce better results.

#### **4.2 Oil and gas refinery applications at high pressures (> 15 bar)**

For example gas separation trains (demethaniser, depropaniser etc.), LNG production, pipelines etc.. For such applications equations of state are required. For the study of contaminants (e.g. H<sub>2</sub>S, CO<sub>2</sub>, mercaptans) the binary interaction parameters are important for an accurate result. For some contaminants generalised binary interaction coefficients are available with hydrocarbons. Do not use these data for high concentrations of e.g. H<sub>2</sub>S.

#### **4.3 Liquid-liquid extraction, sharp distillation and chemicals**

For example Sulfolane extraction, azeotropic distillation etc. Here the interaction between the molecules requires special fitting of binary parameters, with careful checks on ternary/multicomponent or plant data. Also for sharp separations with many trays the accuracy of the basic data is very important to obtain reliable results.

#### **4.4 Special applications**

For example sour water strippers, Amine or glycol package, electrolyte solutions. These packages either use a standard method with a special data base (e.g. SOUR, glycol, alcohol), or use a special developed method (e.g. Hexamer for HF alkylation).

The user may attach his/her own basic data system via a User Added Subroutine. For this the user creates a new PRO/II EXE file. Therefore it is required to have additional UAS files and facilities (from SimSci) and a suitable compiler/linker.

The OLI package for electrolyte solutions was attached to PRO/II. This is a special ELEC version of PRO/II.

### **REFERENCES**

- 1) G. Hajek, "Basic data and phase behaviour methods", DEP 20.00.10.10-Gen., April 1993
- 2) J. de Dood and O. Heijink, "Development of the SMIRK-82 equation of state", AMGR 82.194 and AMGR 82.195.
- 3) H. Meijer and M.A.J. Michels, "Improved mixing rules for two-parameter equations of state", AMGR 84.046
- 4) S. Schreuder and E.M. Hendriks, "Complete description of oil fraction characterisation method FRACHAR", AMGR 92.220.
- 5) J. de Dood and P. Heijink, "Characterisation of C7-plus fractions for physical property calculations: Development of EPSPLIT", AMGR 85.192
- 6) J.M. Broersen, "PRO-II project : physical property prediction methods", Memorandum SIPM-MFT No. (C) 215/93
- 7) H. Meijer, A.R.D. van Bergen, E.M. Hendriks and R.N. French, "Basic data and phase behaviour methods", OP.96.20038

**TABLE 1**  
**Application areas for basic data methods in PROMOTE and PRO/II**

Application	T-max (°C)	P-max (bar)	PRO/II (standard methods)	SPPTS (Shell methods)	Remarks
Atmospheric distillation	260	2	GS	Use GS	
Mild vacuum distillation	360	0.1	GS / BK10	Use GS	
High vacuum distillation	440	0.01	BK10 / GS	Use GS or SMIRK (thermal cracking furnaces)	
Recontacting	50	15	SRK	SMIRK	
Gasoline splitter	160	2.5	GS / SRK SRKM	SMIRK SMIRK	Bulk modelling Benzene content as SPEC
Super fractionation	110	3	SRK	SMIRK	
Gasoil stripper	250	2	GS	-	
Kero stripper	170	2	GS	-	
De-methaniser	-100	100	SRK	SMIRK	T-critical ethane = 32 °C : Discontinuity in SMIRK
De-ethaniser	80	25	SRK	SMIRK	Idem
De-propaniser	100	20	SRK / NRTL	SMIRK / SNRTL	
PP splitter	70	12	SRK / NRTL	SMIRK / SNRTL	Binary data are required !
De-butaniser (stabiliser)	145	15	SRK	SMIRK	

**TABLE 1 (cont'd)**

Application	T-max (°C)	P-max (bar)	PRO/II (standard methods)	SPPTS (Shell methods)	Remarks
Lub-oil distillation	400	0.01	GS / BK10	GS or SMIRK	
Thermal Cracker, TCU	550	5		SMIRK	
Cat Cracker main fractionator	620	2.8	GSE / SRK	GS or SMIRK	
Absorber (FCC)	200	18	SRK / GSE	SMIRK	Use free water option
Bitumen blowing unit	300	1	SRK	SMIRK	
Platforming, TIP units	300	100	SRK	SMIRK	
Hydroconversion HDS / HDT	350	200	SRK	SMIRK	SMIRK Cp's and enthalpies too low for heavy oils
Hydrocracking / HYCON	450	300	SRK	SMIRK	Idem
Sour water stripper	150	10	SOUR	-	
Compressors / expanders	200	300	SRK	SMIRK	
Heat exchangers	420	300	GS / SRK	SMIRK / or GS	
Sulfolane extraction & distillation	300	20	NRTL + UNIFILL	SNRTL	Binary data are required !
Furfural extraction & distillation	300	20	NRTL + UNIFILL	-	Binary data are required !
Propane deasphalting	300	50	-	-	Not accurate for LLE
Idem work up section	300	10	SRK	SMIRK	

**TABLE 1 (cont'd)**

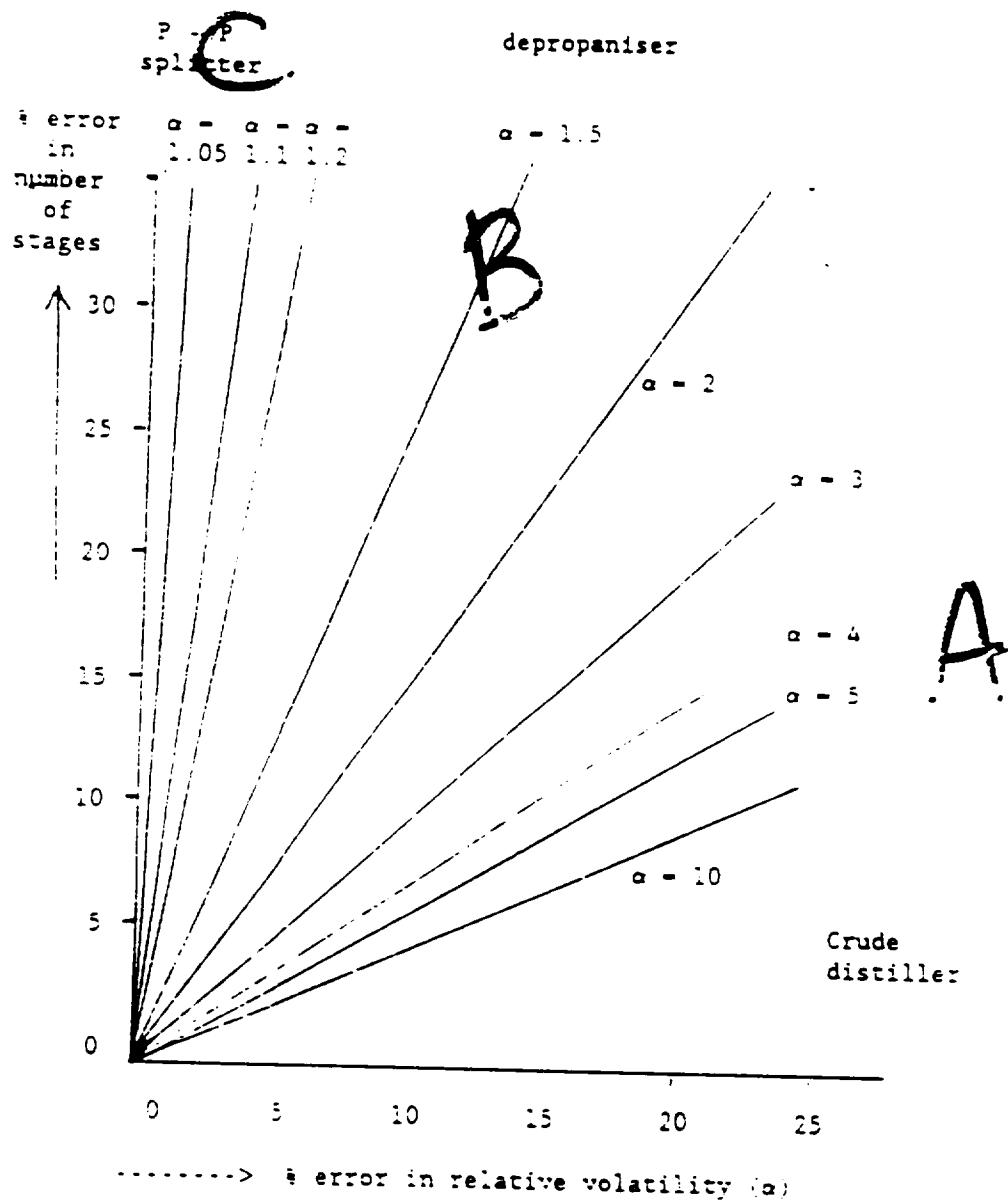
Application	T-max (°C)	P-max (bar)	PRO/II (standard methods)	SPPTS (Shell methods)	Remarks
HF alkylation & distillation	300	50	HEXAMER	-	Binary data are required !
Cryogenic heat exchangers	100	300	SRK	SMIRK	T-critical ethane = 32 °C : Discontinuity in SMIRK
Propane cycle coolers	100	200	SRK	SMIRK	
Air distillation	50	200	SRK	SMIRK	Binary data are required !
Pipeline modelling	200	300	SRK	SMIRK	
Corrosion prediction	150	10	SOUR / ELEC	-	
Hydrate prediction	25	100	PR	Use special STFLASH program	Contact KSLA EE/2
Gas drying with glycols	200	200	SRKM	SMIRK	Binary are required !
Water contamination	300	100	SRKM / SRKKD	SMIRK	Predicts hydrocarbon or contaminants solubility in water
Electrolyte solutions	100	10	ELEC	-	Special OLI package
Water wash systems for salt removal from hydrocarbons	200	10	Not possible in ELEC	-	E.g. Hot or cold pressure separators with water wash

**TABLE 1 (cont'd)**

Application	T-max (°C)	P-max (bar)	PRO/II (standard methods)	SPPTS (Shell methods)	Remarks
Azeotropic distillation	250	300	SRK / SRKM SRKM	SMIRK SMIRK	With L/L separation, e.g. cyclohexane dryers Check binary data !
Three phase column	250	50	SRK SRKM / NRTL	SMIRK SMIRK / SNRTL	Water decanting Rigorous 3 phase VLLE
Chemical reactors	250	50	SRK / SRKM / NRTL	SMIRK / SNRTL	
Reactive distillation	250	50	SRK / NRTL	SMIRK / SNRTL	

FIGURE 1

Sensitivity to error in basic data for distillation modelling



- A.**  $\alpha > 2$  : STND. UNIFAC can be applied
- B.**  $\alpha < 2$  : NRTL, SMIRK must be applied (direct fit from experimental data)
- C.**  $\alpha < 1.2$  : NRTL, SMIRK must be applied (direct fit from experimental data). Confirm against plant data



## Distillation modelling

## Exercise 1

Behaviour of water and methanol as trace impurities in the Depropaniser.

### Purpose of the exercise:

To predict by hand calculation whether water and methanol will end up in the top or the bottom product of the Depropaniser by correcting the ideal VLE calculations and activity coefficients.

The feed to the Depropaniser is contaminated with traces of water and methanol. For the design of treating facilities for the products it is required to know whether water and methanol will end up in the top product with the C3's or in the bottom product with the C4's.

A quick answer (within 30 minutes) is required. So you do not have time to model the column in a flowsheet program.

### Depropaniser data:

- The feed contains only propane, propene and n-butane.
- The pressure is 18 bara.
- The top product is 50% propane and 50% propene.
- The bottom product is n-butane.

### Phase Equilibrium

$$\text{Raoult's law} \quad Y_i P = x_i$$

$$\text{Raoult's law corrected} \quad Y_i P = x_i p_i^0 \gamma_i$$

$$\text{Total pressure} \quad P = \sum (x_i p_i^0 \gamma_i)$$

$$\text{Equilibrium constant} \quad K_i = Y_i / x_i = p_i^0 \gamma_i / P$$

Where:  $Y_i, x_i$  are the molefractions in the vapour and liquid phase resp.

$P$  = the total pressure

$p_i^0$  = the vapour pressure of component  $i$

$\gamma_i$  = the activity coefficient of component  $i$

### Vapour pressures

Find the temperature in the top and bottom of the column from the vapour pressure charts from the Physical and Engineering data book.

Take for propane/butane the molar average.

The  $K$  value (volatility) of butane in the bottom is 1. If the  $K$ -value of water or methanol is  $< 1$ , water or methanol will be leaving the column with the liquid butane. Similarly, if the  $K$ -value of water or methanol is  $> 1$  in the top column, water or methanol will leave the column with the top product.

Find the ideal  $K$ -values (take the activity coefficient = 1).

- Where will water and methanol end up with such  $K$ -values?

## Activity coefficients

The activity coefficient of water can be estimated from the water solubility in the hydrocarbon:

$$\text{Raoult's law corrected: } Y_i P = (x_i p_i^0 \gamma_i)_{\text{in water}} = (x_i p_i^0 \gamma_i)_{\text{in hydrocarbon}}$$

Since the water phase is 100% water:

$$\gamma_{\text{water in water}} = 1 \quad \text{and} \quad x_{\text{water in water}} = 1$$

$$\gamma_{\text{water in hydrocarbon}} = 1 / x_{\text{water in hydrocarbon}}$$

Find the activity coefficient of water in the bottom and top of the column (take the average in propane/propene) and find the non-ideal K-value.

- *Where will water end up with these revised K-values ?*

Since methanol/hydrocarbon dissolves completely at the Depropaniser conditions, the activity coefficient of methanol cannot be calculated from solubility's.  
From binary vapour liquid equilibria the following data were derived:

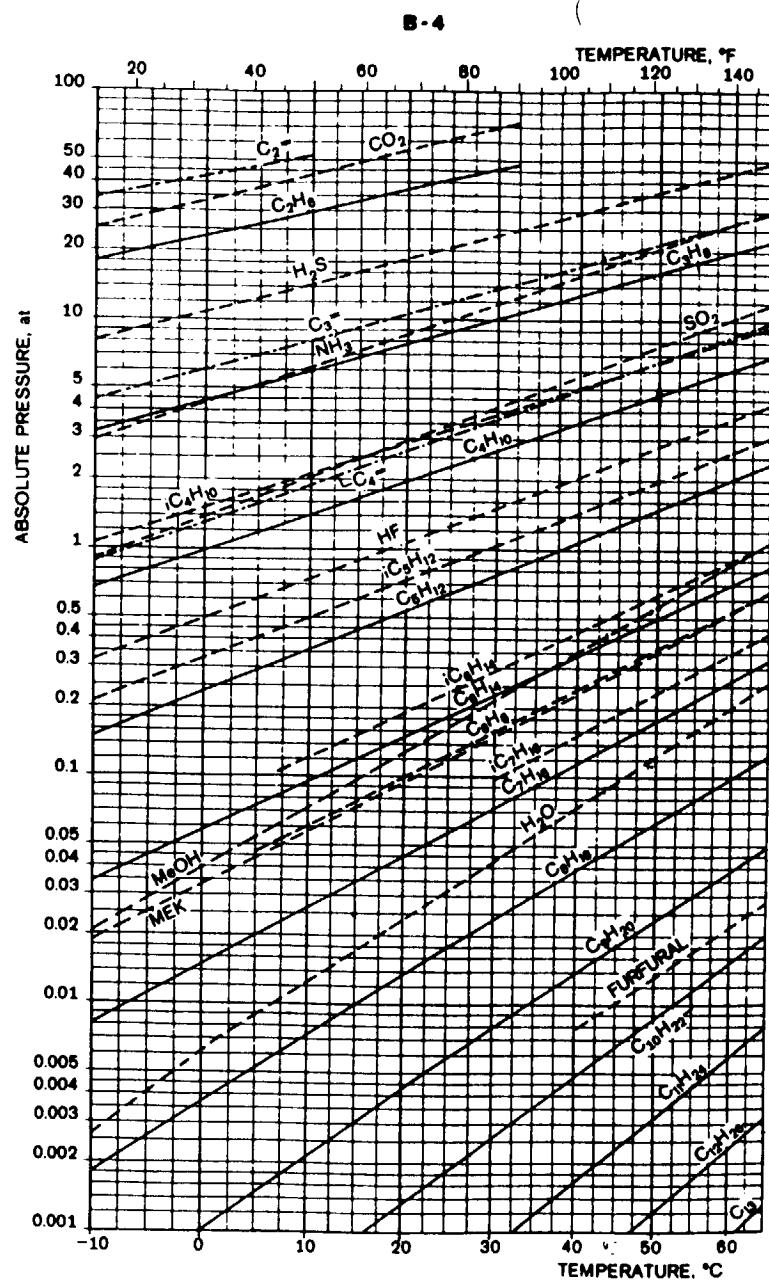
$$\text{methanol in butane at } 110^\circ\text{C} : \quad \gamma = 10$$

$$\text{methanol in propane at } 50^\circ\text{C} : \quad \gamma = 10$$

$$\text{methanol in propene at } 50^\circ\text{C} : \quad \gamma = 5$$

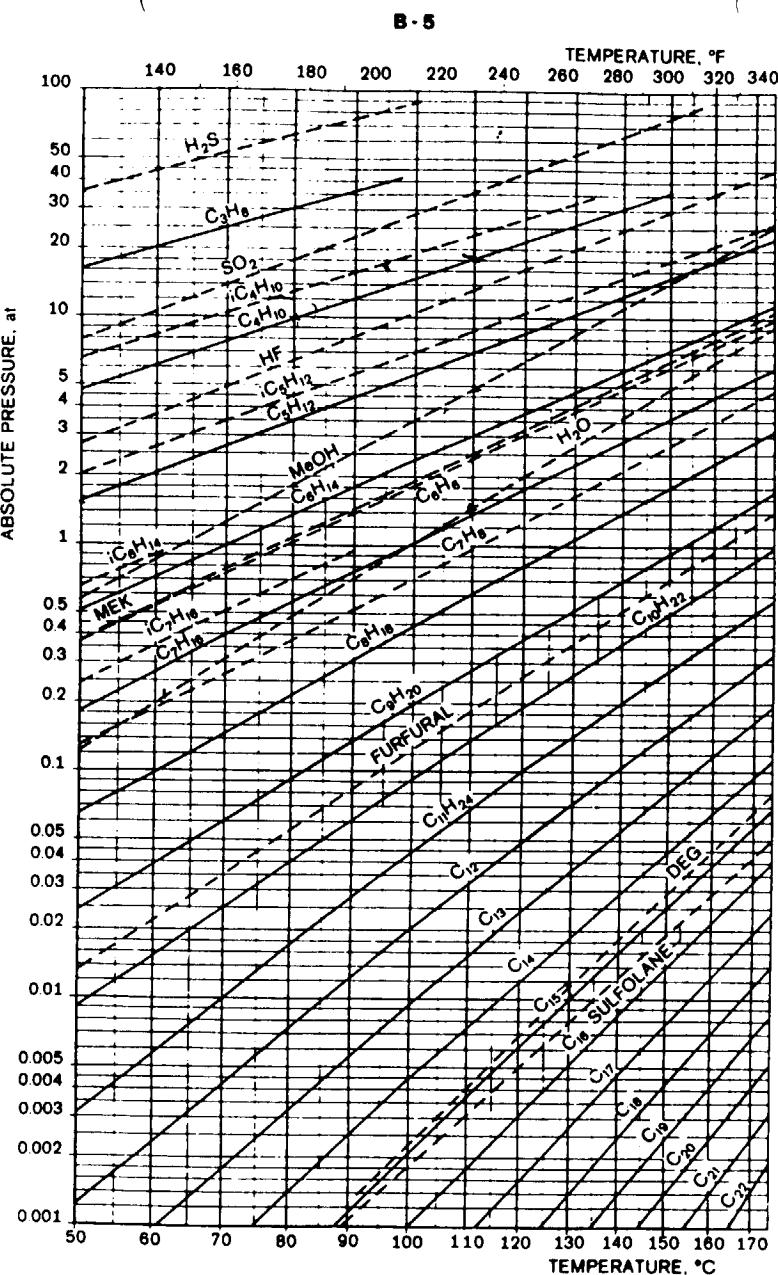
Find the non-ideal K-value of methanol.

- *Where will methanol end up with these revised K-values ?*



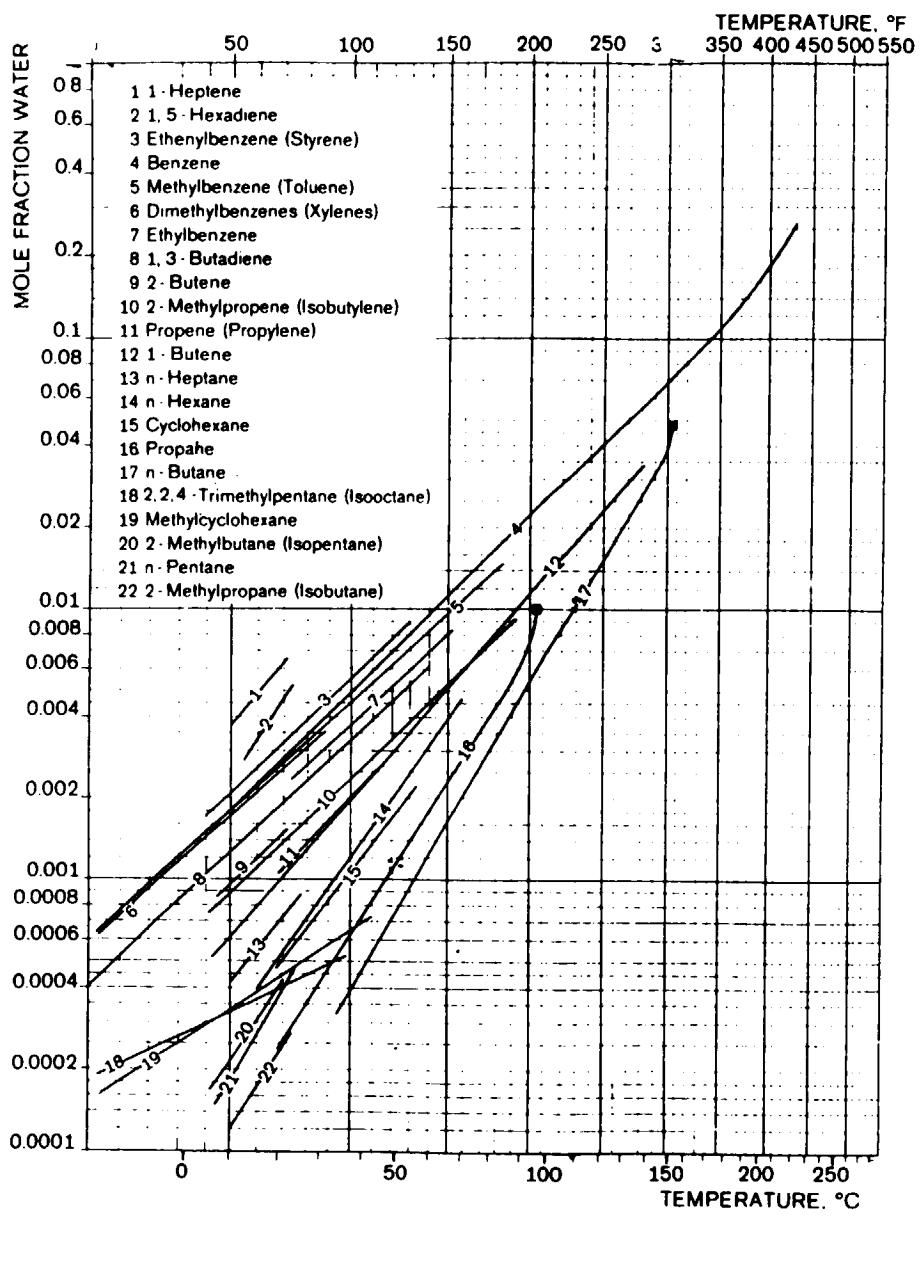
VAPOUR PRESSURE OF HYDROCARBONS  
AND SOME OTHER SUBSTANCES AT  
TEMPERATURES OF -10 TO 65 °C (14 TO 149 °F)

1 atm = 0.98 bar



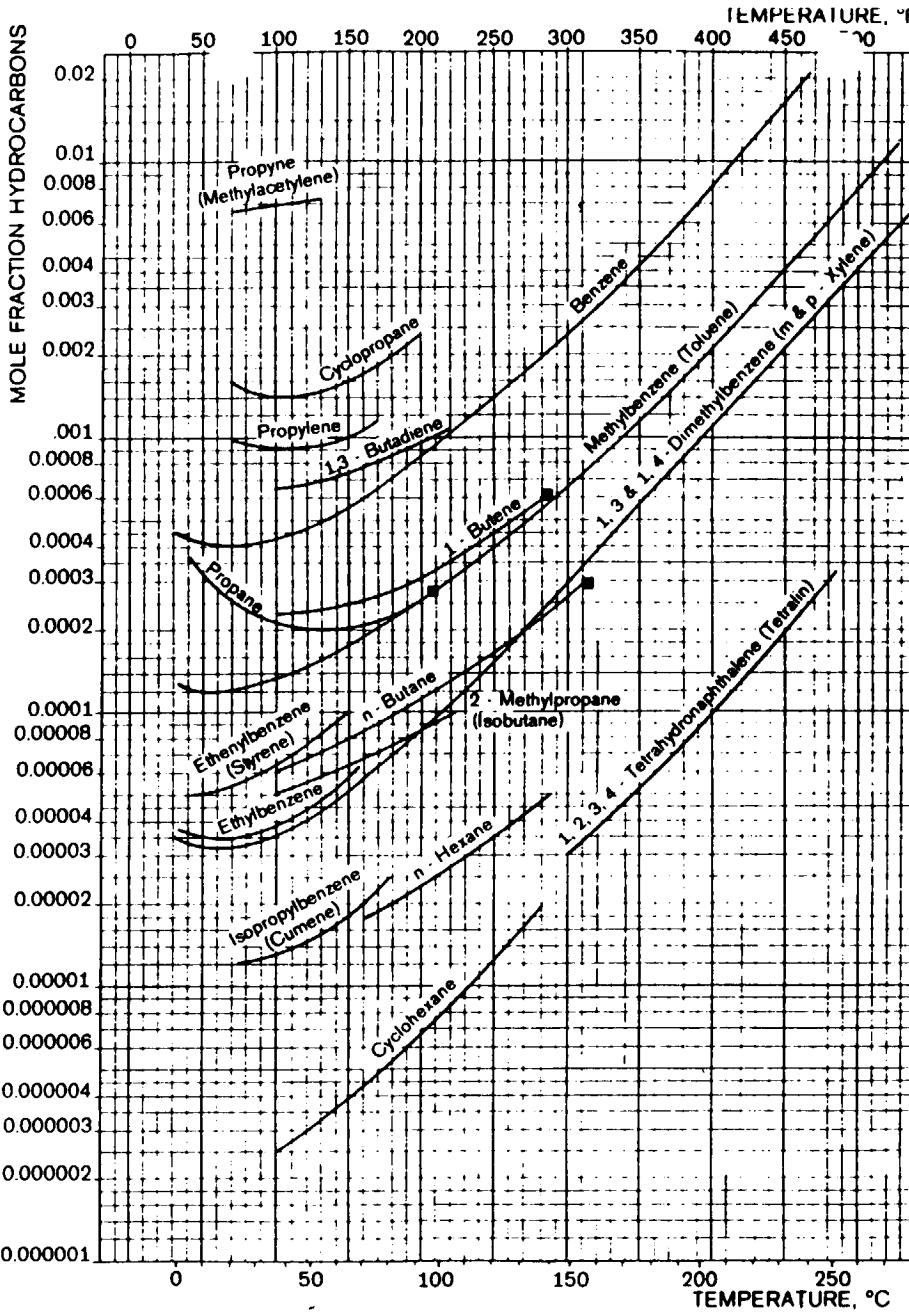
VAPOUR PRESSURE OF HYDROCARBONS  
AND SOME OTHER SUBSTANCES AT  
TEMPERATURES OF 50 TO 175 °C (122 TO 347 °F)

1 atm = 0.98 bar



■ Three-Phase Critical

### SOLUBILITY OF WATER IN PURE LIQUID HYDROCARBONS UNDER VAPOUR-LIQUID-LIQUID EQUILIBRIUM CONDITIONS



### SOLUBILITY OF PURE HYDROCARBONS IN WATER UNDER VAPOUR LIQUID - LIQUID EQUILIBRIUM CONDITIONS

■ Three-Phase Critical

## Exercise 2

Build a PRO-II model

### Purpose of the exercise:

To build a flowsheet model of a Depropaniser column and tune the model to fit the measured plant data.

### Depropaniser

<u>Feed composition</u>	<u>t/d</u>
ethane	2.0
propane	60.0
propene	60.0
iso-butane	39.8
n-butane	161.5
iso-pentane	2.0
n-pentane	1.6

### Feed preheat Heat Exchanger:

Define the heat exchanger as a fixed duty exchanger, with utility is none.

The feed enters on the cold side of the heat exchanger at 45 °C and 18 bara.

The duty of the heat exchanger is 20 ktcal/day.

Pressure drop over the heat exchanger is 0.

### Column data:

The column has 35 theoretical stages (equivalent to 44 trays).

The feed enters the column at stage 12. Note that tray 1 is the condenser.

Total bubble point condenser. Pressure drop over the condenser is 0.5 bar.

The reboiler is a Kettle type.

### Plant data:

Feed quantity	326.9 t/d
Bottom product	204.9 t/d
Pressure at column top	17.8 bara
Pressure at feed inlet	18.0 bara
Pressure at column bottom	18.2 bara
Temperature at feed inlet	81 °C

### Product specifications:

Top product C3 content	95.6 % wt
Bottom C4 content	96.6 % wt

- *Build the model of the Depropaniser*

With ‘view’ you can inspect the curves of the column, e.g. P, T or X, Y profiles.

## Exercise 3

### Test the basic data systems

#### Purpose of the exercise:

To select the proper basic data system for the behaviour of water and methanol.

#### Feed composition

Add to the feed of the Depropaniser:      water = 0.01 t/d  
    methanol = 0.01 t/d

#### Basic data system

Run the problem with two different basic data systems and compare the results.

Select as basic data system in the thermodynamic data section:

System = SRK (Soave Redlich Kwong), the e.o.s. method for higher pressures  
= SRKM (modified SRK) to accommodate polar compounds

Run the program interactively. After convergence click on the column and then on view and on plot. Ask for the methanol and water profile.

Check in the output file whether for water and methanol all binary interaction coefficients with hydrocarbons are available.

- *Find where water and methanol will end up with both these K-value methods*
- *Compare the results with your hand calculations.*

## Exercise 4

### Tray design with PRO-II

#### Purpose of the exercise:

To calculate the tray loading and maximum load factors of the column.

#### The problem

There is an opportunity for a very profitable long term deal, if the refinery could produce more Propane and Butane product. The Platformer stabiliser produces some LPG as top product. The refinery manager approaches you and asks whether this LPG could be processed in the existing Depropaniser. He needs the answer by tomorrow, and it should be a very positive YES or NO.

#### Data

<u>LPG composition</u>	<u>t/d</u>
C2	0.2
C3	65.9
i-C4	10.4
n-C4	29.6
i-C5	1.3
n-C5	0.3

#### Column data

The tray data sheets are appended.

The product and reflux pumps have sufficient capacity for the new condition.

The top and bottom purity should at least be 95.0% wt.

The battery limit temperature of LPG feed is 45°C. The additional LPG is not heated up in the feed heat exchanger.

The column pressure profile will not change with the higher throughput.

The surface tension in the bottom is 3.5 dyne/cm, in the top 4.0 dyne/cm.

#### Exercise:

1. Delete water and methanol from the feed.
2. Change the basic data system again to SRK.
3. To generate transport properties: select in the thermodynamic data section the SRK method with Transport = PURE.
4. Select the critical trays and calculate the % flood on these trays.  
The calculation method is attached.

- *What is the required reboiler duty ?*

## Tray loading calculation summary

The vapour load of a tray is given by  $\lambda_{actual}$  in m/s.

The maximum capacity at which the tray may be safely operated is at 95 % of flood.

$$(\lambda_{actual} / \lambda_{flood}) \leq 0.95$$

The vapour load of the tray is defined as:

$$\lambda_{actual} = (Q_{vap} / A_c) * \sqrt{(\rho_{vap} / (\rho_{liq} - \rho_{vap}))}$$

$Q_{vap}$  is the volumetric flow rate in  $m^3/sec$ .

$A_c$  is the column cross sectional area in  $m^2$ .

$\rho_{liq}$  and  $\rho_{vap}$  are the liquid and vapour densities in  $kg/m^3$ .

The vapour load of a calming section tray at flood is given by:

$$\lambda_{flood} = 0.122 * \alpha * \beta * \gamma / (1 + 1.286 \varphi)$$

$$\alpha = (TS / 600)^{1/3} \quad \text{where TS is the tray spacing in mm}$$

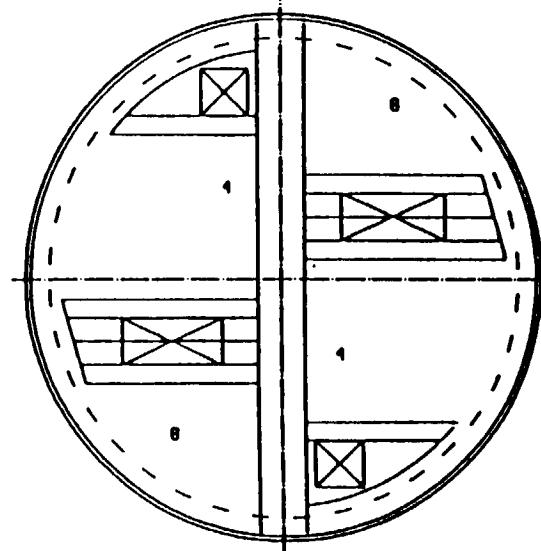
$$\beta = 0.5 + 2.9 \sqrt{\varphi} \quad \text{with } \beta = 1 \text{ for } \varphi > 0.03$$

$$\gamma = (\rho_{vap} - \rho_{liq}) / 470 \quad \text{with } \gamma = 1 \text{ for } (\rho_{vap} - \rho_{liq}) > 470$$

$$\varphi = (Q_{liq} / Q_{vap}) * \sqrt{(\rho_{liq} / \rho_{vap})}$$

Data Requestion Sheet for TRAYS				Design book No Contr Job No MESG No	Page
1 For equipment No C 1111				Specification DEP 31.20.20.31 Gen	
2 No of trays	1-26	21-48			
3 OD ID* of column	1100	1100	mm	MATERIAL SPECIFICATION	
4 Space above tray	500	500	mm	Part	ASTM No
5 <del>VALVE TRAY - BUBBLING AREA</del>				Tray panels, valves	
6 Required net free area				Support beams, hang rods	
7 Basic free area (estimated)				Calming Sections	
8 Diameter of holes <sup>1)</sup>				Weirs, downcomers	
9 VALVE TRAY - BUBBLING AREA				Wedges, clamps	
10 Required net free area	10	10	%	Bolts and nuts	
11 Number of valves per tray <sup>1)</sup>	76	76	:	Gasket material	
12 Type of valve	SHUT	SWING-IN		GENERAL	
13 Diameter of holes	40	40	mm	Welding electrodes	
14 <del>BUBBLE CAP TRAY - BUBBLING AREA</del>					
15 Number of bubble caps per tray <sup>1)</sup>				Corrosion allowance	
16 Type of bubble cap				Reference drawings	T
17 Bubble cap, OD					
18 Height of baffle				TYPE OF SEALING OF CALMING SECTION	
19 <del>GRID TRAY - BUBBLING AREA</del>				A. SEAL PAN	B. OPEN SLOTS
20 Type of bar				C. H.F. SLOTS	
21 Slot width					
22 Maximum free area					
23 To be blanked off to					
24 CONVENTIONAL DOWNCOMERS					
25 Type of downcomer				NOTES	
26 Downcomer area top/bottom				1) Design and construction shall be in accordance with the requirements of DEP 31.20.20.31 Gen.	
27 Height of tray inlet weir				2) Valves/valves/valves shall be evenly distributed; pattern to be specified by manufacturer/patterns as indicated on sheet	
28 Height of tray outlet weir				3) For A, width of seal pan is twice bottom width of Calming Section.	
29 Downcomer clearance					
30 Depth of seal pans					
31 CALMING SECTIONS					
32 Number of Calm. Sect. per tray	4	1			
33 Length at top of weir	450			mm	
34 Width at top of weir	180	220		mm	
35 Height of weir above tray	50	50		mm	
36 Height of baffle above tray	330	370		mm	
37 Depth below tray, excl. slot	300	350		mm	
38 Length at bottom	400			mm	
39 Width at bottom	100	100		mm	
40 Slot height, seal pan depth	25			mm	
41 Type of sealing (A, B or C) <sup>1)</sup>	C	B			
42 Lay-out No	H.F.				
43 INFORMATION TO BE SUBMITTED WITH THE TENDER					
44					
45					
46					
47					
48					
49					
50 REMARKS ON REVISIONS					
51					
52					
53					
54					
55 * Delete what is not applicable					
Mode by	Date	EQUIPMENT DEPRESSURIZER			Rev. letter
Checked by	Date	PLANT DISTILLATION COURSE			Date
Author by	Date	CONSIGNEE PNTK			Sign.
Eng by				Sheet No. 1 cont'd on sheet No	
Principlal				Easement No. C	
				Req. No.	

COLUMN INNER DIAMETER : 1100  
 CALMING SECTION AREA (%) : 23.49  
 CALMING SECTION TOP WIDTH (mm) : 180  
 PANEL WIDTH (mm) : 347  
 TOTAL NUMBER OF SLOTS PER TRAY : 24



NOTES:

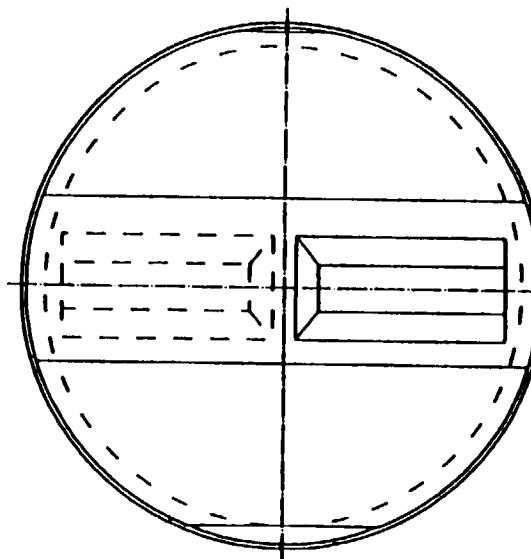
1. ALTERNATE TRAYS ARE MIRROR IMAGE.
2. NUMBERS ON DRAWING DESIGNATE NUMBER OF SLOTS IN GROUPS.

LOCATION : DISTILLATION COURSE  
 COLUMN : DEPROPYRANISER  
 TRAYS : 1-28

Sheet No.	cont'd on sheet No.
Design No.	
Req. No.	

Eng. by :  
 Principal :  
 DEP 101C DC 00 - Conv. sheet 2 2169

COLUMN INNER DIAMETER : 1100  
 CALMING SECTION AREA (%) : 18.42  
 CALMING SECTION TOP WIDTH (mm) : 220  
 CALMING SECTION TOP LENGTH (mm) : 450  
 PANEL WIDTH (mm) : 350



NOTES:

1. ALTERNATE TRAYS ARE MIRROR IMAGE.

LOCATION : DISTILLATION COURSE  
 COLUMN : DEPROPYRANISER  
 TRAYS : 21-48

Sheet No.	cont'd on sheet No.
Design No.	
Req. No.	

Eng. by :  
 Principal :  
 DEP 101C DC 00 - Conv. sheet 2 2169

## Exercise 5

Optimisation with PRO-II.

### Purpose of the exercise:

To select the optimal heat input and the feed tray location(s).

### Data

The reboiler has a maximum capacity of 30 ktcal/day.

The feed preheater heat exchanger has sufficient spare capacity but the maximum outlet temperature is 150°C.

There is some freedom for the location of the feed tray(s).

The maximum condenser duty is 50 ktcal/day.

### Exercise

1. Find the duty of the feed preheater and the feed tray location(s) that minimises the reboiler duty.
2. Optimisation in PRO-II:

The optimiser is a unit operation.

Define the objective function and define the variables and constraints.

The default defines the condenser as heater 1 and the reboiler as heater 2.

Note that the minimum and maximum values are required on constraints and vary's.

- *Which group find the lowest reboiler duty?*



**8**

## **Constraint diagrams**

## **Constraint Diagrams**

### **Introduction**

To explain the various operating characteristics of a distillation column a “straight-run” Stabiliser, also called Debutaniser is chosen as an example. The qualitative trends are general and also applicable to other types of distillation columns.

Typical for a Stabiliser column is that both the top and bottom product should meet a tight specification. This is the major concern for the operation of the column because the product qualities cannot be corrected by the downstream processes.

The LPG top product should fulfil the NGAA specification for pentanes: max. 1.0 vol.% C5 on total C4.

The FR Naphtha bottom product should comply with a vapour pressure specification. This is determined by the vapour pressure specification for the FR Tops fraction (typical RVP of C5-65 tops fraction: 14.0 psia) resulting in an allowable concentration in the stabiliser bottom product of C4 < 0.7% wt.

A distillation column with two product streams has exactly six independent parameters to characterise the process. Two out of these six parameters are assigned to the number of theoretical stages used in the rectifying and stripping sections of the column. So for an existing column with a fixed number of trays there are four process parameters which are used as free variables to operate the column.

The separation between two product streams is characterised by the cut point and sharpness of separation. For the operation of the column these two parameters are typically associated with the reflux flow and reboiler duty. They are used in combination to arrive at the required qualities of top and bottom product. For a Debutaniser it is customary to control the LPG quality with the reflux flow which changes the cut point. Sharpness of separation is then set by a constant heat input to the reboiler to maintain the bottom quality within the specification.

This leaves two independent free variables which may be utilised to optimise the operation of the column.

## **Constraint diagram**

The reflux flow and reboiler duty are normally used to keep the product qualities on specification. There are however other operating variables which are not always recognised as such. They are seldom adjusted since their effect on the column operation cannot directly be assessed. Nevertheless they do have a direct impact on the operating cost of the column.

- When the column and its related equipment has ample spare capacity there may be an excellent opportunity to reduce the operating cost of the column. This holds especially for columns operating at a low throughputs. The aim is to find optimum values for the free variables which minimise the operating costs on heating by utilising these margins.

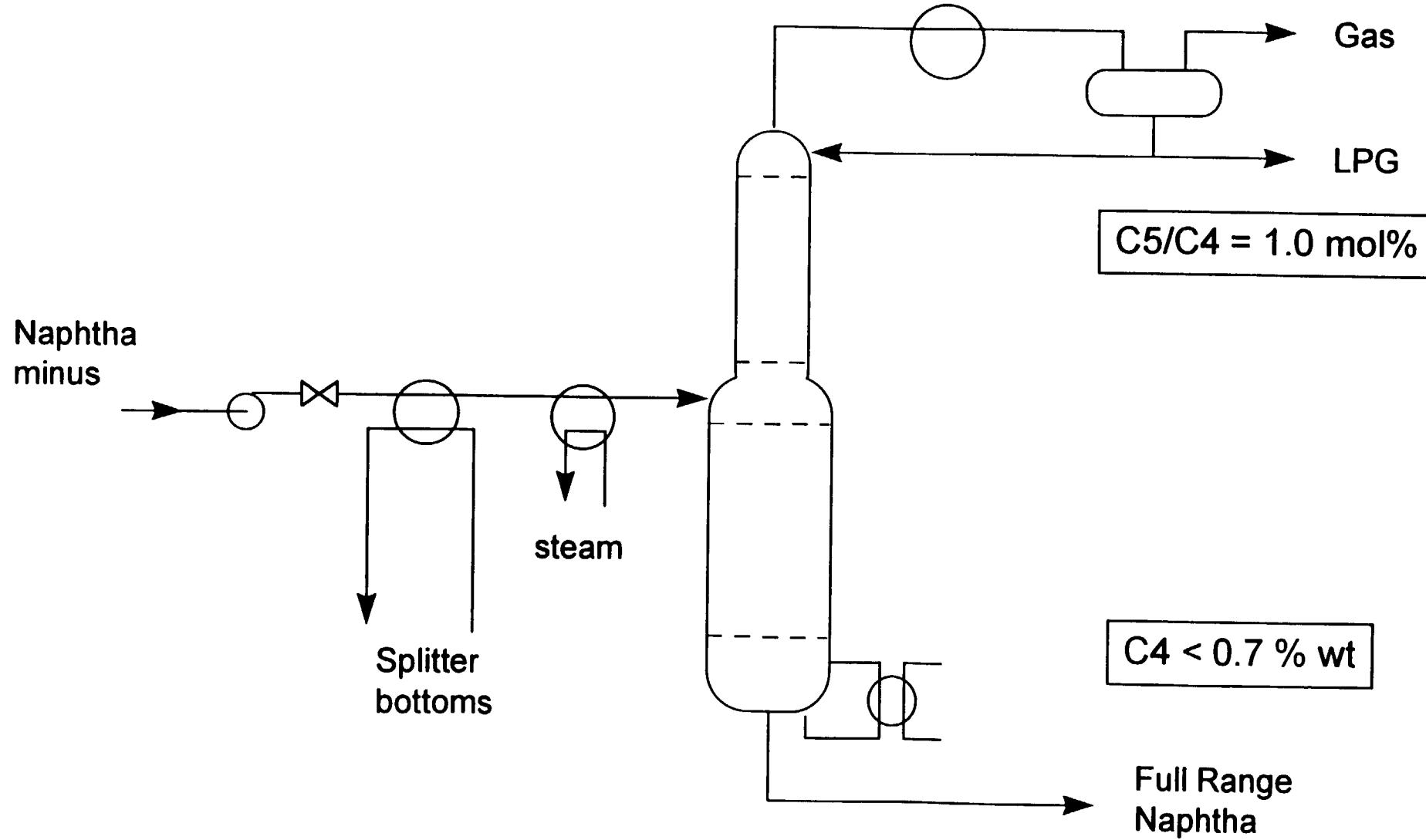
A constraint diagram can be a useful tool to optimise the operation. The diagram can be established by means of flowsheet simulations of the column. It shows the limitation of the main equipment such as:

1. Condenser capacity
2. Reboiler capacity
3. Tray loading in stripping section
4. Tray loading in rectifying section
5. Feed preheat exchanger capacity

The feasible working area of the column is marked by the envelope enclosed by the constraint lines of the equipment. Note that the diagram is only valid for one specific throughput with both top product and bottom product within specification. For higher throughputs the working area becomes steadily smaller and smaller until the point is reached where there is no further operational flexibility. On the other hand, when operating at lower throughputs the working area increases.

In addition to the equipment constraints also lines of constant operating cost can be plotted in the diagram. The operating point at minimum cost is in general at the intersection of two constraint lines; operation by riding two constraints.

# Debutaniser



## Constraint diagram exercise

### Debutaniser operation

The operating limits of the Debutaniser column were scrutinised with the help of a flowsheet simulating column operation.

The quality of the top product was set at 1 vol% pentane on total butane, the specification of LPG final product.

The butane content of the Debutaniser bottoms was set at 0.7 %wt to keep the Reid vapour pressure of the TOPS product within specification.

A set of graphs shows the result of the simulation study.

Fig. 1 Maximum condenser duty

Fig. 2 Condenser duty for 2100 t/d feed intake

Fig. 3 Maximum reboiler duty

Fig. 4 Reboiler duty for 2100 t/d feed intake

Fig. 5 Tray loading stripping section

Fig. 6 Tray loading rectifying section

Fig. 7 Feed inlet temperature for 2100 t/d feed intake

Fig. 8 Heating cost for 2100 t/d feed intake

Fig. 9 Total heating cost

### Exercise 1

What are the free variables to draw the constrained diagram.

### Exercise 2

Construct a constraint diagram for the Debutaniser  
for a feed intake of 2100 t/d

### Exercise 3

Draw the lines of constant total heating cost in the diagram.

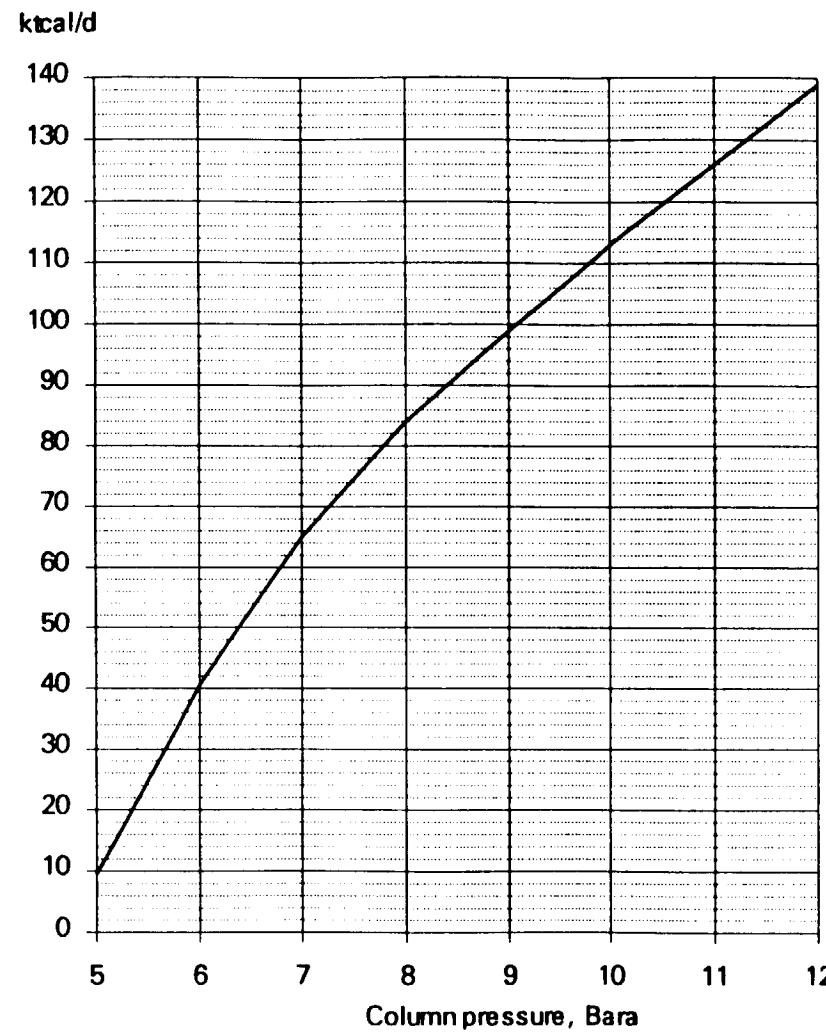


Figure 1 Maximum condenser duty

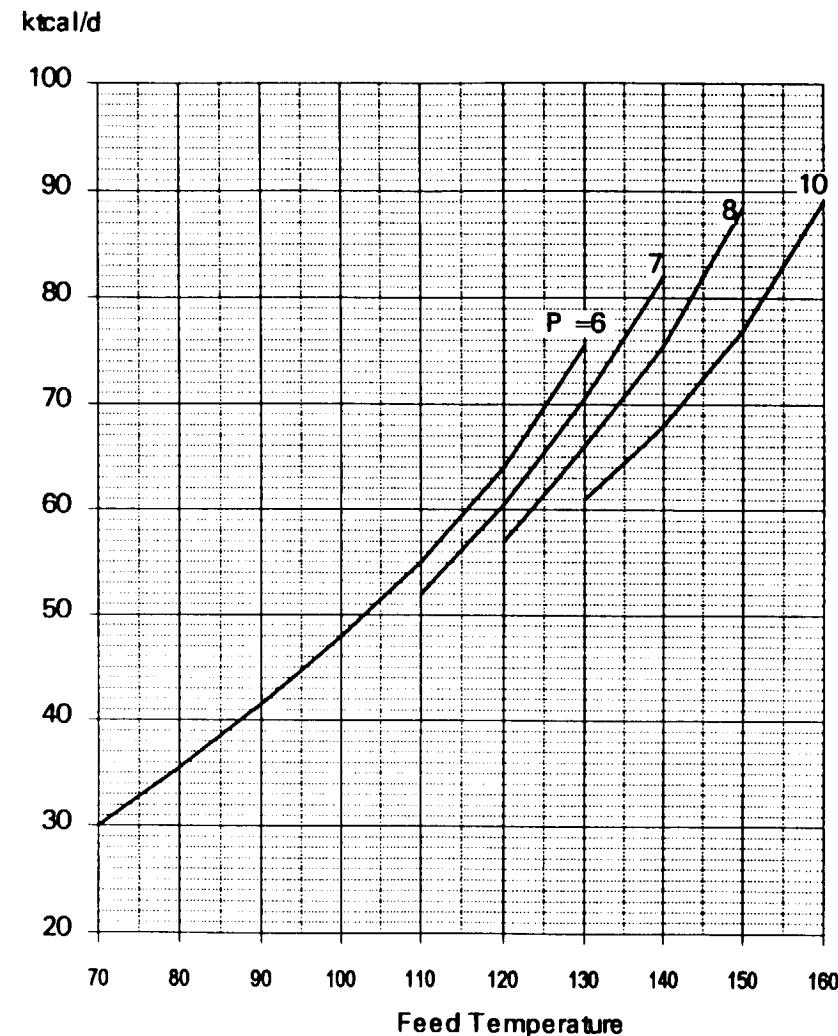


Figure 2 Condenser duty for 2100 t/d feed intake

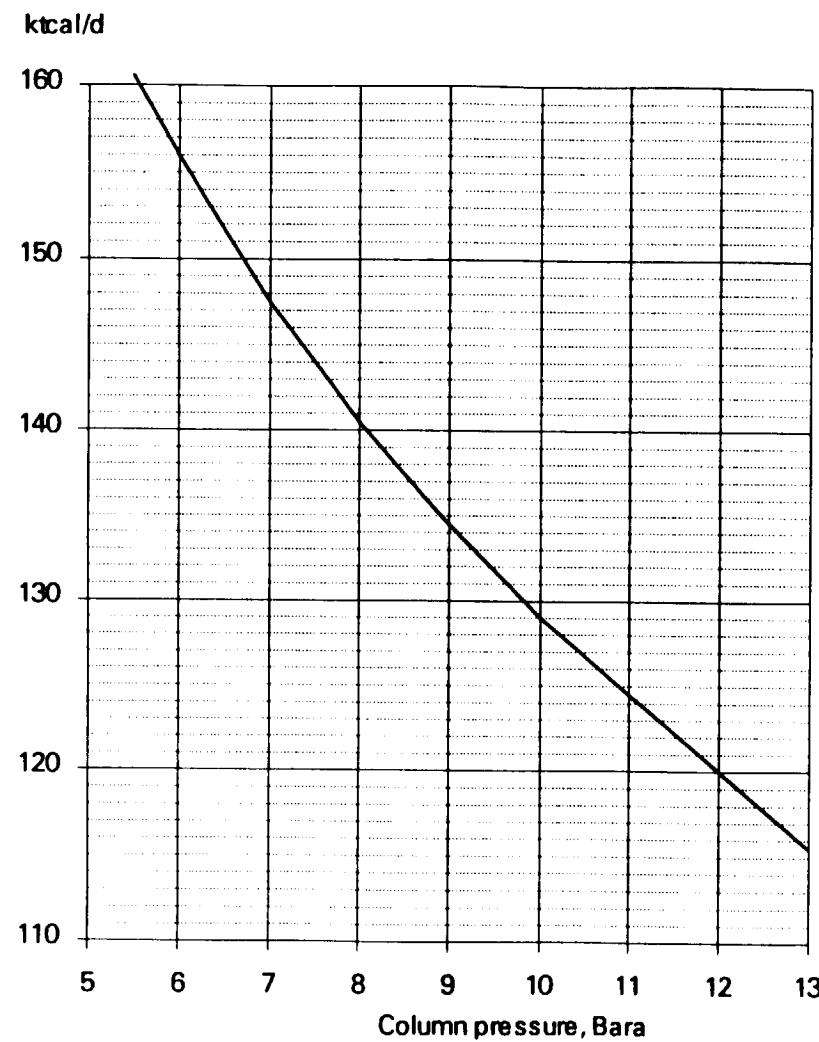


Figure 3 Maximum reboiler duty

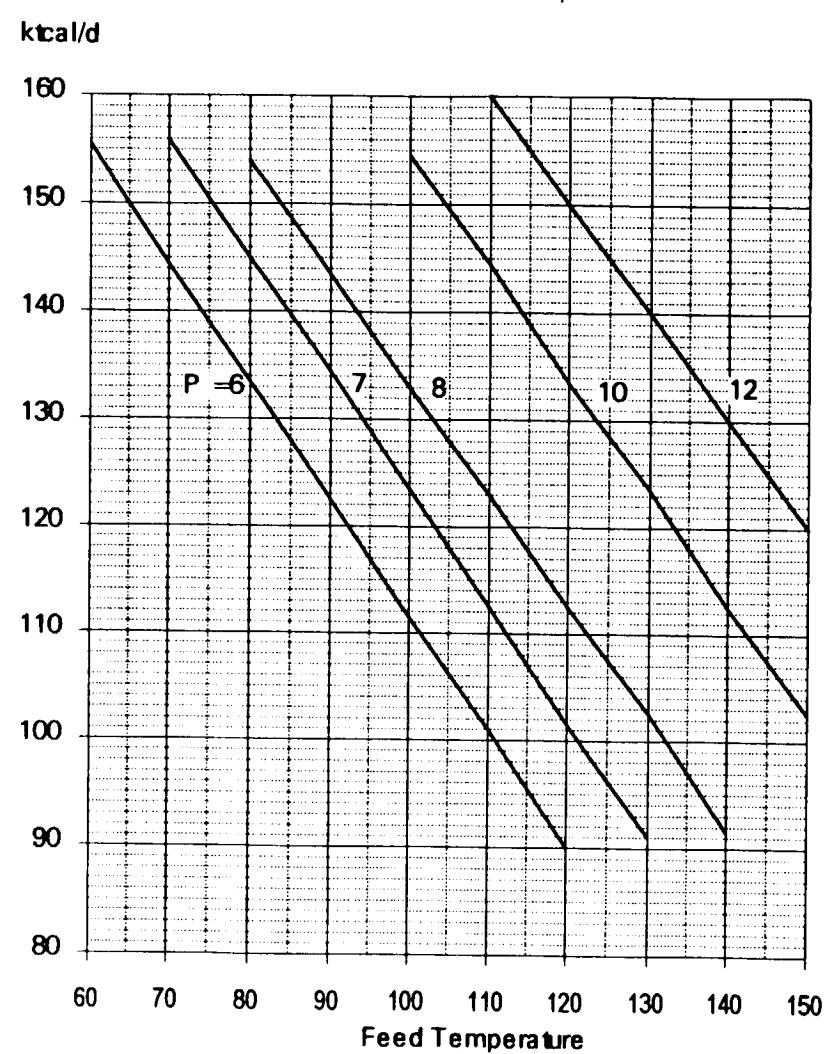


Figure 4 Reboiler duty for 2100 t/d feed intake

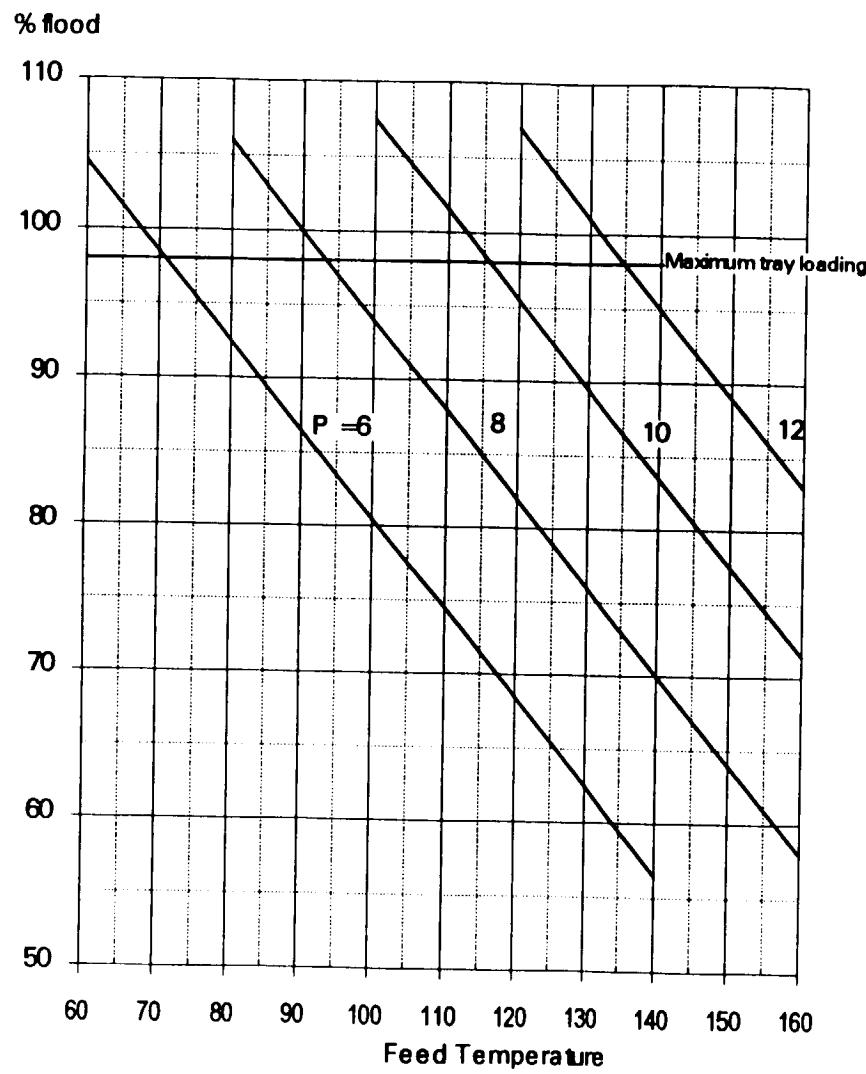


Figure 5 Tray loading stripping section for 2100 t/d feed intake

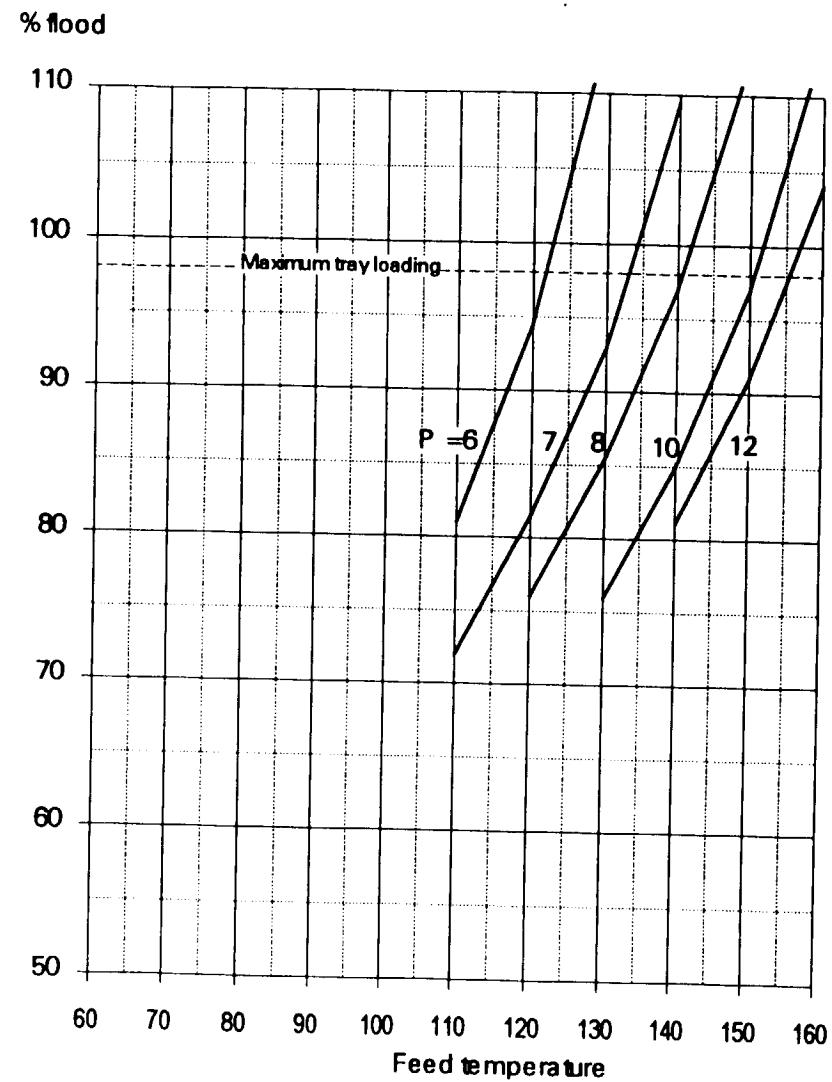
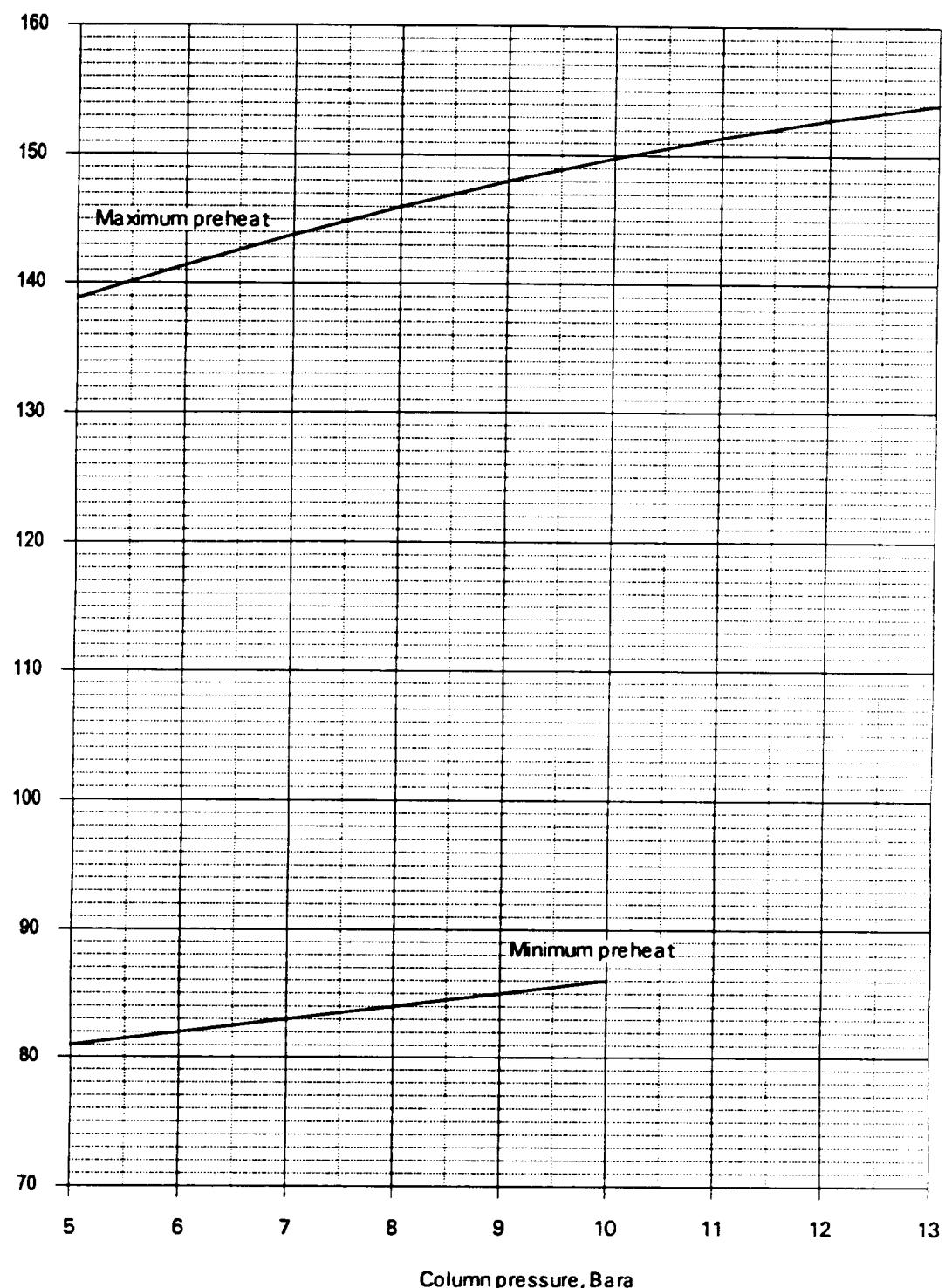


Figure 6 Tray loading rectifying section for 2100 t/d feed intake

**Feed temperature**



**Figure 7 Feed inlet temperature for 2100 t/d feed intake**

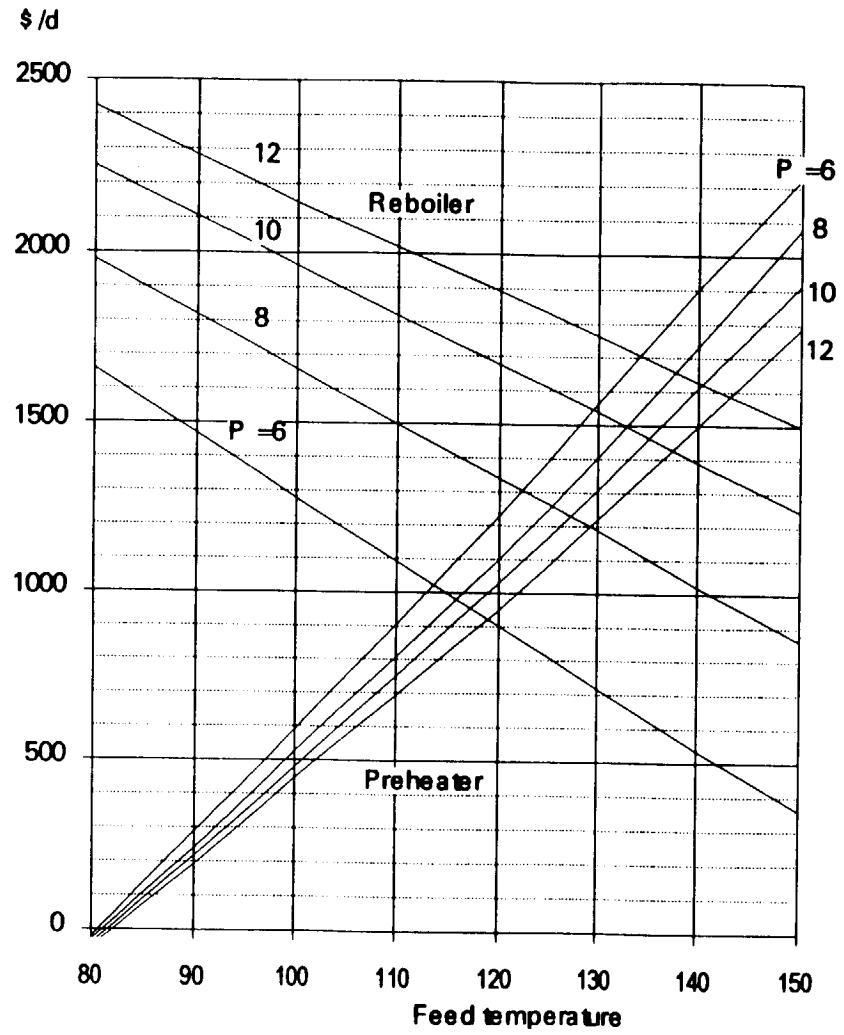


Figure 8 Heating cost for 2100 t/d feed intake

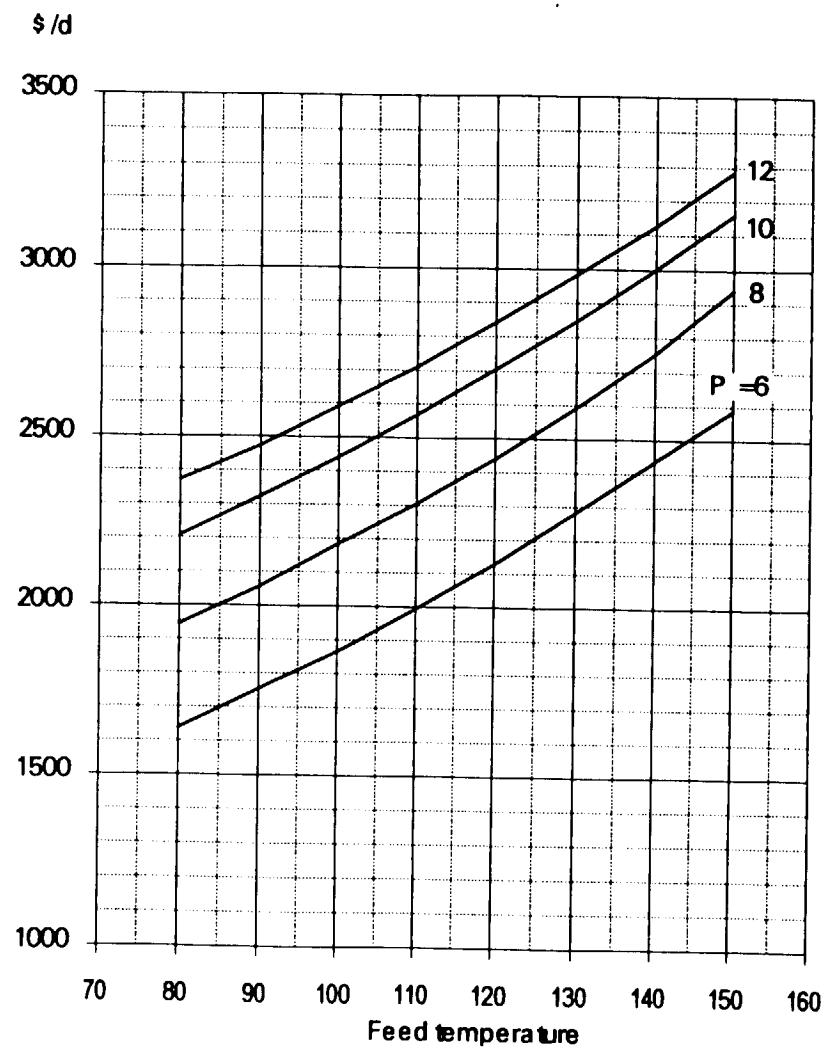


Figure 9 Total heating cost for 2100 t/d feed intake



**9**

## Packing for distillation

## **1. Introduction**

Packings can, depending on the process conditions offer significant advantages over alternative types of column internals. There are two separate types of packings: the random or dumped type and the so called structured type (Mellapak, Montzpak etc). Some types of well known random packings are available from a number of manufacturers (e.g. Raschig and Pall rings) whereas each structured packing is at present proprietary to a single supplier. Many modern random and structured packings offer a very good performance, although experience has shown that the claims and advertisements of manufacturers must be considered with care as they sometimes give a too optimistic view.

In this course manual attention will be focused on the use of packings in distillation columns, with the main purpose to provide sufficient information to enable participants to check or monitor the performance of existing facilities, check new installations and appraise new or modification proposals.

Information on auxiliary equipment such as liquid distributors, support and hold down grids is given a broad brush basis, however enough for the above purposes. For detailed considerations of auxiliary equipment the advice of an expert should be sought. It should be noted that the correct design, installation and operation of auxiliary internals is very important for obtaining the required performance, particularly for high capacity packings which can be sensitive to flow maldistribution.

## **2. Comparison Between Packings and Trays**

The selection between packings and trays is often not a clear cut case. There are various pro's and con's:

### **First the pro's:**

- Packings will give a lower pressure drop per theoretical stage.
- For smaller columns (say below 0.8 m diameter) packing is often more attractive than trays because accessibility constraints in small columns makes installation and maintenance of trays difficult.
- Packing is available in several materials of construction, e.g. plastics and ceramics which might be advantageous in corrosive services where similar material for trays is not available.
- Entrainment from a packed section is, in general, lower than for trays.
- Especially at low internal liquid rates, e.g. in vacuum column, the capacity of packings can be higher than that of trays. In these services packings can have significant advantages over trays.
- Packed columns are less sensitive to foaming than trayed ones.
- Depending on the situation, packings can have a better efficiency i.e. a lower HETP than trays and thus save on column height.

### **Now the con's:**

- Packings are very sensitive to the maldistribution of vapour and liquid. Extra packed height is required to compensate for the particular degree of maldistribution. The proper design, installation and operation of distribution devices is therefore essential.
- Packed columns require significantly more space and height for auxiliary column internals such as distributors, supports and draw-off trays as compared to trays.

- In the case of packed columns visual inspection the column wall is not possible with the packing installed.
- Trays are usually cheaper than packings. However for a complete economic evaluation also the auxiliary equipment and required column height should be valued.
- At high liquid loads the capacity of calming section trays will generally be higher than that of packing.
- Trays must be applied when two different liquid phases occur anywhere in the column.

### **3. Selection of the Packing Size and Type**

Comments on various aspects of different types of packings will be found in the Appendix. The first choice to be made is between random or dumped packings and structured packings. Structured packing consists of a package of corrugated plates ordered in a regular way (see figure 1). Most packings are available in different sizes. Random packings are characterised by the (nominal) size of the packing piece and structured packings by their specific surface area ( $m^2/m^3$ ). The highest efficiency is obtained with the smaller random packings and structured packings with the greater specific surface area. However the higher efficiency packings also tend to have the lower capacities. The optimum choice is therefore a compromise between efficiency and capacity, which will vary from case to case. However it will often be found that either 50 mm Pall rings or Mellapak 250Y, or equivalents, are a good starting point for evaluation.

Due to their regular structure which promotes more efficient wetting and a lower tendency towards channelling and entrainment for a given required efficiency (HETP) the capacity of structured packing is mostly higher than that of random packing. Note however that at very high liquid loads Pall rings have a higher capacity. Structured packings as a general rule will be more expensive than dumped packings.

Several propriety random packings have a lower pressure drop than Pall rings. The capacities however are only slightly higher, and the lower pressure drop packings are more sensitive to flow maldistribution. Therefore if a random packing is chosen it is recommended to apply Pall rings unless the pressure drop is critical. For most applications Intalox/Berl saddles and Raschig rings are no longer a choice due to their poor capacity performance relative to other packings.

The wire gauze structured packings, such as Sulzer BX, offer the advantage of high efficiency at low liquid loads ( $\phi < 0.1$ ). They are however expensive and may be very sensitive to fouling. Sulzer Kerapak has similar characteristics and may be used in ono-aqueous corrosive service. The characteristic size of a packing should not be greater than 10% of the column diameter.

### **4. Performance of Packing**

How a packing performs depends on its capacity, pressure drop and efficiency. In so far as the actual packing is concerned no detailed sizing for liquid or vapour flows is required (as against the case with trays). So the selection of packing type and size is somewhat simpler. The main hydraulic problem lies in the proper fluid distribution, to provide and maintain an evenly distributed liquid and vapour flow through the packing. This is the subject of the next section, in this section we will assume that the liquid and vapour stream are well distributed.

## 4.1 Capacity

There is always a limit to the maximum amount of vapour and liquid that a packed section can handle. By increasing the vapour flow the liquid hold-up will also increase. Consequently because of the increased vapour liquid contact the efficiency will improve but there will also be an increase in the pressure drop over the packed section. At the limit of the vapour flow the liquid hold-up is such that the liquid can no longer flow downwards against the rising vapour and flooding occurs.

Just as is the case with trays, the flooding vapour load  $\lambda_f$  can be plotted against the flow parameter  $\phi$ . In figures 2 and 3 data for various structured and random packings is given. The  $\lambda_f$  from figure 2 or 3 has to be corrected by some system properties to arrive at the maximum allowable  $\lambda_{\max}$ :

$$\lambda_{\max} = \lambda_f * \sqrt{F} * \gamma * \delta$$

The correction factors F,  $\gamma$ , and  $\delta$  are given as follows:

- **Foam factor, F**

The main causes of foaming are surface properties which cannot be predicted in most cases, for this reason the foaming tendency is best judged on the basis of experience of similar systems. Examples of foaming liquids are some crude oils, heavy residues, absorption and extraction solvents. Packings are less sensitive to foaming than trays, so the de-rating factor is the square root of F. In contrast for trays the factor F is used. A list of typical foam factors is given in the section trays. F = 1 for non-foaming systems.

- **Density difference correction factor,  $\gamma$**

The correction factor  $\gamma$  should be applied when the density difference is less than 470 kg/m<sup>3</sup>. It is given by:

$$\gamma = \frac{\rho_l - \rho_v}{470}$$
$$\gamma = 1 \quad \text{for } (\rho_l - \rho_v) > 470 \text{ kg / m}^3$$

- **Liquid viscosity correction factor,  $\delta$**

The liquid viscosity correction factor should be applied when the liquid viscosity  $\mu$  is more than 0.002 Pa.s (2 cP). It is given by:

$$\delta = (0.002/\mu)^{0.05}$$
$$\delta = 1 \quad \text{for } \mu < 0.002 \text{ Pa.s}$$

Finally the design  $\lambda_{\text{des}}$  should be derived from  $\lambda_{\max}$  by:

$$\lambda_{\text{des}} = \frac{\% \text{ fl}}{100} * \lambda_{\max}$$

The percentage of flood is usually taken at 70 to 80%, but may be as high as 90% in revamp cases. From the  $\lambda_{\text{des}}$  the required column diameter can be calculated in the same way as for trays.

## De-entrainment Sections

Low pressure drop packings are frequently applied as de-entrainment devices at low flow parameters ( $\phi < 10$ ), for example in wash oil sections. In such cases the entrainment from the top of the packing at the stated maximum capacity would be significant. This has little effect on mass transfer efficiency but is unacceptable for de-entrainment. For this reason the maximum load factor in the de-entrainment sections is somewhat less than given in figures 2 and 3. The following values for typical packings are given as a guide (for  $\phi < 10$ ):

Packing	$\lambda_{\max}$ (m/s)
Mellapak 125Y	0.14
Mellapak 250Y	0.12
50 mm Pall rings	0.09

The load factor at design may be taken as 85% of the  $\lambda_{\max}$ . Note that the use of spray distributors at load factors greater than 0.10 m/s should be considered with great care in view of the risk of liquid being entrained from the spray before reaching the packing.

In connection (particularly) with de-entrainment sections it is important to note that a minimum packing wetting rate should be adhered to in order to prevent coking/fouling up of the packed bed. For packings typically applied in these sections minimum wetting rate guidelines (superficial liquid velocities) are given below:

Packing	Minimum Wetting Rate (mm/s)
Mellapak 125Y/250Y	0.1/0.1
Pall Rings/IMTP	0.3/0.3

## 4.2 Pressure Drop

At a given flow parameter, the given pressure gradient due to flow in a section of packing is given by:

$$\frac{\Delta P}{H} = 0.54 (\rho_l - \rho_v) \left( \frac{\lambda}{\lambda_{fl}} \right)^2 \cdot \left\{ 1 + \left( \frac{\lambda}{\lambda_{fl}} \right)^3 \right\} \text{ Pa / m}$$

where  $\lambda_{fl}$  is the load factor at flooding, for the given flow parameter, taken from figures 2 or 3. For conversion of the pressure drop to mm of system liquid or mm of Hg the following may be used:

$$\Delta P \text{ (mm system liquid)} = \frac{\Delta P \text{ (pa)}}{\rho_l * 9.81} * 1000$$

$$\Delta P \text{ (mm Hg)} = \frac{\Delta P \text{ (pa)}}{13600 * 9.81} * 1000$$

### **4.3 Efficiency**

The efficiency of packing is given in terms of packing height which is equivalent to one theoretical stage (plate). The HETP values for various type of packing are given in Table 1. They are based on proper flow distribution and stripping factors close to unity.

The required height of a packed bed for a given number of theoretical stages (NTS) is not simply made up of HETP\*NTS, as there are other factors which have to be taken into account:

- Additional packing height for 'end effects'.  
To cope with liquid entrance effects usually 0.5 m height of random packing or an extra layer of about 0.2 m of structured packing is required.
- Additional packing height in the case of low liquid loads.  
For random packing a minimum liquid wetting rate of about 2 mm/s is of importance. If the superficial liquid velocity is lower the efficiency is effected and the bed height should be increased.
- Additional packing height in the case of Marangoni effects.  
The Marangoni effect relates to negative surface tension gradient. If the liquid surface tension decreases by more than 0.32 mN/m per theoretical stage as it passes down the column then special effects can occur.
- Additional packing height needed to account for maldistribution.  
Maldistribution can have a serious effect on packing performance. Some liquid maldistribution will always exist and its magnitude depends on the type of liquid distributor design. Given the inherent distributor performance the correction to the simple calculation HETP\*NTS will depend on the NTS of the particular application. The calculation of these aspects requires careful consideration. As an illustration, a packed bed designed to perform as 5 theoretical stages with a spray type distributor would require about an additional 30% of packing height to account for maldistribution alone.

## **5 Auxiliary Internals**

### **5.1 Liquid Distributors**

A good liquid distribution by which the liquid flow is evenly spread out over the entire cross sectional area is very important if the column is to achieve its expected performance. This is due to the fact that there is very little horizontal mixing, (as against the case with trays) the liquid just flowing downwards. The requirement for good even distribution becomes more stringent the higher the packed bed is or in the case that the liquid load is already low.

In small columns with diameter of below 1.5 m a pan type tray may be used with vapour chimneys and gravity flow devices for liquid distribution. For columns larger than about 1.5 m in diameter a spray nozzle type distributor can be used when less than 4 or 5 stages are required. In other cases a gravity distributor similar in concept to a draw-off tray with troughs and gutters is used, where the trough and gutters contain an arrangement of holes or drip pipes (see figure 4).

### **5.1.1 Spray distributors**

A spray nozzle distributor is suitable for applications in which not more than 4 or 5 stages per bed are required. It is not suitable when the load factor exceeds 0.12 m/s or when entrainment from the spray is to be avoided. An example is shown in figure 5.

- **Layout**

The preferred layout is the “spider” type, with wet nozzles arranged in an essentially triangular pitch, which gives more even wetting than a square pitch. Distance of spray nozzle to top of bed is typically about 500 mm, but should be chosen in conjunction with the spray angle and spacing such that the minimum coverage is 150%.

The diameter of the pipes should be such that velocities are in the range 0.8 to 2.5 m/s. Minimum pipe diameter is usually 1.5".

- **Nozzles**

Most nozzles do not spray properly when the pressure drop is less than 0.5 bar, which is then the minimum that should be applied. Pressure drop at the design point should be in the range 2 - 4 bar. To avoid fouling it is necessary to keep the minimum free passage as large as possible: the minimum acceptable is 4 mm. Generally a 90 degree spray cone is used but 120 degree is acceptable if this is needed to meet the minimum free passage criterion of 4 mm.

To minimise erosion of the nozzles by particles which are inevitably present in process streams (rust, coke etc), the nozzle material should be specified as carborundum or hardened steel or equivalent.

For fouling conditions, special designs should be applied, taking into account experience in similar services and expert advice.

During operation it is useful to monitor the trend of the pressure drop over the spray nozzles vs. flow rate. An increase in pressure drop indicates nozzle blockage. For this reason a pressure gauge should be installed just upstream the column nozzle for the spray distributor.

### **5.1.2 Gravity distributors**

Gravity distributors (see figure 4) can be designed to give a very even distribution of liquid at the top of the bed, but they should be of the “drip point” type, and no the “notched trough” type. As can be appreciated the degree of maldistribution obtained will be a consequence of the care taken with design, manufacture and installation of the distributor.

To obtain proper wetting of the top of the bed it is necessary for the distributor to have sufficient drip points. Furthermore, proper irrigation of the wall zone is also important, so the drip point pattern must continue right up to the column wall. Coarser packings require fewer drip points, as indicated in the table below:

Random packing ring size (mm)	Structured packing specific area (m <sup>2</sup> /m <sup>3</sup> )	Drip point density (points/m <sup>2</sup> )
25	500	120
38	350	80
51	250	60
76	170	40
89	125	30

- The distributor should be of the “drip point” type with an even pattern (square or triangular) of drip points having a density as above. Troughs with weirs over which the liquid is projected sideways to fall on to the packing are not considered acceptable for new designs.
- The drip point may be fed through holes or by notched weirs, see figure 7. The minimum width of a notch is 4 mm and the minimum hole diameter is 8 mm. Notches or holes smaller than 10 mm should not be at floor level where solids can accumulate and cause blockage.
- The minimum liquid head above hole or notched weirs is 30 mm at minimum load. At design load the liquid head should normally be greater than 100 mm. The distributor should not overflow before 110% of column flood is reached.
- The maximum liquid velocity in liquid distribution channels should be 0.25 m/s.
- The vapour riser area should be at least 25% of the column cross-sectional area.
- There must be free communication of the liquid between the various gutters and troughs, to prevent gross maldistribution if part of the predistribution system should block or fail.
- Troughs, gutters and pans must be fixed to the column wall and not simply rest on the packing.

## 5.2 The Packing Support

The support must be strong enough to bear the weight of the packing, a liquid hold up of 50% and any other internals such as a hold down grid that may be resting on the packing.

For random packings a slotted multi-beam or ripple type of support tray is usually applied, see figure 6. For structured packings a self supporting grid (100 - 150 mm high) resting on a ring at the wall is preferred. For both types the minimum free area is 95%, to prevent premature flooding at the support. If required support beams should be of the open structured lattice type to facilitate distribution of vapour.

### **5.3 Hold-down Grids**

A hold-down grid should be installed on top of beds of random packing, to hold the packing in place in the event of a column upset. The grid should not disturb the liquid distribution and must have an open area of at least 95%.

For structured packings a hold-down grid may be needed, depending on the manufacturer's advice, if one is used it will usually have a height of about 50 mm.

### **5.4 Redistribution Trays**

When more than 20 stages are required the bed should be split into sections with redistribution trays in between. Such an internal takes about 2 m of column height. It consists of a liquid collection tray feeding a gravity distributor.

## List of Symbols and Units

Symbol		Units
F	Foam factor	-
HETP	Height equivalent to a theoretical stage	m
N <sub>Rec</sub>	Recommended drip point density	1/m <sup>2</sup>
NTS	Number of theoretical stages	-
ΔP	Pressure drop	Pa
γ	Density correction factor	-
δ	Viscosity correction factor	-
ρ <sub>v</sub>	Vapour density	kg/m <sup>3</sup>
ρ <sub>l</sub>	Liquid density	kg/m <sup>3</sup>
λ	Vapour load factor	m/s
λ <sub>f</sub>	Vapour load factor at flooding	m/s
λ <sub>max</sub>	Maximum vapour load factor	m/s
ϕ	Flow parameter	-
μ	Dynamic liquid viscosity	Pa.s

**Table 1**

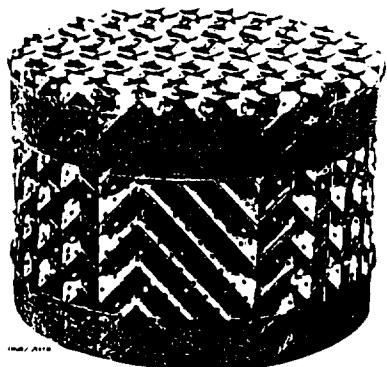
HETP for packings at stripping factors close to unity.

<b>Structured Packings</b>	<b>HETP m</b>
Sulzer BX and Kerapak	0.25
Mellapak 350Y	0.35
Montzpak B1-300	0.40
Mellapak 250Y	0.45
Montzpak B1-250	0.50
Mellapak 250X	0.60
Mellapak 125Y and Montzpak B1-100	0.90
Mellapak 125X	1.20

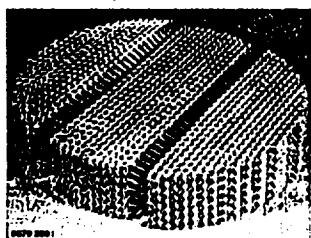
  

<b>Random Packings</b>		
Nominal size	25 mm	0.50
Nominal size	38 mm	0.65
Nominal size	50 mm	0.80
Nominal size	90 mm	1.20

FIGURE 1

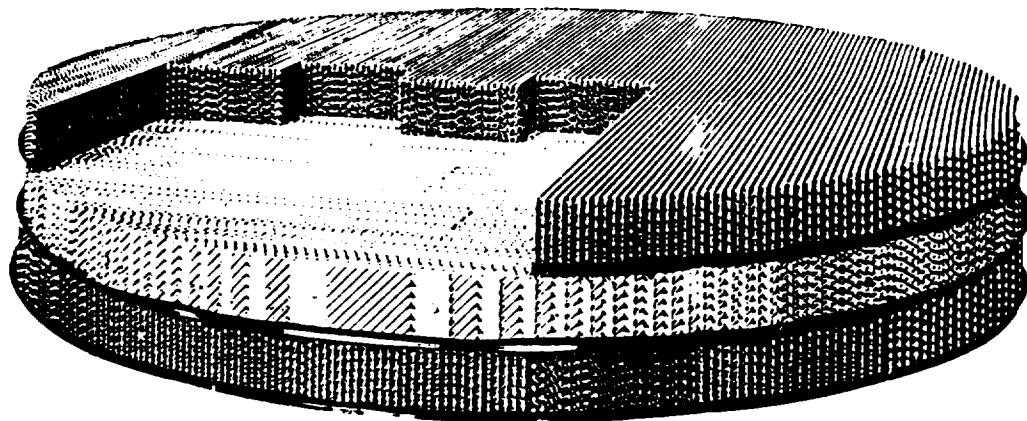
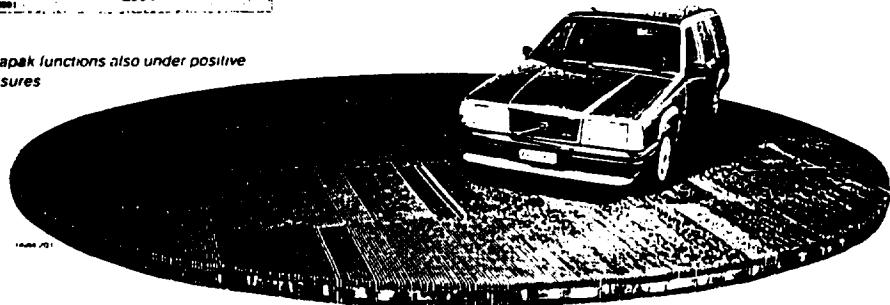


Mellapak in segmented form



MELLAPAK

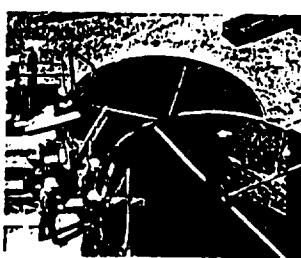
Mellapak functions also under positive pressures



MONTZ-PAK

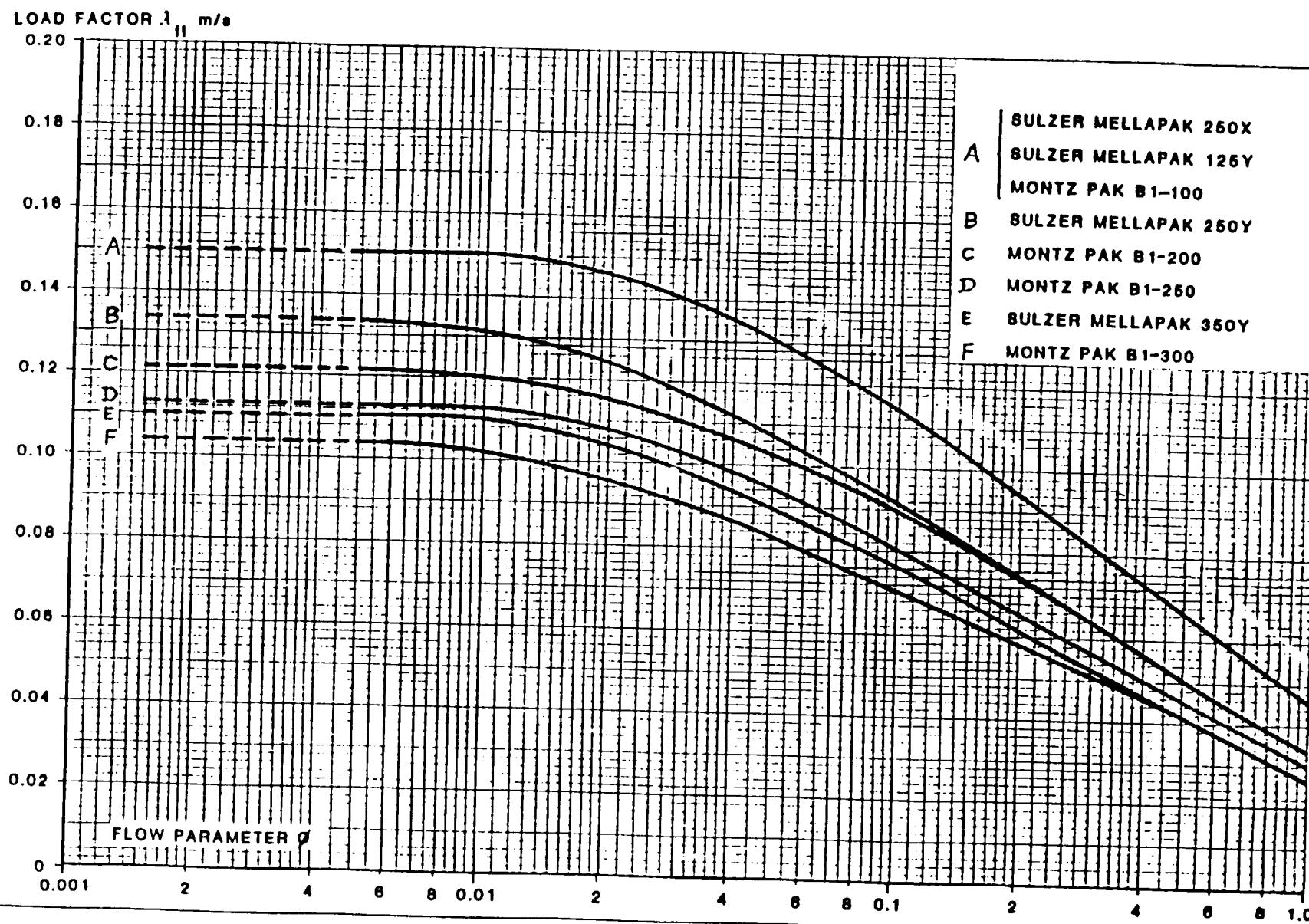


PACKING INSTALLATION  
IN COLUMNS



CAPACITY OF STRUCTURED PACKINGS

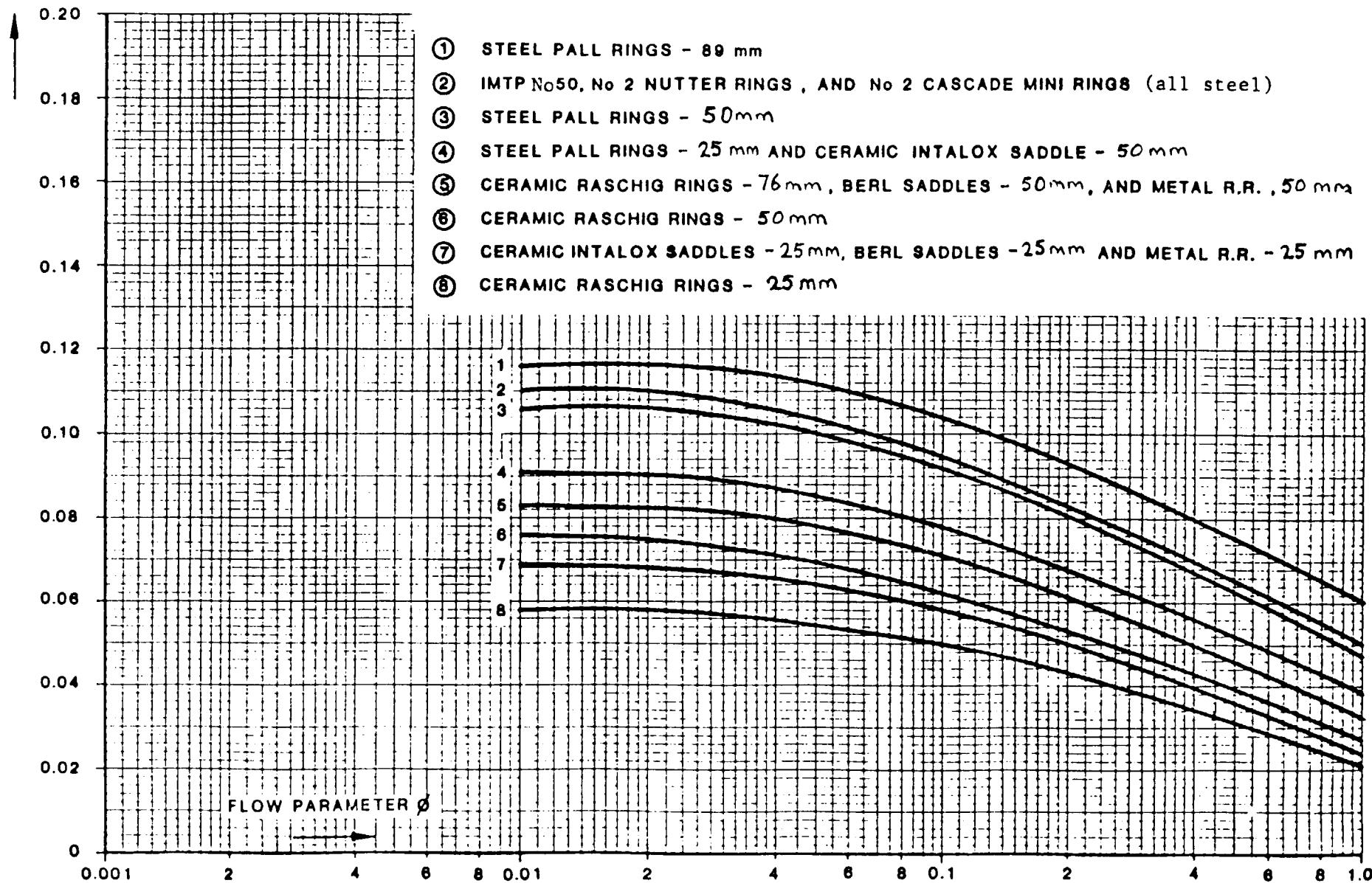
FIGURE 2

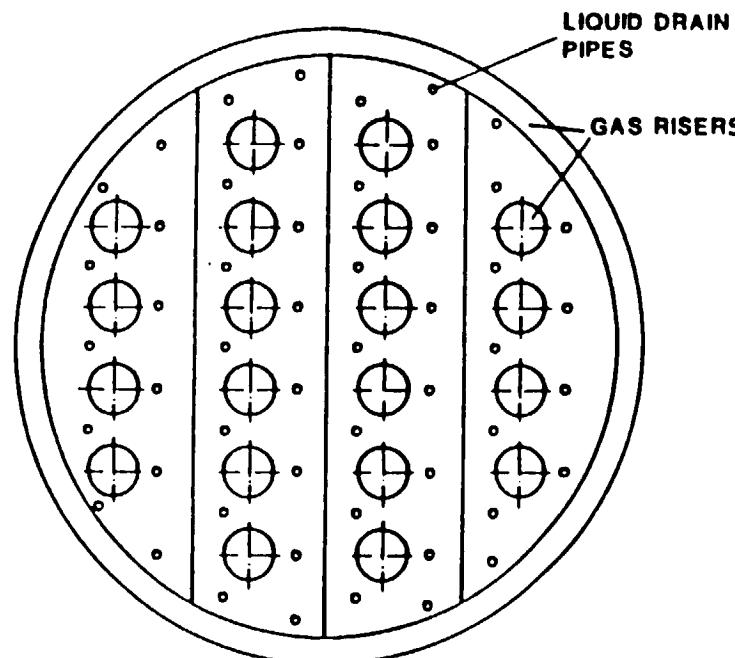


CAPACITY OF RANDOM PACKINGS

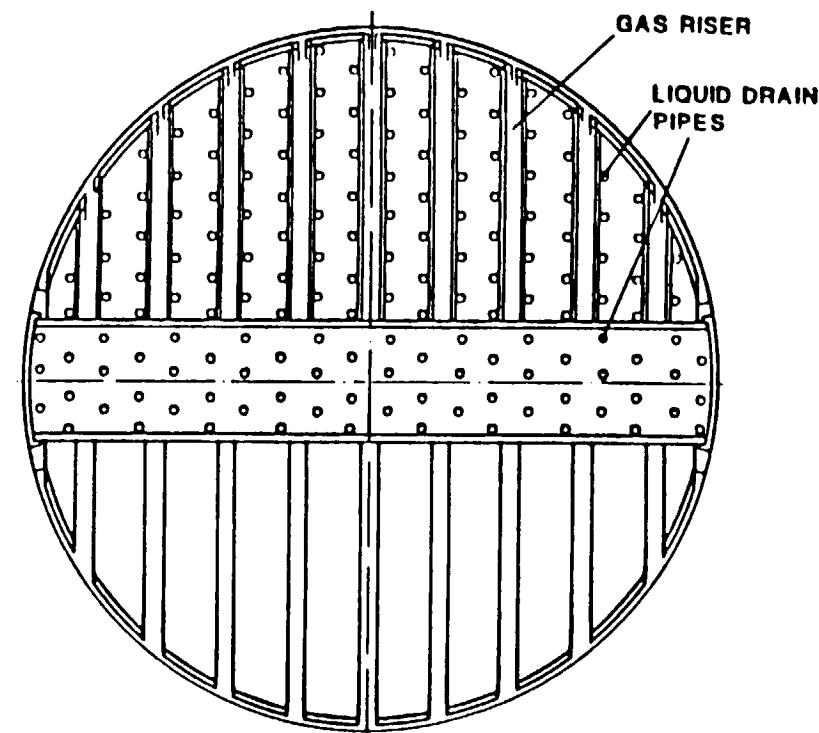
FIGURE 3

LOAD FACTOR  $\lambda_{fl}$  m/s



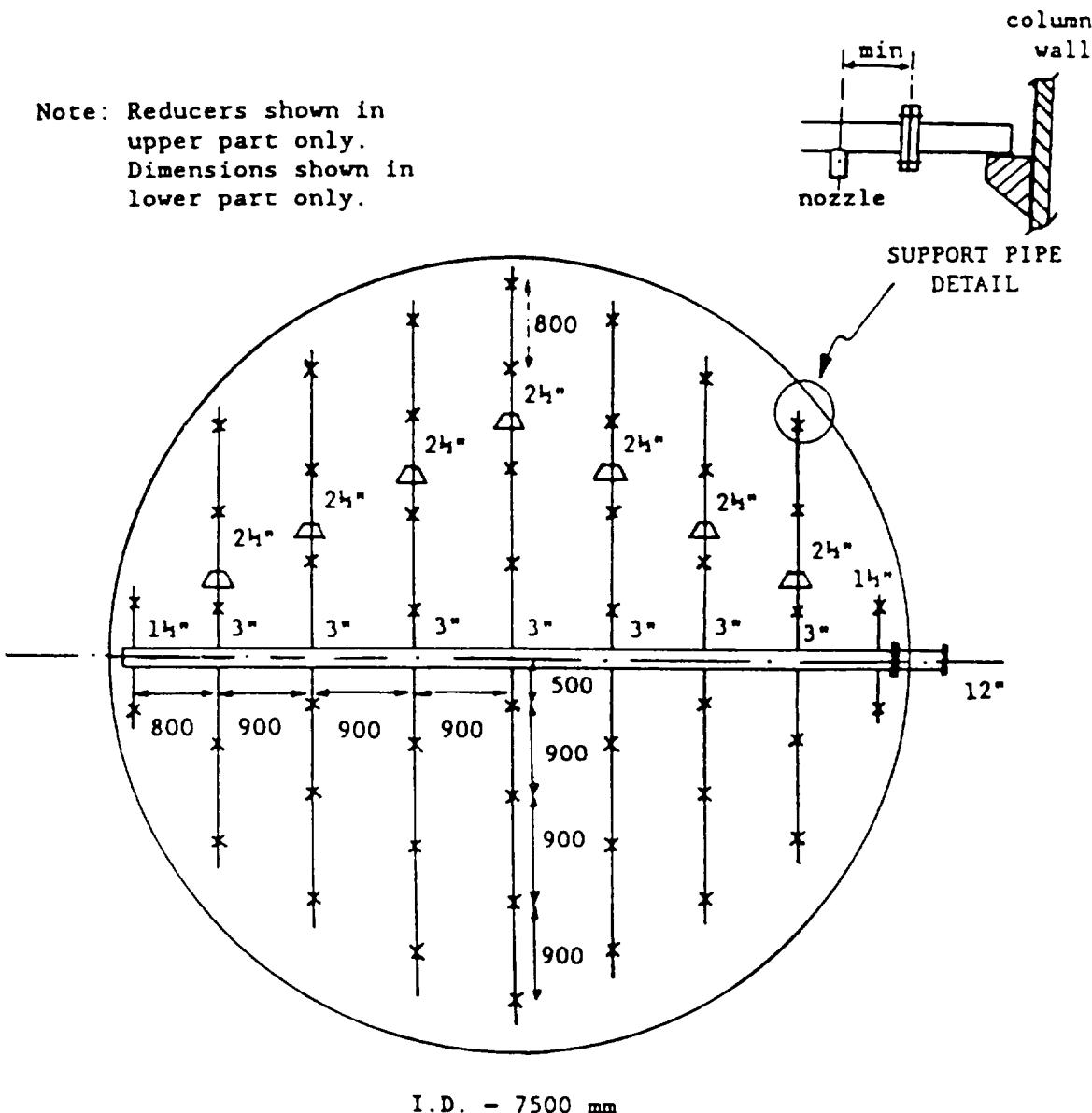


TYPICAL LAY-OUT FOR COLUMN DIAMETERS  
BELOW 1.5 m  
(LIQUID FED TO CENTRE OF TRAY)



TYPICAL DISTRIBUTOR FOR COLUMN DIAMETERS  
ABOVE 1.5 m  
(LIQUID FED TO MAIN CHANNEL VIA SPIDER.)

FIGURE 5 TYPICAL SPIDER DISTRIBUTOR  
WITH SPRAY NOZZLES



#### SPRAY NOZZLES

48 Spray Nozzles

Type: Lechler SZ 5 461.248

120 degrees spray angle

Nozzles spraying downwards

Nozzles are axial spraying

#### SPIDER

Main Header 12"

Side Branches 3" upstream 3 & 4 nozzles

2 1/4" upstream 2 & 1 nozzles

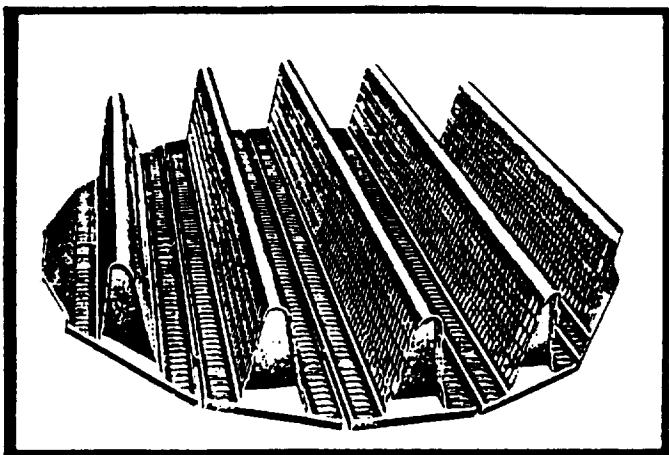
1 1/4" upstream 1 nozzle only

#### Note

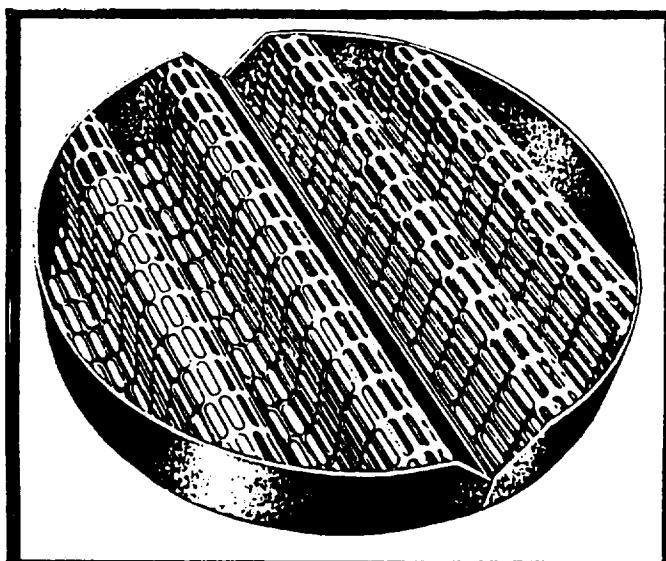
For tangential nozzles, nozzle position should be specified, side branch position suggested.

Nozzle pitch is variable at wall to achieve optimum wetting.

FIGURE 6



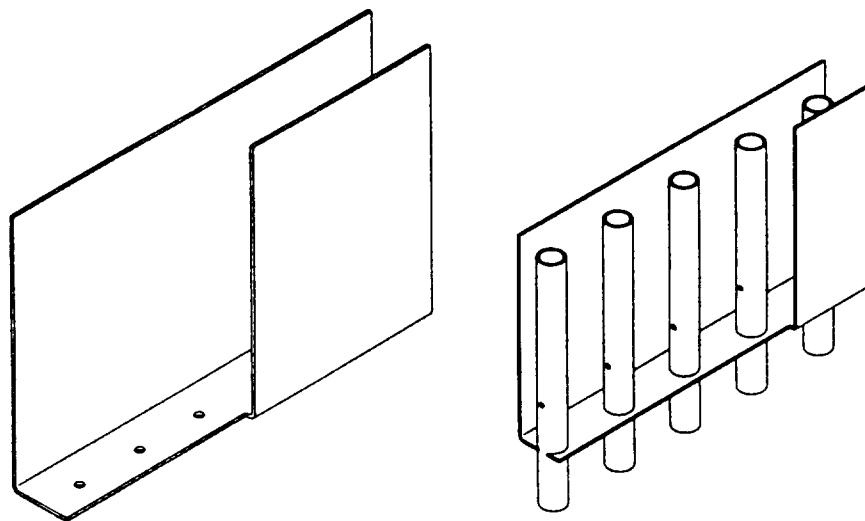
D > 1.2 m  
NORTON MODEL  
804 "MULTIBEAM"  
OR EQUIVALENT  
(height ~ 305 mm)



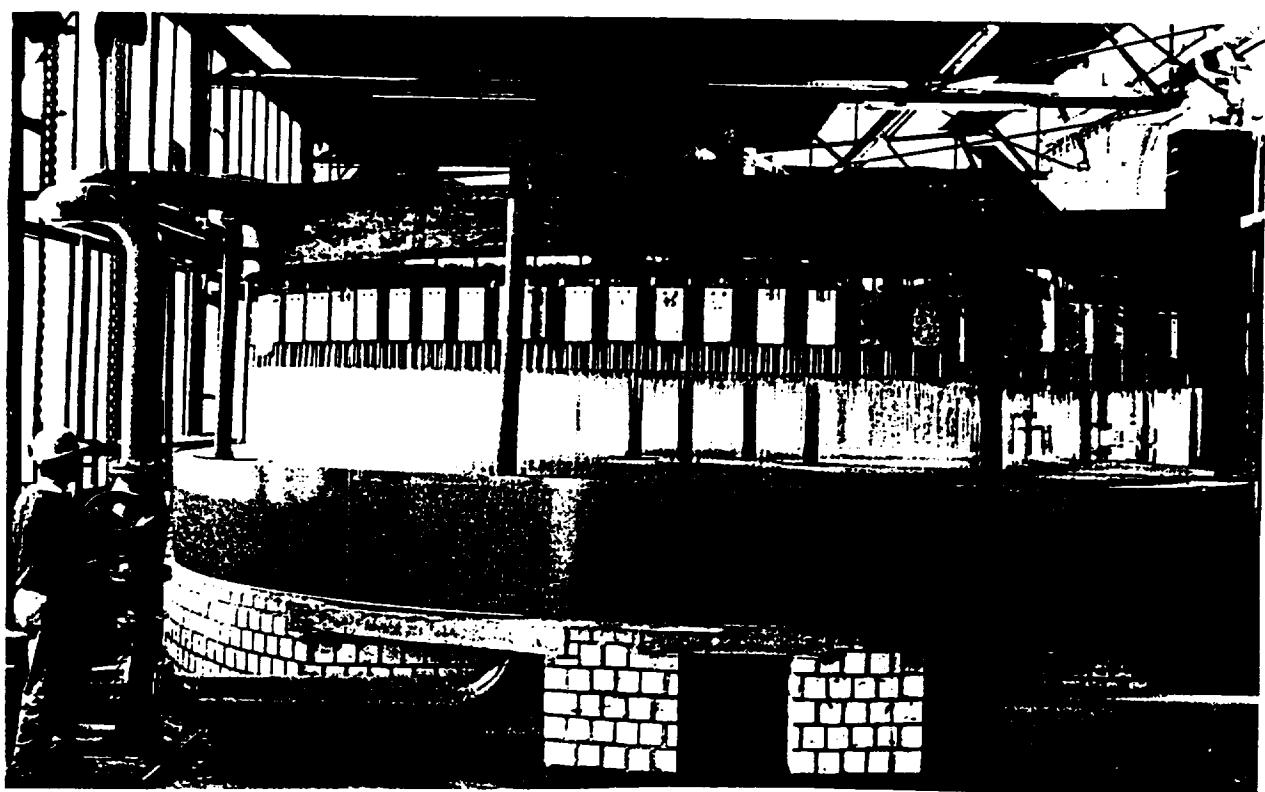
D < 1.2 m  
NORTON MODEL  
818 METAL  
OR EQUIVALENT  
(height = 115 mm)

SUPPORTS FOR RANDOM PACKINGS

FIGURE 7



Discharge systems for liquid distributors.  
Shell systems may have double rows of holes  
or pipes. The pipes may have notches rather  
than holes.



A distributor undergoing test in the manufacturers workshops

## **Appendix**

### **Review of available packing**

In this appendix we give some comments on the more commonly used types of packing.

#### **1 Random Packings**

##### **Pall Rings**

These may be regarded as the industry standard random packing, available from many different manufacturers, and in a variety of materials, plastic and metal. Raschig makes ceramic Pall rings. Although made in many different sizes, the 51 mm ring will often be found to give a good balance between capacity and efficiency.

##### **IMTP**

Intalox Metal Tower Packing is manufactured by Norton. Having the shape of a rather open half-ring, IMTP No. 50 has about 70% of the pressure drop of 51 mm Pall rings at the same vapour rate, and a usable capacity some 5% higher. Efficiency is similar to Pall rings of similar size.

##### **Nutter Rings**

These rings, manufactured by Nutter, are similar in shape and performance to IMTP. Nutter ring No. 2 is equivalent to IMTP No. 50.

##### **Cascade Mini Rings**

A short ring, marketed by Glitsch. The performance of No. 2 Mini rings is similar to that of IMTP No. 50.

##### **Hy Pack**

Manufactured by Norton, No. 2 Hy Pack has a size of 60 mm, and therefore slightly higher capacity and lower efficiency than 50 mm Pall rings, with which it often compared. V-Pak of VFF and K-pack of Koch are identical to Hy Pack.

##### **Raschig Rings**

The classic random packing, having one advantage: it spreads liquid radically faster than the modern packings described above. Its capacity is much lower than that of the equivalently sized Pall ring. Available in ceramic and a wide variety of metal and plastics.

##### **Intalox Saddles and Berl Saddles**

Old packing types illustrated in many text books, they have a significantly lower capacity than that of a Pall ring of the same size. Ceramic saddles are sometime used in corrosive services.

## **2 Structured Packing**

Nearly all structured packings are made from corrugated sheets of material, placed in an ordered fashion so as to offer a large surface area for liquid films, and a high open area for vapour flow. Surface of the sheets is generally roughened, and may be perforated with holes or slits. A bed of structured packing is made up of blocks of a convenient size to handle, the elements being cut carefully to shape by the manufacturer so that they fit perfectly into the round column. The blocks are installed in layers, so that the specified bed height must be an integral number of layer heights. (In exceptional cases layers of half the standard height may be included.)

### **Mellapak**

The most widely used structured packing, made by Sulzer. Mellapak 250Y generally offers a good combination of capacity and efficiency. 350Y offers higher efficiency and 125Y more capacity. The X series (125X, 250X and 350X have channels at 30 degrees to the vertical instead of 45) gives somewhat more capacity at the expense of efficiency. 125 and 250 (X and Y) are available in polypropylene, PFA and PVDF. Layer height 210 mm for Y types, and 225 mm for X types. In Japan Mellapak is made and sold, under licence, by Sumitomo Heavy Industries.

### **BX**

Similar to Mellapak but made of wire gauze. A high efficiency packing from Sulzer. Layer height 171 mm.

### **Kerapak**

Ceramic structured packing from Sulzer, having an efficiency similar to BX. Expensive. Suitable for non-aqueous corrosive applications. Layer height 170 mm.

### **Flexipac**

Mellapak made under licence in the USA by Koch. One difference is the layer height, 267 mm.

### **Montz-Pak**

Manufactured by Julius Montz, and sold in the USA under licence by Nutter. The Montz-Pak B1 series has a somewhat lower capacity for a given efficiency than the Mellapak Y series. Layer height of B1 is 200 mm.

Other packings, superficially similar to Mellapak in appearance (Ralu-Pak or Raschig, Gempak of Glitsch and Norton's Intalox structured packing) can perform well in laboratory tests. However, experience has shown that good plant performance depends on details such as adequate sealing at the wall. The manufacturer's knowledge of such matters, his ability to design and manufacture suitable support plates, liquid distributors and other auxiliary internals, and his ability to give after-sales service should be considered when selecting the packing type.

# **PACKINGS**

**WHAT IS PACKING ??**

-- **FILLING MATERIAL FOR COLUMNS**

**WHAT IS THE OBJECTIVE OF PACKING ??**

-- **GOOD HEAT and MASS TRANSFER AT LOW  $\Delta$  delta P**

**HOW DO WE REACH THIS OBJECTIVE ??**

-- **BY CREATING LOTS OF INTERFACIAL AREA**

**HOW DOES PACKING COMPARE WITH TRAYS ??**

-- **NO GENERAL RULE BUT COMPARE ASPECTS ...**

## **PACKINGS / TRAYS COMPARISON**

**NO OVERALL CRITERIA, BUT THE FOLLOWING IS USEFUL :**

**\*\* = SUPERIOR**

<b>ASPECT</b>	<b>PACKINGS</b>	<b>TRAYS</b>
ENTRAINMENT	**	
CAPACITY AT LOW L LOAD	**	
CAPACITY AT HIGH L LOAD		**
PRESSURE DROP	**	
COLUMN DIAMETERS < 0.8m	**	
CONSTRUCTION	**	
FOULING SERVICE		**
Ht FOR AUX INTERNALS		**
COLUMN WALL INSPECTION		**
MALDISTRIBUTION		**
COSTS		\$\$

**FOR NEW SYSTEMS : DECISION BY OVERALL ECONOMICS**

19.10.90	PACKINGS	
2.1	DESIGN	

## DESIGN ASPECTS OF PACKINGS:

History

Random Packings:

types, main features

Structured Packings:

types, features

main steps in design

pitfalls

Auxiliary internals

19.10.90	PACKINGS	
2.2	DESIGN	

## HISTORY OF PACKINGS

1820: Random Packings – 1" glass balls

1933: Packed column maldistribution

1958: Metal Pall Rings (BASF)

1965: Gauze structured packing (Sulzer)

1979: Sheet structured packing (Sulzer)

19.10.90

2.4

## PACKINGS

## DESIGN

## WHEN TO USE RANDOM PACKINGS?

1. High liquid load
2. Sometimes: fouling service.

In most cases where pressure drop plays a role, it is most economic to choose structured packings

## DESIGN GUIDELINES:

Ref. course manual, design manual.

19.10.90

2.5

## PACKINGS

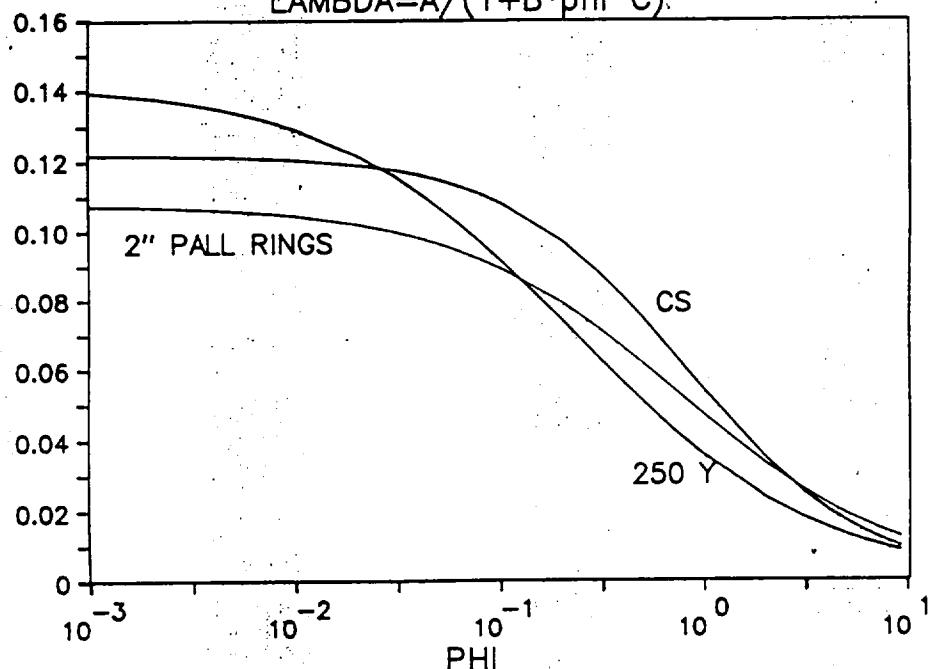
## DESIGN

## CAPACITY CURVES, RANDOM PACKINGS vs OTHERS

$$\Lambda = A / (1 + B \cdot \phi^C)$$

LOAD FACTOR

flood



19.10.90	PACKINGS	
2.6	DESIGN	

### STRUCTURED PACKINGS: TYPES

\* Gauze vs sheet

\* Sheet structure:

Sulzer Mellapak (= Koch Flexipak)

Sulzer Mellapak W

Montz B1

Montz BS

Norton Intalox

Glitsch Gempak

Raschig Ralupak

etc.

\* Specific surface area ( $m^2/m^3$ )

\* Angle of main corrugation: 45 or 60 deg.

Market standard: Sulzer Mellapak 250 Y

Aim: clarify when which alternatives can be used.

19.10.90	PACKINGS	
2.7	DESIGN	

### DESIGN STEPS:

1. Selection

2.  $\lambda$  -flood,  $\lambda$  -max,  $\lambda$  -design, (pressure drop)

3. Packing height

4. Distributors etc.

19.10.90	PACKINGS	
2.8	DESIGN	

## DESIGN STEPS

### 1. SELECTION

\* Aim for minimum volume of packing to achieve required fractionation.

\* Rules of thumb:

- Start with Mellapak 250 Y
- If sharp fractionation required: try 350 Y
- If D is limiting: try 250 X
- If H is limiting: try 350 Y
- If fouling service: 125 Y
- If low liquid wetting: 125 Y

\* Do not specify "OR ALTERNATIVE" but choose alternatives yourself: with care.

19.10.90	PACKINGS	
2.9a	DESIGN	

## DESIGN STEPS

### 2. LOAD FACTOR

Conclusion:

$$\varphi \rightarrow \lambda_{fl} \rightarrow \lambda_{max} \rightarrow \lambda_{des}$$

$$D_{col} = \sqrt{\frac{4 \cdot Q^*}{\pi \cdot \lambda_{des}}}$$

19.10.90 2.10	PACKINGS DESIGN	
------------------	--------------------	--

### DESIGN STEPS

#### 3. PACKING HEIGHT

$$H = H_e + NTS \cdot HETP_o / fw \cdot fm \cdot fr$$

$H_e$  : end effects - packing type, distribution

$HETP_o$  : efficiency

$fw$  : low liquid load

$fm$  : Marangoni effect

$fr$  : maldistribution - NTS, distributor

19.10.90 2.9	PACKINGS DESIGN	
-----------------	--------------------	--

### DESIGN STEPS

#### 2. LOAD FACTOR

Flooding (see curves): pressure drop ↑  
separation ↓

$$\lambda_{max} = \lambda_{fl} * \gamma * \delta * \sqrt{F}$$

$$\gamma = (\rho_l - \rho_g) / 470 \quad \text{for } \rho_l - \rho_g < 470 \text{ kg/m}^3$$

$$\delta = (2 / visco)^{0.05} \quad \text{for visco} > 2 \text{ cP}$$

F = foam factor

$$\lambda_{des} = \lambda_{max} * (P_{fl}/100)$$

NB! in vacuum columns note that:  
between  $\lambda_{des}$  and  $\lambda_{max}$  pressure drop increases,  
consequences for yield and vacuum set!

# PRESSURE DROP

$$\frac{\Delta P}{H} = 0.54 (\rho_l - \rho_v) \left( \frac{\lambda}{\lambda_{fl}} \right)^2 \left( 1 + \left\{ \frac{\lambda}{\lambda_{fl}} \right\}^3 \right)$$

in Pa/m

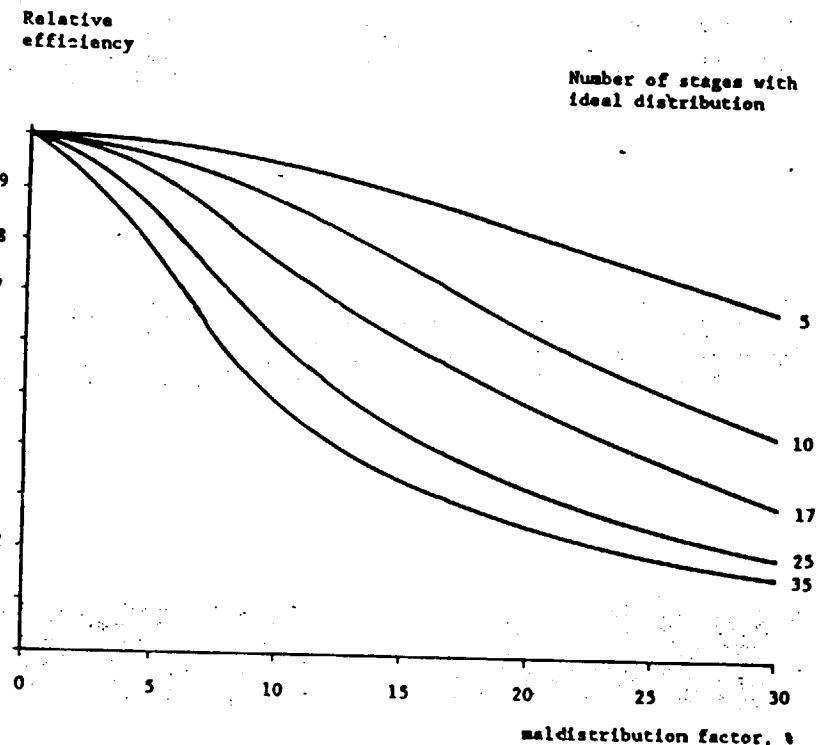
$$\rightarrow \text{mm Liquid : } \frac{\Delta P (\text{Pa}) \times 1000}{\rho_l \times g \cdot 9.8}$$

$$\rightarrow \text{mm Hg : } \frac{\Delta P (\text{Pa}) \times 1000}{13600 \times g \cdot 9.8}$$

-86-

CONFIDENTIAL

Figure 21 Effect of Maldistribution on Packing Performance



## Note

The highest possible distribution quality for liquid distributors which can be obtained in practice gives a maldistribution factor of 4-5 %. This can be achieved with a gravity type distributor, with narrow mechanical tolerances, careful design and hydraulic testing in the manufacturers workshop.

For spray nozzle distributors, 20-30 % is the lowest maldistribution factor which can be achieved.

19.10.90 2.11	PACKINGS DESIGN	
------------------	--------------------	--

## DESIGN STEPS

### 4. DISTRIBUTORS

#### a. Vapour distribution

- avoid local flooding!
- open pipe or schoepentoeter is NOT sufficient
- draw off tray is good distributor, but
  - provide some pressure drop (say 1 mbar)
  - avoid jetting through vapour risers
  - new type of double decker tray
- manufacturers are starting to use min. no. of risers per m<sup>2</sup>

19.10.90 2.12	PACKINGS DESIGN	
------------------	--------------------	--

## DESIGN STEPS

### 4. DISTRIBUTORS

#### b. Liquid distribution

##### SPRAYS:

- poor distribution, so for low NTS only
- low pressure drop (vacuum columns)
- risk of plugging: dramatic loss in separation!
- risk of entrainment

##### GRAVITY DISTRIBUTORS

- frequent cause of mal-performance!
- not all manufacturers fulfill requirements
- good performance obtainable if well designed and tested

## **MINIMUM WETTING RATE**

**\* EFFICIENCY**

**\* FOULING**

**Typical minimum wetting rate:**

**Random Packing = 0.30 mm/s**

**Structured Packing = 0.10 mm/s**

19.10.90 2.14	PACKINGS DESIGN	
<b>DESIGN STEPS</b>		
<b>4. DISTRIBUTORS</b>		
<p><b>d. Redistributions</b></p> <p>Needed if NTS &gt; 20 (typ.)</p> <p>Consult the "expert"</p>		

19.10.90 2.13	PACKINGS DESIGN	
<b>DESIGN STEPS</b>		
<b>4. DISTRIBUTORS</b>		
<p><b>c. Feed inlets</b></p> <p>Combination of:</p> <ul style="list-style-type: none"> <li>- inlet device (e.g. schoepentoeter)</li> <li>- vapour distributor</li> <li>- liquid distributor</li> </ul>		

## ● CAPACITY

- Loadfactor

$$\text{Lambda} = Qv \cdot \text{Sqrt}(Rho_v / [Rho_l - Rho_v]) / A_c$$

- Flowparameter

$$\Phi = Q_l / Q_v \cdot \text{Sqrt}(Rho_l / Rho_v)$$

## ● EFFICIENCY

Tray efficiency

Height Equivalent to Theoretical Plate

- Design rules for auxiliary tray-column internals

Memorandum CMFE No. C55/81

by J. van der Harst

Date of issue: 11 February 1982

- Gas/liquid separators

Type selection and design rules

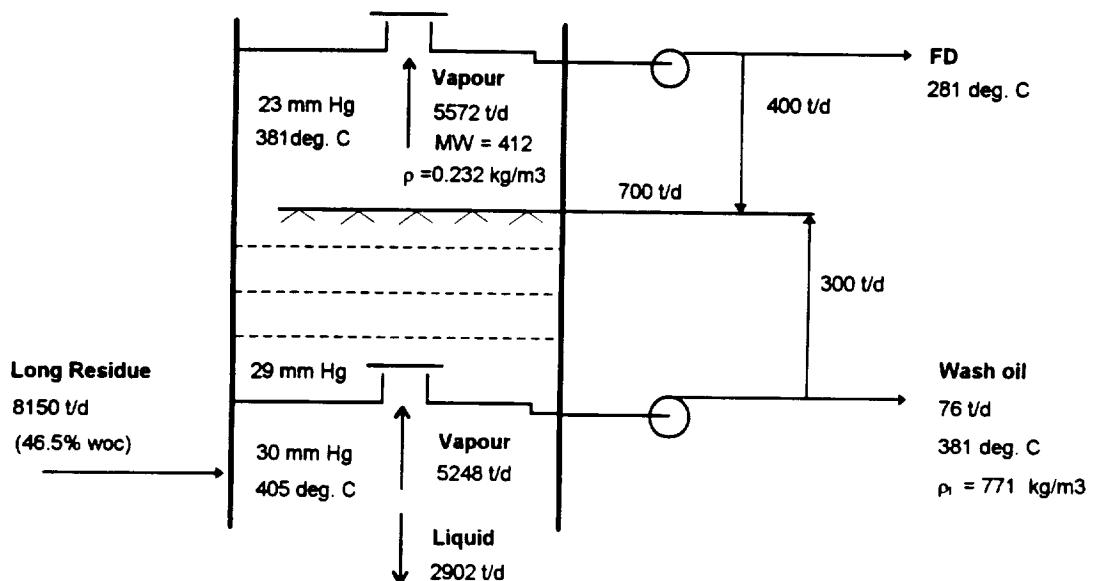
DEP 31.22.05.11-Gen.

Date of issue: February 1991

## Application of Structured Packing

A High Vacuum Unit produces flashed distillate (FD) as feedstock to the cat. Cracker unit. The 8.5 m diameter column has trays in the wash oil section above the flash zone.

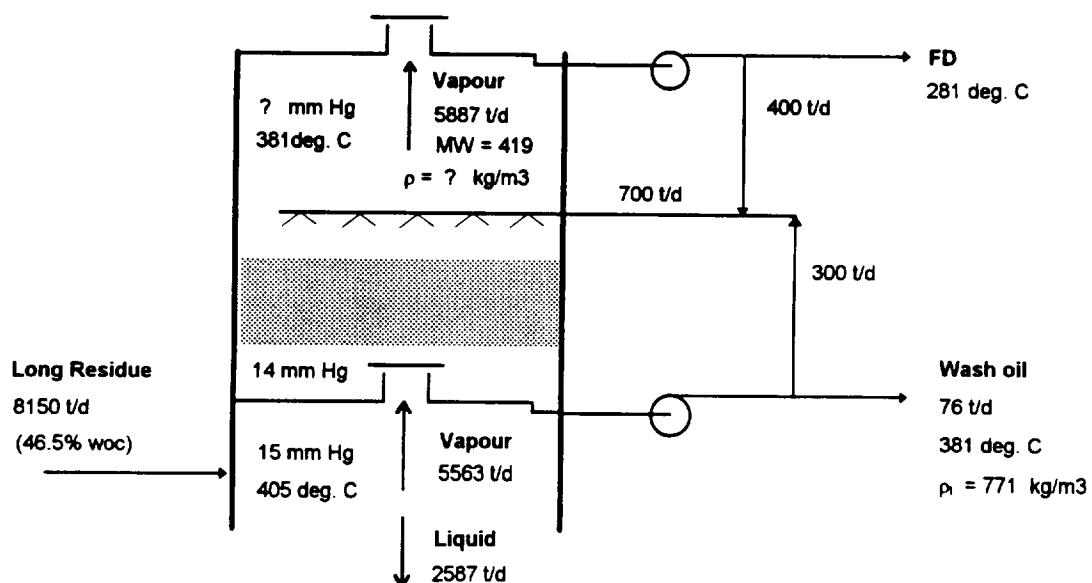
During a recent test run the Long Residue feed intake was 8150 t/d and the WD production 5172 t/d. The flash zone conditions were 30 mm Hg and 405°C and the total pressure drop measured over the wash oil section was 7 mm Hg. Other test run data of the wash oil section is shown below.



There is a demand for more FD feedstock. A simulation study shows that more FD can be produced from the same Long Residue intake. With a flash zone pressure of 15 mm Hg the FD will increase to 5487 t/d.

The mol weight of the distillate vapour flow from the wash oil section increases to MW = 419. The other simulation results are shown below.

Note that the pressure drop over the wash oil draw-off tray remains 1 mm Hg.



## **Exercise**

To reduce the flash zone pressure it has been decided to study the possibility of replacing the (relatively high pressure drop) internals currently installed in the wash oil section with more modern, lower pressure drop ones. As part of this study the replacement of the trays in the de-entrainment section (with much lower pressure drop) packing needs to be considered. For this study a number of questions have to be addressed:

1. Knowing that the wash oil section represents only one theoretical stage, which type of packing would you suggest and how much?  
Although in a full study the choice may be between several different packings, for the purposes of this exercise we will limit our choice to consider the structured packings Sulzer Mellapak 125Y and 250Y only.
2. Assuming (to be checked later) that under the new conditions the packing is loaded to about 80% of flood, what will be the pressure drop over the new packed bed and how does it compare with the existing pressure drop.
3. How does the vapour load factor at the highest loaded part of the wash oil section change as a function of pressure, taking into account the lower vapour density in that part and the extra 1.8% WOC distillate?
4. Considering the maximum practical allowable capacity of the packing can 15 mm Hg in the flash zone be achieved and is the assumption about the packing loading OK?
5. What other modifications to the HVU may have to be made to achieve the higher distillate yield objective?



**10**

## **Fractionator operation**

## Exercise

### Fractionator operation

Although a Fractionator is basically a combination of a number of two-cut splitter columns the actual operation to control the product qualities can be rather complex.

The purpose of the exercise is to understand the steady state responses of a typical Fractionator column.

We will consider the operational changes shown on the attached worksheets. The first worksheet gives the changes on product specs. The second sheet is about the column operation.

For each case we will identify the effect on the main variables.

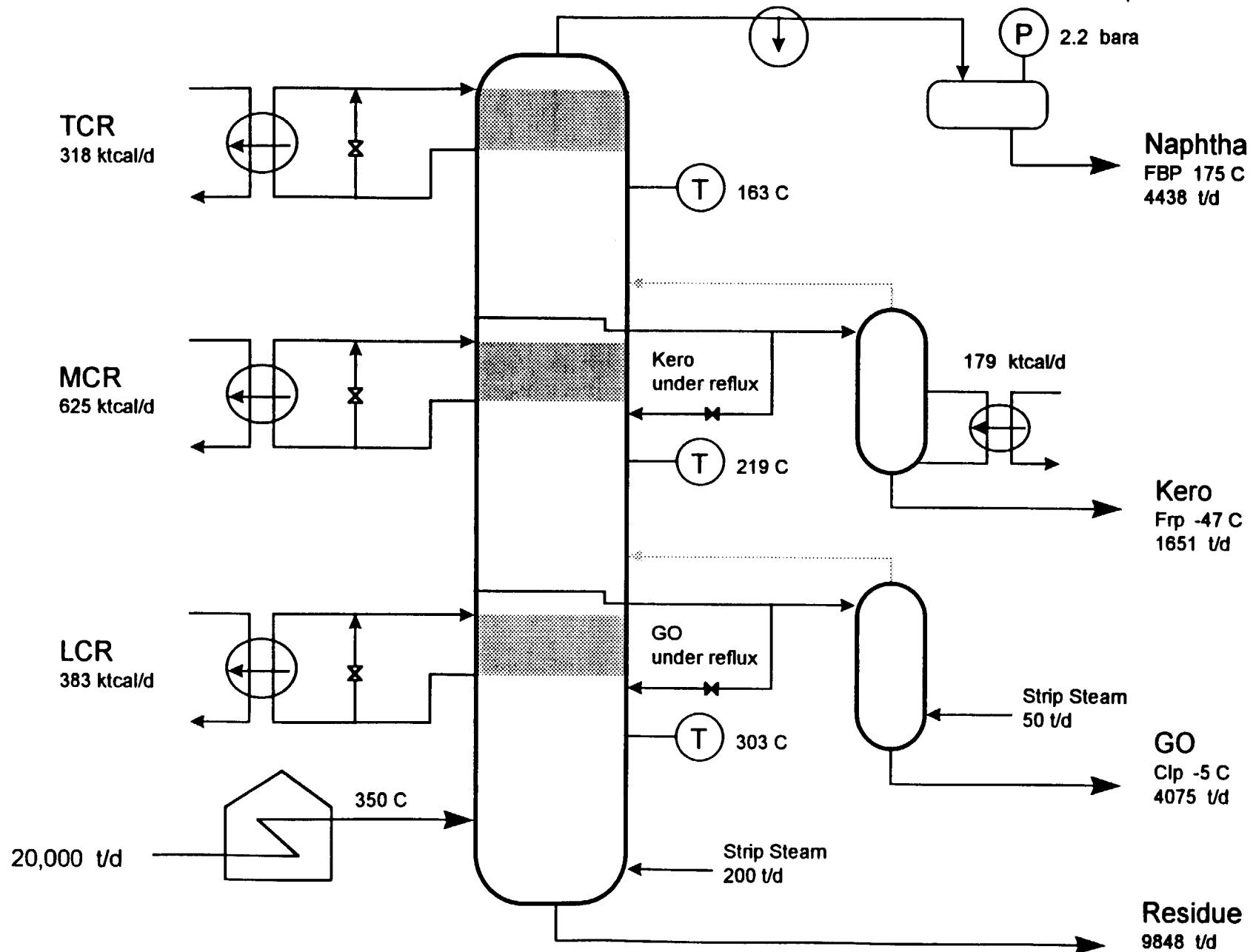
- Tray temperatures
- Circulating reflux duties
- Product yields and qualities

#### Exercise

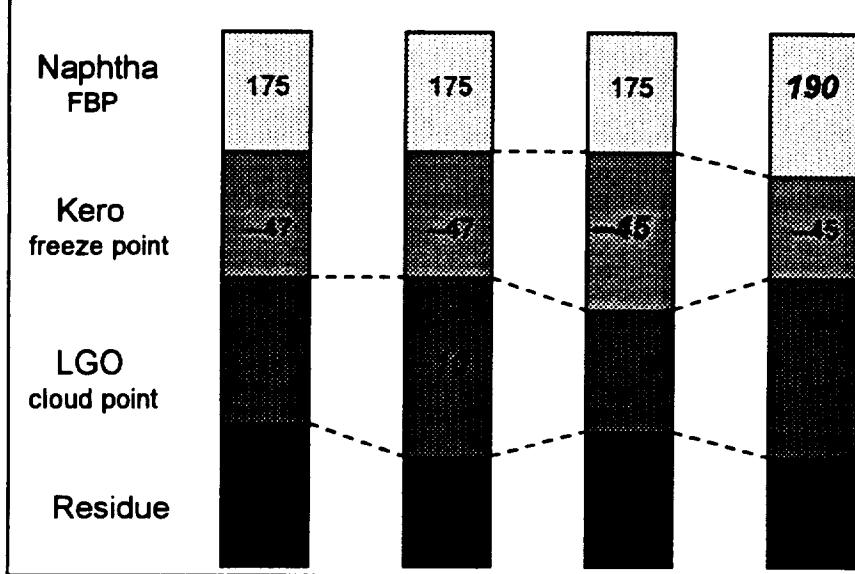
Consider the cases one by one with the following steps.

1. Indicate the responses of the main variables on your work sheet.  
It is sufficient to identify the direction of change.  
Jot a + on your worksheet when you think the variable increases or a - sign when it decreases.
2. Update and run the flowsheet model. Don't forget to reset the previous change !  
Generate an output report.  
Look for the results in the output listing under 'TABLE'.  
Note that the *Tray temperatures* and *ASTM's* are given as  $\Delta T$ .  
The *Duty's* and *Yields* are given in %.
3. Check your worksheet results against the TABLE.  
When you notice a difference, explain why your prediction was not correct.  
Don't take it for granted. It is vital to improve your basic understanding.

# Fractionator (base case)



## Concertina effect



Cloud point is the cold flow property for gasoil fractions. In fact it is the temperature at which the crystallisation of the liquid starts.

For Kero the cold flow property is called Freeze point.

The Cloud point and Freeze point depend on both the front end and the heavy tail of the fraction; more light ends in the fraction will decrease the cloud point temperature while a longer tail will increase the temperature.

## Fractionator operation exercise

	Base case	GO cloud point	Kero freeze point	Naphtha FBP	Kero flash point	Combined
<i>Temperatures</i>						
Top control tray	163					
Middle control tray	219					
Lower control tray	303					
Furnace outlet	350	350	350	350	350	350
Column (bara)	2.2	2.2	2.2	2.2	2.2	2.2
Kero under reflux	closed	closed	closed	closed	closed	closed
GO under reflux	closed	closed	closed	closed	closed	closed
<i>Strippers</i>						
Kero reboiler (kcal/d)	179	179	179	179		179
GO strip steam (t/d)	50	50	50	50	50	50
Res strip steam (t/d)	200	200	200	200	200	200
<i>Circ. reflux (kcal/d)</i>						
TCR	318					
MCR	625					
LCR	383					
<i>Product yields (t/d)</i>						
Naphtha minus	4438					
Kero	1651					
GO	4074					
Residue	9848					
<i>Product qualities</i>						
Naphtha FBP	175	175	175	190	175	190
Kero freeze point	-47	-47	-45	-47	-47	-45
flash point	54				48	
IBP	173					
FBP	239					
GO cloud point	-5	-10	-5	-5	-5	-10
IBP	194					
FBP	354					
Residue IBP	272					

## Fractionator operation exercise

	Base case	Kero under refl.	GO under refl.	GO strip steam	Furnace temp.	Residue strip steam
<i>Temperatures</i>						
Top control tray	163					
Middle control tray	219					
Lower control tray	303					
Furnace outlet	350	350	350	350	360	350
Column (bara)	2.2	2.2	2.2	2.2	2.2	2.2
Kero under reflux (t/d)	closed	1500	closed	closed	closed	closed
GO under reflux (t/d)	closed	closed	1500	closed	closed	closed
<i>Strippers</i>						
Kero reboiler (ktcal/d)	179	179	179	179	179	179
GO strip steam (t/d)	50	50	50	100	50	50
Res strip steam (t/d)	200	200	200	200	200	300
<i>Circ. reflux (ktcal/d)</i>						
TCR	318					
MCR	625					
LCR	383					
<i>Product yields (t/d)</i>						
Naphtha minus	4438					
Kero	1651					
GO	4074					
Residue	9848					
<i>Product qualities</i>						
Naphtha FBP	175	175	175	175	175	175
Kero freeze point	-47	-47	-47	-47	-47	-47
flash point	54					
IBP	173					
FBP	239					
GO cloud point	-5	-5	-5	-5	-5	-5
IBP	194					
FBP	354					
Residue IBP	272					



**11**

**Crude Distilling Unit**

## Crude oil

- **Hydrocarbons**
  - PIANO
  - asphaltenes
- **Organic compounds**
  - Sulphur, Nitrogen, Oxygen
  - Vanadium, Nickel
- **B S & W (Bottoms Sediment & Water)**
  - formation water from the production field
  - transportation ballast water
  - soluble salts: NaCl, CaCl<sub>2</sub>, MgCl<sub>2</sub>
  - sand
  - scale (FeS; pyroforic)

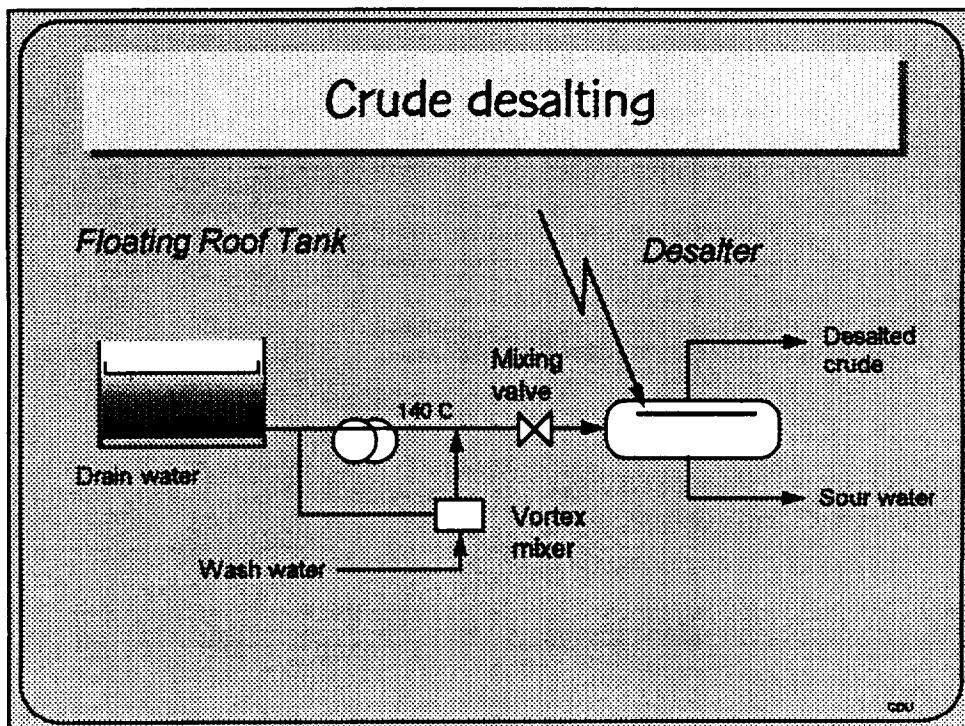
CDU

Hydrocarbons come in the form of Paraffin's, Iso-paraffins, Aromatics, Naphthenes and Olefins.

Crude oil also contains asphaltenes, very large molecules with a complex structure. They give the crude its deep brownish colour and are present in all black products.

Sulphur is the most important organic compound. In the form of mercaptan it smells.

Water in crude contains soluble chlorine salts which causes fouling and corrosion.



Crude parcels are stored in floating roof tanks where most of the water is drained off.

To lower the salt concentration the crude is mixed with wash water. The mixing with crude should be sufficient to wash out the salts but should not be perfect to avoid forming of an water crude emulsion. A pressure drop of 0.5 bar over a globe valve is sufficient. An additional static mixer may be used for low crude intakes.

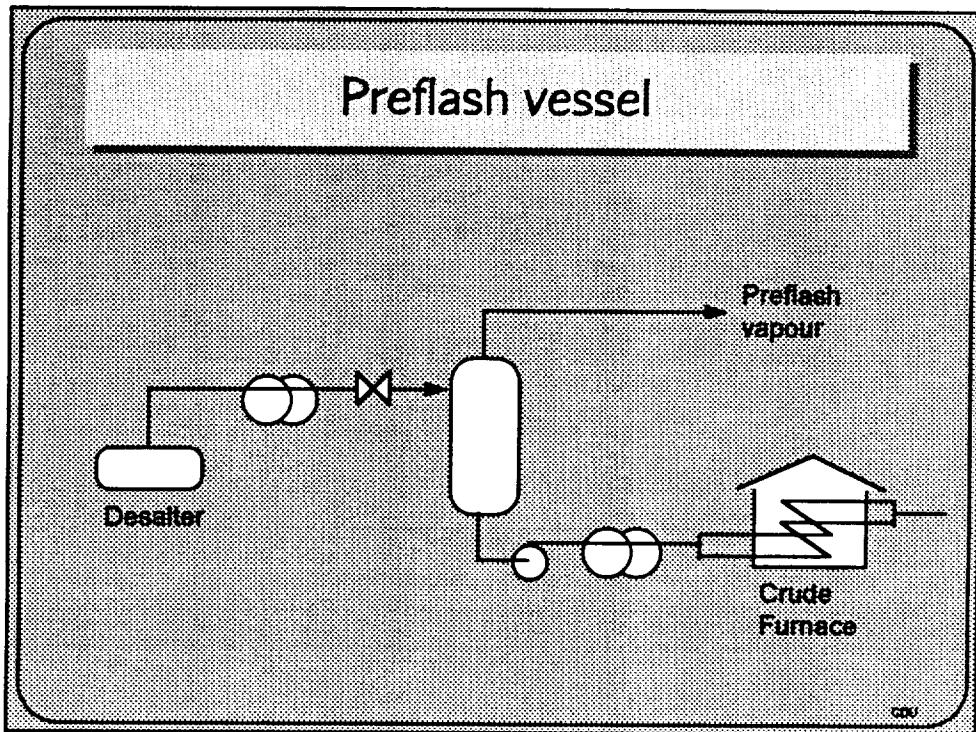
The wash water is premixed with a slipstream of crude. The crude is heated to about 140 C to lower the viscosity. A high pressure keeps the water in the liquid phase.

In the desalter the water settles out. To promote the settling out of the water (which is polar) an electrical field is generated with electrodes inside the desalter vessel.

Desalted crude still contains some salt in the dissolved water.

To convert the Calcium and Magnesium chlorides into more stable Sodium chloride some Caustic soda (NaOH) may be injected after the desalter.

## Preflash vessel



The main purpose of preflash vessel is to keep the heat exchangers and the crude furnace coils free of salt deposits. With crude preflashing all the water evaporates and the soluble salts will deposit in the preflash vessel.

The second purpose of the vessel is to flash off the light HC's. For equal flow distribution over the furnace coils the preflashed crude should be kept in the liquid phase. With pre flashed crude the pressure at the furnace inlet can be relatively low.

## Foaming pre-flash vessel

- **foaming tendency**
  - $H_{foam} = F(V_{visco}, V_{inlet})$
- **operational effects**
  - sudden loss of liquid level in vessel
  - carry over of black HC to column
- **foam suppression**
  - injection of an anti foam agent
  - *reduce pressure* preflash vessel

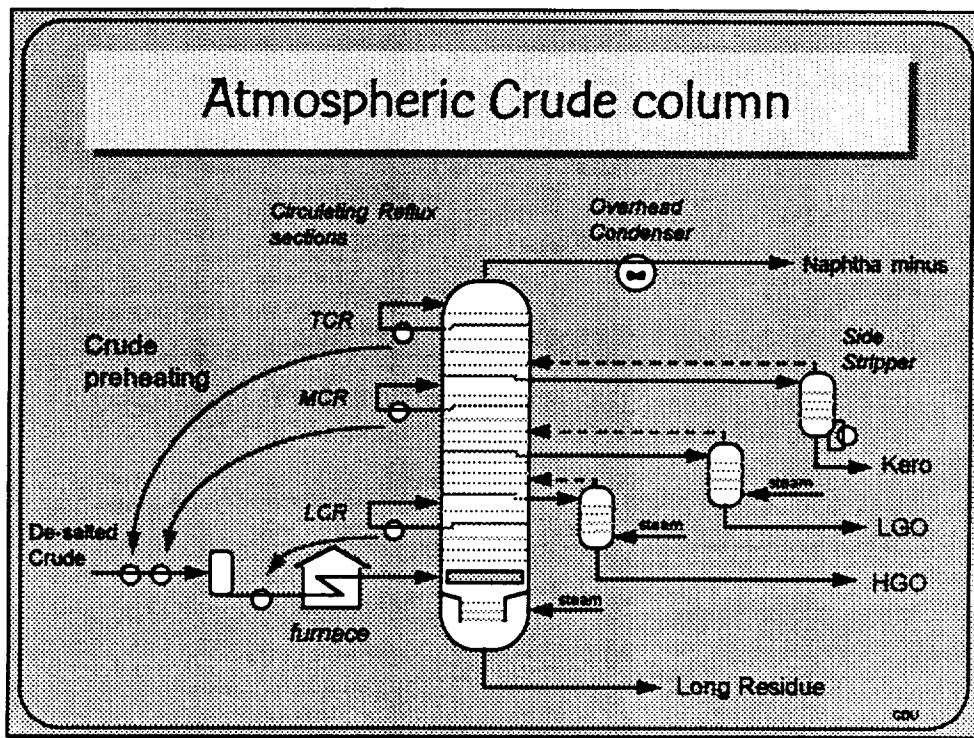
CDU

Crude may foam when it flashes in the preflash vessel. This can happen suddenly without any pre-warning. Foaming starts when the vapour does not separate from the liquid; the liquid level in the vessel is lost.

When foaming occurs the main problem is to keep a positive NPSH for the bottom pump.

An additional problem with foaming is the carry over of black residue material with the preflashed vapour.

Foaming should be suppressed immediately whenever it occurs. This can be done either by injecting an anti-foaming agent or by breaking the foam with a high velocity at the inlet of the vessel. The inlet velocity can be increased with a lower preflash vessel pressure; flashing off more vapour.



The crude is heated in the crude furnace to about 360 C, about the maximum temperature to avoid cracking.

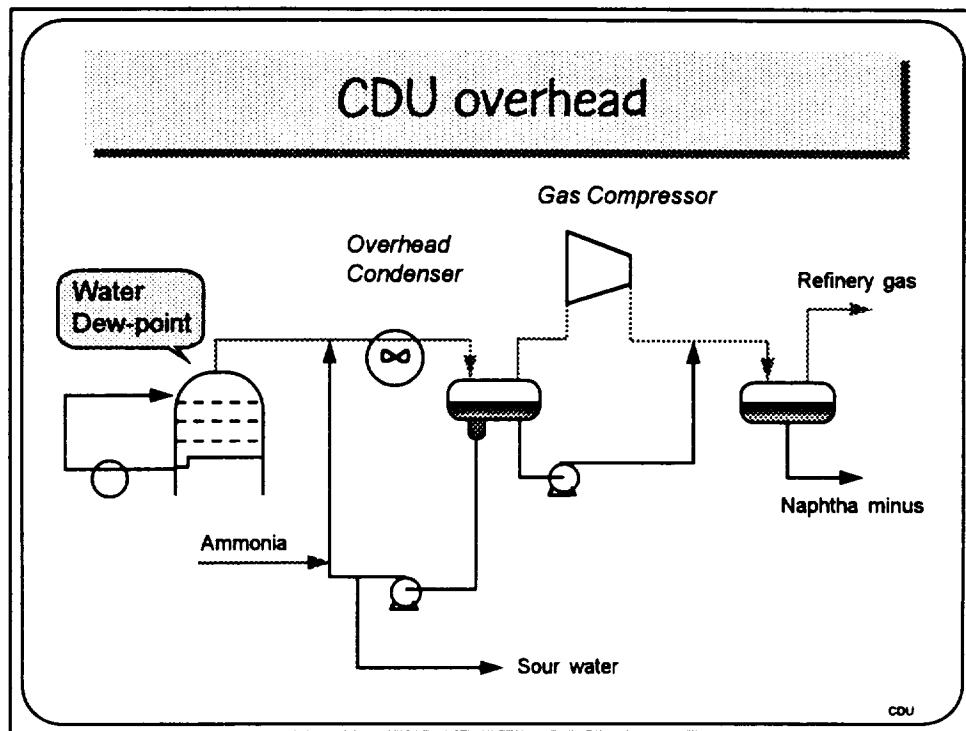
The heated crude is routed via the transfer line and flashed in the flashzone section of the atmospheric crude column. The flashed vapours are fractionated to produce the straight run fractions:

- Naphtha minus as an overhead product
- Kero, LGO and HGO as side cut fractions.

The heavy part of the crude, liquid which is not flashed, is steam stripped in the bottom section of the column and run down as long residue.

A Fractionator column has circulating reflux sections. Liquid is drawn from the column, cooled against crude inlet and returned to the same column section. With the subcooled liquid part of the vapour is condensed in this column section to generate internal reflux. The circulating reflux sections can be seen as internal partial condensers.

Separation into the various fractions is done in the fractionating sections. These sections are located between the circulating reflux sections.



Free water is very corrosive in the presence of HCl gas. The HCl originates from the Ca and Mg chlorine salts dissolved in the water which come with the crude.

The return temperature of the TCR should be above the water dew point temperature otherwise strip steam will condense and form free water in the top of the column.

Water dew point corrosion may occur at cold spots (non-insulated vent valve) on the top of the fractionator.

Water with a high HCl concentration may be present in the condenser tubes at the point where the very first few droplets of water do knock out.

To prevent corrosion a water is circulated through the overhead condenser. The circulation water should be sufficient so that the bulk of the injected water remains in the liquid phase; this will reduce the HCl concentration in the water.

The acidity of the circulation water is neutralized with ammonia.

## **Crude Distilling Unit**

### **1. Process description**

The refinery crude intake is processed in the Crude Distilling Unit (CDU). The main purpose is to separate the crude in various fractions with a specific boiling range: Naphtha minus and three middle distillate fractions Kero, Light Gas Oil (LGO) and Heavy Gas Oil (HGO). The remaining heavy part of crude is run down as Long Residue (LR).

The Naphtha minus fraction is routed via a Hydrotreater unit to the Debutaniser column. In the Debutaniser the Naphtha minus is stabilised by removing gas (including H<sub>2</sub>S) and LPG. The gas is used as refinery fuel gas and the LPG is a final product. The stabilised full range Naphtha is subsequently separated in a Splitter column to TOPS and Naphtha.

The middle distillates are final products or products that will only be hydrotreated.

### **2. Crude preheat section**

Crude from tankage is routed through the first set of heat exchangers. At a temperature of about 140 C, the viscosity of the crude is low enough to mix it with wash water. The soluble salts in the crude are wash out with the injected water. In the Desalter the effluent water together with the salts is removed from the crude. A full description is given in the Crude oil desalting chapter.

The desalted crude is routed through a second set of heat exchangers. At a temperature of about 180 C the light hydrocarbons and the dissolved water are flashed off in the preflash vessel.

The preflashed crude is heated further in a third set of heat exchangers and transferred to the crude furnace. The furnace inlet temperature is about 250 C depending on the overall efficiency of the crude pre heat train. With the crude furnace the preflashed crude is heated to about 360 C. The furnace effluent and the preflashed vapours are routed via the transfer line to the flash zone of the atmospheric crude column.

### **3. Atmospheric crude column**

The main column consists of a number of fractionation sections, circulating reflux sections and a stripping section in the bottom of the column. These column sections are all stacked on top of each other. A typical crude column is shown in the Figure.

The pre flashed crude enters the column in the flash zone where the distillates are flashed off. The middle distillate fractions are condensed and drawn from the column as side cuts. These streams are transferred to side stripper columns where the volatile components are stripped off before they run down as final product. The light ends fraction leaves the column as an overhead vapour stream.

### 3.1 HGO/LR separation

The separation between the HGO and LR is made in the fractionation section above the flash zone. The internal reflux in this section is generated in the Lower Circulating Reflux (LCR) section.

Liquid from the feed flash together with the internal reflux from fractionation section runs through the stripping section in the bottom of the column. Here it is steam stripped to recover the gasoil components from the LR.

### 3.2 Kero/LGO/HGO separation

The separation between Kero and GO is made in two fractionation sections. The lower section separates LGO and HGO while the higher section separates Kero and LGO. The internal reflux for both sections is generated in the Mid Circulating Reflux (MCR) section above.

### 3.3 Naphtha/Kero separation

The separation between the Naphtha and Kero is made in the fractionation section above the MCR. The internal reflux in this section is generated in the Top Circulating Reflux (TCR) section in the top of the column.

## 4. Overhead Recontacting Section

The overhead vapours are condensed in the overhead condenser and routed to the accumulator where the three phase are separated: gas, Naphtha minus liquid and sour water. The gas is compressed and recontacted with the Naphtha to recover the propane components. The remaining gas is routed to the refinery fuel gas pool.

## 5. Side-strippers

The middle distillate fractions drawn from the crude column are not suitable as final product. They require an improvement on flash point by stripping of the volatile components. Kero will generally be stripped in a reboiled side-stripper while LGO and HGO are stripped in side-strippers operating with steam. The stripped off vapours are routed back to the main column.

Steam stripped Gasoil is saturated with water. The free water is removed in a dryer column where the water is flashed off under vacuum conditions.

The Kero product from the reboiled side-strippers is already free of water.

## 6. Product Qualities

### 6.1 Naphtha

The maximum FBP of Naphtha is limited to about 190 C depending on the type of Platformer process down stream.

The minimum FBP is limited by the flash point specification of the Kero product. The flash point of Kero of 38 C will generally be reached when the Naphtha FBP approaches 140 C.

### 6.2 Kero

The specification of Kero flash point is typically 38 C. This is a front end quality of the fraction. It is controlled with the Kero side stripper.

When Kero is produced for Avtur product there is a freeze point specification of -47 C. The freeze point property depends on both the front end and the heavy tail of the Kero fraction.

Additionally Avtur may have an ASTM FBP specification of 300 C maximum.

### 6.2 Gasoil

The flash point specifications of LGO and HGO are typically around the 66 C. This is a front end quality which is controlled with the GO side strippers.

The cutpoint between LGO and HGO is arbitrarily but should be consistent over a period of time. The actual value depends on the requirements for final product blending. A sharp fractionation between these products is not necessary since both fractions ultimately end up in the same Gasoil pool.

LGO is often used as a Fuel Oil blending component. A V50 viscosity may be specified.

For HGO there is normally a Cloud point spec. Additionally there may be an ASTM 90% spec, also determined by the final product blending requirements.

GO as final product should meet the commercial specification: 'free of water at ambient conditions'. Because the LGO and HGO are stripped with steam these products are saturated with dissolved water when they leave the side strippers. The water is subsequently removed under vacuum condition in Dryer columns.

#### 6.4 Long Residue

In case the LR is routed to an HVU the separation between the HGO and LR will be determined by the maximum quantity of gasoil components that the HVU can handle. This is normally specified by a maximum of 7% wt of 350 minus components in the LR.

### 7 Operating conditions

#### 7.1 Column pressure

The operating pressure of the atmospheric crude column is set as low as possible to maximise the HGO yield. However the pressure at the suction of the overhead compressor should always be above the atmospheric pressure. This is for safety reasons to avoid ingress of air.

#### 7.2 Furnace

The furnace outlet temperature (FOT) is typically 360 C. The temperature is limited to avoid cracking reactions. Too much cracking gives fouling in the furnace coils by coke deposits.

Coke deposit is suppressed with a high velocity in the coils. Cross over steam may be used.

#### 7.3 Residue stripping

The amount of strip steam in the bottom of the column ranges from 1 to 3 % wt. on LR. Higher strip steam rates are not effective.

#### 7.4 Circulating refluxes

The pump around flow of circulating refluxes should be kept constant otherwise critical dynamic effects will upset the column operation. The condensing duty should be regulated with temperature only !

With the column flash zone conditions fixed,

The separation between HGO and LR originates from the over flash; typically in the range from 0.5 to 1.0 %woc. The flash zone conditions (column pressure, furnace outlet temperature and cross over steam) are normally not changed. The cutpoint between the HGO and LR is controlled with the LCR duty. The LCR is therefore the key control parameter for the HGO cloud point and ASTM 90% specification.

The cutpoint between Kero and LGO is controlled with the internal reflux in this column section. The MCR condensing duty is the key control parameter for the Kero freeze point.

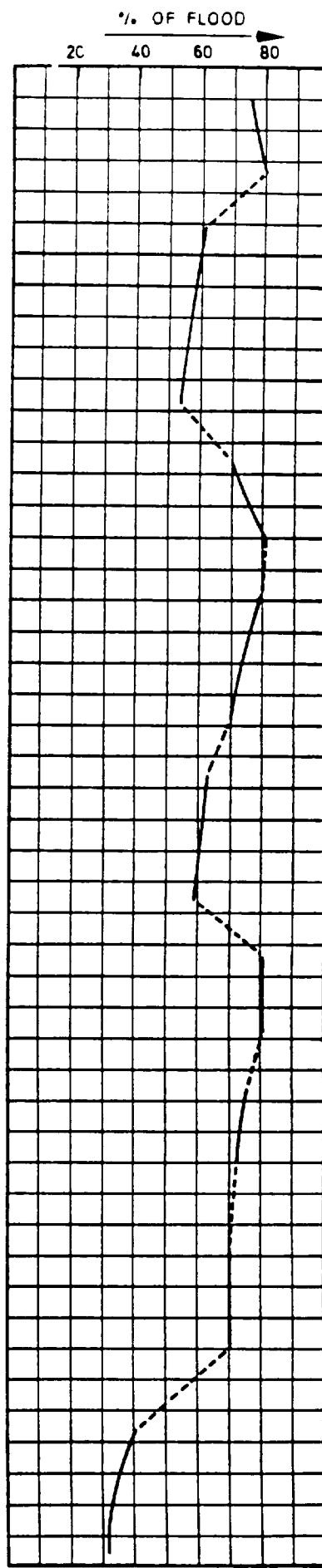
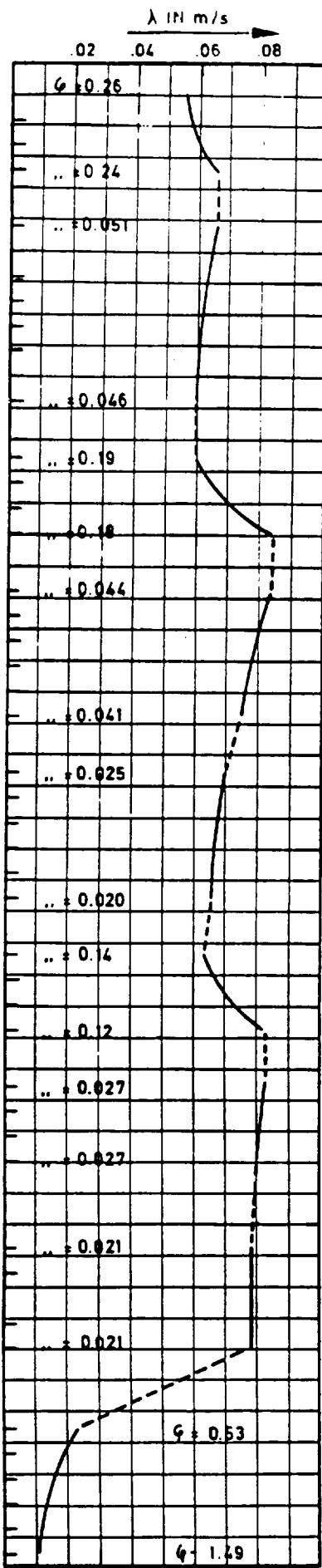
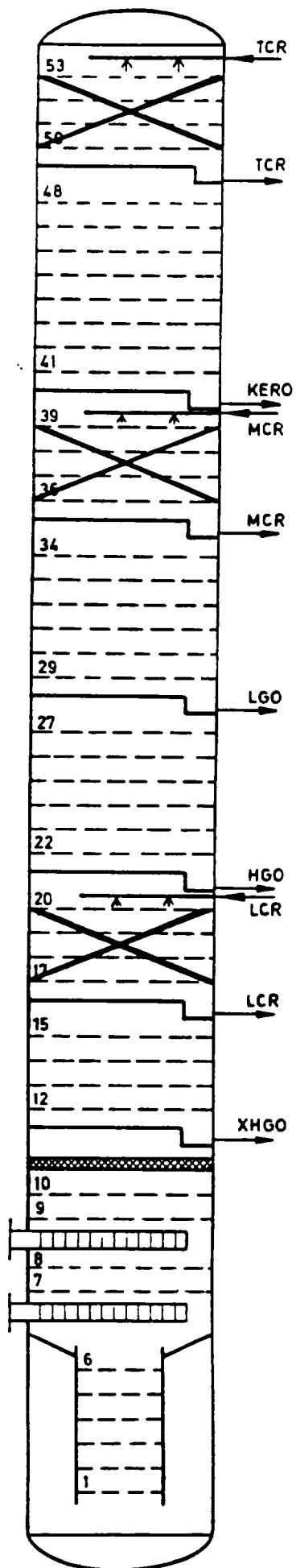
The separation between Naphtha and Kero is set by the TCR duty. Hence the Naphtha FBP is controlled by the TCR.

The heat balance between the circulating reflux sections depends on the yield structure of the distillates. The condensing capacity of the circulating refluxes is in general limited by a minimum temperature of about 50 C below the bubble point draw off temperature. Consideration should be given to the return temperature of the TCR. Too low a temperature will create free water in the top of the column, which is very corrosive or may cause fouling by ammonium chloride salt deposits.

#### 7.4 Under refluxes

With opening of the under refluxes it is possible to transfer internal reflux from one column section down to the section below. Condensation duty is then shifted to the section above. This will increase vapour/liquid traffic in the upper sections of the column. The result will be an improved separation in the upper section at the expense of a lower heat recovery from the lower section.

Normally under refluxes are used to accommodate changes on distillate yield e.g. when processing different crudes. They are not suitable for cut point control.



LOAD PATTERN OF A CRUDE DISTILLATION COLUMN

## Crude oil desalting

### 1. Introduction

Crude oil as it arrives at the refinery carries some water and solid particles like silicates (sand) and FeS (scale). The water which originates from the formation water and the ballast water during transportation of the crude contains water soluble salts such as sodium, magnesium and calcium chlorides. In large quantities they can be harmful in the refinery process:

- Furnace and heat exchanger fouling as a result of salt deposition.
- Column overhead corrosion and fouling because of chlorides.
- A high sodium content in the Fuel oil.
- Residue conversion units (see Appendix)

The purpose of crude desalting is to remove the solids and the water soluble salts before the crude is processed in the CDU.

### 2. Desalting process

The basic steps of the desalting process are:

1. Heating of the crude oil to lower its viscosity
2. Effectively mixing of fresh wash water with the crude to dilute the salt concentration of the water contained in the crude.
3. Removal of the water soluble salts by separation of water from the oil in the desalter vessel. The effluent water is drained from the bottom and routed to the CPI. The desalted crude leaves the vessel over the top.

### 3. Desalting efficiency

Prior to desalting a lot can be gained by proper dewatering of the crude upon arrival at the refinery. A residence time of at least 24 hours in the crude tank reduces the salt content of the crude already with some 50 % by settling out part of the water. The water should be drained carefully before the tank is lined up to the CDU.

The desalting depends on the mixing efficiency of wash water and the subsequent dehydration in the desalter vessel. The desalting efficiency is normally better than 80%. For two stage desalting it can be as high as 97%.

The salt contained of the desalted crude can be estimated by:

$$S = S_{\text{crude}} * W_d / (W_c + M * W)$$

where:

- |       |   |
|-------|---|
| S     | = Salt content                            |
| $W_c$ | = Free water in fresh crude               |
| $W_d$ | = Free water in desalted crude            |
| W     | = Wash water                              |
| M     | = Mixing efficiency (ranging from 0 to 1) |

#### 4. Wash water

The amount of wash water is typical 4 to 5 % on crude throughput. For very light crudes with viscosity V<sub>50</sub><12 it may be less because of an better mixing efficiency.

The quality of the water should preferably not exceed the following limits:

pH value between 7 and 8, measured at 40 deg C

[NH<sub>3</sub> + NH<sub>4</sub><sup>+</sup>] content less than 200 ppm and

[H<sub>2</sub>S + HS<sup>-</sup>] content less than 50 ppm.

Stripped sour water from the SWS is preferably used because such re-use limits the amount of effluent water to be treated. Other sources can be low contaminated process water such as condensate water from the CDU overhead, industrial water or clarified river water.

Before the water is injected in the main crude stream it may be pre-mixed with a small slipstream of crude to facilitate the mixing efficiency. The pre-mixing is done in a cyclone mixer where small droplets of water are dispersed through an oily phase.

#### 5. Operating temperature

The crude is pre-heated to 125 - 140 deg C. The temperature has a significant influence on the desalting process, because the crude viscosity decreases with higher temperature. A low viscosity improves the mixing and the water separation. This is especially important when heavy crudes are desalted. For example, in a heavy crude ( $d_{15/4} = 0.92$  and  $V_{50} = 30$ ) the water settling velocity at 100 deg C is only a third of that at 140 deg C.

#### 6. Mixing

Wash water and crude are mixed with a globe valve to bring the wash water in contact with the water droplets contained in the crude. To avoid formation of a water/oil emulsion the mixing energy should not too high. The pressure drop over the mixing valve is preferably in the range 0.4 - 0.8 bar.

Crude properties that are indicators for emulsion formation and thus require special attention with regard to mixing include:

- presence of small particles like rust (iron oxides and sulphides), clay, coke
- high naphthenic content
- high asphaltene content

The mixing efficiency can be substantially improved and simultaneously the emulsion formation tendency reduced, by installing a static mixer downstream the mixing valve. The residence time in the mixing zone is extended (longer with more mixing elements) and the mixing is also more gentle. An equilibrium is reached between break-up and coalescence of water droplets. This results in a more narrow water droplet size distribution than obtained over a mixing valve only. Consequently there remains only a small fraction of very fine (emulsified) droplets that cannot be separated in the desalter vessel.

#### 7. Desalter vessel

In the desalter vessel the salt containing water is separated from the crude. The Stokes parameter is a measure for the settling time of water droplets through the crude phase, defined as the ratio between density difference of crude and water over the crude viscosity at

operating temperature and pressure.

The vessel is operated at high pressure to keep the water in the liquid phase. The water settles out by gravity. Deep dehydration of the desalted crude is achieved by coalescence of small (polar) water droplets with an electrical field. The electrical field is generated with a high voltage electrode grid inside the vessel.

- Steady oil/water interface level control is essential to keep the electrical grid in the oily liquid phase. When the water reaches the grid the deep dehydration stops because of an electrical short circuit. A breakthrough of the oil phase will contaminate the effluent water.

Emulsion formation is by far the most frequently problem, observed as short circuiting of the electrode grid and subsequent power trip. It can be suppressed with an oil soluble demulsifier. The injection point is upstream the crude charge pump or upstream the mixing valve. The dosage rates range from 2 to 12 ppm on crude. Processing slops, especially slops containing cracked materials is likely to lead to emulsion formation. During slops runs the demulsifier injection rate can be increased but overdosing should be discouraged since it may in fact promote again emulsion formation.

Sludge deposition (solid particles in the crude) will accumulate at the desalter bottom. Excessive sludge build-up may reduce the performance by reducing hydraulic residence time for the water phase. A pressure water jetting system can be used for on-line flushing of sludge and sand. To prevent compacting of the sludge, this operation should be carried out at regular intervals (e.g. once per shift for 5 to 10 minutes). Preferably recycled wash water is used for jetting.

Precautions should be taken when cleaning out the desalter vessel during a maintenance stop since it may contain pyroforic FeS fines.

Appendix**Residue conversion units****Thermals Cracking Units (TGU)**

The salt content in thermal cracker feedstock should be limited to 50 ppm wt NaCl. Higher salt levels will increase the coke formation in the cracker furnace, and above 100 ppm wt NaCl severe coking can be expected. The level of salts that can be accepted in the crude after desalting are thus dictated by the LR and SR yield: for LR-feed usually between 8 -20 ppm (average 15 ppm NaCl), for SR-feed in general around 8 ppm NaCl.

**Fluidised Catalytic Cracking Units (FCCU)**

In the case of flashed distillate FCCU's no salts are carried with the feed. However, when residue is co-processed, there is a number of problems to be avoided:

- excessive coke formation on the catalyst
- catalyst poisoning with metals (V, Ni)
- salt deposition (Na, Ca, Mg), which deactivates the catalyst by sintering the pores.

A maximum of 5 ppm wt of Na (i.e. appr. 13 ppm of NaCl) in total FCCU feed will usually maintain equilibrium Na<sub>2</sub>O levels below 1.1 % wt, which is the approximate level for the onset of sintering problems. These requirements, in combination with the LR yield and the proportion of LR processed in the FCCU, lead to a NaCl specification on desalted crude, which in most cases will lie around 5-10 ppm wt NaCl, requiring two-stage desalting.

**Hycon**

In this process short residue is hydroconverted, where V and Ni are removed in the first HDM reactors. In order to prevent blocking of the catalyst bed by salts and iron, deep desalting is required. The salt content of the SR is limited to 6 ppm wt Na. For high sulphur crudes this translates in a salt specification on desalted crude of appr. 1.5 ppm Na (including any possible caustic addition for crude neutralisation). As for catcrackers, this stringent specifications calls for a double stage desalting process.

**Shell Gasification Process (SGP)**

The Shell Gasification Process processes the most viscous residues available on the refinery, such as deep thermally cracked flashed residues and C4/C5 asphalt's. To prevent fouling of the SGP waste heat boiler and scrubber, the following specifications on feed quality are set:

- Na + K < 30 ppm wt
- Ca + Mg < 20 ppm wt (up to 80 ppm wt with anti fouling additives)

Depending on the residue yields, a typical specification on desalted crude would thus be 3 ppm wt Na + K and 2 (max. 8 ) ppm wt Ca + Mg.

**Delayed coker**

For SR processed by delayed cokers usually a 20 ppm Na spec is set, in order to limit fouling. From the survey it appeared that for most refineries that quoted strict specifications on desalted crude these are governed by quality requirements for either LR (typically 10 -30 ppm Na) or for SR (typically 20 - 50 ppm).



**12**

**Revamp study**

## **Revamp study of a crude distilling unit**

### **1. Summary**

We will consider a small crude distilling unit which produces the following straight run distillate products:

Gasoline
Kero
Gasoil

Initiated by developments on the oil market the plan is raised to produce more distillate products. The current crude distiller has a maximum capacity of 1500 t/sd primarily because some severe bottlenecks of the crude furnace and the overhead condenser. This was confirmed by a recent testrun. The intention is to revamp the unit and increase its capacity to 2000 t/sd crude intake.

In this exercise we will do some process calculations necessary to check the feasibility of the envisaged capacity increase. Further we will develop revamp options.

### **2. Introduction**

The crude distiller is an old unit build in 1948. It consists of a main atmospheric crude column with two steam side-strippers, a crude furnace and a pre-heat train with pre-flash vessel as shown in Figure 1. The overhead vapours are condensed with a water cooler to produce top reflux and Gasoline product. Both, the main column and the side strippers are equipped with bubble cap trays, original from the design.

The total utility consumption of the unit (fuel and steam) is high due to the poor heat recovery for crude pre-heating. From the inspection reports it is evident that the heat exchangers of the crude preheat train, including the steam heater are in a bad state and require a lot of maintenance. It was decided that all exchangers have to be renewed.

Due to developments on the market, demand for oil products has increased spectacularly and will continued to grow for the coming years. This provides a strong economic incentive for increasing the refinery throughput and thus to raise the crude distilling capacity to the maximum possible extent. As a first step of the revamp project a testrun was conducted to establish the maximum crude intake of the current unit and to identify the limiting constraints.

### 3. Maximum capacity testrun

Before the testrun was held the main instruments were checked and burners were cleaned. Precautions were taken to avoid tank switches and any other external disturbance which may effect the testrun results. Furthermore the specifications for the straight run products were checked and agreed with E&S department.

**Product specs**

ASTM (% vol.)	Gasoline (degr C)	Kero (degr C)	Gasoil (degr C)
10	-	176	241
50	110	186	278
90	148	209	358

The testrun was performed on Kirkuk crude. Initially the crude feed rate was gradually raised while the column operation were adapted to keep the products on specification. At an intake of 1500 t/sd the increase was stopped. At this feed rate both the crude furnace and the overhead condenser appeared to be the constraints of the unit. The following bottlenecks were observed:

- |                           |   |
|---------------------------|---|
| <u>Crude furnace</u>      | Maximum capacity reached. Fuel supply to the furnace could not be raised any further. |
| <u>Overhead condenser</u> | The cooling water supply valve was fully open.  |

During 24 hours the crude intake was maintained at the maximum capacity of 1500 t/sd. and the plant conditions were kept as steady as possible. The measured flows, temperatures and pressures are shown in Figure 1 and the overall mass balance is given in Table 1.

Analytical data obtained from the product samples taken during the testrun are listed in Table 2. From the results it is evident that the product qualities were kept within specifications. Especially the separation between Kero and Gasoil is better than required.

#### 4. Revamp feasibility study

From the testrun results it is clear that the feed rate could not be raised any further than 1500 t/sd due to limitations of the crude furnace and overhead condenser.

To improve the refinery economics it is the intention to revamp the crude distiller with the following objectives:

- 1- Increase the maximum capacity of the unit to 2000 t/sd crude intake.
- 2- Maximise the yield of distillate product in %woc.
- 3- Reduce the specific utility (fuel and steam) consumption of the unit.

Preparing a complete and detailed revamp design is in general a very time consuming exercise. It requires a thorough inspection of every piece of equipment, including pumps, line sizes, control valves and a review of the safety aspects such as safety relief cases . . . ect. So before embarking on such a project, first the technical feasibility (and economics) of the revamp should be confirmed by a broad brush check of the main pieces of equipment.

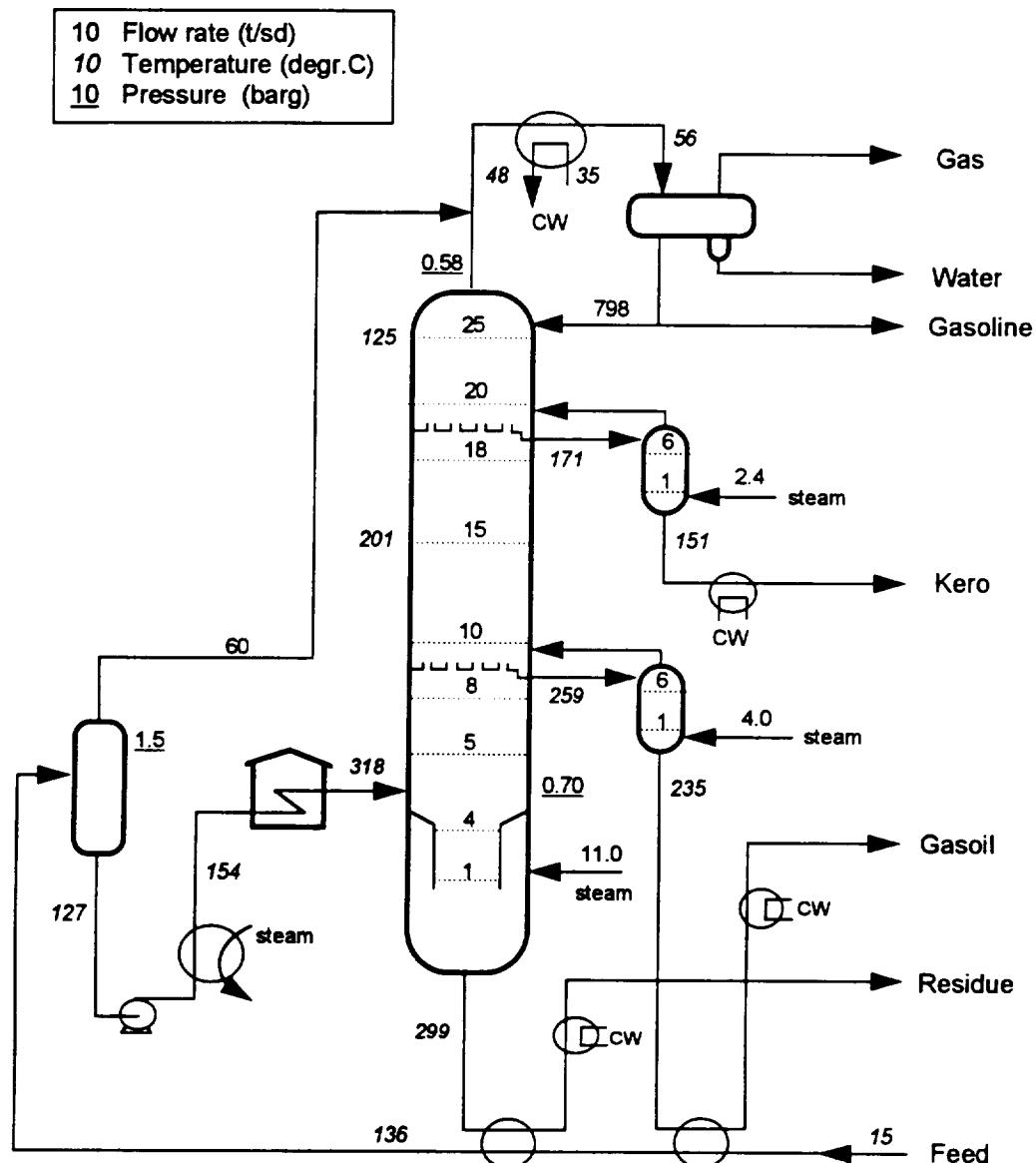
In view of the available time we have to restrict ourselves to the main items.

TABLE 1, OVERALL MASS BALANCE

Inlet	t/sd	Products	t/sd	% woc
Crude intake	1505	Gas	51	3.4
Strip steam main column	11.0	Gasoline	291	19.3
Strip steam Gasoil stripper	4.0	Kero	159	10.6
Strip steam Kero stripper	2.4	Gasoil	344	22.9
		Residue	660	43.9
Total	1522.4	Water	17.4	
		Total	1522.4	

TABLE 2, ANALYTICAL DATA TESTRUN

	Crude	Gas	Gasoline	Kero	Gasoil	Long Residue
ASTM			(degr C)	(degr C)	(degr C)	
IBP			42	167	214	
10 %vol			62	175	246	
20			76	179	255	
30			88	182	263	
40			98	185	271	
50			108	188	279	
60			117	192	290	
70			127	196	305	
80			137	201	325	
90			147	208	357	
95			155	215	378	
FBP			166	218	393	
d 15/4	0.846		0.712	0.788	0.841	0.950
UOP-K			12.5		12.0	
MW (dry gas)		70	108			



Main column: 2740 mm ID  
Insert: 1207 mm ID  
Sidestrippers: 905 mm ID  
Tray spacing: 28"

Figure 1 Crude distilling unit

## Exercise

### 1.1 Test run results

Considering the process scheme and the testrun data what are your observations.

E.g. can you comment the following points before doing any process calculations:

- flooding of trays
- crude preheat train
- crude preflash
- furnace temperatures
- condenser temperatures
- strip steam flow rates

### 1.2 Crude data

Attached is the COED (crude oil evaluation data sheet) for Kirkuk, the test run crude.

- If you compare the potential distillate yield with the testrun result can you explain the difference.
- Is the analytical data in line with the COED ?



COED

CONFIDENTIAL

KIRKUK

IRAQ

GENERAL INFORMATION				LIGHT HYDROCARBONS (% M/M)							CRUDE OIL CHARACTERISTICS				
THIS EVALUATION WAS CARRIED OUT ON A SAMPLE COLLECTED FROM ACTUAL PRODUCTION.				C2 MINUS	<.01						DENSITY AT 15 C	KG/L	.845		
THE ANALYTICAL RESULTS ON THIS PAGE CAN BE REGARDED REPRESENTATIVE OF THE PRESENT QUALITY OF THE CRUDE OIL.				C3	.11						API GRAVITY 60 F	DEGR.	35.9		
A STANDARD SIMPLE ASSAY CARRIED OUT ON A SAMPLE TAKEN 01/85 HAS NOT SHOWN QUALITY DEVIATIONS OF ANY SIGNIFICANCE.				IC4	.35						BARREL FACTOR BBL/T		7.457		
N O T E : THIS PAGE REPLACES THE ONE ISSUED: 01/72 SAMPLE: 08/71 ASSAY: 12/71 ISSUED: 03/87				NC4	.85						KIN.VISC. AT 100 F	MM2/S	4.61		
				IC5	1.25						SULPHUR CONTENT % M/M		2.06		
				NC5	1.30						REID VAPOUR PRESSURE PSI/KPA		5.5 / 38		
				CYCLO C5	.09						POURPOINT (ASTM-MAX) EXIST.	DEG C	-36		
				NC6	1.50						H2S CONTENT MG/KG		< 1		
											POTENT. H2S CONTENT MG/KG		64		
											CALC.GROSS CAL.VALUE KJ/KG		44890		
FRACTIONS	IBP C FBP C	CRUDE OIL	ATMOSPHERIC DISTILLATION										FLASHING		
			C5 65	65 100	100 150	150 200	200 250	250 300	300 350	350 370	370 +	370 495	495 +	370 510	510 +
YIELD ON CRUDE OIL % M/M			4.7	5.3	8.8	9.7	9.5	9.3	9.0	3.2	39.2	19.2	20.0	21.0	18.2
YIELD ON CRUDE OIL % V/V			6.2	6.4	10.0	10.5	9.9	9.4	8.8	3.1	34.4	17.9	16.6	19.4	14.9
POSITION IN CRUDE OIL % M/M	1.3	6	0	-11	3	-20	1	29	8	39	3 - 48	6	- 57	60.8 - 80	0 - 100
MID YIELD ON CRUDE % M/M			3.7	8.7	15.7	25.0	34.6	44.0	53.1	59.2	80.4	70.4	90.0	71.3	90.9
DENSITY AT 15 C KG/L		.845	.644	.701	.746	.782	.807	.835	.864	.884	.962	.909	1.020	.913	1.028
REL.DENSITY AT 25/25 C KG/L											.929	.875	1.017	.987c	1.025
DENSITY AT 70 C KG/L															.995c
ASTM-DISTILLATION: IBP DEG C	39			75	115	161	209	260	307			Vac.	Vac.		
05% V/V RECOVERED AT DEG C	74			78	118	164	212	263	311			423	397		
10% V/V RECOVERED AT DEG C	198			79	119	165	213	264	312			411	399		
30% V/V RECOVERED AT DEG C	184			82	122	167	216	267	315			450	416		
50% V/V RECOVERED AT DEG C	289			84	126	170	219	269	317			504	440		
70% V/V RECOVERED AT DEG C				88	129	174	223	273	321			472	474		
90% V/V RECOVERED AT DEG C				93	137	182	231	279	327			512	519		
END POINT DEG C				97	147	189	238	285	333						
KIN.VISC. AT 100 F MM2/S	4.61						1.68	3.06	6.25	11.1		280	22.7	20100	25.2
KIN.VISC. AT 140 F MM2/S												43.9	7.23	870	7.79
KIN.VISC. AT 210 F MM2/S							4.1	9.7	14.9	18.4		33.8	25.5	42.0	25.9
VISCOOSITY V50 VALUE	12.8														48820
MERCAPTANS (AS S) MG/KG			257	314	337	167									1561
SULPHUR CONTENT MG/KG			333	903	1810										43.2
SULPHUR CONTENT % M/M	2.06														
TOTAL NITROGEN MG/KOH/G	<.05		<.05	.1	0	0	.30	.85	1.60	2.17	4.00	2.30	5.35	2.35	5.69
TOTAL ACID NR/VALUE MG KOH/G	<.05		<.05	<.05	<.05	<.05	.05	.05	.149	.388	.10	.915	.15	.10	.15
REID VAPOUR PRESSURE PSI/KPA	38		96	52	38	< 38									
RESEARCH ON-CLEAR			73	60C	47C										
RESEARCH ON+.15 G PB/L (TEL)			81C	65C	54C										
RESEARCH ON+.4 G PB/L (TEL)			86C	65C	54C										
RESEARCH ON+.5 G PB/L (TEL)			88	67	56	44									
RESEARCH ON+.5 G PB/L (TML)			88	63	53	40									
TOTAL PARAFFINS % M/M			97	75	62	52									
TOTAL NAPHTHENES % M/M			3	23	26	25									
TOTAL AROMATICS % M/M			0	2	12	23									
FIA-AROMATICS % V/V							20	17							
SMOKE POINT MM							24	22	19	< .5	< 1	< 1.5			
COLOUR (AFTER AGEING)							30	19	< .5	< 1	< 1.5				
ANILINE POINT DEG C							27	16	< .5	< 1	< 1.5				
ANILINE POINT 150/250 DEG C							54.8	66.1	71.8						
FREEZING POINT DEG C							61.0	-->							
FREEZING POINT 150/250 DEG C							< -60	-41	-16						
							52	-->							
CLOUDPOINT DEG C															
CLOUDPOINT 300/370 DEG C															
CLOUDPOINT 250/370 DEG C															
CLOUDPOINT 200/370 DEG C															
POURPOINT (ASTM-MAX) DEG C															
POURPOINT 250/370 DEG C															
POURPOINT 200/370 DEG C															
CETANE NUMBER															
CETANE INDEX IP-218/84															
WAX CONTENT (SHELL) X M/M	3.9						46	53	58	62	60				
CONGEALING POINT WAX DEG C	52							52	55	56	56				
PARAFFIN WAX (DIN) X M/M									8.0	13	9.0	13			
CONG.POINT PAR.WAX DEG C									32	41	53	52			
REFRACTIVE INDEX ND20C			1.3918	1.4162	1.4369	1.4492	1.4649	1.4805	1.5928			1.4883	7.00	1.4901	
REFRACTIVE INDEX ND70C			5.50	5.80	6.10	6.10	6.20	6.50	6.60			392	423		
C/H RATIO M/M			89	115	129	122	223	262	309						
MOLECULAR MASS															
UOPK-FACTOR 12.01															
C7-ASPHALTENES % M/M		1.5										3.6	< .04	< .06	
CONC.CARB. RESIDUE % M/M		3.8										10.0	.60	.70	
ASH CONTENT % M/M													< .001	< .001	
VANADIUM MG/KG															
NICKEL MG/KG															
SODIUM MG/KG															
PENETRATION AT 25C MM															
SOFTENING POINT DEG C															
C value calculated * value estimated or derived from other assay												300	35	188	40

## Exercise 2      Equipment constraints

Calculate the duties in tcal/sd of the equipment which created the bottleneck:

2.1 overhead condenser

2.2 crude furnace

To facilitate the calculations you may use the control surfaces shown in Figure 2.  
The enthalpies of the streams given below.

### ENTHALPIES

Streams	Flow (t/sd)	Temperature (degr C)	Enthalpy (tcal/t)
Preflash vapour	60	127	150
Column OVHD (dry)	1080	125	140
Column OVHD ( $H_2O$ )	17.4	125	650
Gas	51	56	113
Gasoline	291	56	30
Reflux	798	56	30
Kero	159	151	83
Gasoil	344	235	136
Residue	660	299	165
Strip steam (total)	17.4	243	715
Water	17.4	56	56
Flashed crude	1445	154	80

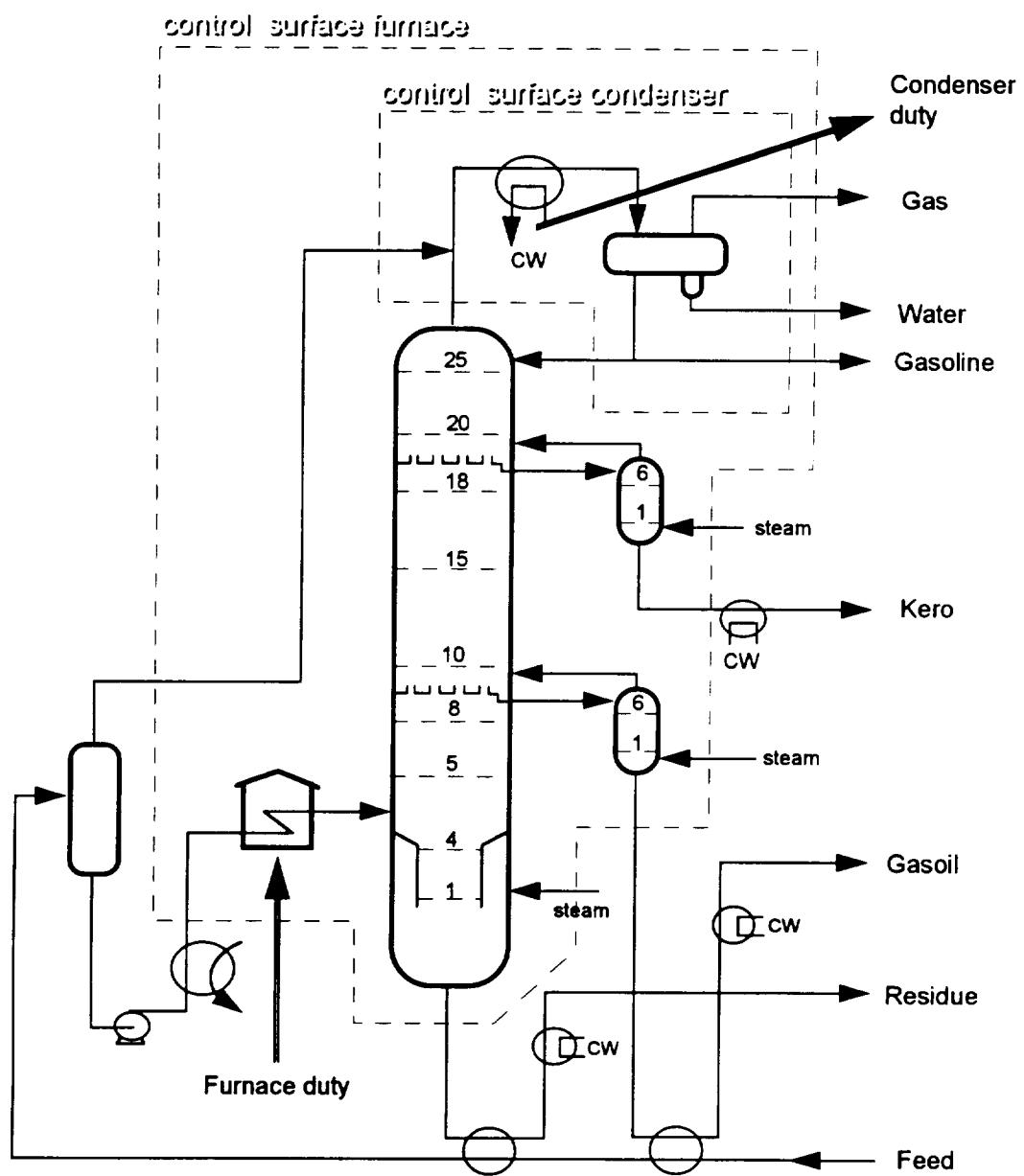


Figure 2 Control surfaces

### Exercise 3      Column loading

We will do a broad brush check of the column capacity to see if it can handle 2000 t/sd crude intake. What you need to know is the tray loading at the test run condition and scale them up proportionally to the new feed intake.

In general it is not necessary to do a check for all the trays. Inspection of the most critical trays is sufficient. In this case the highest loading is at the:

- top tray of the rectifying section; *Tray 25*
- top tray of the steam stripping section; *Tray 4*

The capacity curves for the installed bubble cap trays are given in Figure 3.

To calculate  $\varphi$  and  $\lambda$  the internal vapour and liquid flows are required. As usual they are not included in the test run data. However with some imagination you may be able to derive them from the given test run data using the attached correlation graphs.

*Note that high accuracy's are not required in this stage.*

#### 3.1    Tray 25

Calculate the loading of Tray 25 (at test run condition) in % flood.

The tray loading is based on the liquid *from* the tray and vapour flow *to* the tray. So the first step is to calculate these flows: Draw a control box around tray 25 and make a mass and enthalpy balance.

The temperature of vapour from tray 24 can be estimated by a linear interpolation between tray 20 and tray 25.

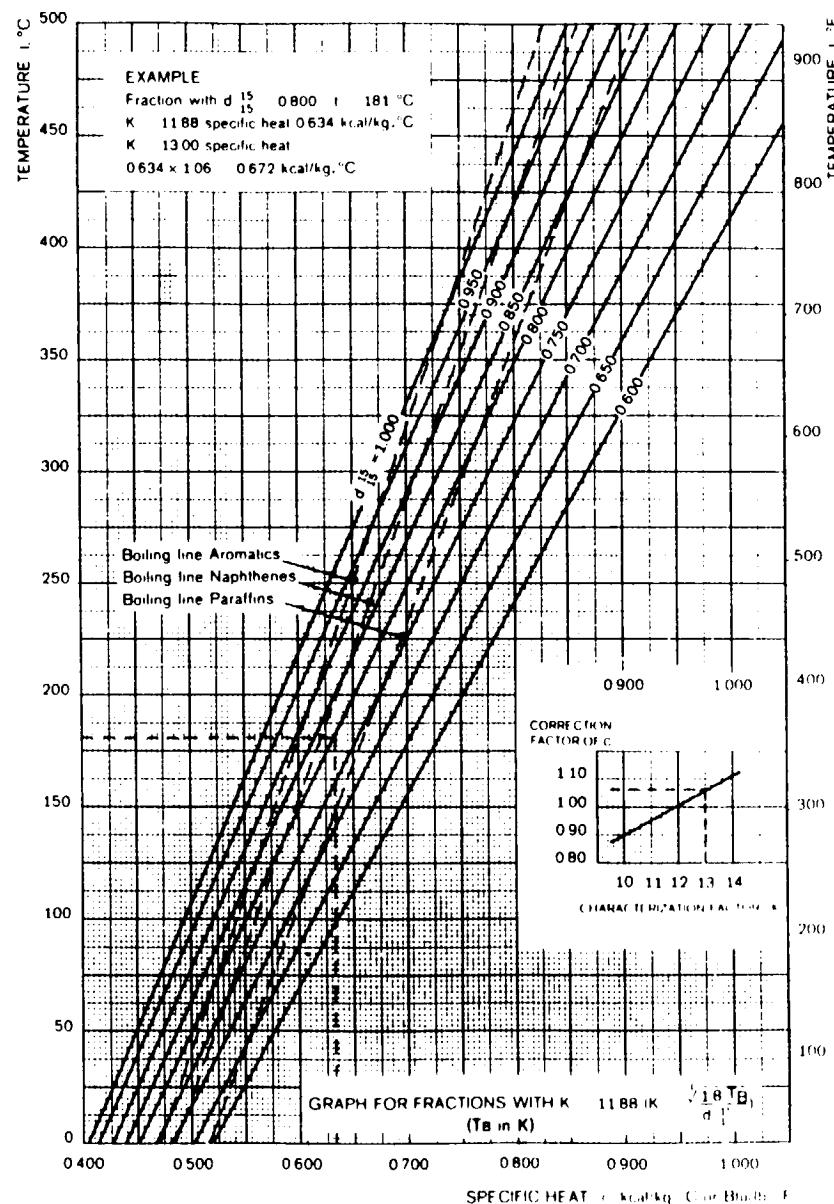
#### 3.2    Tray 4

Calculate the loading of tray 4 (at test run condition) in % flood.

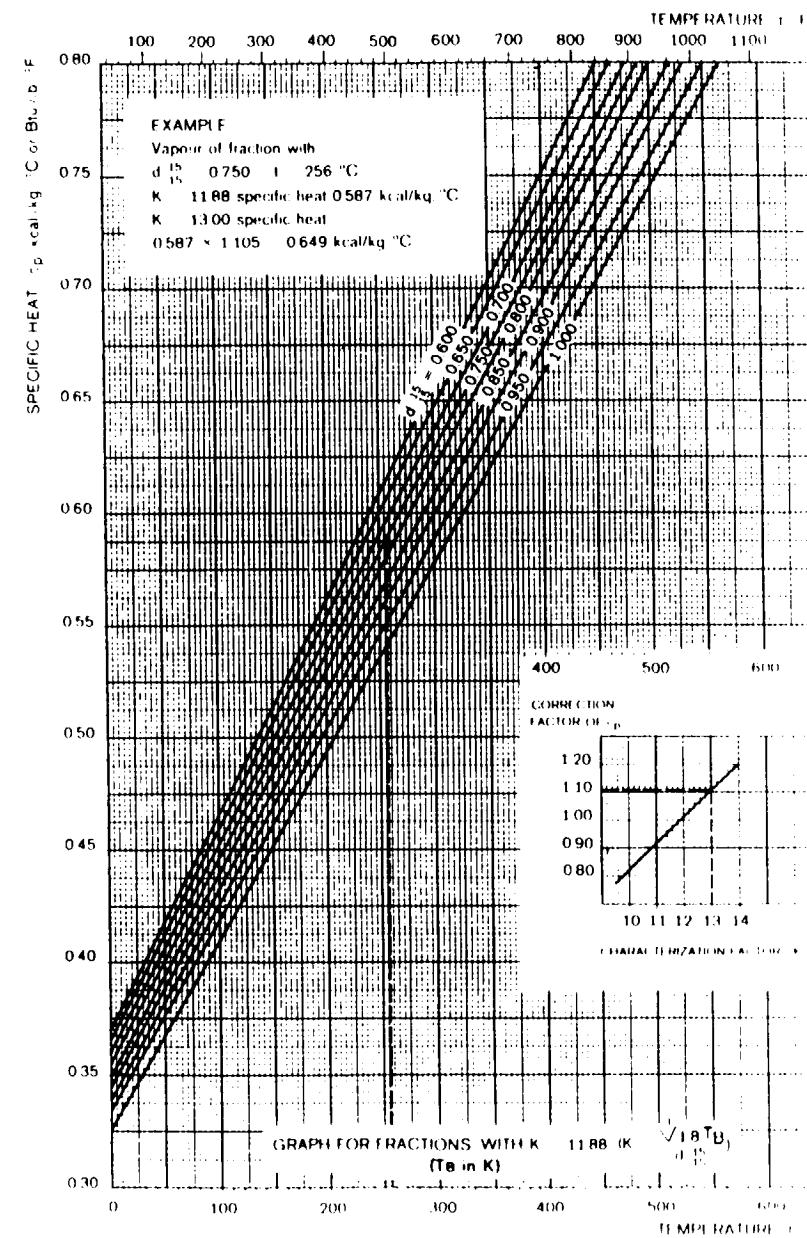
Note that for a steam stripping tray the loading is based on the liquid and vapour *from* the tray. This is to compensate for the increase of HC vapour on a steam stripping tray.

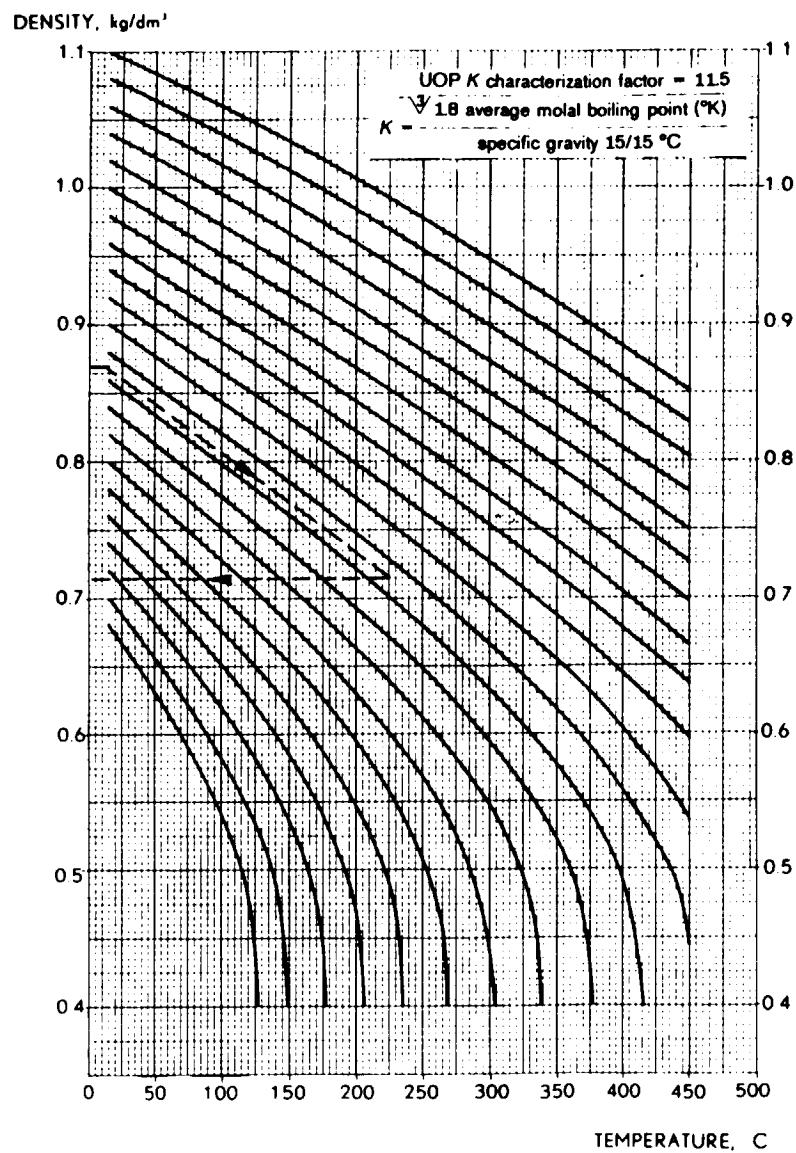
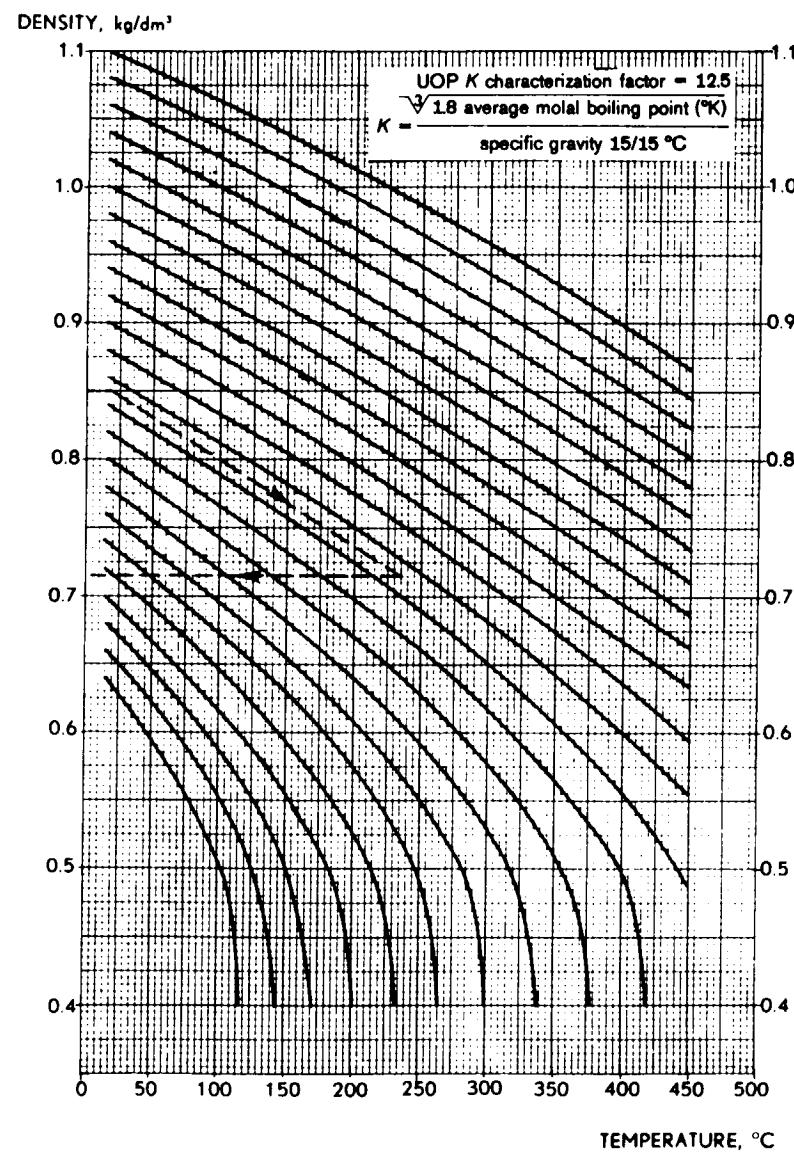
Use a foaming factor of  $F = 70\%$  for residue steam stripping.

D-1



D-2



**C - 2****DENSITY OF LIQUID HYDROCARBON MIXTURES (K = 11.5)****C - 3****DENSITY OF LIQUID HYDROCARBON MIXTURES (K = 12.5)**

D - 14

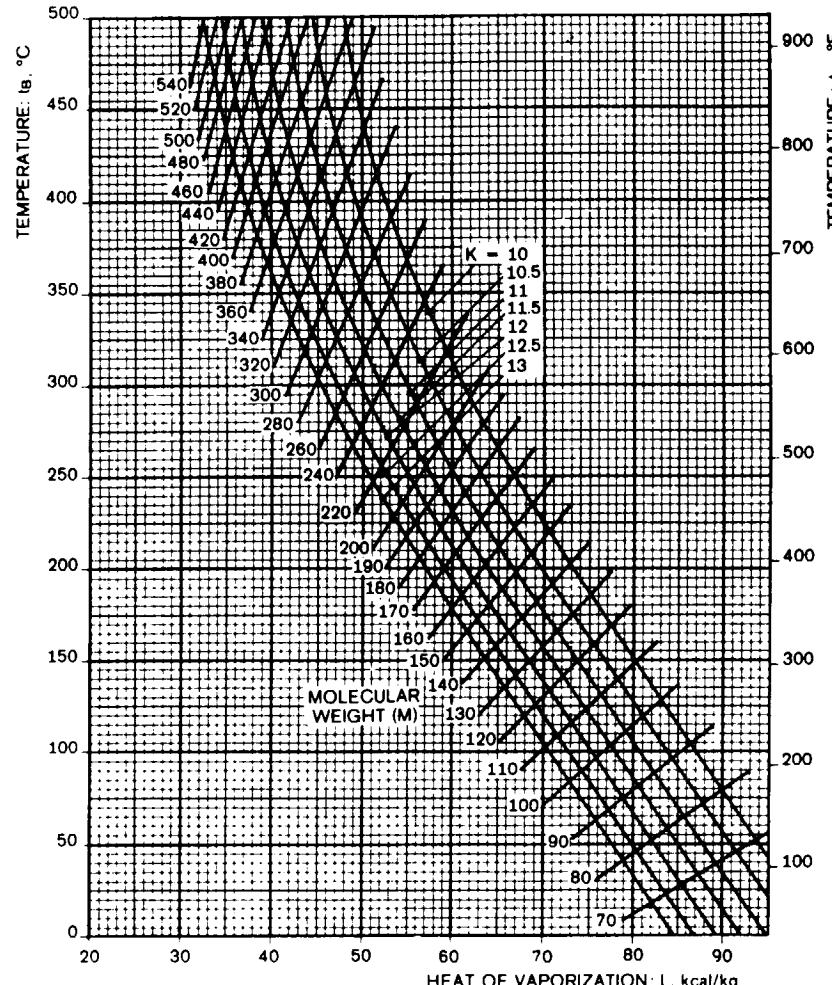
°C	H <sub>2</sub>	N <sub>2</sub>	O <sub>2</sub>	CO	H <sub>2</sub> O	°F
0	3.40	0.248	0.219	0.248	0.443	32
25	3.42	0.248	0.219	0.249	0.444	77
100	3.45	0.250	0.223	0.250	0.450	212
200	3.47	0.252	0.230	0.253	0.462	392
300	3.48	0.256	0.238	0.258	0.475	572
400	3.50	0.261	0.245	0.264	0.491	752
500	3.51	0.267	0.251	0.270	0.506	932
600	3.53	0.272	0.257	0.277	0.522	1112
700	3.57	0.278	0.260	0.282	0.539	1292
800	3.62	0.282	0.263	0.287	0.556	1472
900	3.66	0.287	0.266	0.291	0.572	1652
1000	3.71	0.290	0.268	0.294	0.587	1832
1250	3.84	0.298	0.274	0.302	0.621	2282
1500	3.96	0.303	0.278	0.306	0.650	2732
1750	4.07	0.307	0.282	0.310	0.673	3182
2000	4.16	0.310	0.287	0.313	0.691	3632

°C	CO <sub>2</sub>	SO <sub>2</sub>	H <sub>2</sub> S	NH <sub>3</sub>	AIR	°F
0	0.196	0.145	0.245	0.491	0.240	32
25	0.202	0.149	0.247	0.500	0.240	77
100	0.220	0.159	0.254	0.527	0.241	212
200	0.238	0.171	0.266	0.570	0.245	392
300	0.255	0.180	0.278	0.615	0.250	572
400	0.268	0.188	0.290	0.654	0.255	752
500	0.278	0.193	0.303	0.700	0.261	932
600	0.287	0.198	0.315	0.740	0.266	1112
700	0.294	0.201	0.328	0.778	0.271	1292
800	0.300	0.203	0.336	0.812	0.276	1472
900	0.305	0.205	0.344	0.844	0.280	1652
1000	0.309	0.207	0.352	0.873	0.283	1832
1250	0.316	0.209	0.367	0.932	0.290	2282
1500	0.322	0.212	0.379	0.979	0.295	2732
1750	0.325	0.213	0.387	1.013	0.299	3182
2000	0.328	0.214	0.393	1.040	0.303	3632

For heat content calculations use the mean specific heat

SPECIFIC HEAT  $c_p$  OF GASES AT 1 atm, kcal/kg. °C or Btu/lb. °F  
 1 kcal/kg. °C = 4186 J/kg. °C

D - 8



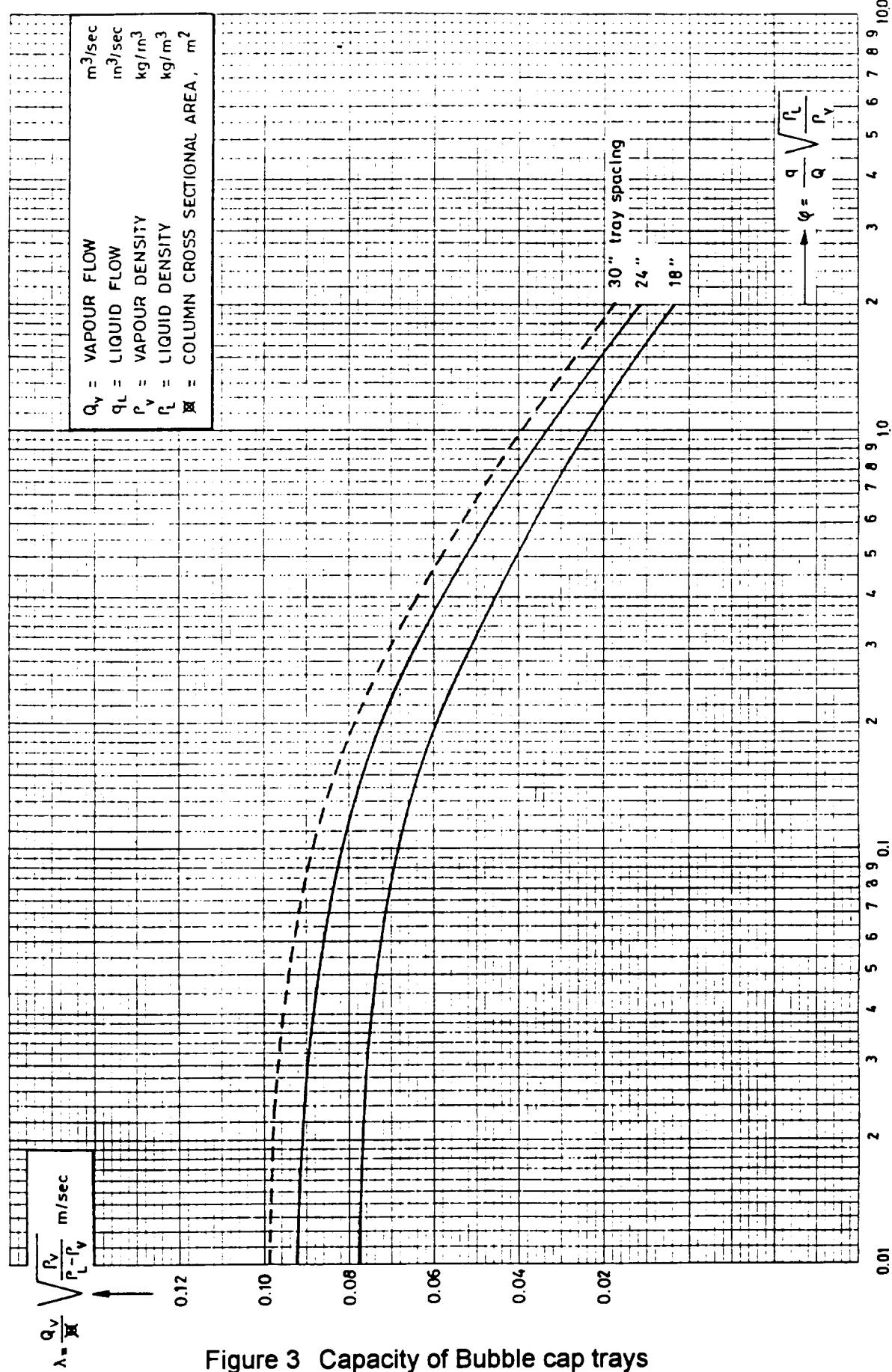
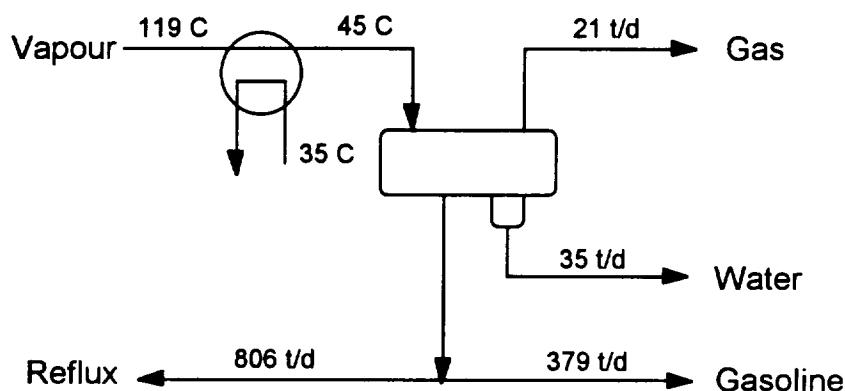


Figure 3 Capacity of Bubble cap trays

## Exercise 4      Overhead Condenser

The gas flow to fuel gas will increase to 73 t/sd for the revamp case, if we keep the current condenser. Part of this gas is potential Gasoline. Unfortunately the refinery cooling water system is at its maximum capacity. The cooling water supply to the condenser can not be increased. However when we install a new condenser (duty 154,000 tcal/d) we may be able to reduce the gas rate to 21 t/sd and recover 52 t/sd more Gasoline.

$$Q = 154,000 \text{ tcal/d}$$



### 4.1 What is the annual saving of the recovered Gasoline.

Unit availability 345 stream days per year.

The price differential between Gasoline and fuel gas is 20 \$/t.

### 4.2 Calculate the overall heat transfer coefficient K.

The condenser is a single shell exchanger with one tube pass.

The effective heat transfer surface area of current condenser  $A = 380 \text{ m}^2$ .

The heat duty is given by:  $Q = K * A * \text{LMTD}$

$$\text{LMTD} = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln \frac{T_1 - t_2}{T_2 - t_1}}$$

### 4.3 What is the installed cost of a new condenser.

Specific heat for water  $c_p = 1.0 \text{ tcal/t.C}$

Use the cost relation: Installed cost =  $100,000 + 300 * A^{1.2}$  (\$)



**13**

## **Economics on capital investment**

## Economics on capital investment

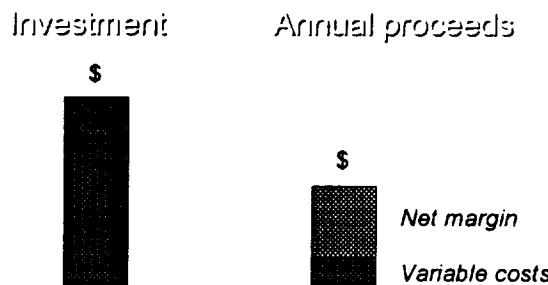
### Objectives

- Explain the discounted cash flow calculations
- Understand the various profitability indicators
- Be able to use capital charge to evaluate investment options
- Be better equipped to translate technical expertise into company profitability

Economics

## Economic evaluation

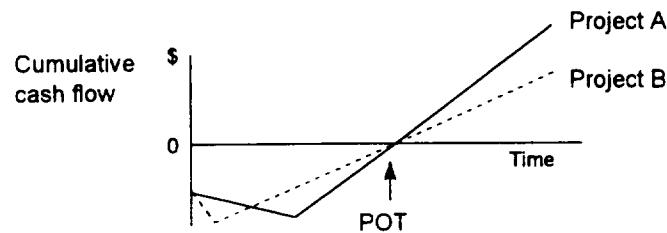
- How to appreciate a capital investment against the proceeds



Economics

## Pay-Out Time

The time necessary to recover the investment  
(plus half the building time)

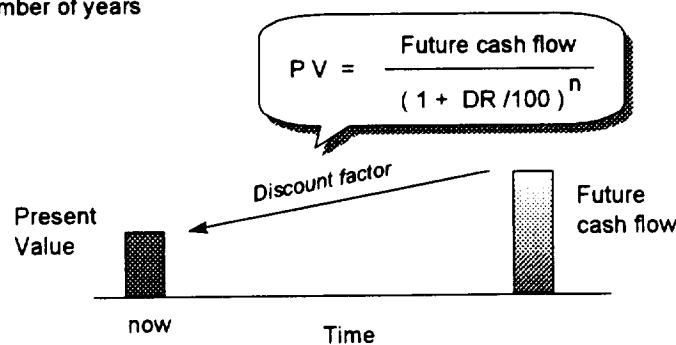


- Presents the exposure time money is at risk
- Used for small investments with quick return

Economics

## Discounted cash flow

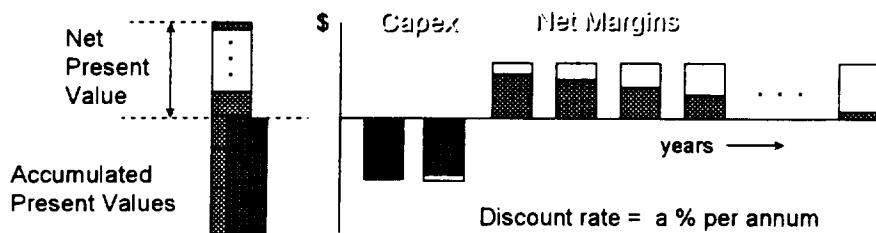
- Money in the future has a *lower* value
- Discount factor depends on
  - Discount Rate ( % / a )
  - Number of years



Economics

## Net Present Value

Net result of all future discounted cash flows

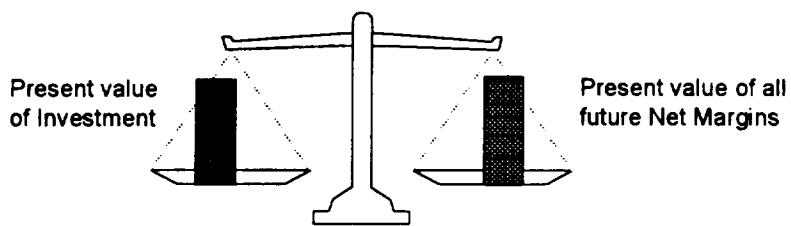


- Indicates the capacity to generate cash
- But has no relation to the investment

Economics

## Earning Power

EP = Discount Rate  $\longrightarrow$  when NPV = 0



- Measures the return on investment  
ie. the efficiency of using the capital

Economics

## How does an investment contribute to the business

- Economic & Fiscal environment
  - Taxation level
  - Asset depreciation
  - Inflation rate
- Project characteristics
  - Capex phasing
  - Economic life time of investment
  - Future cash flow profile
- Business targets
  - Real Term Earning Power (RTEP)
  - Return On Average Capital Employed (ROACE)

Economics

## Discounted cash flow

Year	Investment		Proceeds		
	Capex	Discounted	Net margin	Tax	Discounted
1	(0.8 C)	(0.741 C)			
2	(0.2 C)	(0.171 C)			
3					
10					
<b>Present Value</b>		<b>( 0.912 C )</b>			<b>2.893 P</b>

*Cash profile*

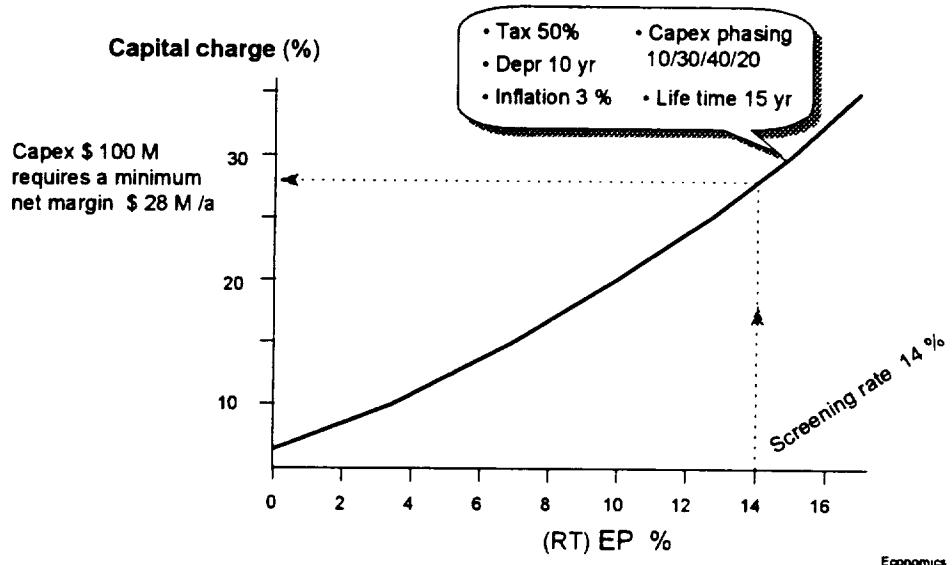
*Discount rate*

$$P = 0.315 C$$

**Capital charge = 31.5 %**

Economics

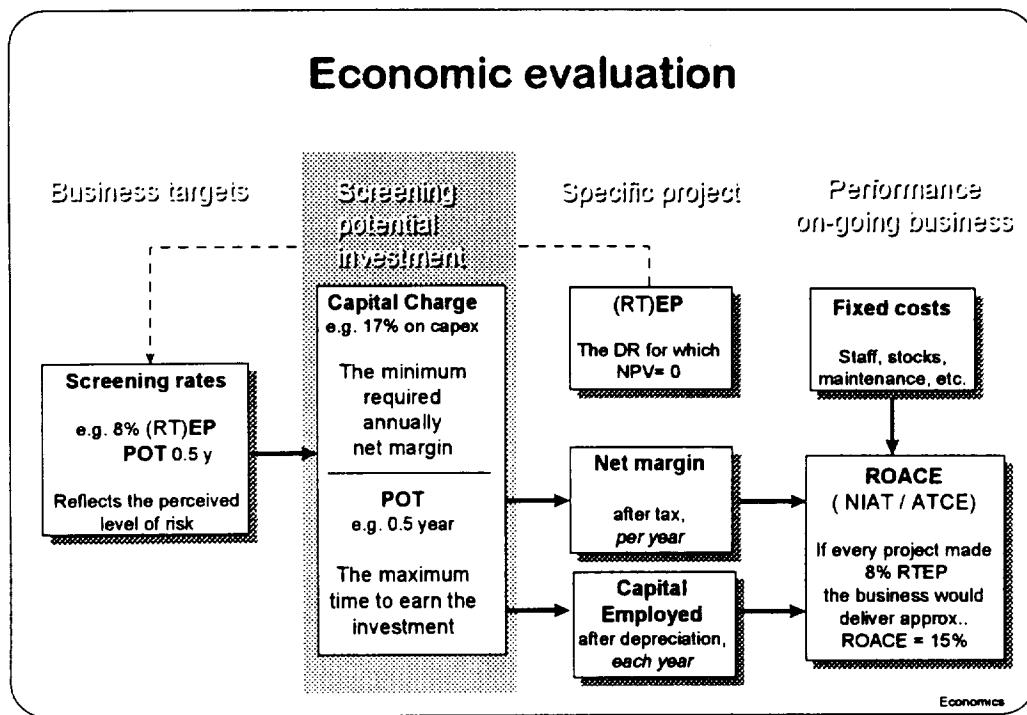
## Capital charge vs EP



## Capital charge

- Net Margin Before Tax (NMBT) that gives the required return-on-investment
- Can be viewed as the posted price to lease capital
- A convenient method to screen the viability of a potential investment
- Allows a quick ranking of investment options

## Economic evaluation



## Exercise

### **Screening projects**

1. Suppose that the estimated capital investment to recover an additional 30 t/sd distillate is \$ 800,000. The unit has 345 stream day per year and the price differential of the distillate is 20 \$/t. Capital charge for this project is 20 %.
  - Is this a viable investment.
  
2. With the installation of an additional heat exchanger we may be able to reduce the fuel consumption with 10 t/sd. The unit has 300 stream day per year and the value of the refinery fuel is \$ 80 per ton. Capital charge is 20%.
  - What is the maximum installed cost to justify an additional exchanger.

### **Ranking projects**

There are three options to increase the capacity of a unit.

The required capital investment and the NMBT of the projects are given below.

Project	Capex k\$	NMBT k\$/a
A	1,500	500
B	2,500	700
C	4,000	1,100

- I. Suppose that the capital charge is 20%, which project would you propose.
  
- II. What is the ranking of the projects if the capital charge is increased to 25%.



14

## Heat integration

## Pinch Technology

### Objectives:

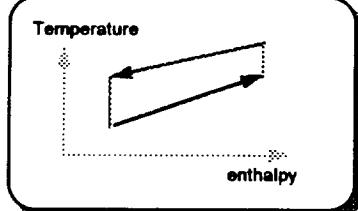
- The significance of the 'Pinch' in the heat balance of the process
- How is the capital investment of heat exchangers is optimized against the utility heating cost
- To set-up a heat integration scheme

## Heat exchangers

- counter current flow

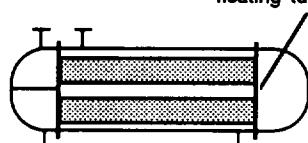
$$H = k * F * A * LMTD$$

$$F = f(TEMA, T) > 0.8$$



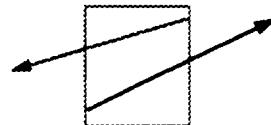
Tema type = BES

floating tube sheet



## Trade off between capex and fuel cost

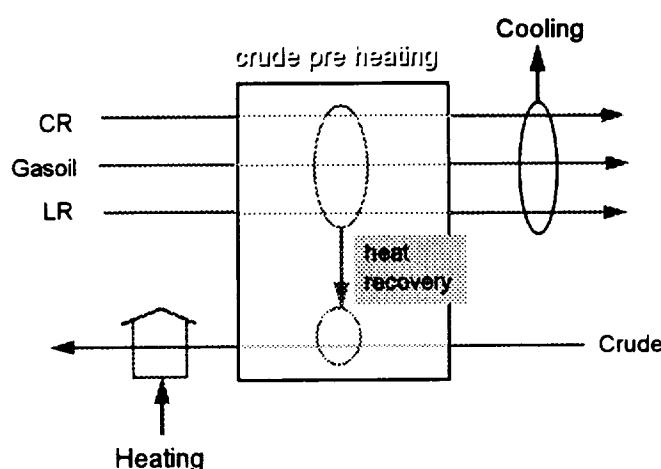
- small heat exchanger
- large LMTD
- low capital cost



- large heat exchanger
- more heat recovered
- less fuel cost



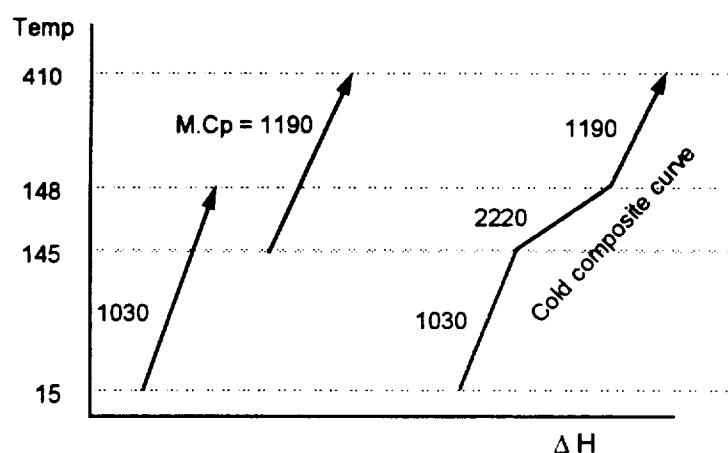
## The heat balance



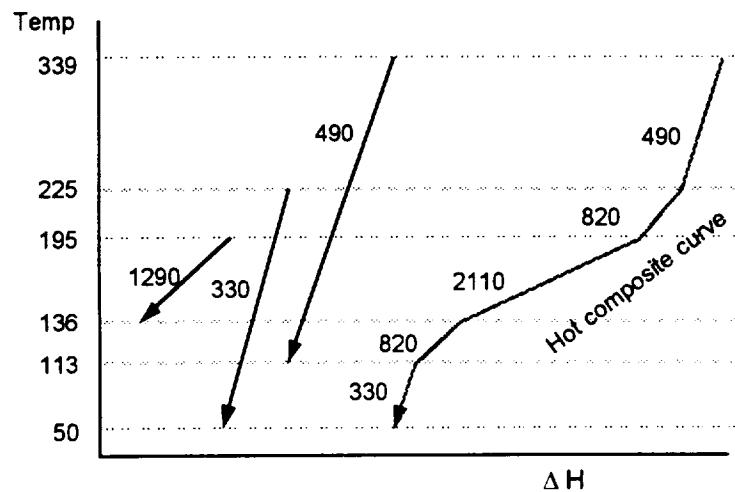
## Stream data

stream	start (C)	target (C)	duty ( tcal/d )	M.Cp ( tcal/d / C )
Hot streams	Circ reflux	195	136	76000
	Gasoil	225	50	57700
	LR	339	113	110700
Cold streams	Crude	15	148	137400
	Pft crude	142	410	318500

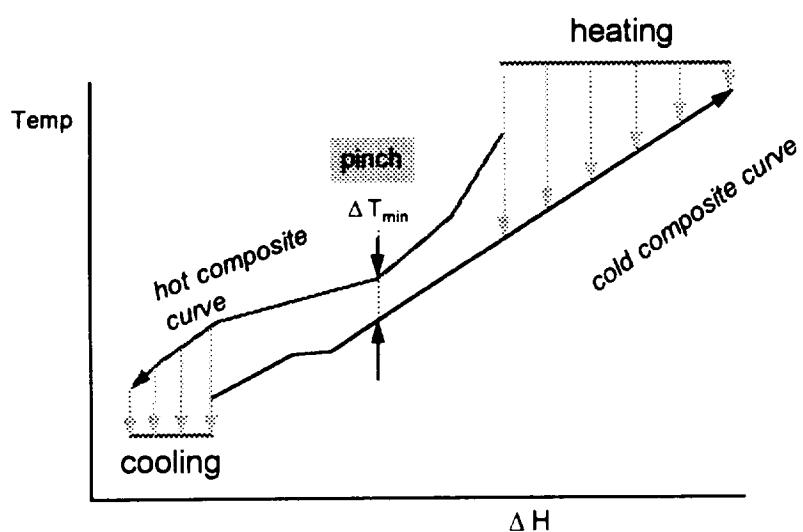
## Cold composite curve



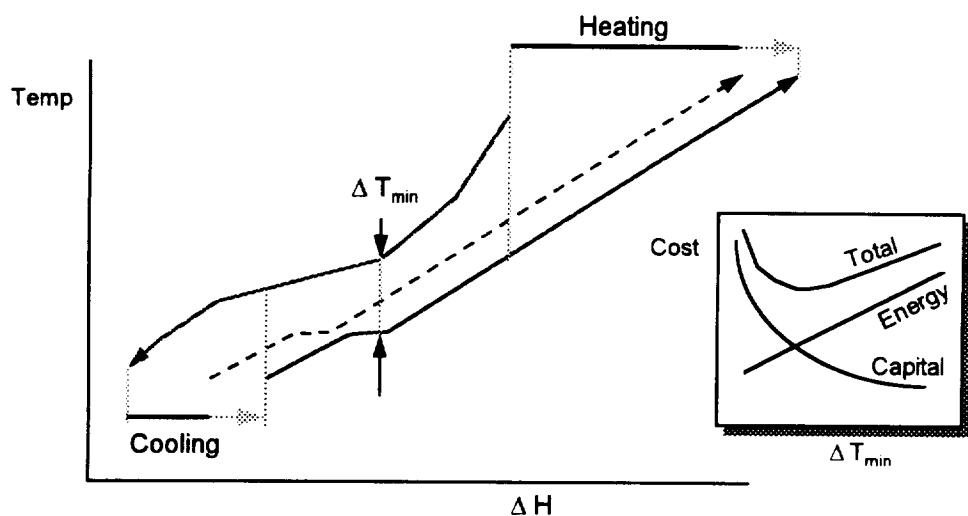
### Hot composite curve



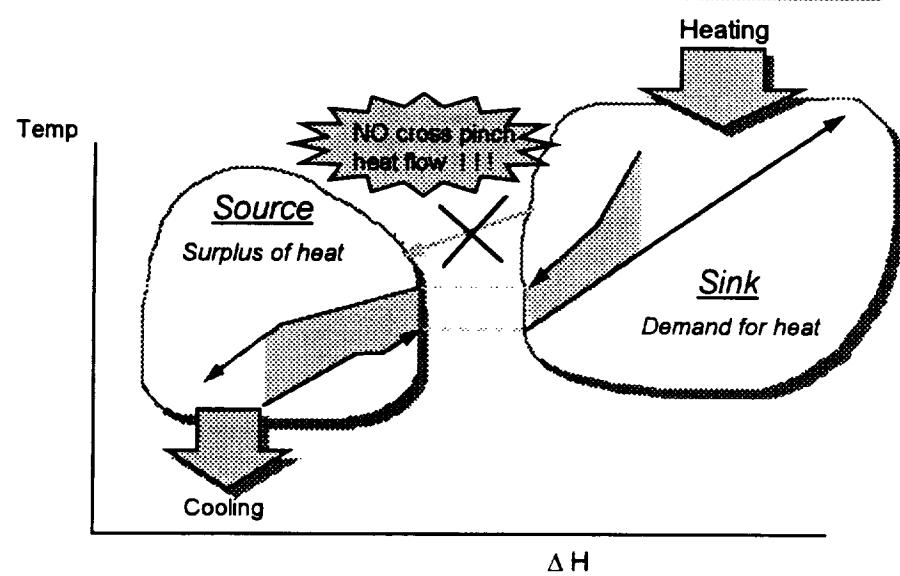
### Composite curves



## Optimum $\Delta T_{\min}$



## Pinch principle



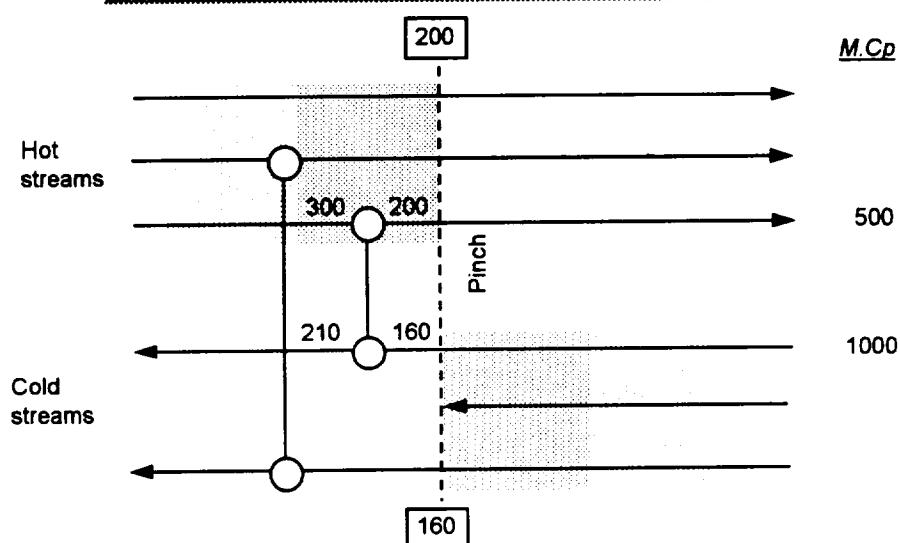
## PINCH rules

- The Pinch sets the optimal heat recovery
- Divides the problem into a **SOURCE** and **SINK**
- The regions are independent

Remember this . . .

- NO cooling to the **SINK** above the pinch
- NO heating to the **SOURCE** below the pinch
- NO heat recovery across the pinch

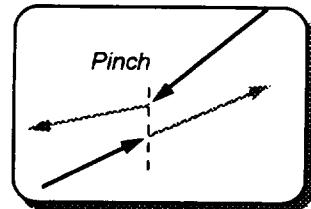
## Grid diagram



## At the pinch

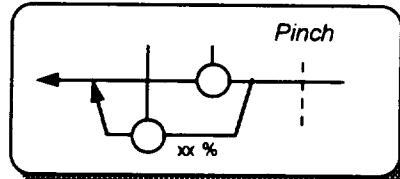
- Start with the big stream flowing **TO** the pinch

$$(M \cdot C_p)_{To} < (M \cdot C_p)_{From}$$



- Split a stream flowing from the pinch when

$$N_{To} > N_{From}$$



## Heat integration exercise

We have seen that the present furnace capacity is insufficient. Suppose that it is possible to increase the maximum duty with a minor revamp of the furnace. The question is what is the required furnace duty to process 2000 t/sd.

Here we can use the Pinch method. Based on an economic evaluation, the Pinch analysis provides us the minimum requirements of the network including the attainable furnace inlet temperature. However the technical feasibility of these results needs to be confirmed.

With a grid-diagram of the heat exchanger network we should demonstrate *how* it can be achieved.

Consider the following streams for heat integration:

### Cold streams

The feed is pre-heated in two steps:

- First it is heated to the target temperature of 148 C.
- Then the pre-flashed crude at 142 C is heated further to the maximum possible furnace inlet temperature.

### Hot streams

These are the streams that need to be cooled down:

- Long Residue rundown at 339 C. Target temperature 113 C.  
A lower temperature is not acceptable due to the high viscosity.
- Gasoil rundown at 225 C. Target temperature 50 C.
- Circulating reflux stream at 195 C.  
The required duty is 76,000 tcal/d. We are basically free to choose the flow rate. But for an efficient transfer of heat, the M.Cp should not be lower than the M.Cp of the cold streams.  
For a flow rate of 2000 t/d the resulting target temperature is 136 C.

For simplification of this exercise we will ignore the Kero and Gasoline rundowns. The heat contents of these streams are small anyway.

The stream data is summarised below:

	Crude	Prefl. Crude	CR	Gasoil	LR
flow (t/d)	2000	1927	2000	577	806
Tin (C)	15	142	195	225	339
Tout (C)	148	410 *	136	50	113
Duty (tcal/d)	137400	318500	76000	57700	110700
M.Cp (tcal/K)	1030	1190	1290	330	490

\*) Apparent temperature without vaporisation.

### Pinch analysis results

- Given the given specific cost on utility and equipment the optimum pinch is 30 C. The hot pinch temperature is at 195 C and the cold pinch temperature at 165 C.
- Utility cooling is not required to reach the target temperatures of the hot streams.
- At least six S/T heat exchangers are required for the network. The estimated total heat exchanger service area is 750 m<sup>2</sup>.
- The maximum possible furnace inlet temperature is 233 C. The resulting furnace duty is than 211,000 tcal/d.

### Exercise

Construct a heat exchanger network to confirm the pinch analysis results.

Use the attached grid-diagram to show the heat transfer from the hot to the cold streams. Calculate the temperatures of the intermediate streams between the heat exchangers.

The Pinch rules are:

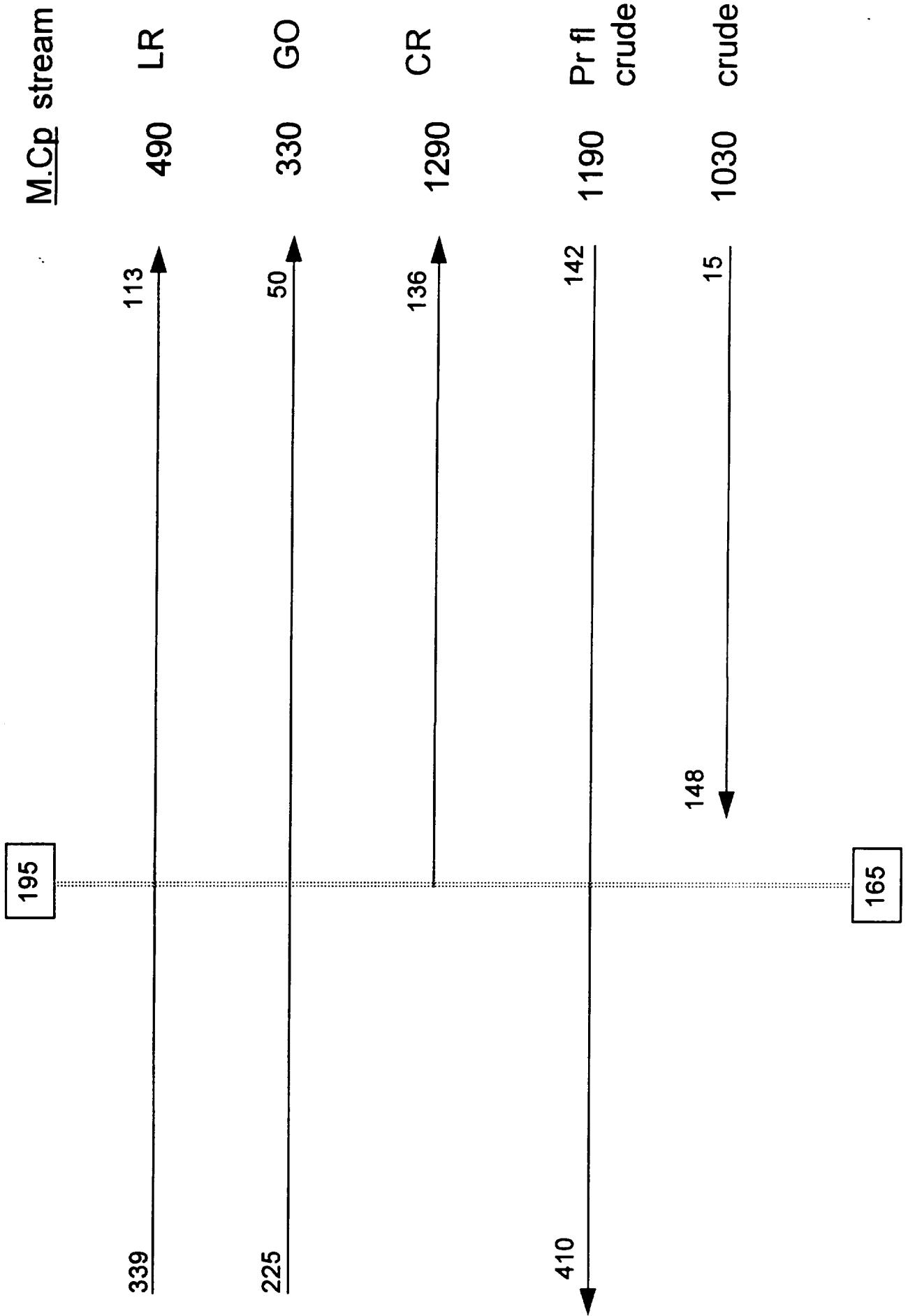
1     Start at the pinch !!!

The most difficult streams to match are those that flow towards the pinch, in particular the stream with the largest M.Cp.  
Don't transfer heat across the pinch.

2     Consider stream splitting:

When the number of streams flowing from the pinch is less than the number of streams towards the pinch make them equal by splitting a stream.

When the M.Cp of the steam towards the pinch is too large reduce it by splitting this stream.





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## **Optimisation of CDU operation**

# Optimisation of CDU operation

## 1 Introduction

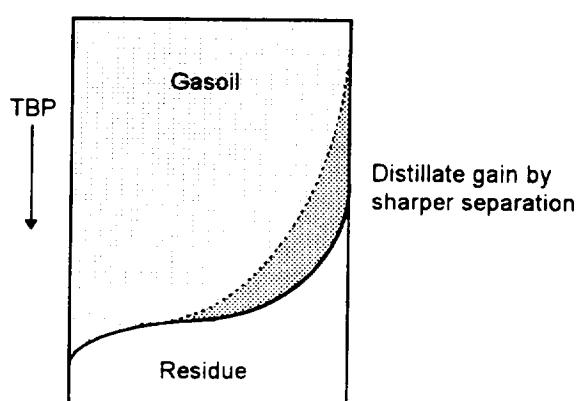
Crude oil distillation is the primary refinery process that produces the straight run middle distillate fractions, Kero and Gasoil. With an energy consumption of about 30% of the total refinery fuel bill, the operation of the primary distillation has a significant impact on the refinery operating cost. Therefore efficient operation of the CDU is of paramount importance for the economics of the refinery.

Efficient operation in this context means maximising the amount of distillates at a minimum cost on energy. This condition is achieved with an optimal combination of process variables such that the available spare capacity of the equipment is utilised to the ultimate limits. Starting with some basic principles on fractionation it is shown that the total distillate yield is essentially determined by a limited number of operating variables. Next we will see what impact this has on the fuel consumption and net margin. Finally some conclusions are drawn how to optimise the CDU.

## 2 Main Operating Variables

Maximising the yield of middle distillates is a matter of sharp separation between gasoil, the most heavy distillate fraction and residue. Since this separation is achieved in the lower sections of the main column the focus is primarily on the flash zone, the fractionation section above the column feed inlet and the residue stripping section in the bottom of the column. The separation performances of the other column sections to cut off the lighter distillate fractions have no direct effect on the total distillate yield.

Quality of gasoil such as cloud point is mainly determined by the length of its heavy tail. The effectiveness of rejecting the residue components depends on the reflux ratio in the fractionation section. More vapours from the lower column sections allow a higher reflux ratio. This improves the sharpness of separation. The figure below illustrates the net effect on distillate yield for a constant quality of gasoil.



Non-symmetric separation between gasoil and residue is very characteristic. The heavy gasoil tail is much shorter than the front end of the residue. Sharper fractionation, with a fixed gasoil quality, effects primarily the separation of the light components. Changes on the effective cut point (ECP) and separation index (SI) of the heavy gasoil tail are relatively small. The distillate gain arises mainly from a shorter front end of the residue.

Vapours to the fractionating section originate from two different sources. In the first place they are generated in the flash zone by flashing the feed at the column inlet. Most of these vapours pass through to the higher column sections and eventually end up as distillate product, but a minor part condenses in the lower circulating reflux (LCR) section to make reflux liquid. The observed feed overflash is shown in Figure 1. It is defined as the surplus of flashed vapours from the feed, including the crude pre-flash, which is in excess of the total amount of distillates. The feed overflash, normally given as a percentage on crude intake is an important operating variable.

Additional vapours to the fractionation section are generated in the stripping section where gasoil components are stripped off from the residue. Note that these vapours are *not* included in the overflash although they do contribute to the total distillate make.

The amount of reflux can be derived from the material balance around the fractionation section. It is equal to the sum of the overflash and the stripped vapours from the stripping section. Hence it is very well feasible to operate a column with a slightly negative overflash. In that case the stripped vapours are more than sufficient to make the required reflux with the LCR.

The effect of steam stripping is of a somewhat different nature than the feed flash. The injected steam promotes the evaporation of relative volatile gasoil components in the residue. Since the heat of vaporisation is drawn from the liquid, the temperature decreases proportionally with the amount of stripped vapours. This temperature effect reduces the stripping efficiency to recover gasoil.

### 3 Distillate yield

There are various options to increase the distillate yield. The operating variables which can be put to use are all related to the lower sections of the main column. However when the gasoil is maintained at a constant quality there are only two key variables: the feed overflash and residue stripping.

- **Feed overflash**

The combination of furnace outlet temperature and column pressure determines the overflash. Both variables can be adjusted to a certain extent during operation. The effect on distillate yield is given in Figure 2. Apparently it does not make too much a difference at what combination of temperature and pressure the overflash is achieved. The corresponding yield of distillates is about the same.

- **Residue stripping**

Figure 3 shows the effect of stripping. Note the similarity with pressure in Figure 2. The only difference between the two operating variables is the overflash; it is hardly effected by stripping.

Although stripping steam increases the distillate yield, the stripping efficiency decreases at higher steam rates. This is induced by the temperature drop across the stripping section. Since the equilibrium temperature at the top of the stripping section is difficult to measure we will consider the temperature drop between furnace outlet and residue bottom stream. This relation is shown in Figure 4. From an operational point of view this temperature drop is an excellent measure to assess the amount of potential gasoil that is left in the residue, almost independent of furnace outlet temperature.

#### 4 Energy Requirements

Crude oil fractionation requires direct heat of approximately 940 MJ per ton crude intake to heat up the crude from ambient temperature to about 350 C and to vaporise the distillate fractions. A considerable part of this heat is obtained by preheating the crude against the circulating reflux flows and products run down streams. Only the remaining part, about 55% of the required heat is supplied by the crude furnace.

From Figure 2 we see that a higher furnace temperature increases both the distillate yield and the overflash. More overflash requires more reflux in the fractionation section to keep the gasoil at constant quality. This additional reflux is generated with the LCR that exchanges heat against the crude. The subsequent increase of furnace inlet temperature partly compensates the fuel that is required to raise the furnace outlet temperature. When we take the thermal efficiency of the crude furnace at 90%, Figure 5 shows the net effect on marginal fuel consumption. It may happen that the LCR heat duty has reached its maximum capacity and the additional reflux has to be made with the bypass reflux from the section above. In that case there is no compensation on furnace inlet temperature by recovery of LCR heat. This has a clear effect on the marginal fuel consumption as shown by the broken lines.

From Figure 5 it is apparent that operating at the lowest possible pressure is always profitable, even when the LCR heat is not recovered. For example, a reduction of 0.5 bar provides a distillate gain ranging from 0.6 to more than 1 % woc for the same amount of fuel, depending on the degree of heat recovery.

The incentive for increasing the distillate yield with a higher furnace temperature depends on the refinery economics. It is obvious that it must have a positive contribution to the net margin; the payoff between distillate gain and incremental cost on fuel. Since the additional distillate is gasoil, the value of the distillate gain is in fact the marginal value of gasoil that can be derived from the price difference between gasoil and residue.

When we consider a fuel price (say \$100 per ton SRF) of twice the marginal value of gasoil (say \$50/t) we can plot lines of constant net margins in Figure 5. The

slope of these lines indicate that it is profitable indeed to aim for a high temperature to raise the distillate yield. A reduction of the furnace outlet temperature for the sake of fuel saving, has a negative effect on the net margin.

A comparison between Figures 2 and 3 reveals that both stripping steam rate and column pressure are equally effective to increase the distillate yield. The difference is that stripping steam adds up to the operating cost, identical to fuel. If we take a conversion factor of 15 ton steam per ton SRF, the actual consumption on equivalent fuel is shown in Figure 6. It is apparent that LCR heat recovery increases the stripping steam efficiency. At full recovery for example, a stripping steam rate of 3 %wor  $\equiv$  1.5 %woc requires about 0.035 %woc on fuel, much less than the conversion factor.

The net margin lines in Figure 6 indicate that residue stripping is always profitable. Initially stripping steam is very effective to increase the yield but the increment on distillate gain reduces as more steam is applied. For practical reasons the amount of stripping steam is normally limited to maximal 3 % wt. on residue.

## 5 Synergy with HVU

The objective of maximising the total distillate yield should be seen in a broader context. In complex refineries the residue from the CDU is processed in a High Vacuum Unit to recover a waxy distillate (WD) fraction and the remaining gasoil (VGO). These vacuum distillates should also be included to complete the survey of total distillates. For the purpose of maximising the distillate yield from crude the CDU and HVU can be considered as one integrated unit.

For the performance of the HVU there is a strong incentive to increase the gasoil content of the residue feed. The presence of these relatively light gasoil components has a positive effect on the feed flash. They cause a lifting effect in the flash zone that increases the yield of WD.

So leaving some extra gasoil in the residue will improve the total distillate yield on crude. The maximum amount of gasoil is of course limited by the capacity of the HVU.

There are two options to adapt the gasoil content of the residue: decreasing the furnace temperature or reducing the stripping steam rate. As we have seen in Figure 4 both are equally effective. The remaining question is which variable is the most suitable to reduce the operating cost. Figure 7 shows the effect on the marginal fuel consumption. Apparently the largest fuel saving can be achieved by decreasing the furnace temperature, leaving the stripping steam rate constant. Observing the differences at reduced temperatures it may look as if LCR heat recovery requires more fuel. This is not correct. What is shown is the effect on marginal fuel for the two cases. The *absolute* fuel consumption with heat recovery is always less than without heat recovery.

## 6 Conclusions

It is shown that for crude distillation it is far more profitable to optimise the distillate yield than try to reduce fuel consumption. Maximising the overflash and residue stripping have a positive effect on the net margin.

In practice it may be difficult to demonstrate the acquired distillate gain from plant readings because the relative effects are small, say in the order of 1 % woc.

However in terms of actual tons of additional gasoil the result can be significant.

With the aim of maximising the distillate yield, the column operation should be adjusted with the following variables, in decreasing order of priority:

- **Reduce the column pressure to its minimum.**

The normal limitation is the overhead condensing capacity. But there may be other constraining factors such as tray loading of the main column, recontacting compressor capacity or circulating reflux duties due to lower equilibrium temperatures.

A lower column pressure also shifts the vaporisation profile in the furnace and care should be taken not to have vapour in the convection bank with a risk of tube failure.

In any case it is essential for safety reasons to maintain a positive gauge pressure in the accumulator.

- **Increase the furnace to its maximum admissible outlet temperature.**

Note that at higher temperatures the furnace pressure drop increases and this may limit throughput. Also the effect on vaporisation profile may need consideration.

Do not compromise on temperature to save fuel.

If the furnace duty is limited, don't raise the inlet temperature by means of residue circulation. It will have a negative effect of the feed flash and therefore on distillate yield.

A lower furnace outlet temperature can be acceptable when the residue is processed in a HVU. A longer residue helps to increase the WD yield and the gasoil left in the residue is recovered as VGO anyway.

- **Set the stripping steam rate at 3 % wt. on residue.**

A higher steam is not practical since it may create corrosion and fouling problems in the top of the Fractionator. Moreover it boosts up sour water effluent.

- **Observe the temperature drop across the stripping section.**

It is an excellent tool to assess the amount of potential gasoil that is left in the residue.

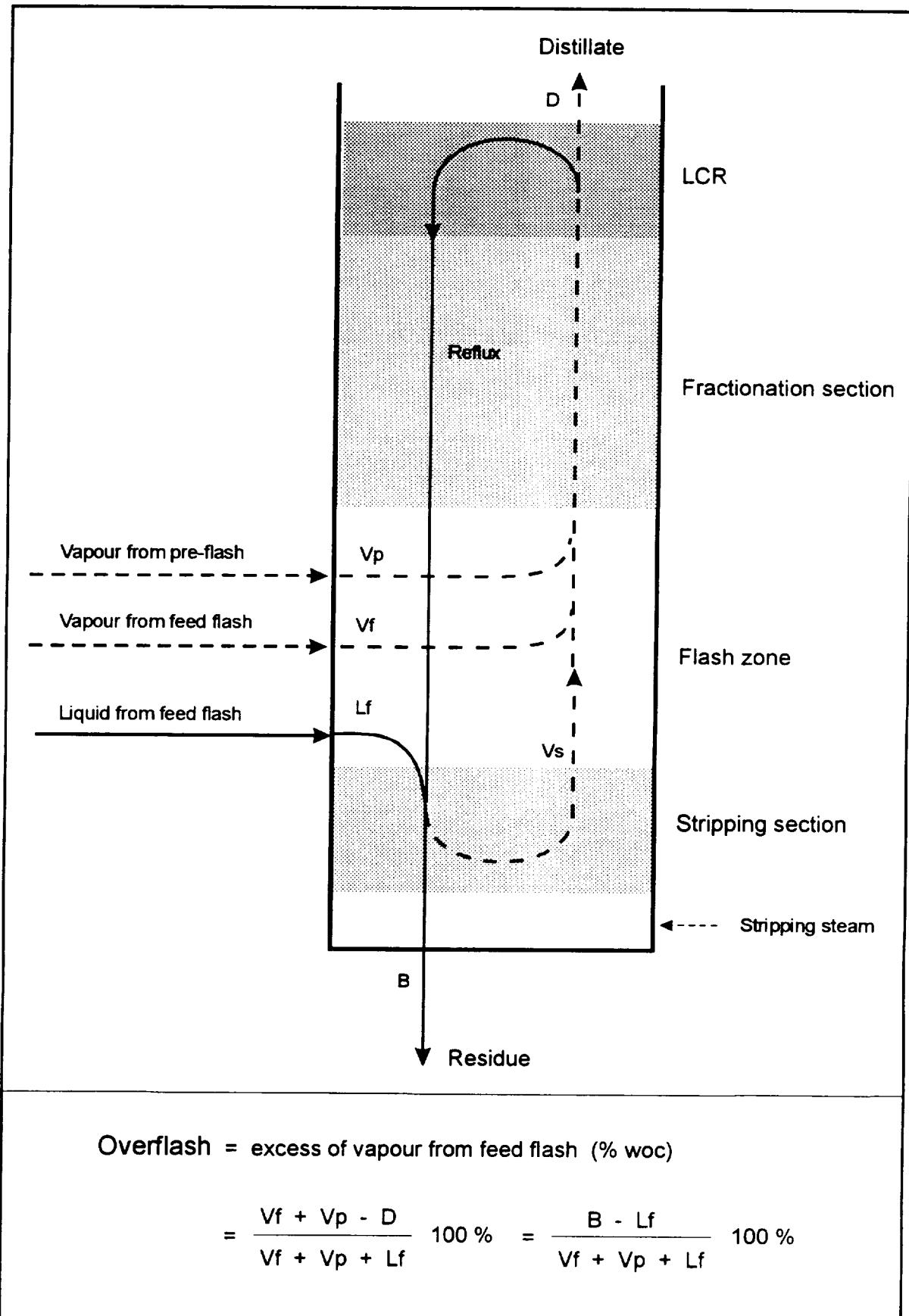


Figure 1. Fractionator bottom sections

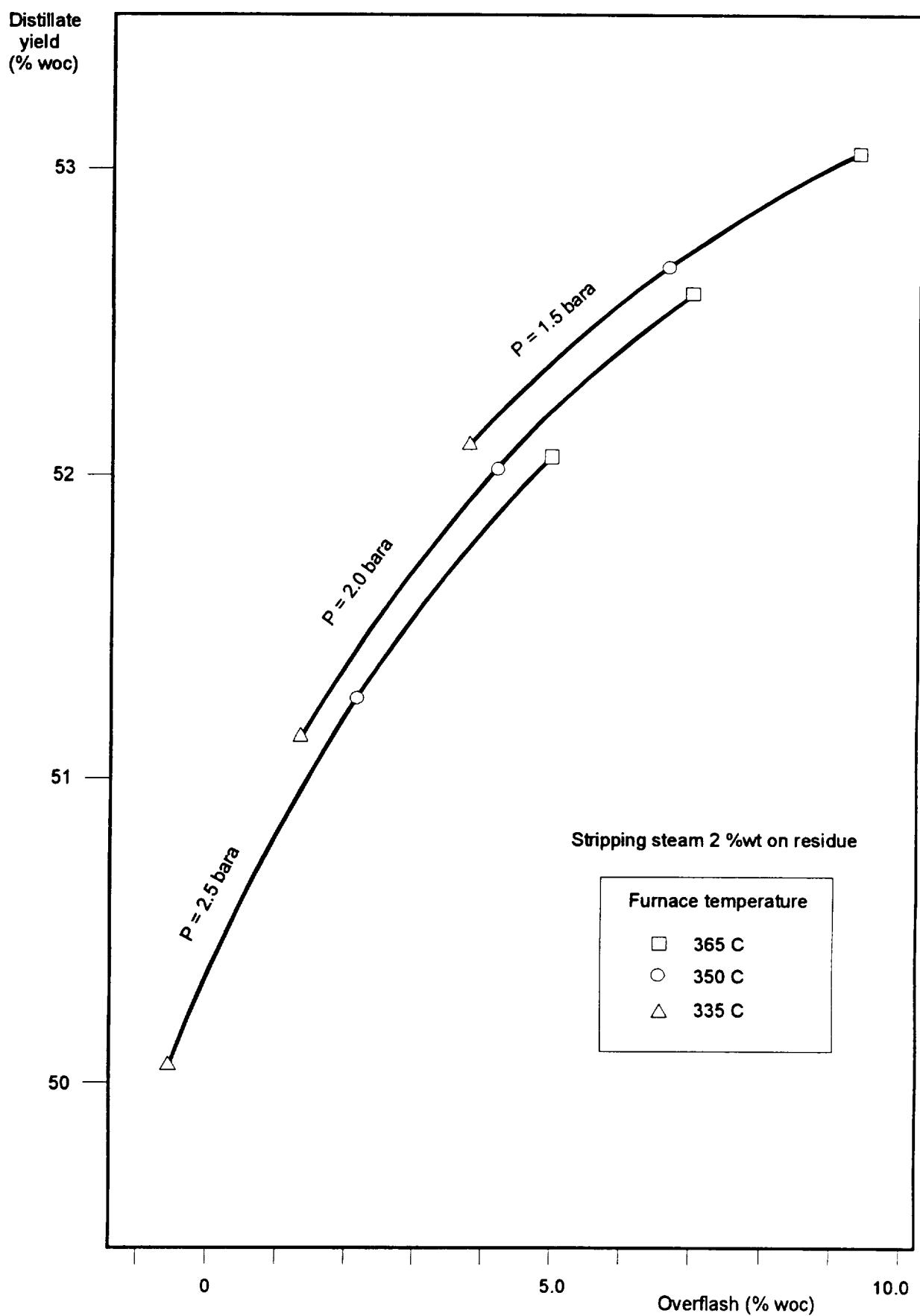


Figure 2 - Distillate yield as function of overflash

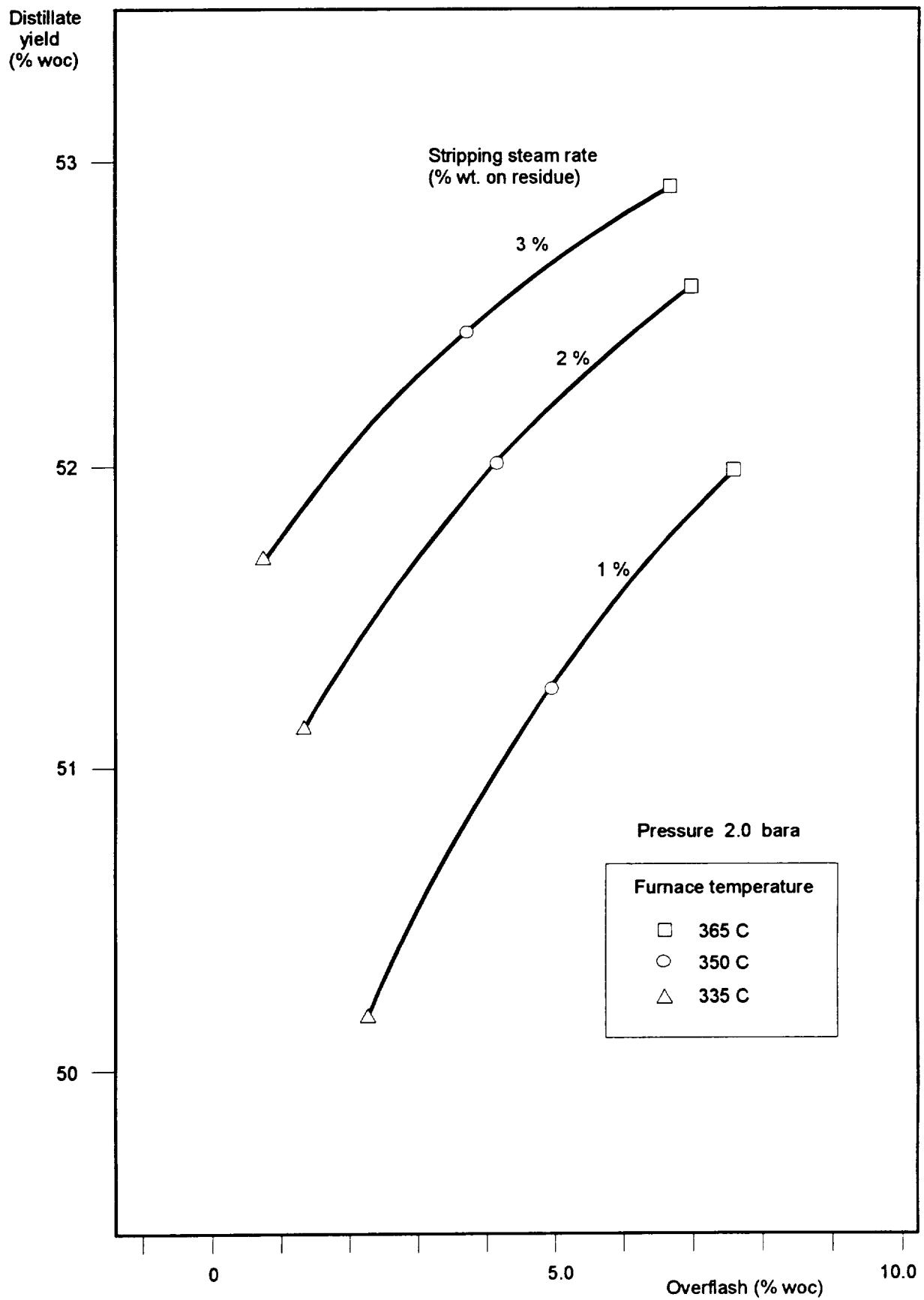


Figure 3 - Distillate yield as function of stripping steam

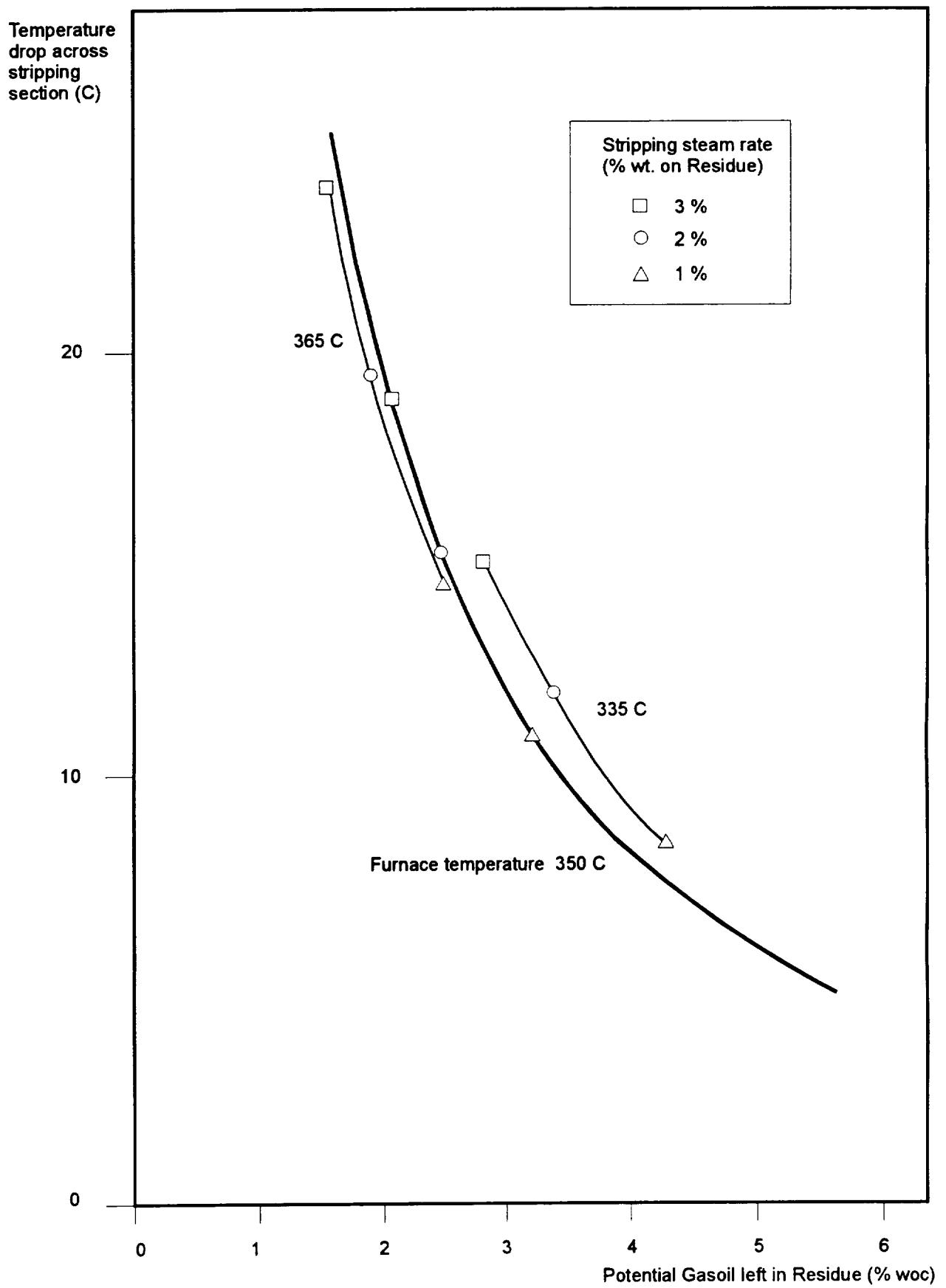


Figure 4 - Temperature drop stripping section

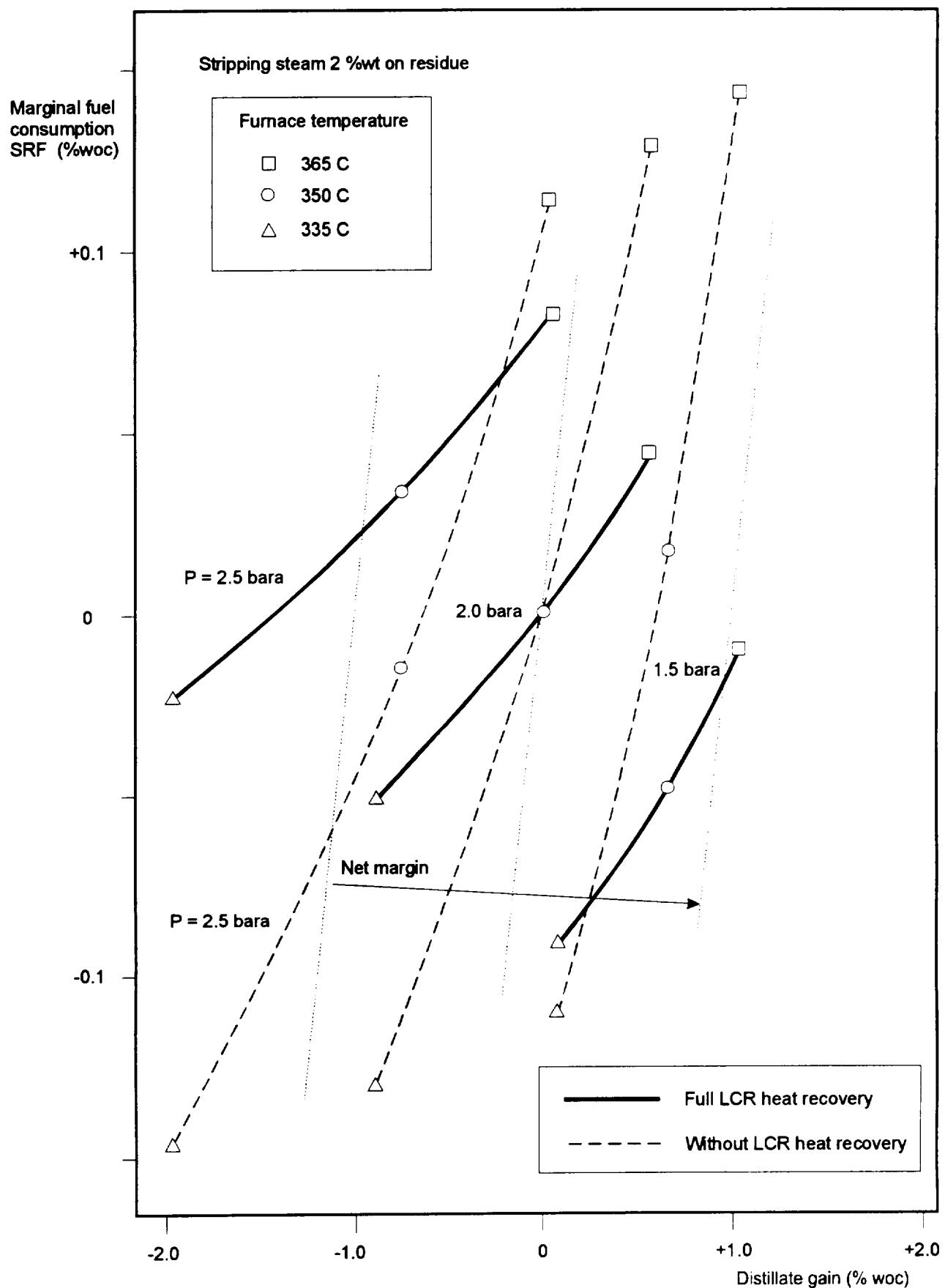


Figure 5 - Marginal fuel consumption crude furnace

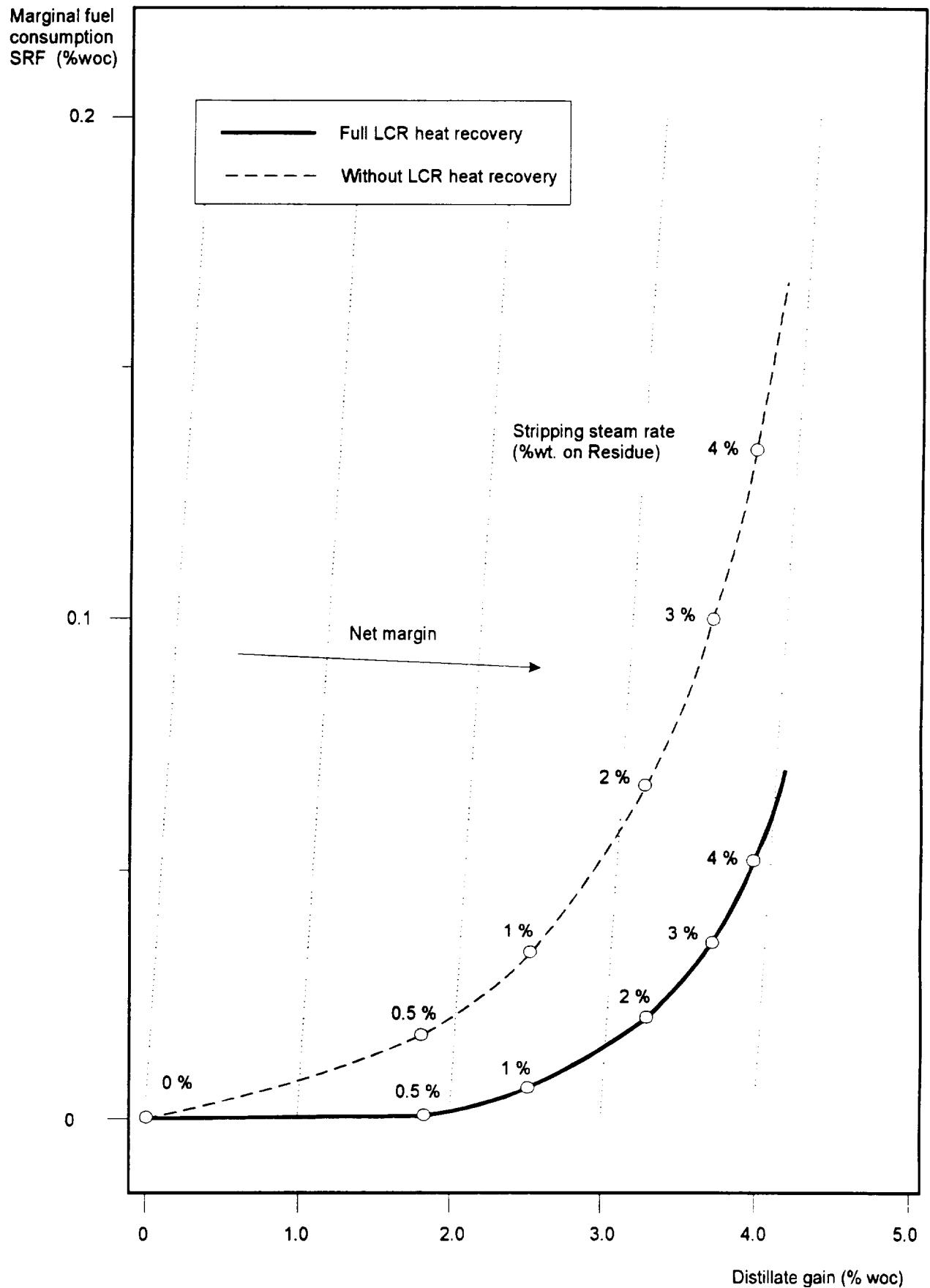


Figure 6 - Marginal fuel consumption stripping steam

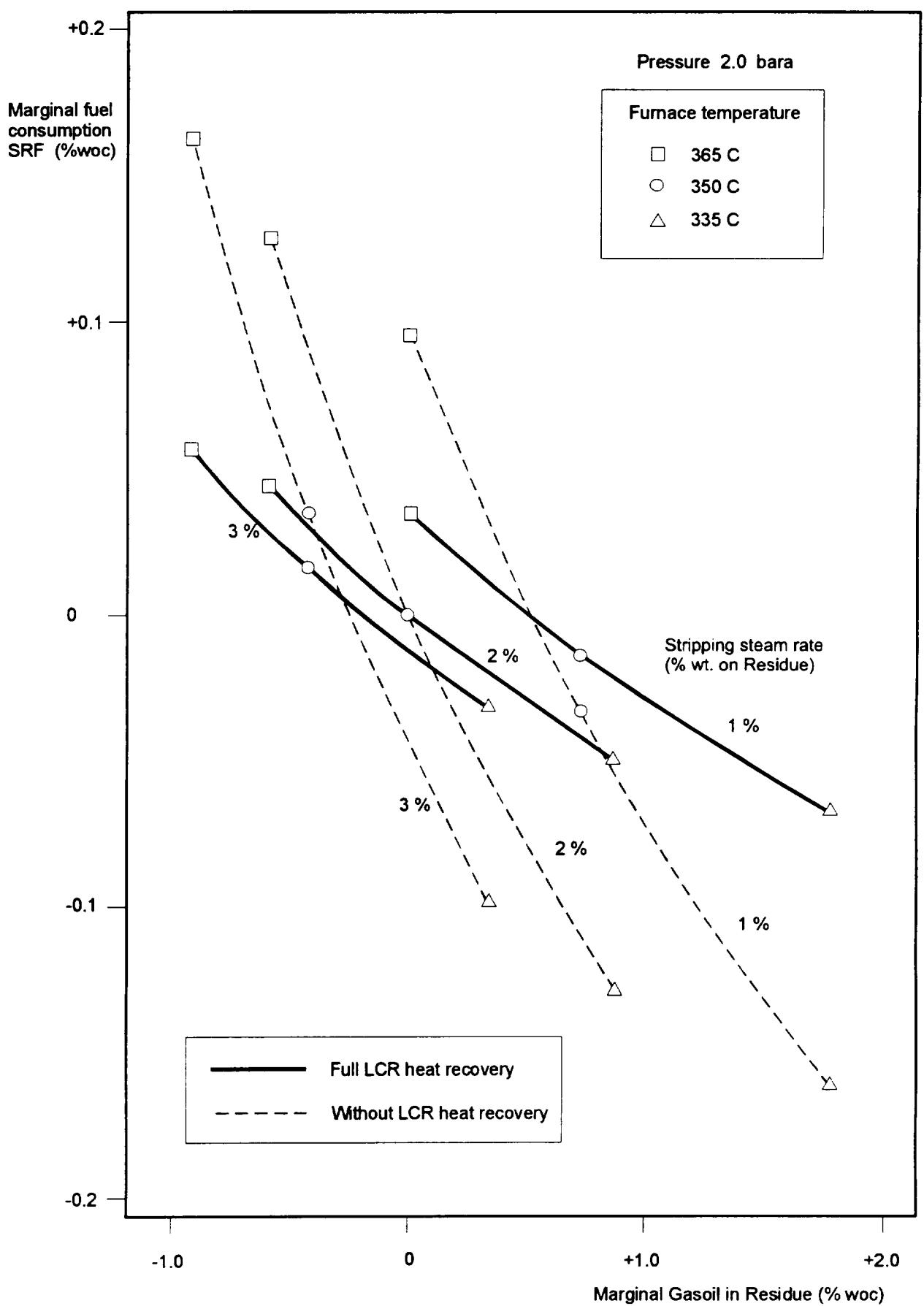


Figure 7 - Fuel saving when leaving Gasoil in Residue



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## **High Vacuum Unit**

# **High Vacuum Units**

## **1 Types of High Vacuum Unit**

### **1.1 Feed preparation vacuum unit**

Feed prep type of High Vacuum Units, also called vacuum flashers, are designed to recover a waxy distillate (WD) fraction from Long Residue (LR) feedstock. The WD is used as feedstock for downstream conversion units such as Cat Crackers and Hydrocrackers. The relative small amount of 360 C minus components contained in the LR may be recovered as a separate Vacuum Gasoil (VGO) fraction.

Although a sharp separation between the WD and the Short Residue (SR) bottom products is not required it is essential that the heavy asphaltenes components are rejected from the WD fraction. This is a key aspect of the process since the metal compounds of the asphaltenes may poison the catalysts in the conversion processes.

A simplified flowscheme of a feed prep. Vacuum Unit is shown in Figure 1. The unit comprises three major sections:

#### **Furnace**

The LR feed is in general directly routed hot from the crude distiller. When the LR is drawn from storage it is first preheated in a system of heat exchangers before it is fed to the furnace. In the furnace the LR is heated typically to temperatures in the order of 400°C depending on the type of unit. It flows usually via several “passes” within the furnace which recombine at the furnace outlet into a transfer line which will carry the partially vaporised mixture of SR and distillates into the column.

#### **Vacuum Column**

The partially vaporised mixture of SR and distillates usually enters the column via a vane type inlet device, a so-called “Schoepentoeter”. This type of device helps to efficiently separate the vapour and liquid streams and reduce entrainment of residual material into the rising vapours.

The liquid part of the feed (SR) is withdrawn via the bottom of the main column and cooled via a series of heat exchangers before going to storage or further processing. Some lower temperature short residue is recycled upstream of the bottoms (SR) pumps in order to reduce the suction temperature of these pumps and therefore reduce any tendency for cavitation.

The feed vapours rise through a so-called de-entrainment section. The function of the de-entrainment section is to catch and wash down any entrained residual material which if allowed to rise further and enter the upper parts of the column would contaminate the waxy distillate stream with adverse effects in the downstream conversion units e.g. poisoning of CCU catalyst. In the example shown in Figure 1 the de-entrainment section consists of a vane pack device and a packed bed (usually structured packing such as Mellapak) i.e. it is a two stage de-entrainment section, other units may only have a single stage section.

In order to work efficiently and to prevent fouling/coking up etc, the de-entrainment devices (vane pack, packed bed etc.) must be washed with at least a minimum

amount of reflux, or so-called wash oil. If this wash oil flow falls below a minimum flow (which is set by the type of device and size of the system) then it will only be a matter of time before serious fouling of the section occurs, leading to loss of de-entrainment efficiency, and a much increased pressure drop over this part of the column.

In the system shown in Figure 1 the vane pack is washed with the wash oil drawn off from below the packed bed above. The liquid from the bottom of the vane pack is drawn off from the column in order to monitor its flow and such that it may either be routed to the SR rundown system or recycled back to the furnace inlet. In some locations this material may also be used as flux oil in bitumen manufacture. The packed bed is washed with WD condensed in the section above. Note that in the example shown in Figure 1 both the vane pack and the packed bed are supplied with wash oil via spray distributor systems in order to give adequate liquid distribution and that filters are provided upstream the spray nozzles to prevent plugging of the (relatively) small spray nozzle openings.

Above the de-entrainment section there is one or more condensation sections to condense the WD. In older units these consisted of conventional trayed circulating reflux type systems. Today however the move is towards applying so-called spray sections in this service, due to their much lower pressure drop characteristics. As shown in Figure 1 liquid is drawn off from a draw-off tray which for a spray section will be a so-called "double decker" tray. It is circulated over a series of heat exchangers for heat removal, part is run down as waxy distillate product but most is recycled via a filter and spray distributor back to the column.

Some feed prep. units will only have WD condensation sections in the upper part of the column, while others as shown in Figure 1 will include facilities for VGO recovery. If VGO recovery is applied then a refluxed fractionation section between the WD and VGO will be used to enhance separation between these products. VGO will be used to enhance the separation between these products. In older units trays were used for this service, but structured packings are replacing trays in more modern units and in revamps because of their lower pressure drop and higher efficiency. Above the WD/VGO fractionation section a VGO condensation section will be included to condense the VGO and provide reflux for the fractionation section. A so-called "cold front" reflux is provided at the very top of the column to condense the lightest VGO components and reduce hydrocarbon carry over to the vacuum system. Again in modern units these condensation sections will consist of spray sections with "double decker" draw-off trays.

### Vacuum System

Any vapours that are not condensed within the column go overhead to the vacuum system. These vapours usually consist of cracked gassed which might originate from thermal cracking in the HVU furnace, air which might have leaked into the system and any purge gases from various instrument tappings. The purpose of the vacuum system is to maintain a low pressure in the column by removing these vapours.

Figure 1 shows a typical high vacuum system line up. The vapours from the column are initially sucked by two stage steam-driven ejector sets with condensers. Condensed water and hydrocarbons flow into the separating vessel. The non-

condensed vapours from the first ejector are sucked by a second ejector and those from the second ejector are sucked by a liquid ring pump (LRP).

Water is circulated over the LRP to provide the liquid ring seal and to cool the pump. The compressed vapours are routed to the waste gas disposal system, usually via a system of seal and knock out vessels.

## 1.2

### Luboil High Vacuum Units

These units are different from feed preparation type HVU's. Luboil vacuum units produce waxy distillate fractions with specific boiling ranges. They are used for luboil manufacturing feedstock. The product specifications are viscosity and volatility.

A typical simplified flowscheme of Group luboil HVU is given in Figure 1.2. Two columns are used to recover the products with one furnace for the long residue (LR) feed and one for the intermediate residue (IR).

Long residue is fed to the first furnace and heated typically to about 380 - 385°C. This temperature is somewhat lower than used in feed prep. units and is limited in order to prevent thermal cracking which in turn would lead to degradation of the quality of the luboil products.

Steam (so-called crossover steam) is injected into the furnace coils and the partially vaporised mixture is fed to the first vacuum column.

The first column separates Vacuum gasoil (VGO), Spindle oil (SPO) and Light machine oil (LMO). Two circulating refluxes are provided to condense the VGO, SPO and LMO and the first column "overflash". Fractionation sections for VGO/SPO SPO/LMO are provided. The system has side strippers to strip the SPO and LMO products and also various under or bypass refluxes will be provided to adjust the cut and fractionation between the products. In addition some so-called "buffer cuts" will be drawn off from the columns e.g. from column 1 circulating reflux to give additional flexibility and control over the main stream qualities. A steam stripping section is provided in the bottom of the column to recover VGO/SPO/LMO components from the intermediate residue.

The intermediate residue is reheated in the second furnace and fed to the second column which separates Light medium machine oil (LMMO) and Medium machine oil (MMO) from Black oil (BO) and Short residue (SR). Again crossover steam is used in the second furnace.

In the second column two condensation sections are provided to condense the LMMO, MMO and the "overflash" which gives rise to the Black oil. Again fractionation is catered for by providing stages and reflux between the BO and MMO draw-offs. A stream stripped section in the bottom of the second column is again for the recovery of light components from the short residue, and to prevent excessive slippage of lighter components to the downstream processing units (de-asphalting).

As luboil HVU's require many more column internals than feed prep. type units, the columns exhibit higher pressure drops and this sets the need for the large amounts of steam that are required to reach adequate hydrocarbon partial pressures at the

temperatures acceptable in the flash zones. The use of such quantities of steam however then also limits the column top pressure to some 50 - 60 mm Hg (see section 3.4).

## 2

### The Development of Feed Prep Units

Five types of HVU mark the development of this process to date. The simple flowschemes of these five types of units are given in Figures 2.1 to 2.5 and are described below:

- Type 1      The first generation of vacuum units were simple residue flashers. They had a internal heat exchangers/condenser system. About 12 of these units were built up to the early 1950's.
- Type 2      The second generation were wet units with overhead pre-condensers. The column had a large cyclone compartment with a tangential feed inlet and a separate residue steam stripper. The column sections were provided with four state of the art grid trays. Seven of these units were built in the late 1950's.
- Type 3      These units were built during the 1960's specifically for bitumen manufacture. They operate dry with one main condensing section. The separate run down of dirty wash oil was used as flux oil, a diluent to correct the bitumen for viscosity.
- Type 4      By the late 1970's the considerable advance on the in-house developed internals technology allowed the application of spray sections, trays, packing and schoepentoeter. The flashed distillate yields improved with achieved low flash zone pressures.
- Type 5      This is the latest type of HVU. The features of these units are discussed in more detail in section 4.

The main characteristics of the different types of units are summarised in Table 1. A comparison of the typical performance data of these HVU's given in Table 2 clearly shows the considerable improvements in distillate yields.

**Table 1** - Types of feed preparation unit in the Shell Group

Type	Operation	Column inlet	Wash-oil section	Column internals
1	Dry	Cyclone	Grid trays	Internal condenser
2	Wet	Cyclone	Grid trays	Grid trays
3	Dry	Vane	Grid trays	Grid trays
4	Dry	Vane	Packing	Valve trays, Spray sections
5	Semi-wet	Vane	Vane Pack packing	Spray sections

**Table 2** - Typical Feed Prep. HVU performance for Arabian Light LR

Type of unit		1	2	4	5
Top of column	(mm Hg abs.)	10	60	9	9
Flash zone	(mm Hg abs.) (°C)	50-60 395	80-90 395	25-30 395	15 420
Operating mode		Dry	Wet	Dry	Semi-wet
Cut point WD/SR * (°C)		505	525	545	585
WD 95% ASTM (°C)		500	530	550	590
Length of SR	(% woc.)	26	22	17	10.9

- \*) Note that these cut points are the “equilibrium flash vaporisation” temperatures. They are obtained with a laboratory-type flasher at vacuum condition. The measured equilibrium temperatures are converted to temperatures at atmospheric pressure with the Maxwell/Bonnell correlation. The “equilibrium flash vaporisation” cut-point should not be confused with the Effective Cut Point (ECP), as can be illustrated as follows: The ECP of Arabian Light LR (46.4 % woc.) corresponding to a SR length of 13 % woc. is 603°C, whereas the “equilibrium flash vaporisation” cut-point is only 570°C.

### 3 Operational aspects of Feed Prep units

#### 3.1 Flash zone temperature

The effect of raising the flash temperature at constant pressure for a given crude is shown in Figure 3.1. It may seem that the higher the flash zone the lower the short residue yield. Higher flash zone temperatures require higher furnace outlet temperatures with the associated aspects of higher crack gas yields and coke laydown in the furnace tubes. However, with steam/air coking and coil balancing facilities now more commonplace the trend is towards higher and higher flash zone temperatures to realise the distillate benefits. However it appears that a practical maximum cut point temperature of about 600°C exists due to vaporisation of metals containing material.

#### 3.2 Flash zone pressure

The effect of lowering the flash zone pressure at constant flash zone temperature is shown in Figure 3.2. The flash zone pressure is set by the vacuum system inlet pressures plus the pressure drop over the column internals.

The minimum vacuum system inlet pressure may be set by the first inter-cooler cooling water temperature as for moderate and tropical climates the cooling water temperature is 20 - 30°C. This results in an inter-cooler condensation temperature for water of 30 - 40°C which will result in a condensation pressure for water of 30 - 50 mm Hg. Given a typical ejector pressure ratio of about 5 this leads to a minimum ejector inlet pressure of some 5 to 10 mm Hg.

### **3.3 The length of the Long Residue feed**

The length of the long residue feedstock is an important process parameter. Roughly speaking, each ton of heavy gas oil left in the long residue in the CDU so enhance flashing at the same pressure and temperature that about 1.2 to 1.3 tons more vapour are produced, i.e. one ton of gas oil and 0.2 to 0.3 extra tons of waxy distillate (see Figures 3.1 and 3.2).

In order to recover the gas oil, trays or a packed section have to be present between the gas-oil and waxy-distillate condensation sections. It should be borne in mind that the separation in the CDU must be kept the same when lowering the HGO yield. Bad stripping or too small an overflash usually leads to an overloading of the ejector set, with a consequent loss in flash zone pressure.

### **3.4 Amount of Steam Injection**

The application of steam either as stripping steam or as crossover steam enhances vaporisation. Crossover steam, which is intimately mixed with the HC vapour/liquid mixture, acts in such a way that the total mixture can be regarded as being in equilibrium, i.e. on theoretical stage. If an efficiency of 20% is taken for steam-stripping on calming section trays, then the application of steam as stripping steam is more efficient if more than five actual stripping trays are installed. Steam is usually applied only when the steam can be condensed in a pre-condenser, which, in turn, results in higher flash zone temperatures, i.e. 50 - 60 mm Hg abs. instead of 15 - 25 as typically found in modern dry HVU's. However, depending on refinery conditions, the extra ejector drawing steam required during operation under deep flash conditions is sometimes unwarranted. The effects of steam are shown in Figures 3.2 and 3.3.

### **3.5 Process Optimisation**

Together with and in addition to the above aspects, the following measures can be taken in present day operation to optimise gasoil/flushed distillate recoveries:

- Keep the furnace outlet temperature as high as possible, preferably with a computerised coil-balancing system. Monitor the coil inlet pressures downstream of the control valves to check on coke deposition. A 1°C drop in temperature is about 0.2% WOC distillate.
- Check the pressure in the flash zone and top of the column on a daily basis and try to run at a minimum.
- Check for fouling of the pre-condenser, where applicable. If a second element is available, evaluate the cleaning cost as a distillate gain. Ascertain whether in-situ cleaning via the air-rumbling or Taprogge system is possible.
- Review crossover and/or stripping steam rates.
- Check the CDU operation regularly.
- If a gas-oil recovery system is available, evaluate and check the performance of the CDU and HVU in combination by slipping HGO ex the CDU to the HVU.
- Check the viscosity of the short residue regularly to detect any leakage from the wash oil section.

With regard to the flash zone pressure, it should be borne in mind that a reduction of pressure will result in a higher column loading which may lead to overloading and heavy entrainment. The higher loading is because of a lower density of the vapour and also because more vapour will flash off from the feed, although this second effect depends on the furnace. When the furnace is maintained at fixed duty the amount of vapour is almost constant.

## 4 Developments on HVU's

### 4.1 Introduction

In the early 1980's Group Refineries outside North America were processing about 33 million tons per annum of long residue to yield about 16 million tons of flashed distillate as feedstock for conversion units, mainly cat crackers.

From the second half of the 1970's it has become clear that there was a definite economic incentive for producing more flashed distillate feedstocks at the expense of residue even within the constraints of the then available cat cracker catalysts. This being part of the wider concept known as "cutting deeper into the barrel". Attention was therefore focused on flashing of long residue under more severe conditions than had previously applied.

### 4.2 Flash Zone Temperature

The design philosophy for feed preparation units up to and including type 4 in Table 1 was based on the concept of run lengths of at least two years, in view of possible coke deposition. At temperatures above 400°C, as existing in the tubes of radiant cells, in combination with residence times for the non-flashed part of the feed of the order of 10 - 20 seconds at these high temperatures, a certain degree of thermal cracking will always occur. Thermal cracking results in the formation and subsequent deposition of coke on the inside of furnace tubes, thus limiting the run length of the unit.

It had been found that a design value of 395°C for the flash zone temperature would, in the case of vertical SIPM-type furnaces, normally result in run lengths of at least two years. From experience with other type of furnaces, it became apparent that higher flash zone temperatures could be acceptable.

### 4.3 Flash Zone Pressure

The pressure in the flash zone is governed by the pressure created in the top of the column by the vacuum-pulling equipment and the pressure drop over the column internals located between top and flash zone. Starting with a minimum pressure in the top of the column, a pressure of 15 mm Hg abs in the flash zone can be reached with spray sections in combination with structured packing in the wash oil section.

In the selection of more severe flash zone conditions for the deep-flashing feed preparation unit, the degree of thermal cracking was considered to be the ruling yardstick. In this respect a new concept was developed for the temperature/residence time relationship of a hydrocarbon feed on its route through the tubes of an HVU furnace. Based on this concept, a furnace design philosophy was developed aiming at a minimum degree of thermal cracking by creating temperature and flow pressure profiles along the tubes such that the resulting flow regime was not of the

undesirable mist-flow type but of the plug-slug and annular-ring types in which less thermal cracking takes place.

## 4.4 Process Engineering

### 4.4.1 Furnace

Deep flashing requires a high furnace temperature. Figure 4.3 shows a flash curve for Arabian Light long residue. It can be seen that the change in vaporisation rate is considerable, viz. from 57.5% wt for 395°C/25 mm Hg abs. to 75% wt for 420°C/15 mm Hg abs. As a result of the larger vapour flow, the two-phase flow regime and its corresponding pressure drop, the inside-tube heat transfer mechanism and the furnace tube diameter are greatly influenced.

#### Flow Regime

In conventionally designed HVU furnaces, where vaporisation takes place mainly in the last furnace coils, a higher vaporisation rate means a shift of a prevailing flow regime towards the annular-mist type. To predict a certain flow pattern under given conditions, a flow map is usually used such as, for example, the one shown in Figures 4.4A and 4.4B. The transitions are plotted against the superficial velocities of liquid and gas. At low vaporisation rates the flow is of the bubble type. At intermediate gas velocities various types of slugging flow occur, especially at the lower liquid velocities (below 1 m/s). The two-phase flow becomes unstable if low superficial vapour velocities (1 - 3 m/s). The combined shaded areas in Figure 4.4 represent the unstable region at which pressure pulsations may interfere with the stable and harmless operation of the equipment.

Another unfavourable flow regime exists at higher vaporisation rates and subsequent velocities. In annular flow a thin liquid film covers the tube wall, while vapour containing entrained liquid droplets flows through the centre of the tube at a high velocity. At the increased gas velocity most of the liquid is carried by the gas in the form of fine droplets and "burn-out" of the tube wall may take place, with a subsequent failure of the heat transfer. The mechanism of "burn-out" may occur in the annular-mist flow regime, where the film of liquid is lost as a result of entrainment and vaporisation, the tube wall subsequently becoming dry.

On the flow chart the burn-out regime has been indicated. This is the regime that must be avoided at all times in designing high vacuum furnaces.

#### Heat Transfer

In the first part of the tube coil up to the point of vaporisation (the "bubble point") the heat transfer mechanism is mainly dictated by forced convection. From this bubble point to the end of the coil the heat transfer mechanism is called "vapour-induced forced convection" such as nucleate and film boiling.

In the case of bubble/froth patterns, the heat transfer coefficient is slightly higher than that for single-phase liquid flow. A maximum occurs just before the transition from annular flow to annular-mist flow. Figure 4.5 shows quantitatively how the heat transfer coefficient varies with the flow conditions. A considerable drop occurs in the mist/annular flow regime.

Towards the end of the coil vaporisation and temperature have increased, and if the flow regime is of the annular-mist type:  
the heat transfer over the internal film decreases (see Figure 4.5);  
heavy material concentrates in the liquid phase; this is usually material reactive with respect to coke formation;  
the residence time inside the liquid film increase owing to a viscosity increase.

Under these unfavourable conditions the liquid in the film undergoes chemical reactions which may result in the formation of solids such as coke. As coke is deposited on to the tube wall in the oil film, the temperature in the tube wall rises further and further until the furnace has to be shut down for de-coking when the maximum allowable temperature is reached. Hence, annular-mist flow regimes should be avoided in order to reduce the risk of coke formation.

To prevent the occurrence of undesirable flow regimes, a furnace design concept has been developed which is based on the suppression of vaporisation inside the furnace tubes. This suppression is obtained by creating a certain pressure drop over both the individual passes outside the furnace and the common transfer line.

Under the adiabatic conditions existing in the furnace outlet coils and transfer line, the enthalpy does not change and, as a consequence, the process bulk temperature at the furnace outlet will be higher than in the case of non-suppressed vaporisation. However, owing to the more favourable flow regime the film temperature is lower and the residence time of the still liquid feed fraction (at that film temperature) is shorter, both factors leading to reduced coke formation, which, in turn, will result in a longer run length before de-coking is required.

Apart from the above-mentioned process factors, coke formation is also affected by factors relating to equipment design which play a role by means, for example, of the heat flux, particularly the peak heat flux.

Various items have to be taken into account when establishing furnace design criteria aiming at an optimum between the residence time and the temperature of the liquid film, such as heat flux, geometry of the radiant cell and tube layout

Figure 4.6 shows a typical layout of the radiant cell of a HVU furnace. The main feature is a tube layout such that the feed flow passes first through tubes and along two refractory walls and, subsequently, through tubes in the middle which are fired from both sides. In this way four square radiant sub-cells are created in which the same peak heat fluxes are obtained for each of the four phases.

### Transfer Line

The transfer line between the furnace outlet and the vacuum column inlet is designed such that the pressure drop across it is enlarged, thereby suppressing the vaporisation inside the furnace. For the Pernis HV-8 there are two radiant cells (Figure 4.7) and the common transfer line is situated in between these two radiant cells at a level equivalent to the top of the cell roof. Each individual furnace outlet pass is connected to the common transfer line via a loop which is extended above the top of the radiant cell. The tube diameter of each individual loop increase from the outlet of the radiant cell to the common line. This line-up of the transfer line meets the requirements of both pressure drop and allowable force at the outlet nozzles of each furnace pass at the radiant cell roof.

#### **4.4.2 Feed Inlet Device**

The partly vaporised feed from the furnace has to be separated as efficiently as possible into a liquid stream which will leave the column via the bottom outlet and a vapour stream which will flow upwards. The standard feed inlet device applied in the past consists of a downwards-blowing schoepentoeter in combination with an M-cap deck for catching the entrained droplets of residual material.

Simulation tests revealed that the vortices in the bottom compartment resulting from the performance of the schoepentoeter are more favourable in terms of entrainment for the sideward-blowing schoepentoeter. However, it became clear that significant velocity differences occur. The presence of a draw-off tray with a typical pressure drop of 0.5 mm Hg will reduce these velocity differences to an acceptable level.

#### **De-entrainment section**

A wash oil section is usually installed immediately above the flash zone in order to catch the residual material still present in the rising vapour stream. From the pressure profiles over the column measured during test runs it is well known that the wash oil section could contribute considerably to the total pressure drop between the top of the column and the flash zone due to the fouling of the trays or packed bed in this section.

Although the sideward-blowing schoepentoeters have a good separation efficiency, a de-entrainment device is still necessary. A vane pack is the preferred device to catch the entrained droplets from the flash. The vane pack should be wetted with a relatively coarse spray to flush the entrained liquid droplets, thereby preventing coke formation. The sprayed liquid may be dirty wash oil originating from the wash oil section located just above the vane pack.

The total draw-off tray below the vane pack has two purposes:

- To correct the maldistribution of the rising vapour flow from the flash zone.
- To collect liquid from the vane pack and route it either to the furnace inlet or to the short residue rundown steam.

#### **4.4.3 Wash Oil Section**

The required wetting of the packed bed is obtained by spraying the top of the packing with a mixture consisting of roughly 50% (fresh) heavy FD and 50% recycled dirty wash oil, which is drawn off from the column by means of a total-draw-off tray located below the packed bed. The pressure drop over a packed bed of 900 mm height and a total-draw-off tray will be approximately 2 mm Hg.

As already mentioned, the condensing sections, both for the FD and for VGO, are for deep-flashing feed preparation units of the spray type having a typical pressure of less than 1 mm Hg. The condensing section of a deep-flashing feed preparation unit may consist of two or three parts equipped with sprays and double-decker total-draw-off trays. The choice of two or three sections is based on an optimisation study of the heat recovery for feed preheat and steam generation.

Owing to the large temperature differences which may exist between the different sections of a column, hot vapour may sometimes condense against the cold parts of a draw-off tray containing cold liquids, which is known as a "wild reflux". Several methods of thermal insulation have been devised and designed for. The method currently preferred is to install double-wall gutters.

If the feed contains a VGO fraction, the preferred fractionation equipment is a structured packing, which has a more efficient mass transfer and even a lower pressure drop than random packing. Trays applied in the past are unattractive. They have a relatively large pressure drop, approximately three to four times as high as structured packing for the same performance.

#### **4.4.4 Bottom Compartment**

The phenomenon of the after-cracking of liquid from the flash zone is considered to be one of the main causes of cavitation problems with vacuum column bottom pumps. To reduce the residence time of hot liquid, the SR is quench cooled in the bottom compartment with cold recycled short residue.

A radiation screen separates the flash zone and bottom compartment. This screen avoids the loss of potential distillates by condensation of hot flashed vapours against the relatively cold liquid surface.

#### **4.4.5 Column Overhead System**

Some thermal cracking will inevitably occur at the prevailing conditions at the furnace. This results in a substantial amount of cracked gas which is removed from the column together with some air leaking into the vacuum system.

Disposal of the vacuum waste gases is preferably by combustion in the radiant cell of the furnace. In the event of a furnace trip the cracked gas production will rapidly diminish and routing the waste gases to safe location will be less of a problem.

However, since the waste gas contains H<sub>2</sub>S burning of the untreated gas adds up to the refinery sulphur emission.

The vacuum-pulling equipment usually consists of three stages, see Figure 2.5. The first stages are two steam-jet ejectors in series. The third stage is a liquid ring pump using condensate from the ejectors as sealing medium.

### **5 Feed Prep. HVU - Revamp opportunities**

Over the years many vacuum units have seen some revamp activity. The objectives were in general to increase the amount of waxy distillate feedstock for the conversion processes.

#### **5.1 Column pressure drop**

The conversion of trays in the condensing sections to spray zones will give lower pressure drops. However, with spray zones the draw-off temperatures will be slightly lower. The consequences of this for the heat removal capacities has to be checked. Spider plugging/blockage might occur in hot environments when liquid velocities are low. It is recommended to keep velocities above a minimum of 0.5 - 0.75 m/sec by tapering the side branches of spiders. Dead end should be kept to a minimum. In fouling service special devices are applied.

Figure 5.1 shows that structured packing has a far lower pressure than trays or dumped packings. Therefore the replacement of trays or dumped packings (in fractionation and/or de-entrainment sections) by structured packings will result in lower flash zone pressures.

A low pressure drop in the column can have a disadvantage which may influence the performance of packing. Vapours not evenly distributed will, when approaching a structured packing section, not be damped out, i.e. the maldistribution remains nearly as it is and, hence, higher than expected/allowable velocities may occur. It is, therefore, essential to have a good vapour distribution, which can be achieved by installing a double-decker draw-off tray below the packing. The omission of a draw-off tray or any other pressure drop device between, for instance, a vane-type feed inlet and structured packing will almost certainly lead to maloperation of the packed section.

Liquid distribution is also important, especially when large numbers of theoretical stages have to be realised. This is seldom the case in HVU operation, but care should be taken, as, otherwise, unpleasant surprises will occur. In nearly all cases HVU design liquid to a packed section is supplied via a spray system. In order to prevent/minimise plugging and coking-up of nozzles, the minimum number of spray nozzles should be applied, preferably with the largest possible clearance for liquid passage.

## 5.2 Distillate Capacity

Increasing the amount of distillates implies a higher vapour loading in the column. This effects particularly the wash oil section, the most heavily loaded section of the column. When the flash zone pressure decreases, the vapour velocity increases with the ratio of  $p_{old}/p_{new}$  for the same amount of distillates. With additional distillate that will be produced because a lower flash zone pressure the vapour velocity will increase even further.

For vacuum columns structured packing is superior to trays or random packing. As shown in Figure 5.2 the maximum capacity at lower flow parameters region is higher.

## 5.3 Better de-entrainment of Wash Oil section

There are several options to reduce the entrainment of residual material into the distillate sections. In older units, an M-cap deck with grid or calming section trays are installed in combination with a de-mister mat. Under the prevailing wash oil section conditions, i.e. high vapour loads and very low liquid loads, trays suffer from considerable entrainment.

The typical requirements for various type of internals are summarised below:

Variable	Trays	Random Packing	Structured Packing	Spray section
max. vapour load, $\lambda$ (m/s)	0.06	0.085	0.11	0.04
min. wetting rate (mm/s)	0.3	0.3	0.1	0.3
column height (mm)	2000	800	420	1500

From these data it is obvious that structured packing is performing well as compared to trays and random packing. It has a higher capacity and requires less wetting which reduces the loss of potential distillate. A second advantage of the is that no de-mister mat is required. The entrainment with structured packing is low for  $\lambda < 0.11$  m/sec.

When trays are applied a de-mister mat is required. The drawback is that it may coke-up after a few months of operation.

#### **5.4 Waxy Distillate Quality**

WD feedstock for the HCU normally has a specification in terms of 95% ASTM or TBP cut point with SR. So the distillate yield can be increased with sharp separation. Increasing the number of theoretical stages in the wash oil section from one to four, generates approximately 1.5% woc. of additional feedstock with the same specification.

In general, it can be said that a packing height of 2 m Mellapak 250Y is equivalent to four theoretical stages, provided that both vapour and liquid distribution is adequate. For high vapour load a less denser packing like 125Y is required. However, this type of packing demands more column space for the same number of theoretical stages.

#### **5.5 Vacuum Gas Oil separation**

When WD is used as feedstock for the HCU a good separation of vacuum gas oil (VGO) is of prime importance. Many older units have trays in the VGO fractionation section. Replacement of these trays with structured packing will yield the benefits of their lower pressure drop and higher efficiency. Calculations show that usually six theoretical stages are sufficient for a very good separation.

### **6. Luboil HVU - Revamp opportunities**

#### **6.1 Pernis-type modification**

A conventional luboil HVU consists of two columns equipped with trays and trayed circulating reflux sections. In Pernis the two column luboil HVU has been revamped, retaining the two column concept but replacing trays in the VGO/SPO, LMO/MMO and MMO/BO fractionation sections with structured packing. Figure 5.3 shows a simplified flowscheme of the modifications.

Application of the packing can give an increased yield of luboils at constant quality due to the enhanced efficiency of packing trays, in this case typically 3% wt more luboil distillates on feed can be recovered by increased separation between vacuum gasoil/spindle oil and between medium machine oil/black oil.

Better energy efficiency can also be obtained due to the lower flash zone pressures resulting from the packing application (some 20- 25 mm Hg lower pressure drop). Lower flash zone pressure requires lower flash zone temperatures (lower furnace duties for equivalent performance).

#### **6.2 Further revamp options**

The shortcomings of a conventional type two-column vacuum unit can be summarised as:

- The residue is stripped twice. This requires relative large amounts of steam.

- High entrainment and poor separation in the trayed sections above the flash zone initiated by a very low liquid load and high vapour velocities.
- Pre-condensers of the second column are often blocked due to wax depositing.
- Yield losses inherent to the production on the LMMO. This fraction is not used for luboil manufacture.

Some of the disadvantages of the conventional system can be alleviated by implementation of structured packing, resulting also in a significant increase in luboil distillate upgrading, and lower overall energy consumption.

Implementation of the modifications as discussed in 6.1 can result in about 5% decrease in energy consumption accompanied by additional 8% gain in luboil distillates. Due to the higher separation power of the structured packing substantial amounts of spindle oil are recovered from the vacuum gas oil without increasing its volatility.

Processing more long residue to increase production of luboil distillates, however, is not possible since the remaining trays are constraining. This can be dealt with either by operating at higher pressures in the flash zone (at the cost of more energy) or by replacement of all try by the structured packing.

Further decrease in the pressure drop can be achieved by replacement of the trayed circulating refluxes with spray sections. In this way throughput can be increased by making use of both high capacity of Mellapak and operation at higher pressures at a minimum energy consumption.

The packed column philosophy would ultimately lead to a one column concept, either on wet or dry operation. An additional improvement in terms of energy consumption and yields is possible due to further decrease of the pressure drop and better heat integration as compared to the conventional HVU. Furthermore higher capacity and fractionation power of the structured packing enable an increase in luboil distillate recovery: 9 - 10% more luboil distillate is produced with 10 - 12% less direct fuel and 75% less steam.

The drastic reduction in steam consumption results from re-using the steam from the first column, strippers and ejectors in the second column which is now used as a simple short residue stripper to recover luboil components.

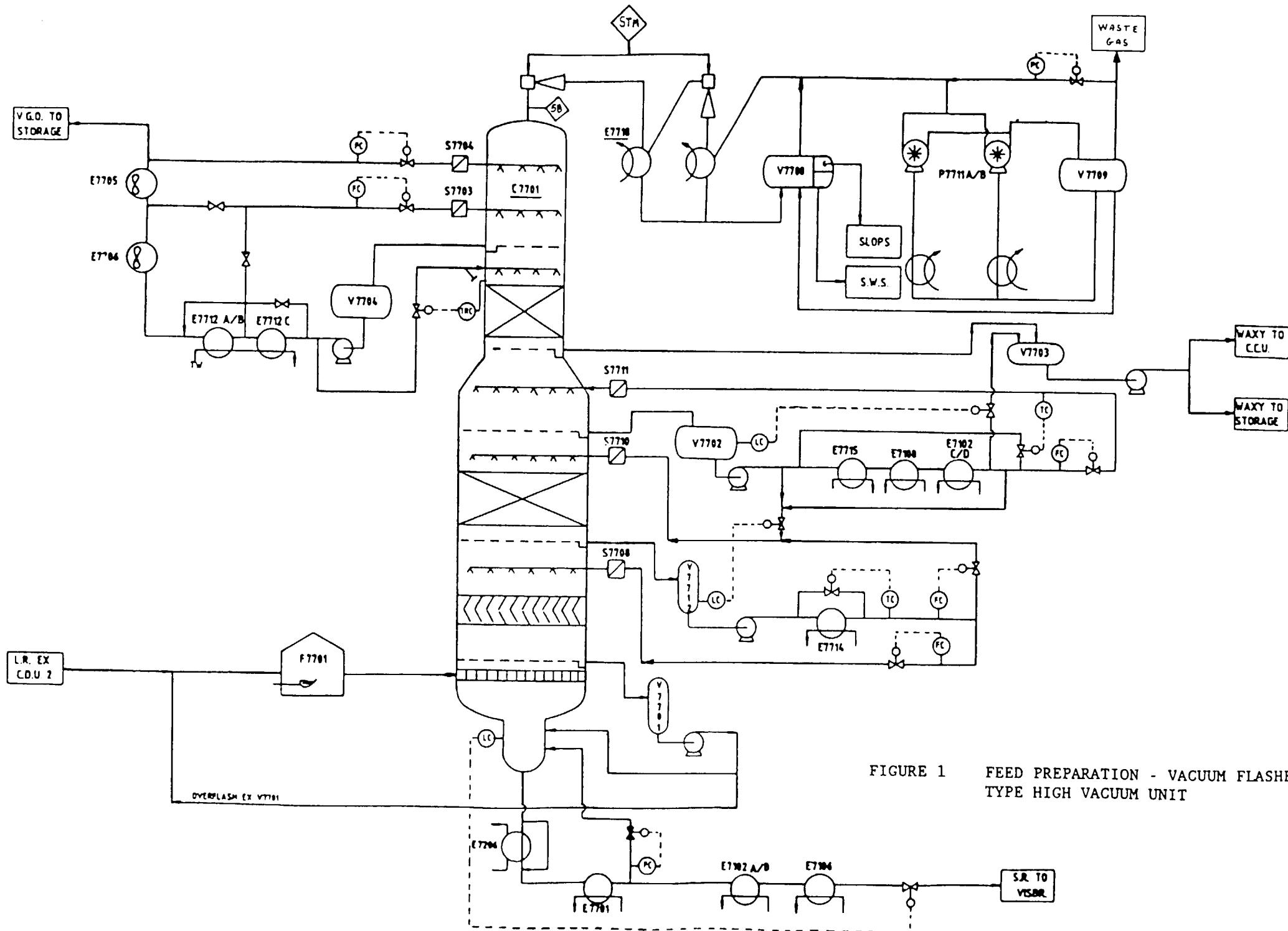


FIGURE 1 FEED PREPARATION - VACUUM FLASHER  
TYPE HIGH VACUUM UNIT

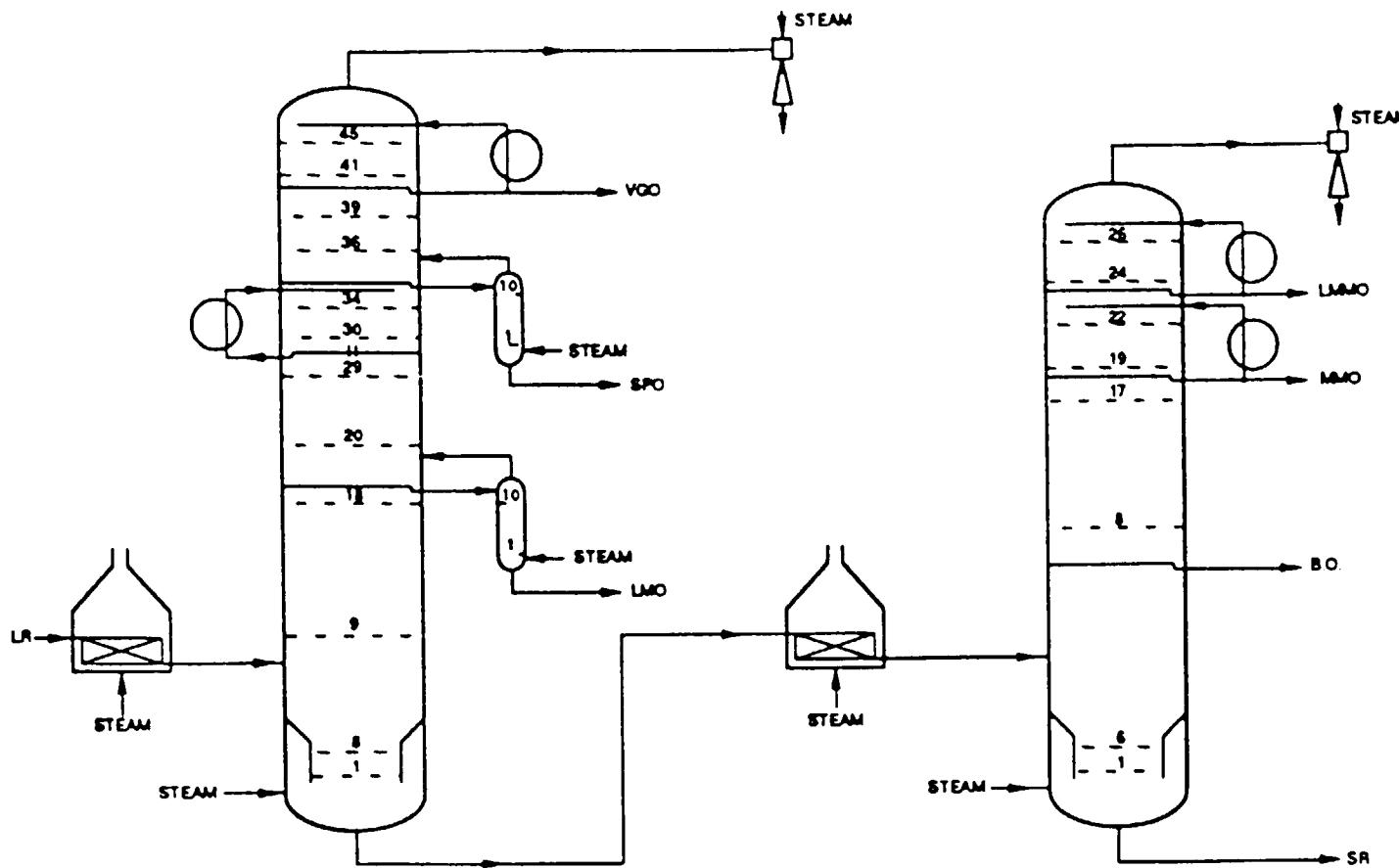


FIGURE 1.2 CONVENTIONAL LUBOIL HVU

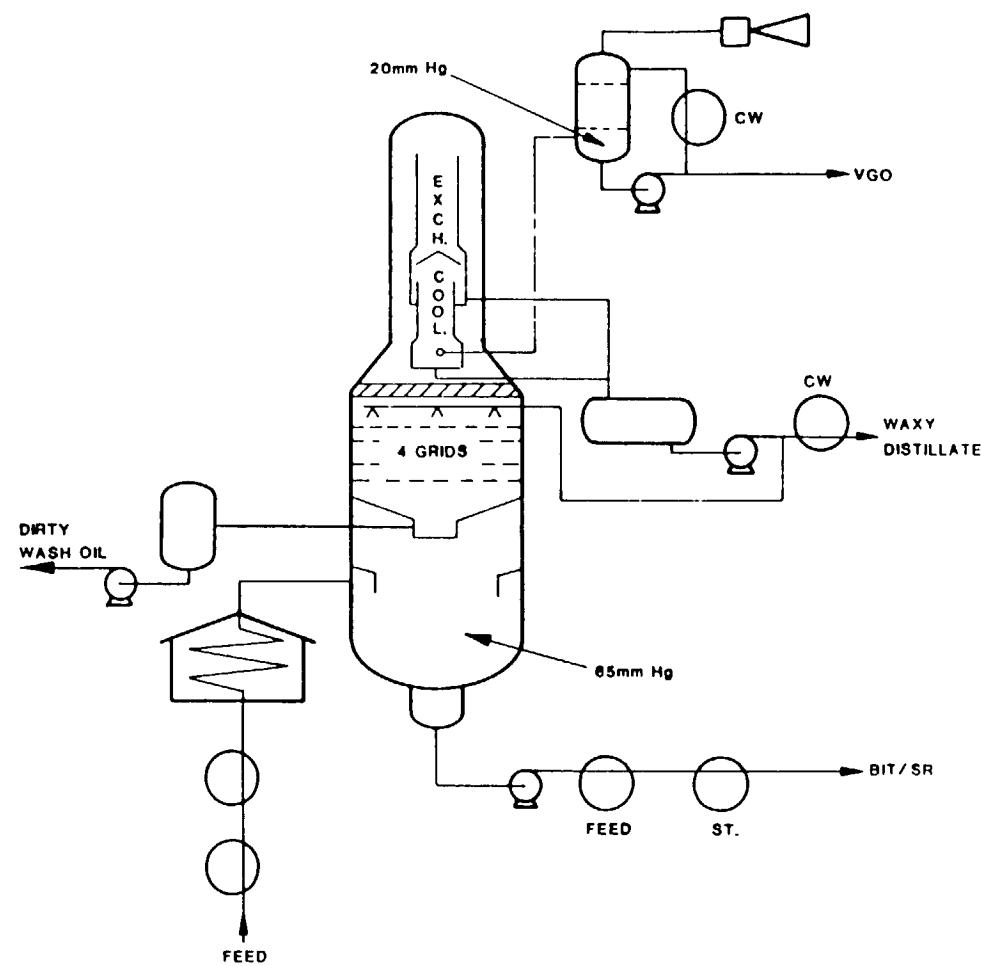


FIGURE 2.1 TYPE 1 HVU

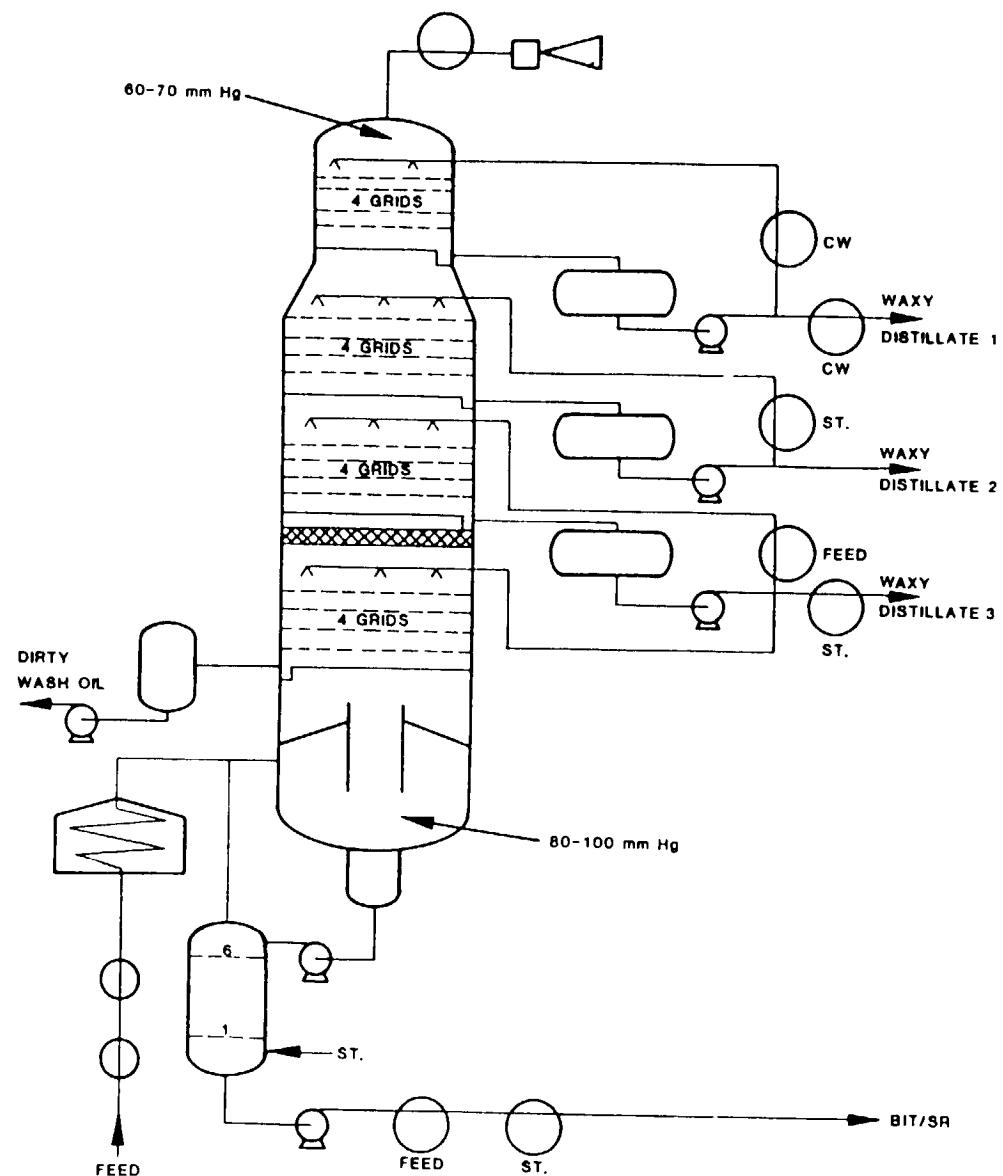


FIGURE 2.2 TYPE 2 HVU

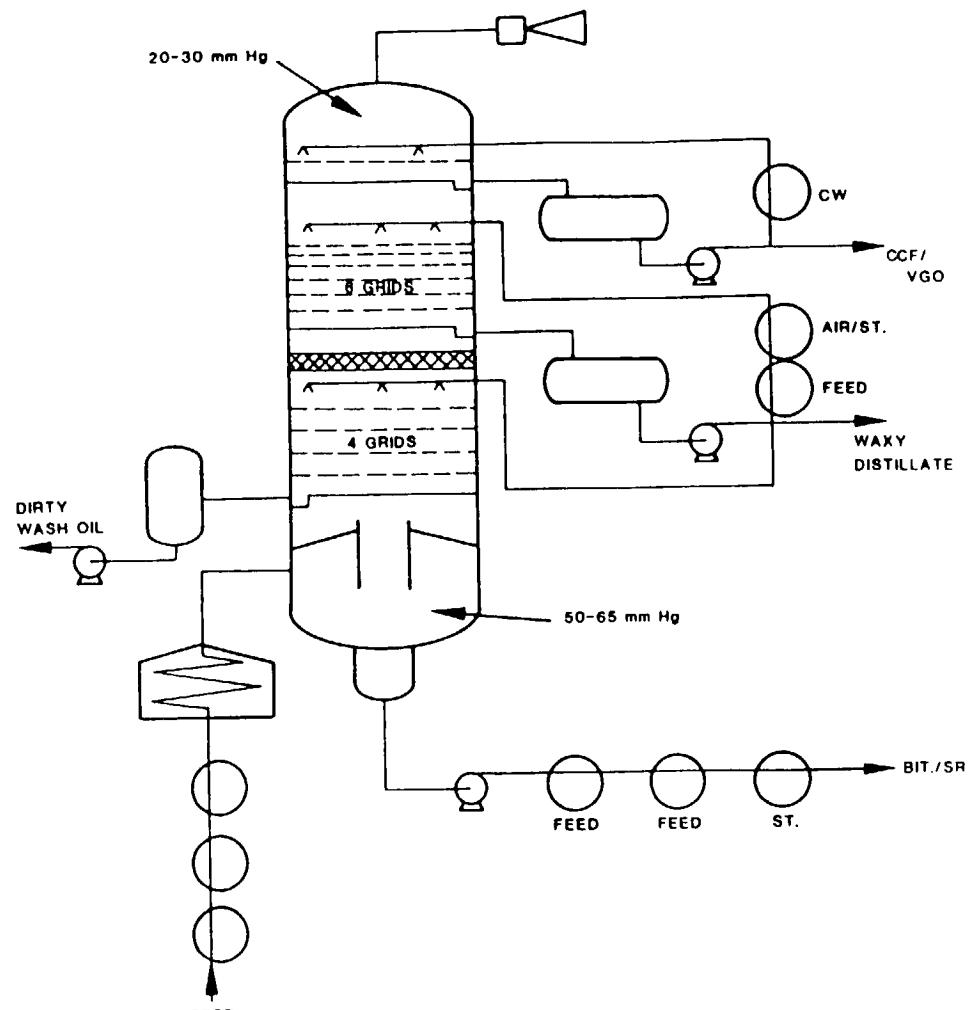


FIGURE 2.3 TYPE 3 HVU

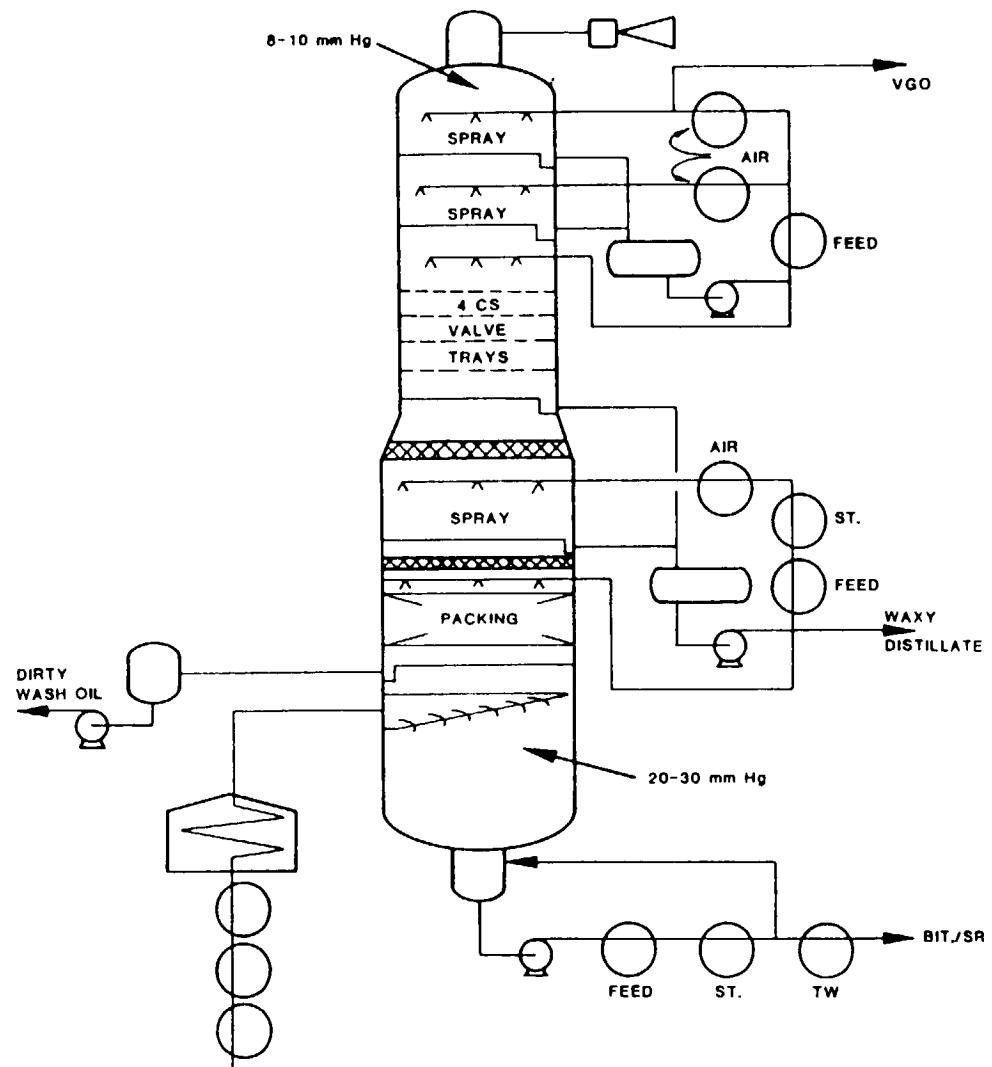


FIGURE 2.4 TYPE 4 HVU

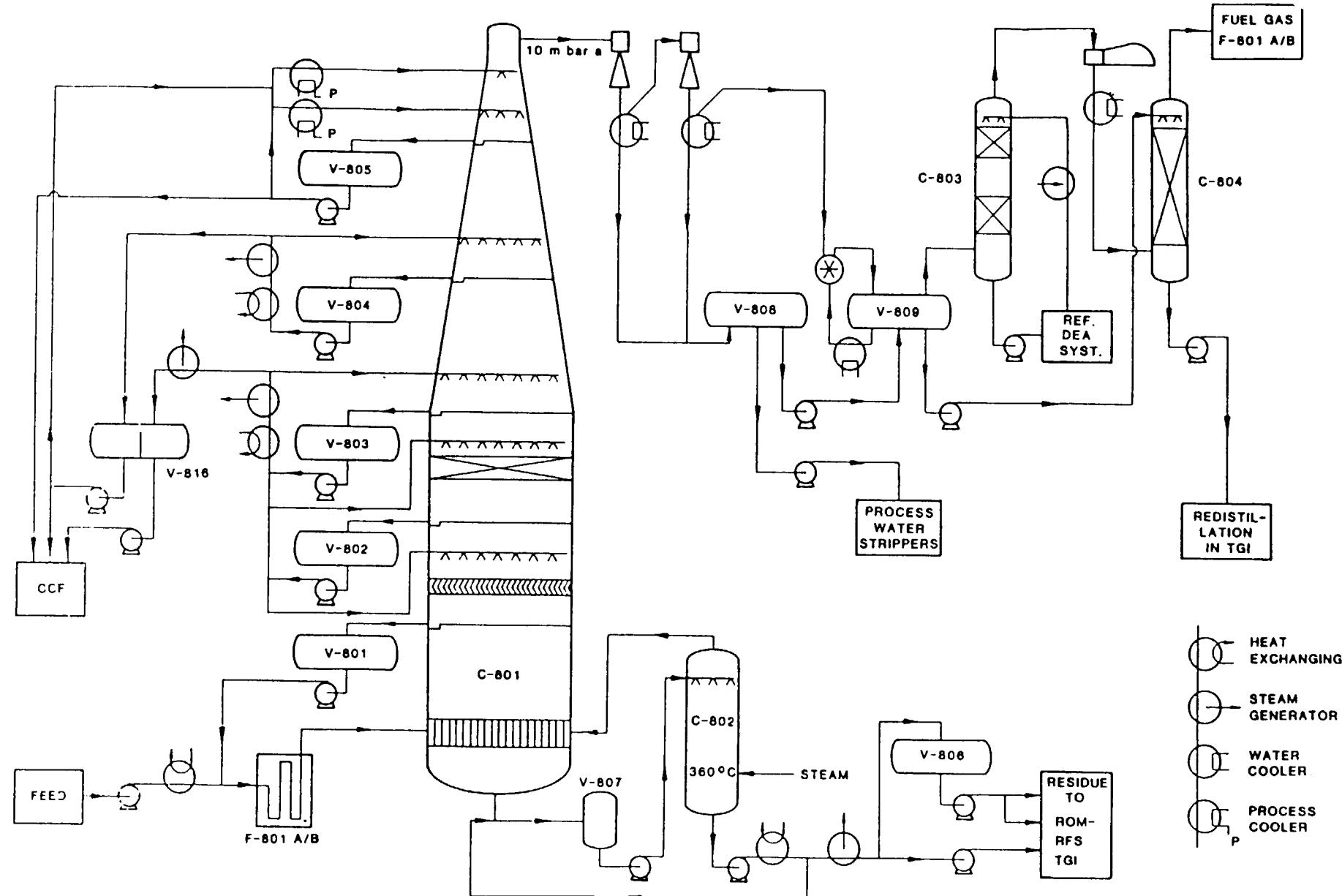
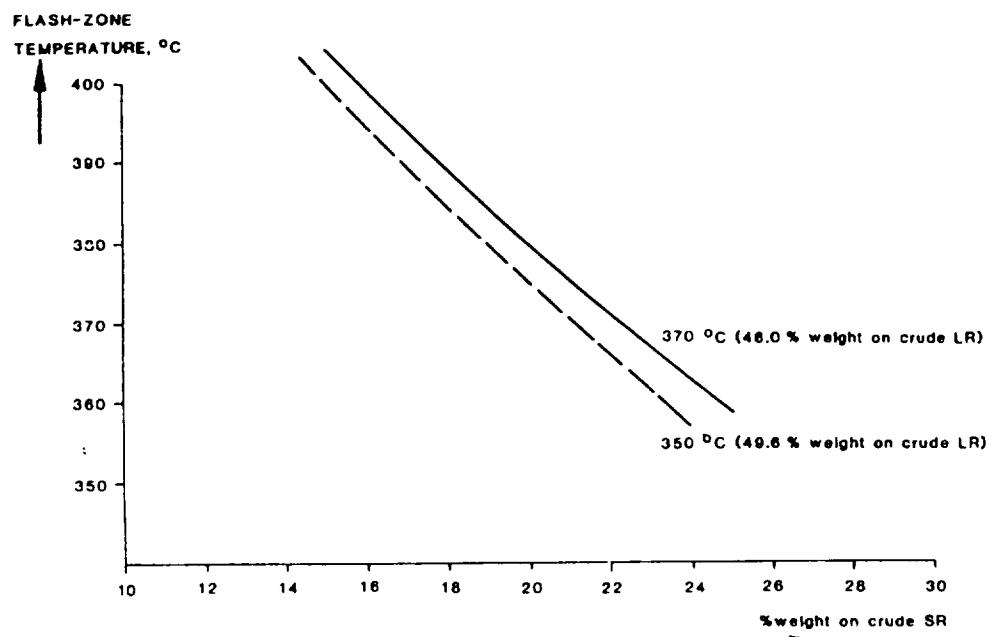


FIGURE 2.5 TYPE 5 HVU



PRESSURE = 25mm Hg abs  
CRUDE: ARABIAN LIGHT

FIGURE 3.1 EFFECT OF RAISING FLASH-ZONE TEMPERATURE ON SHORT RESIDUE YIELD AT CONSTANT PRESSURE

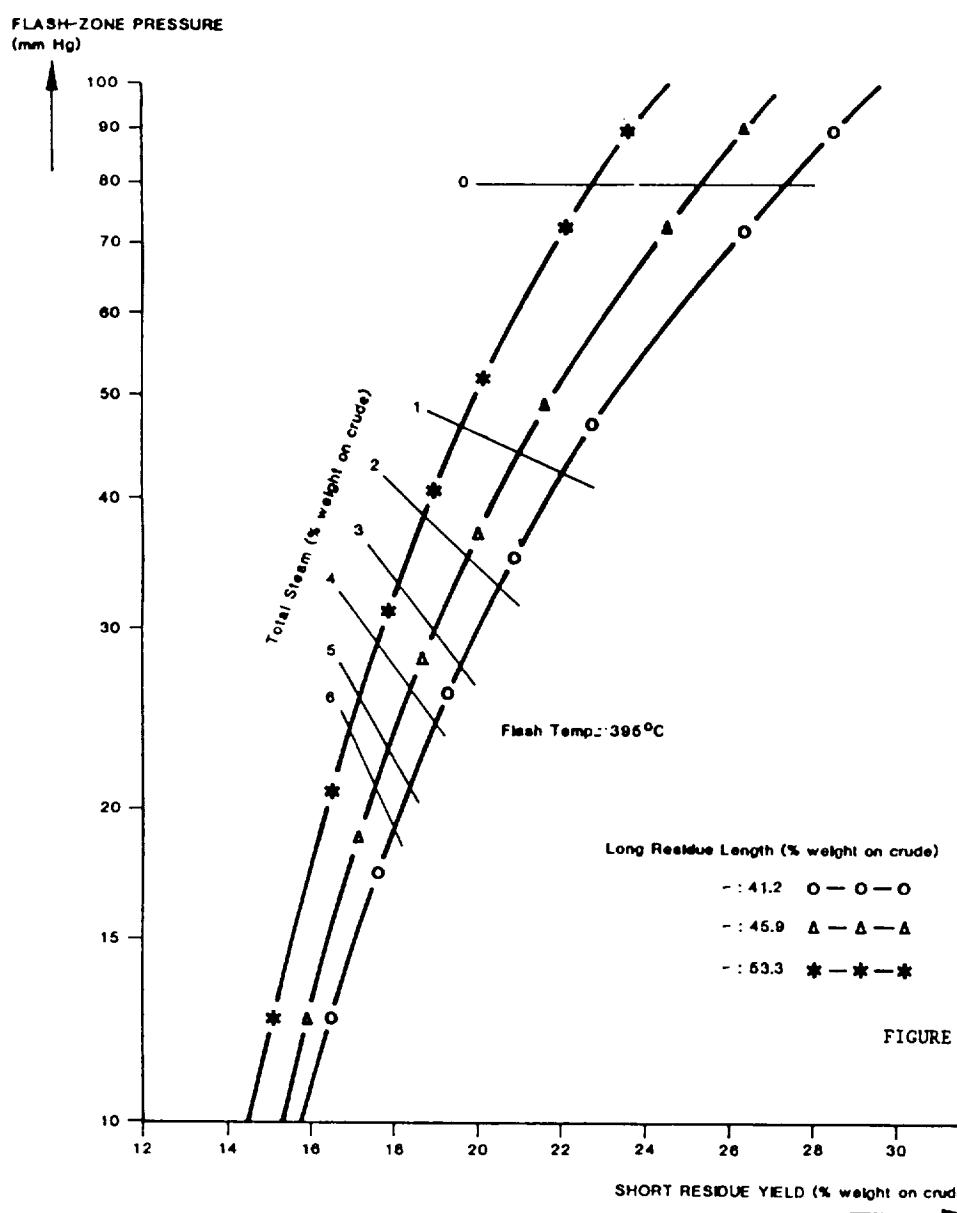


FIGURE 3.2 EFFECT OF TOTAL STEAM ON DISTILLATE YIELD AT CONSTANT FLASH-ZONE TEMPERATURE

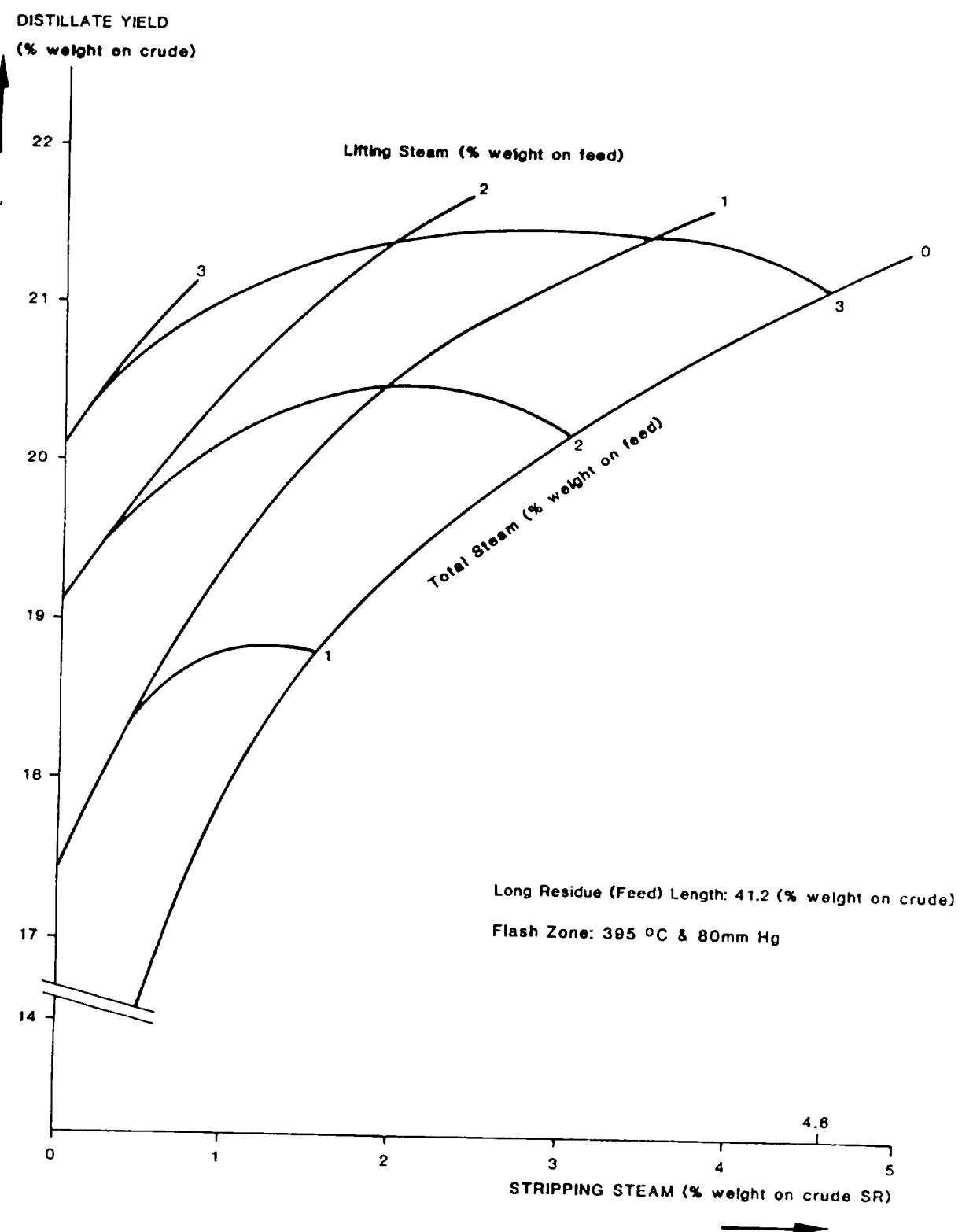


FIGURE 3.3 EFFECT OF STRIPPING STEAM ON DISTILLATE YIELD

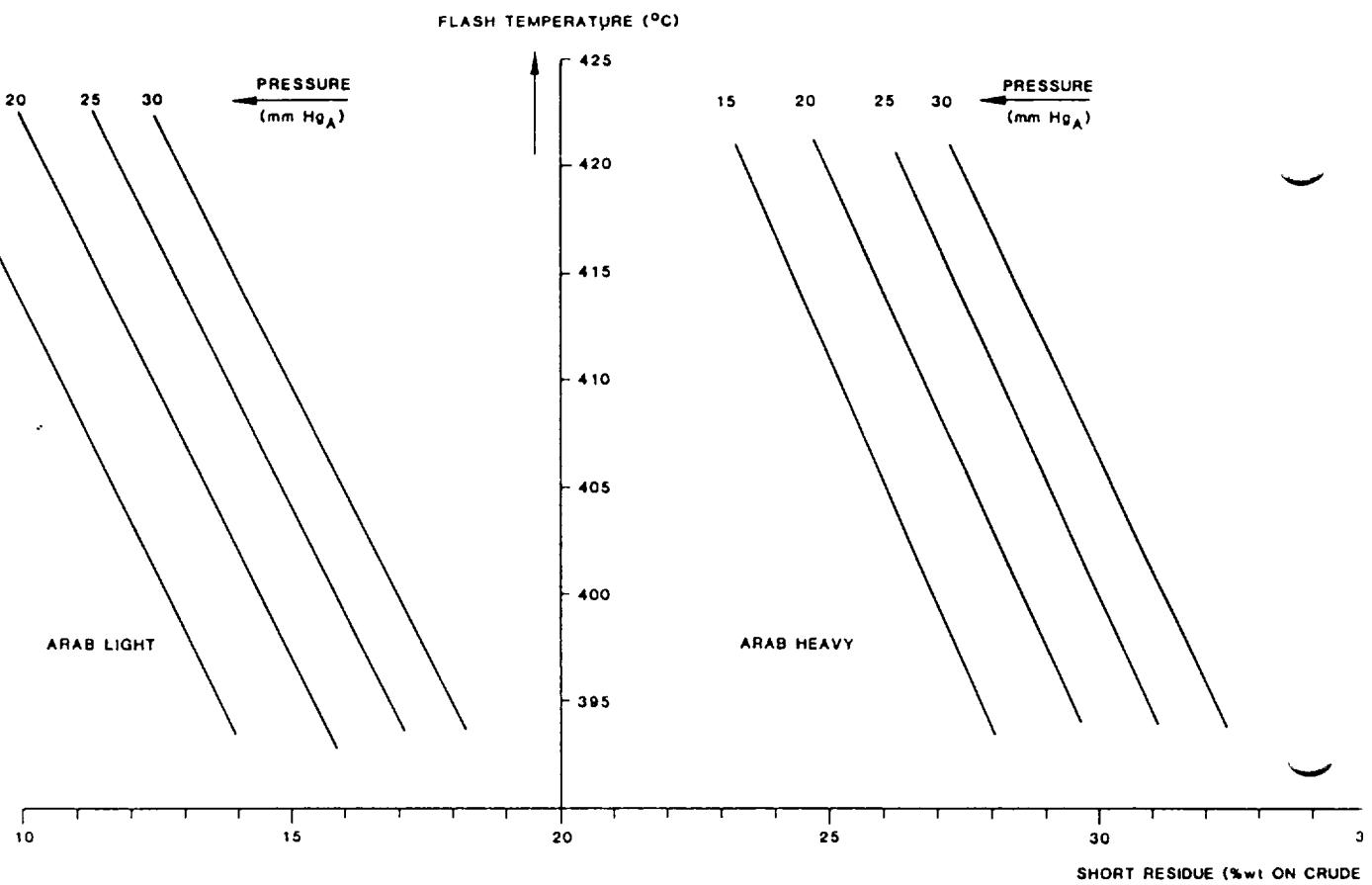


FIGURE 4.1a FLASH CURVES

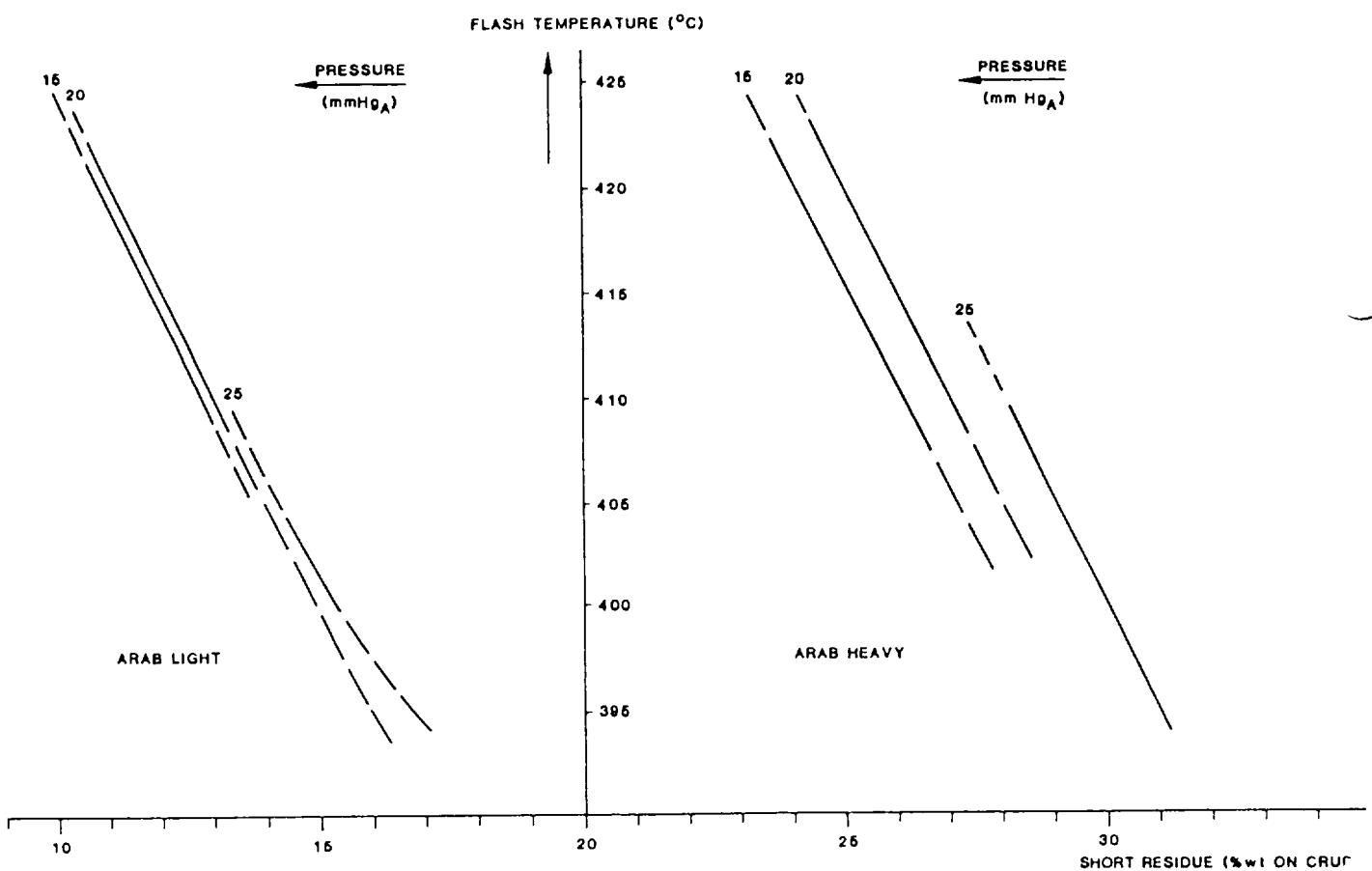


FIGURE 4.1b COMMERCIAL DEEP FLASHER  
(TABLE 3 DATA)

ARAB LIGHT

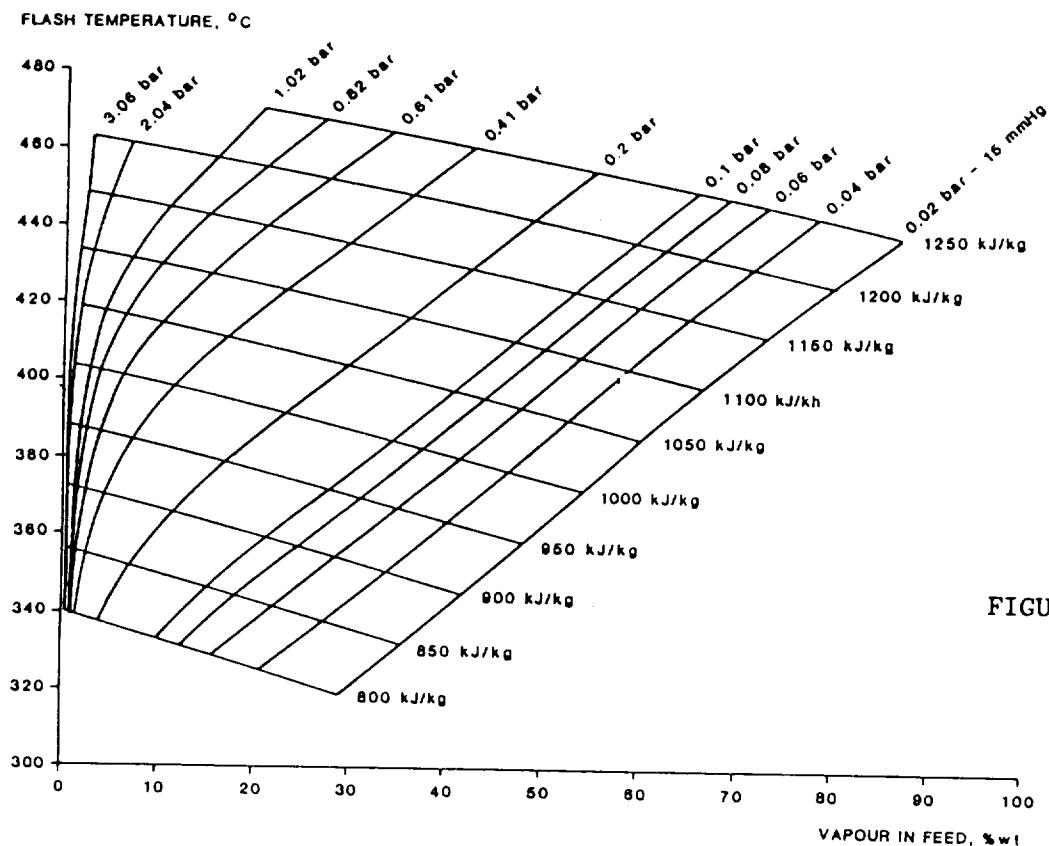
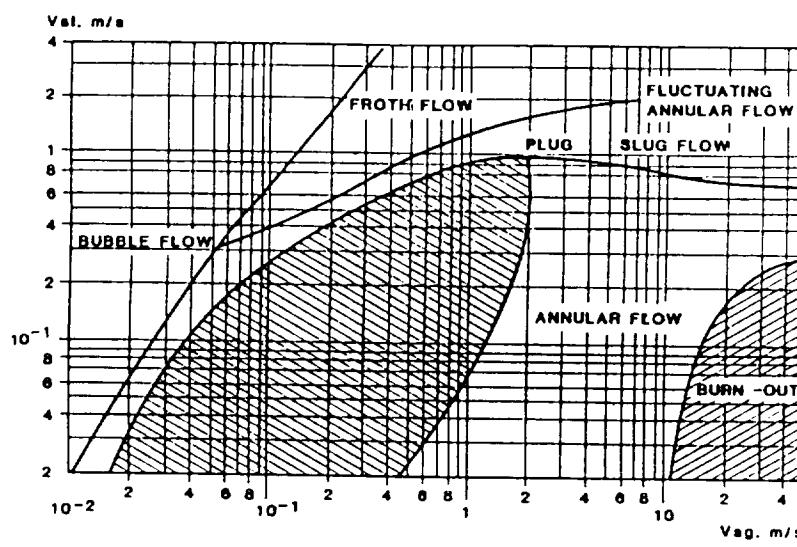
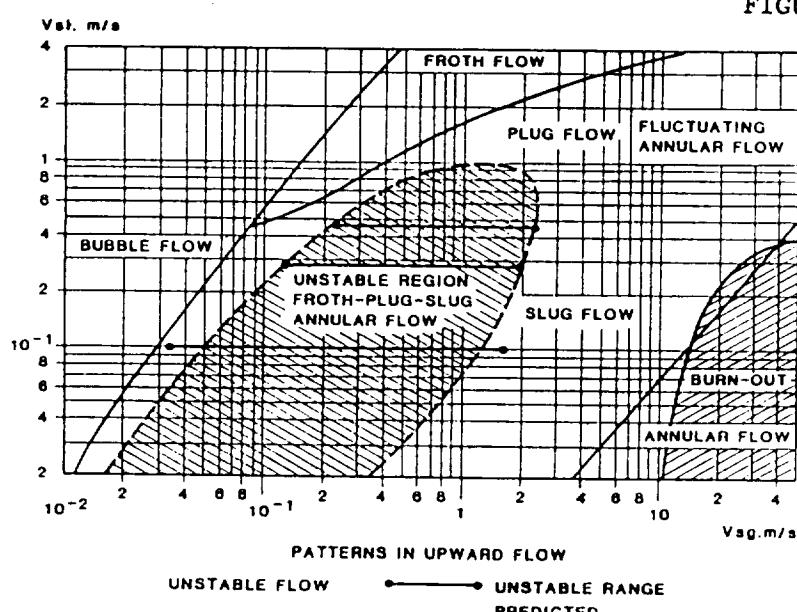


FIGURE 4.3 FLASH CURVE



Patterns in downward flow



FIGURES 4.4 a+b FLOW OF GASOIL AND AIR IN  
140 mm ID PIPES

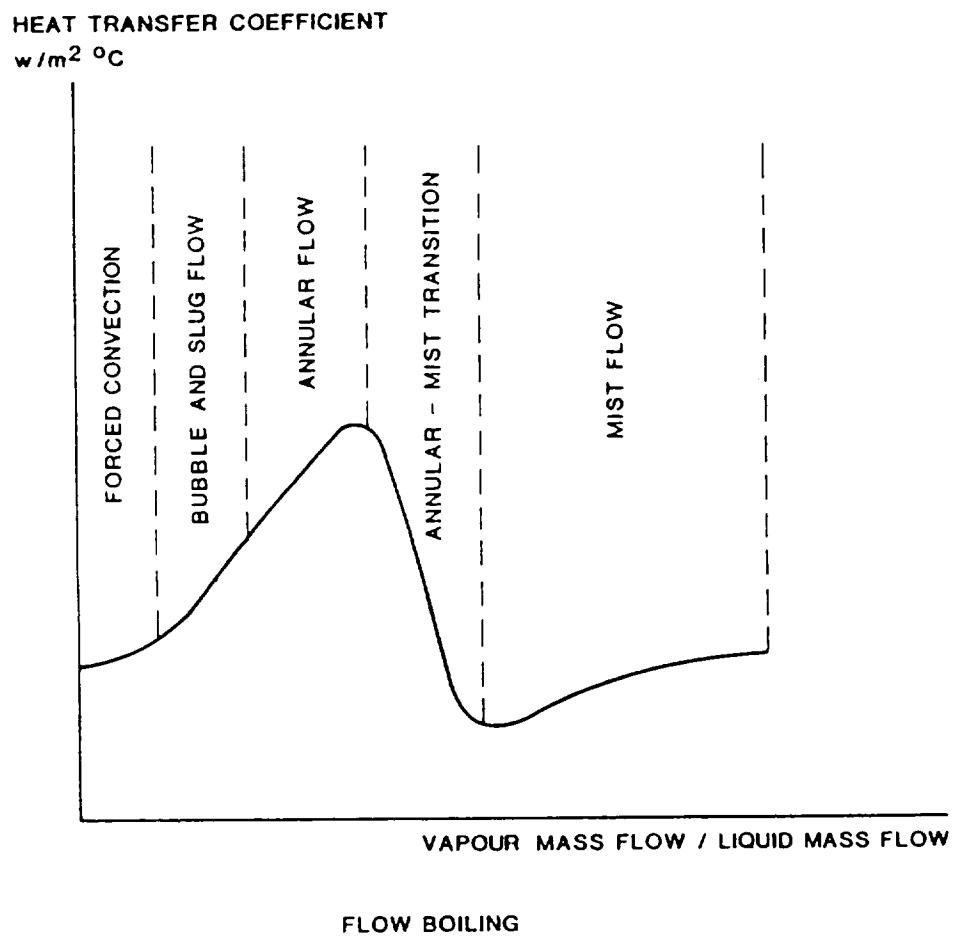


FIGURE 4.5 HEAT TRANSFER COEFFICIENT FOR TWO PHASE FLOW

FIGURE 4.7 LAYOUT OF TRANSFER LINE

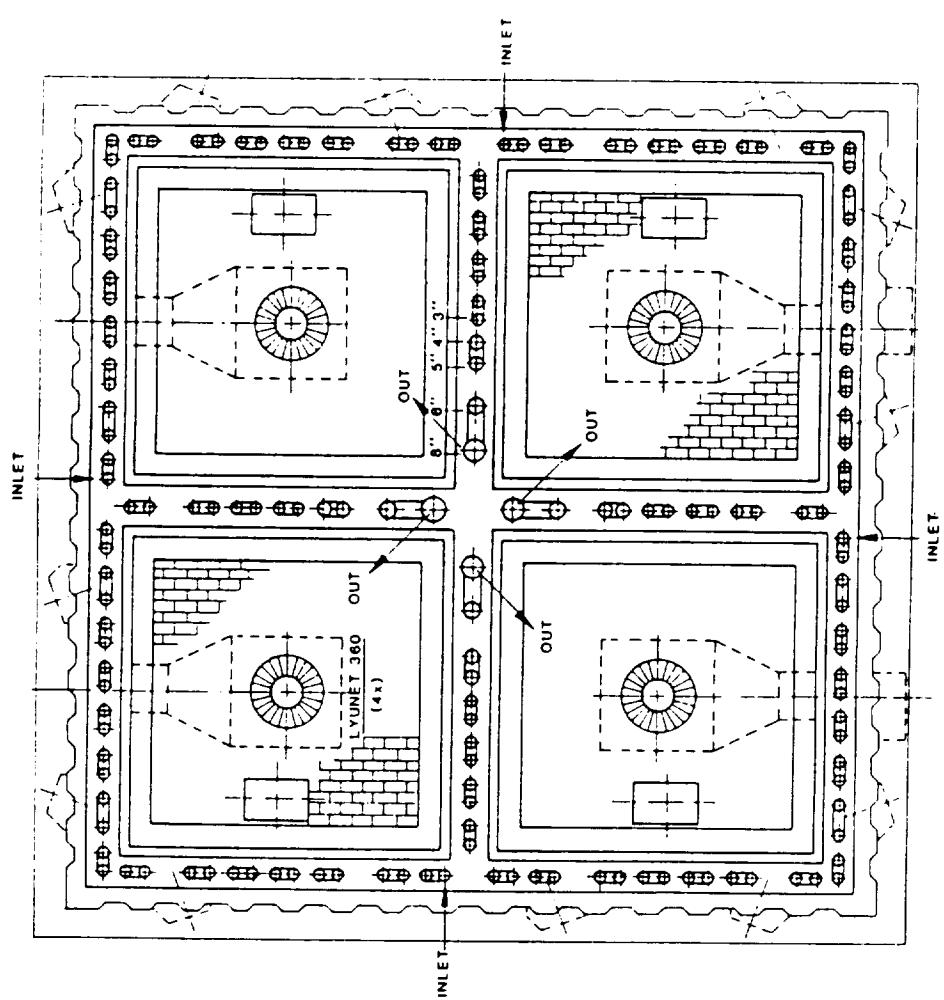
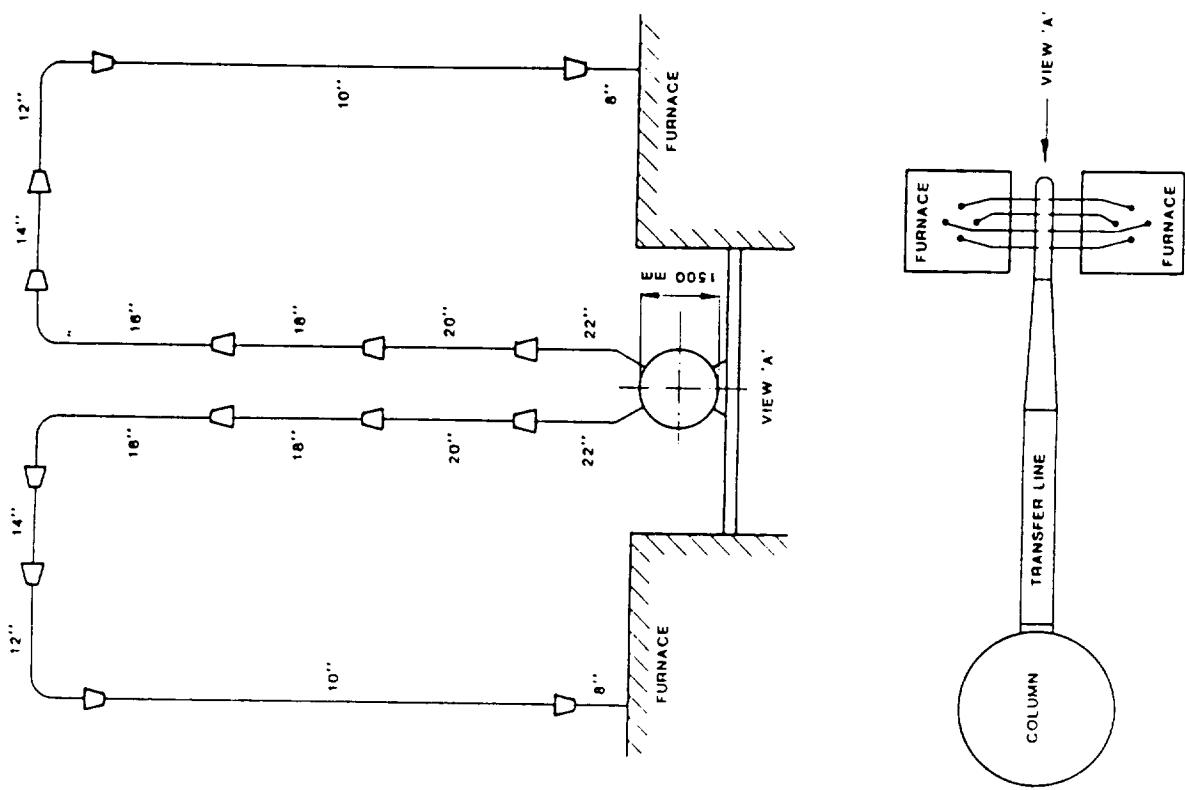


FIGURE 4.6 LAYOUT OF MODERN HVU FURNACE

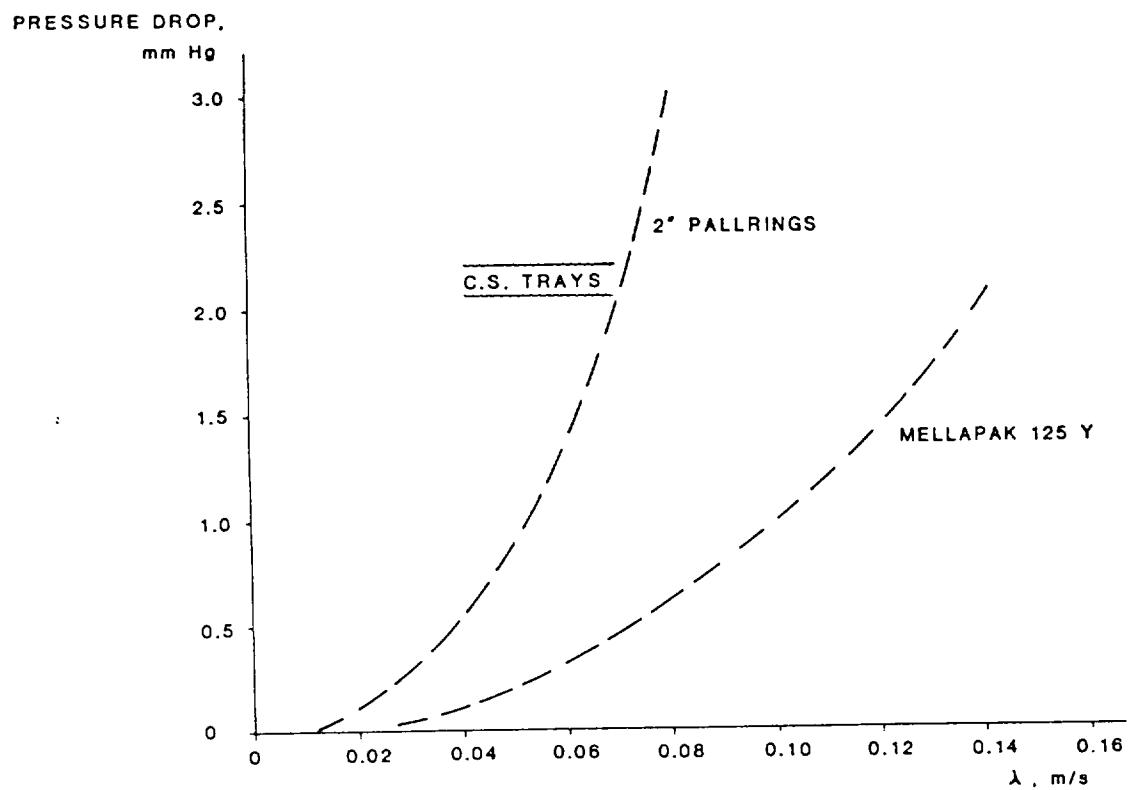


FIGURE 5.1 PRESSURE DROP TRAYS/PACKINGS

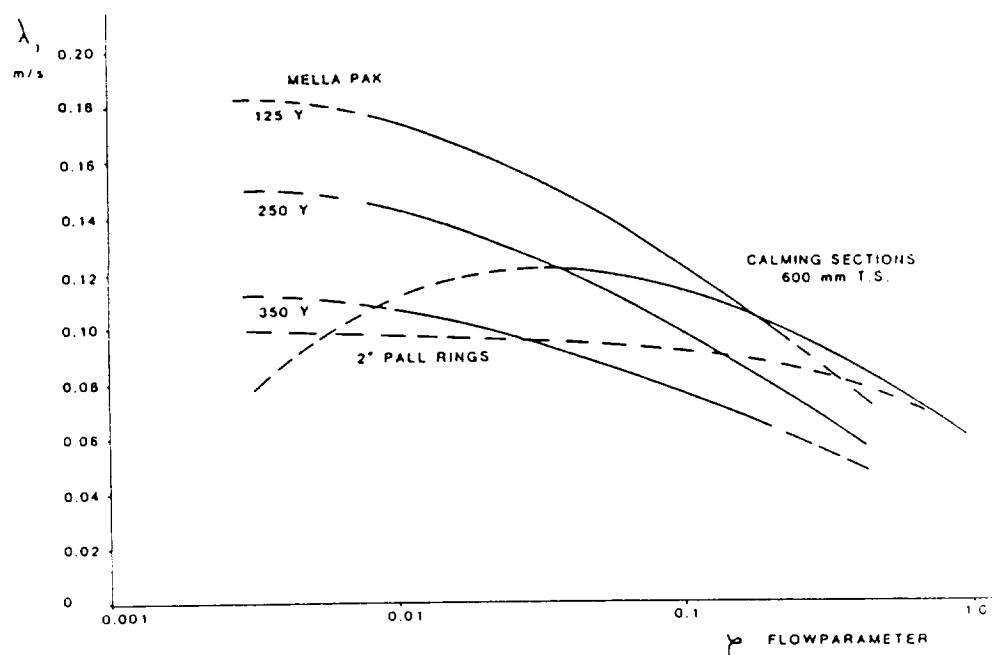


FIGURE 5.2

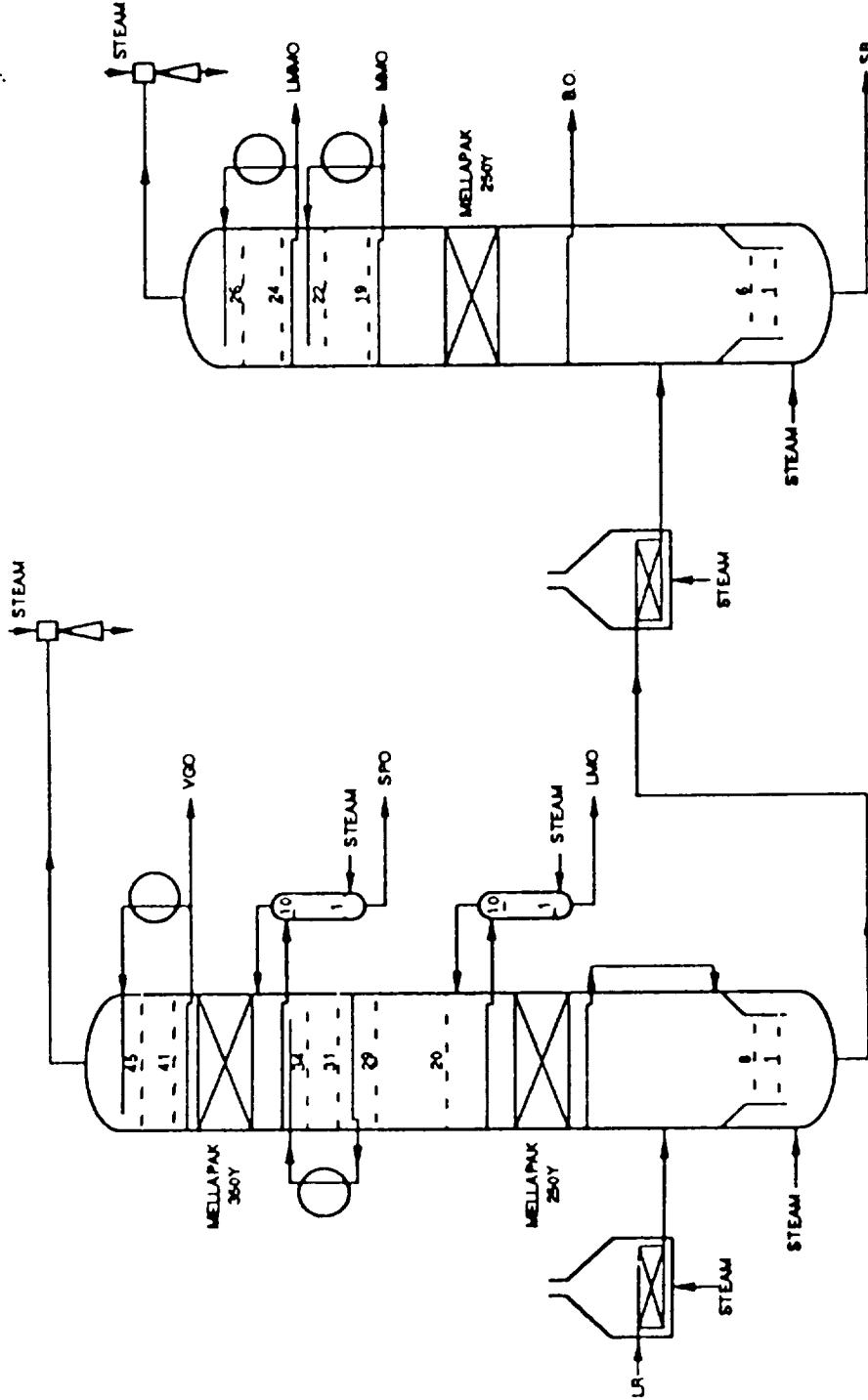


FIGURE 5.3 LUBOIL HVU WITH PERNIS-TYPE MODIFICATIONS

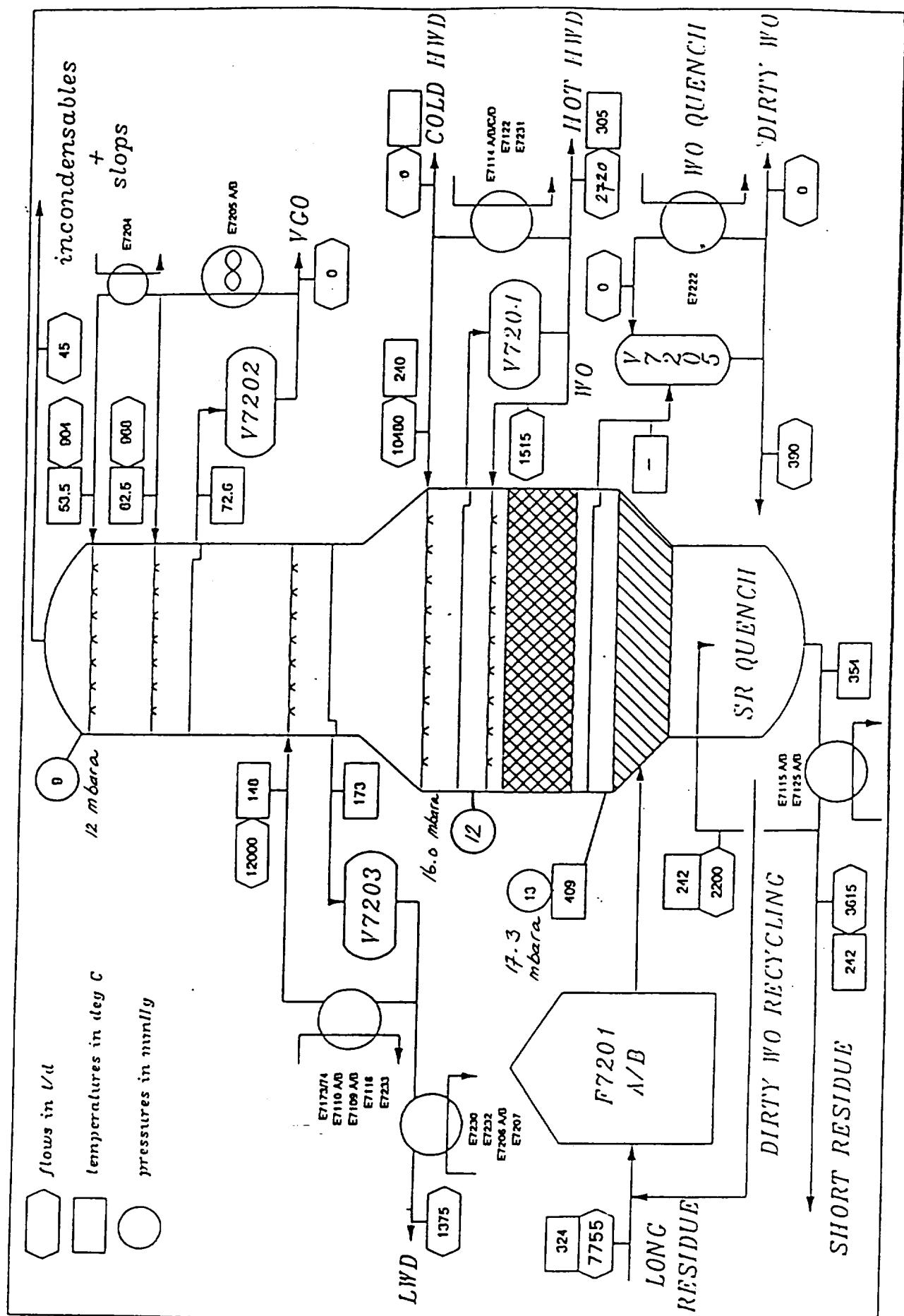
## HVU CASE STUDY: TEST RUN EVALUATION

Conducting regular (mini) test runs is a powerful tool to monitor the performance of a unit in a systematic way. In this case study, you will calculate the key performance monitoring parameters of a Feedprep HVU test run in order to evaluate its performance.

The results of the Feedprep HVU test run are summarised in attached figure (Appendix 1), showing the relevant flows and operating conditions of the unit. The lab. results and some key hardware information (transfer line and column diameter) are presented in the attached EXCEL spreadsheet (Appendix 2). Formulas which you may need are given in Appendix 3.

### Questions:

1. What is the Maxwell-Bonnell (MB) equivalent atmospheric temperature of this unit. Use the attached page B-2 of "Suffertje" (MF 77-500) or formula in Appendix 3 (use UOP-K of 12.0). Is the actual MB temperature higher or lower than the design value of 580 °C.
  
- 2.a What is the SR content in the DWO leaving the bottom of the Wash Oil (WO) bed (in t/d). Make a Vanadium (V) balance for the DWO.
- 2.b Calculate the clean waxy distillate in the DWO (in t/d) and calculate the wetting rate at the bottom of the WO bed in mm/s (exclude the SR content of the DWO). Use a liquid density of 750 kg/m<sup>3</sup> for the DWO.
- 2.c Is the actual wetting rate in line with the recommended values:
  - FZT < 380 °C: 0.05 mm/s
  - FZT > 400 °C: 0.08 mm/sWhat action is needed to correct the actual wetting rate.
  
- 3.a What is the transfer line vapour velocity. Make a mass balance over the unit. For calculating of the vapour density and mol.weight, see formula's in Appendix 3.
- 3.b What is the sonic velocity (formula in Appendix 3). Is the actual vapour velocity higher or lower than 70 % of this sonic velocity.
  
- 4.a Calculate the vapour loading ( $\lambda$ ) at the entry of the LCR (= top of WO bed) by making a mass balance. For formula's, see Appendix 3. Use the following values  $\rho_l = 750$  kg/m<sup>3</sup>, temperature at top of bed is 370 °C and same molweight as calculated under 3.a. Compare the actual vapour loading with the maximum  $\lambda$  value of 0.10 m/s.
- 4.b Is there scope to lower the vacuum in the HVU (assuming that the ejector set has spare capacity)



INPUT PANEL 1 - FEED BACK PLANT OPERATION FEEDPREP HVU										
Location: Case study		Unit: HVU		Test run date: 15/11/95						
Hardware information:		Column internals information:								
Column diam. bottom (ID)	11.00	m	Deentrainment section: Mellapak 125Y - 4 layers							
Column diameter top (ID)	7.55	m	Feed inlet type: Schoepentoeter - Side							
Transferline diam. (ID)	3.00	m	Condensation section: Spray sections							
Mechanical design temp.:										
Furnace	NA	°C								
Bottom column	445	°C								
Material balance (t/d):		Actual	Design	Feed information:		Actual	Design			
Waste gas	15	0	0	Long residue origin:		Iran Heavy				
Overhead slops	30	0	0	Feed rate		7755	8658 t/d			
VGO	0	0	0	Inbalance		0	-57 t/d			
Light Waxy Dist.	1375	3325								
Heavy Waxy Dist.	2720	2245								
DWO product	0	0	0							
Short residue (excl. DWO)	3615	3145		Sulphur content LR		3.3	% wt			
Total products	7755	8715		Total distillate yield (including DWO)		4140	5570 t/d			
Product:	Overh.	VGO	LWD	HWD	DWO	SR				
Analysis:	slops									
S.G.-d15/4		0.7513	0.9289	0.9292	0.9459	0.9824	1.0440			
S.G.-d70/4										
Viscosity:										
cSt at °C 100		6.6	7.6	18.7	106					
cSt at °C 130					1730					
TBP - 0 °C		193	239	358	413	412				
10 °C		354	370	429	512	574				
30 °C		405	415	471	558	638				
50 °C		433	440	503	585	690				
70 °C		461	466	540	614	729				
90 °C		508	508	591	676					
% at 620										
% wt										
Additional information:		LWD	HWD	DWO	SR					
Vanadium ppm wt			0.5	100	313					
Nickel ppm wt			0.7	31	94					
Asphalt. % wt										
V-50		25.3	30.1		47.3					

## 1. MAXWELL-BONNELL TEMPERATURE

The Maxwell-Bonnell temperature can be calculated with the help of the nomogram on page B-2 of the "Suffertje" (MF 77-500) or with the following formula:

$$T_{at} = (748.1 \cdot A) / (0.3861 \cdot A + 1 / (T_p + 273.15) - 0.0005161) + B - 273.15$$

$$A = (3.877 - \text{Log}(P/1013)) / (2387.262 - 95.76 \cdot \text{Log}(P/1013)) - 0.0002867$$

$$B = 2.5 \cdot (\text{UOPK-12}) \cdot \text{Log}(P/1013) / 1.8$$

where: UOPK = Watson Characterisation Factor

P = Absolute pressure in mbar(abs), and  $2 < P < 1013$ :

$T_p$  = Boiling point at pressure P in °C

$T_{at}$  = Atmospheric boiling point in °C, Maxwell-Bonnell temperature

## 2. MOLWEIGHT ESTIMATE OF FLASH ZONE VAPOURS

$$\text{Molweight Hydrocarbon vapours} = 123 + 0.521 \cdot \text{MB temp} - 145 \cdot \text{VGO fraction}$$

Where: MB = Maxwell-Bonnell temperature in °C

$$\text{VGO fraction} = \text{VGO} / (\text{VGO} + \text{WD})$$

When applicable, a correction should be made for the cross over and stripping steam, if injected.

## 3. VAPOUR DENSITY CALCULATION

With the molweight known, the vapour density can be calculated with the Ideal Gas Law:

$$\rho_v = 12.03 \cdot P \cdot MW / (T + 273) \quad (\text{Ref. page C-8 of "Suffertje"})$$

where:  $\rho_v$  = density in kg/m<sup>3</sup>

P = pressure in bar(abs)

MW = molweight

T = temperature in °C

## 4. SONIC VELOCITY FORMULA

$$V_{sonic} = \sqrt{\kappa R T / MW} = \sqrt{FZP / \rho_v} \quad \text{with } \kappa = 1.0$$

Where:  $V_{sonic}$  = sonic velocity in m/s

R = gas constant (8314 J/kmol.K)

T = temperature in K

MW = molweight of vapour phase

FZP = Flash Zone Pressure in Pa

$\rho_v$  = vapour density in kg/m<sup>3</sup>

## 5. VAPOUR LOADING FORMULA

$$\text{Vapour loading: } \lambda = Q_v / A_c \cdot (\rho_v / (\rho_l - \rho_v))^{0.5} \quad (\text{in m/s})$$

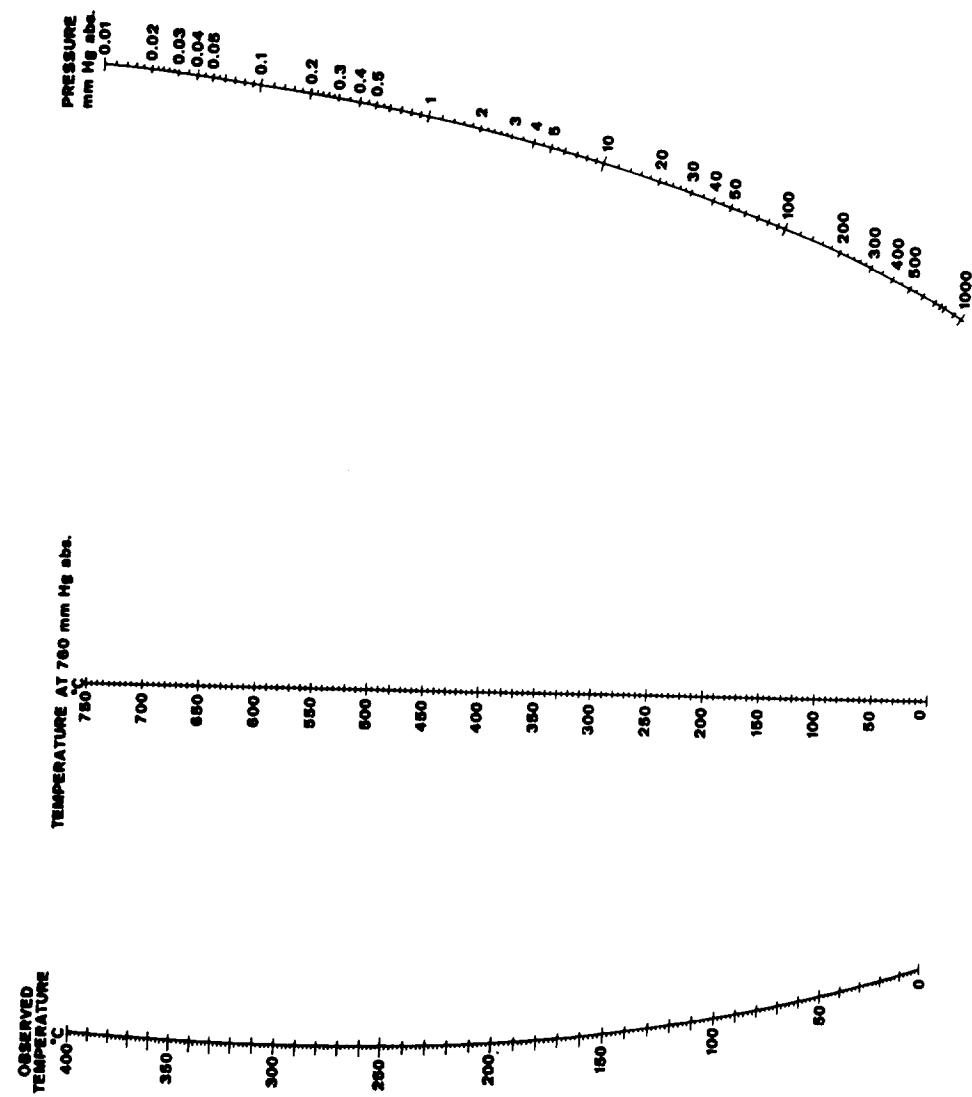
Where:  $Q_v$  = vapour flow in m<sup>3</sup>/s

$A_c$  = column area in m<sup>2</sup>

$\rho_v$  = vapour density in kg/m<sup>3</sup>

$\rho_l$  = liquid density in kg/m<sup>3</sup>

B - 2



**TEMPERATURE-PRESSURE CONVERSION FOR PETROLEUM HYDROCARBONS**  
1 bar = 750 mm Hg



17

**FCC Main Fractionator  
and Gas Plant**

## **Cat. Cracker Main Fractionator and Gasplant An Overview**

The distillation facilities associated with a Cat. Cracker Unit (CCU) can be as extensive as a whole Crude Distillation Unit (CDU). A typical facility may consist of a main fractionator column with two side strippers, an absorber column, a debutaniser, depropaniser and gasoline splitter. This gives a total of 7 columns all with their associated equipment (pumps, heat exchangers, vessels, compressors etc). Many of the aspects of operation and design of these systems are the same as with other columns as discussed during the course. There are however some aspects not covered elsewhere in the course and the following sections are devoted to the design and operational aspects of CCU distillation facilities. This overview consists of 4 sections:

- **Design and operational aspects of Cat. Cracker Fractionator.**  
This section is based on papers from recent cat. cracking and alkylation meetings and discusses the general design and operational aspects of cat. cracker fractionators. Aspects of the design and operation and of the desuperheating and TCR sections are particularly focused on. Guidelines on how the main operating variables effect the main product yields and properties are then discussed.
- **Developments of Cat. Cracker Fractionator.**  
This section is based on a paper from the 1986 distillation meeting. It highlights design developments and differences as compared to older units.
- **Rectifying Absorber**  
This section gives a basic introduction to absorber columns as applied in CCU gasplants.
- **Cat. Cracker Absorber operation**  
This section covers options for improvement for CCU gasplant absorbers. Alternative process line-ups and the application of chilled water cooling are discussed.

# **Design and operational aspects of Cat. Cracker Fractionator**

## **1. Introduction**

The degrees of freedom for the design of new facilities, revamp of existing units and actual operation of Cat. Cracker Fractionator facilities are limited as compared to, for instance, a CDU because the feed to the main column, the feed temperature and operating pressure are governed by the operation of the Cat. Cracker reactor/regenerator section. Therefore the scope of design and operation presents special problems. On the basis of recent design experience and computer simulations of a typical unit guidelines for the main design and operational aspects are given. In addition some data on refinery experience with respect to the de superheating and TCR sections is presented.

## **2. Design Features**

Feed to a cat. cracker fractionator differs from feed to, for instance, a crude distiller because it is superheated and contains a large amount of steam and incondensables. The bottom and top sections of the fractionator are therefore the interesting sections having specific problems.

Figure 1 shows the lay-out of a typical CCU fractionation column. The fractionator consists of one desuperheating/slurry condensation section, two condensation section (TCR and MCR) and two fractionation sections for the separation of LCO/HCO and LCO/Gasoline.

### **2.1 Desuperheating Section**

Superheat is removed and recovered in a circulating system via heat exchange with, for instance column reboilers, CCU feed and steam generation. A minimum amount of slurry has to be condensed, depending on the acceptable concentration of catalyst fines in the slurry, at the same time observing an acceptable maximum bottom temperature in view of possible coke formation. Shell-designed fractionation columns are normally equipped with round bar grid trays. UOP-designed columns, for instance, as well as the very old Shell-designed columns have baffle-type trays (segmental baffles or disc and donut trays). Both types of tray are reasonably robust to fouling and they both have approximately the same max. vapour handling capacity, although capacity independence of liquid load is better for baffle-type trays. Shell prefer grid trays because of their lower entrainment. In order to enhance the performance of grid trays especially in larger columns radial baffles are installed on the trays. For de-bottlenecking purposes, the desuperheating section capacity can be increased by installing radial baffles, increasing the NFA and the tray spacing and reducing the circulating flow rate. For the latter, additional heat recovery may be required.

A low heat and mass transfer efficiency is a common feature of grid and baffle trays. High efficiencies are not needed, because, ideally, the desuperheating section should represent only about one theoretical stage. A higher efficiency than that assumed in the design can result in considerably more than one theoretical stage. This can give rise to the requirement to increase the amount of slurry condensed (at the expense of HCO), in order to keep the column bottom temperature within acceptable limits. This change will result in an increased desuperheating section duty at the expense of the duty from the MCR loop. If a certain MCR duty is required for, e.g. reboiling this will obviously give problems in those particular columns. Also the reduced MCR duty means less

internal reflux over the trays governing the LCO/HCO separation, hence there will be a reduction in the LCO/HCO fractionation.

Figure 2 show the amount of slurry produced as a function of the bottom temperature, simulating the desuperheating section with one theoretical stage and by two theoretical stages with 50/50% heat withdrawal. As can be see, the amount of slurry at a certain bottom temperature almost doubles. Figure 3 shows the respective heat duties of the slurry loop and the mid-circulating reflux for both cases. A substantial shift in heat duty from the MCR loop to the slurry loop can be observed. Shell-designed units have four to six grid trays. From plant data it can be concluded that four grids already represent more than on theoretical stage, so that in the case of difficulties with regard to HCO yield or MCR duty the reduction of trays to four can be considered.

In modern designs a “two stage” system is generally installed as shown in Figure 4. This two stage system offers the flexibility to vary the overall efficiency of the total desuperheating section by varying the flows of the two slurry returns. The flow of the top return should however always be sufficient to maintain enough liquid wetting for the upper trays. The bottom quench finally enables the independent reduction of the bottom temperature in the case of a temperature limitation to the fractionator bottom or slurry loop (equipment protection).

## 2.2 Top Circulating Reflux Section

The top circulating reflux section of a cat. cracker fractionator exhibits the combined feature of a high, often throughout-limiting, loading with the problem of possible water condensation, entailing corrosion and tray plugging by salt deposition. To prevent water condensation, the following two requirements can be formulated:

1. The temperature of the overhead vapours should be above the dewpoint temperature of steam in the hydrocarbon vapour leaving the column. This requirement sets the ultimate minimum naphtha FBP, which can be achieved in the fractionator.
2. The reflux return temperature should be above the dewpoint temperature of the steam in the hydrocarbon vapours leaving the column.

Figure 5 shows the relationship between the FBP of LP distillate, the temperature of overhead vapours and the dewpoint temperature of steam in the overhead vapours for a typical case. As can be seen, the minimum FBP is around 165°C. Whereas for new designs the TCR return temperature can be selected to be above the water dewpoint temperature, this is not the case with many existing units. Instead the risk of some local water condensation/evaporation with the accompanying possibilities of salt deposition and corrosion is accepted and monitored. In some cases facilities exist or have been added to enable water washing of the trays in the top circulating reflux to remove salt deposits.

## 2.3 Example of Revamp Possibilities

As an example of the type of changes that can be made when debottlenecking or revamping a unit the following steps could be considered if one of the revamp objectives was to maximise LCO yield whilst also increasing throughput. These two requirements are inter-related and cannot be met simultaneously. A certain required throughput sets the minimum achievable naphtha FBP and vice versa.

On the basis of a minimum TCR return temperature at the lower FBP, the circulating flow has to be increased to withdraw the required heat duty. As the maximum vapour loading of the TCR trays is lower at a higher circulation rate (higher  $\phi$ ), the throughput has to be reduced. Figure 6 shows the relationship between throughput and minimum achievable FBP established for a recent de-bottlenecking exercise.

To remove the column constraint in the TCR section the following solutions could be considered:

- 1) Retract with the highest capacity trays and increase the tray spacing by sacrificing one tray, accepting that the draw-off temperature will decrease. Alternatively, remove the trays and install a spray section. The draw-off temperature in the design with a spray section will decrease significantly due to the reduction in "staging" as compared to a conventional circulating reflux. The spray section option could offer the possibility to operate wet at very low naphtha FBP's. Tray fouling/corrosion would be less of a problem as there is only a draw-off tray in a spray section. The chance of water flooding as with trayed TCR's (where water condenses in the upper part but evaporates in the hotter, lower parts) is much less due to the much lower temperature difference across a spray section. However, the water in the TCR would have to be removed e.g. via a settling vessel.
- 2) Remove water vapour from the fractionator feed, thereby reducing the dewpoint temperature of steam in the overhead vapours and the total vapour load in the TCR. This could be accomplished by routing off-gas from the cat. cracker stripper to a separate scrubber.
- 3) Off-load the top circulating reflux section by running down a slip stream from the mid-circulating reflux section to the LCO stripper up to the FBP specification of LCO or install a new condensation section above the existing LCO draw-off. There would be some loss of fractionation between LCO and gasoline but this would probably be acceptable.
- 4) Accept a high FBP for the gasoline in the fractionator and make use of existing (or provide) facilities to split the gasoline downstream such that the heavy product is used as an LCO blending component.
- 5) Install a top section with a larger diameter.
- 6) Accept frequent problems due to tray-plugging and corrosion and apply frequent water-washing

## 2.4 Heat Integration

In normal designs sufficient flexibility is incorporated to cater for the various operating modes of the cat. cracker and the fractionation plant. Balancing the heat removal from the three circulating reflux loops can be achieved by MP and LP steam generation and by air-cooling. The aim should be to allow a certain flexibility to shift the heat duty such that the sharpness of fractionation is not affected, especially the LCO/HCO separation, i.e. the heat consumers in the gas plant should not govern the heat removal from the various loops in the fractionator.

Heat from the TCR can be used in the gas plant, for BFW heating or possibly as heating medium for a chilling unit. The heat load of the absorber and the stabiliser reboiler can be reduced by using LPD as lean oil. Additional heat becoming available in the slurry/MCR loop can be used in some cases to increase the CCU feed preheat, thereby reducing the coke make and improving the NDG/coke ratio.

In de-bottlenecking exercises where slurry loop duties and slurry rundown flows are reduced, care should be taken to ensure that liquid velocities through heat exchanger tubes do no drop below acceptable limits (typical: 1.2 m/s tubular velocities), as excessive fouling may be experienced in such a case.

### 3. Operating Variables

The main operating variables which play a role in a CCU main fractionator are discussed in the following sections. For all quantitative information given below and in the figures only one operating variable at a time has been changed.

This quantitative information is based on computer program simulations, using for example SMBP or Hyperpace to simulate the column operation whilst processing a typical feed. In this case this feed consists of a fresh waxy distillate feed intake of 2300 t/d, combined with a recycle of 460 t/d of HCO (CFR - 1.2). For a simplified process flow scheme of the fractionator studied reference is made to Figure 1. Various flow data are given in Appendix I.

#### 3.1 Cut Points

The main fractionator incorporates three circulating reflux sections: the desuperheating, MCR and TCR loops. The amount of duty removed sets the cutpoints between slurry and HCO, HCO and LCO and LCO and gasoline minus respectively. Figures 2 and 3 show the slurry yield and desuperheating duty as a function of the fractionator bottom temperature. For quantitative information about the influence of the remaining cutpoints on the product properties see Figures 7 and 8. For the relation of the TRC temperatures (cutpoint control) to LCO ASTM 95 and gasoline FBP respectively, see Figures 9 and 10.

##### Slurry/HCO Cut Point

From a review of information given in various reports it appears that most of the main fractionators are operating around the line of two theoretical stages with 50/50% heat-removal (see Figure 2). The Table below shows data gathered from some refineries with respect to the actual fractionator bottom temperature.

**Table 1 Typical Fractionator bottom temperatures**

Bottom temp. range (°C)	No. of locations *	Average bottom temp. (°C)	Max. bottom temp. (°C)
350	11	332	347
350 - 360	6	356	360
360 - 370	4	367	370

\*) Some locations are mentioned twice

With regard to the operating temperature of CCU fractionator bottoms a too high temperature (depending on the feed being processed and the configuration of the slurry loop system) can lead to excessive coke formation. Other limitations such as a maximum operating temperature of downstream equipment or a certain minimum amount of slurry which needs to be produced for an acceptable concentration of catalyst fines in the slurry may set a lower bottom temperature.

#### **HCO/HLO Cut Point**

Sometime heat consumers (e.g. gas plant reboilers) in the MCR loop dictate the MCR heat removal and thus indirectly the LCO ASTM 95%. If too much heat is required from the MCR-loop, this will result in a low LCO ASTM 95% (quality give-away), see Figure 7.

### **3.2 Stripping Steam Amounts and LCO Reboiler Duty**

In Figures 11, 12 and 13 quantitative information about the influence of stripping steam on combined slurry + HCO, HCO and LCO flashpoints and yields respectively is given. Several refineries have facilities for reboiling the LCO-stripper, for instance, if there is maximum water specification on the LCO product.

### **3.3 Extra-heavy Gasoline Draw-off**

In several refineries the possibility exists of taking an extra-heavy gasoline draw-off from the top circulating reflux section. The typical boiling range of this gasoline is ASTM IBP 50 - 70°C and FBP 215 - 250°C. This extra heavy gasoline in most locations bypasses the whole gasplant and is blended into the heavy gasoline ex the gasoline splitter. The advantage of such an extra-heavy gasoline draw-off is the off-loading of the gas plant. The disadvantages are rather low IBP and, hence, high RVP and a give-away of light gasoline is sometimes fed to the gasoline splitter when spare capacity is available. A typical composition for an extra-heavy gasoline is given below.

**Table 2: Typical Composition of Extra-heavy Gasoline**

Fraction	t/d
C4 and lighter	1.5
C5 - 100°C TBP	14.5
100 - 165°C TBP	39
165 - 221°C TBP	95
Total	150

Fresh feed: 1800 t/d

### **3.4 Duty Shifts**

In the case of where the TCR heat exchangers cannot remove all the required duty, an auxiliary reflux from the first overhead accumulator may be helpful. If auxiliary reflux is used to take the place of some of the TCR duty then of course duty is shifted from the TCR loop to the overhead condensers

By opening the bypass reflux the LCO draw-off line (and keeping the HCO/LCO cutpoint constant) duty is shifted from the MCR to the TCR section. As in most cases (especially at low LCO/gasoline cutpoints), the TCR section is already heavily loaded, this bypass reflux is normally closed. However, an advantage of using the bypass reflux is an increased sharpness of separation between the LCO and gasoline, although at the expense of the temperature level at which the heat is recovered.

Increasing wash oil flow results in a duty shift from the desuperheating loop to the MCR section. In most cases the capacity of the wash oil recycle system is limited and the duty shift which can be obtained in actual operation will be rather small.

#### 4. Top Circulating Reflux Section: Operational Aspects

The following operational problems in the top circulating reflux section of cat. cracker fractionators have been reported at low temperatures:

- Tray and pump filter plugging by salt deposition.
- Corrosion.
- Unstable operation from tray plugging and from the presence of free water.

##### Salt deposition

Salts are formed from reactions of hydrochloric acid and H<sub>2</sub>S with ammonia, all potentially present in the feed from the cat. cracker. To describe the deposition of these salts the following two mechanisms can be considered:

1. Direct desublimation of NH<sub>4</sub>Cl and, less likely, NH<sub>4</sub>HS from the vapour phase at the prevailing low temperatures in the top section. Free water need not be present, see Appendix II.
2. Assuming that free water is present, ammonia, H<sub>2</sub>S and hydrochloric acid will dissolve in the free water as NH<sub>4</sub>, HS-, H<sub>2</sub>S and Cl-. Towards the draw-off tray, where higher temperatures prevail, free water evaporates and a small amount will dissolve in the circulating reflux stream up to the solubility limit. As the free water disappears it leaves salts behind. During this process the presence of water can give rise to corrosion.

##### Free water

Free water occurs if any temperature in the top circulating reflux action is below the dew point temperature of steam, namely if the top temperature (= the top tray temperature) and/or the reflux return temperature are below the dew point temperature. The water dew point temperature depends on the column pressure, the respective amount of steam and hydrocarbons in the fractionator overhead stream and can be calculated approximately using the following equation:

$$Td = 100 * \left[ \frac{P_c}{1 + \frac{18 * HC}{St * MW}} \right]^{\frac{1}{4}}$$

in which:

Td = dew point (°C)

Pc = column pressure (kg/cm<sup>2</sup> abs.)

HC = dry vapour (t/d)

St = steam flow (t/d)

MW = mol weight of hydrocarbons

The risk of water condensation is higher if the dew point temperature is higher, (i.e. at a high column pressure, a high amount of steam) or at a low gasoline/LCO cut point (i.e. low top temperature). The formation and behaviour of free water in the top circulating reflux section depends on the prevailing operating conditions. Two cases can be distinguished:

- a) If the temperature is below the dew point temperature, massive condensation of free water in the bulk can be expected. What will happen with the free water, once formed, depends on the temperatures of the trays towards the draw-off tray and is a matter of heat and mass transfer.

If the draw-off temperature (equal to the bottom tray temperature) is low, water will be drawn off together with the circulating reflux stream and could be separated in a settling vessel. In essence, we have here a “wet system”.

Fouling/tray plugging will probably not occur but corrosion can be expected. If the draw-off temperature is above the dew point temperature, the free water will evaporate and a small amount will be dissolved in the descending reflux stream up to the solubility limit at the prevailing temperature. On spots where free water disappears, salt deposits can be expected. Water condensation will continue in the upper part of the top circulating reflux section and eventually free water will build up and result in unstable operation (flooding of the trays in the top circulating reflux section).

- b) If the top temperature is above the dew point temperature but the TCR return temperature is below the dew point temperature, local condensation on colder surfaces can be expected. Towards the draw-off tray this free water will dissolve/evaporate again. The building-up phenomenon described above, though less likely is still possible, depending on process conditions.

It is clear that if there is no desire to operate the TCR system wet, a top temperature below the dew point temperature has to be avoided at all times and a safe margin of say 15°C between top temperature and dew point temperature should be observed. In the case of the reflux return temperature this is a different matter. Whereas for new designs the reflux return temperature can be conservatively chosen above the dew point temperature, existing plants normally operate some 20 to 30°C below the dew point temperature of 80 to 90°C, apparently without great operational problems. Some plant data are given in the table below:

**Table 3: Temperatures** (Dew point temperature of steam between 80 and 90°C)

Number of locations	Column top temp.		FBP heavy gasoline		TCR return temp.		Aux. reflux temp.	
	average	min.	average	min.	average	min.	average	min.
3	146	126	217	211	<i>No TCR</i>		59	53
8	121	100	208	177	65	30	<i>No aux. reflux</i>	
4	108	80	178	165	54	49	48	36

Whether low top and return temperatures will really create operational problems depends heavily on the chloride content of the feedstocks being processed. The safe

Whether low top and return temperatures will really create operational problems depends heavily on the chloride content of the feedstocks being processed. The safe combination of minimum TCR return temperature and top temperature can only really be set for each unit based on its particular operating experience. It is possible that up to now not too many problems have occurred because the FBP's of naphtha are still relatively high. However, if there is a tendency in the future to run at low FBP's, maintaining low top circulating reflux return temperatures and if feedstocks are processed with high chloride contents, refineries will probably have to face more frequent operational problems in their fractionator top circulating reflux sections. For processing of heavier feedstocks (including long residue) good salt removal for this aspect is important.

Given the above it will then be advisable to review/monitor operating conditions in the fractionator and check whether adequate facilities have been installed and procedures have been established to carry out water washings as a routine operation if necessary.

## 5 Optimising LCO Yield

If refineries wish to maximise the LCO yield at the expense of the gasoline yield, the constraint for the light end of the LCO is the flash point specification. For this study a flash point specification of 68°C (184°F) has been used.

At an average LCO stripping steam percentage of 2.6% wob it can be seen from Figure 8 that the minimum gasoline-minus amount at which the minimum flash point specification is met is 1030 t/d. The LCO yield is then 2050-1030=1020 t/d and the matching gasoline FBP is 167°C. By increasing the percentage of LCO stripping steam to 4.0% wob it is possible to increase the LCO yield by 24 t/d (1.2% wof); see right hand axis of Figure 13. The application of 70 tcal/ton of additional reboiler duty will increase the LCO yield by another 29 t/d (1.2% wof). This increase is at the expense of  $70 \times 1072 = 75,000$  tcal/day high-level heat from the desuperheating loop (equivalent to MP steam). About 55% of this reboiler temperature is increased as well in the reboiler case. The value of this recovered heat depends on the actual line-up of the LCO in the heat recovery scheme. Depending on the price differential between LCO and gasoline, it should be possible to justify the application of LCO reboiler duty. The remaining part of the LCO reboiler duty should be removed in the TCR section. For the above-mentioned case, this is 33,000 tcal/day or an increase of the TCR duty by approx. 14%. As the TCR duty is already high, due to the low gasoline cut point operation, the problems in the TCR section are increased further (see also Section 4).

Another possibility for increasing the sharpness of separation between the LCO and gasoline is the opening of the bypass reflux in the LCO draw-off line, as discussed in section 1.6. However, for this type of operation the MCR duty is shifted to the TCR loop and has, therefore, the same disadvantage as the above-mentioned reboiler option.

The sensitivity of the LCO upper cut point to the LCO flash point is less than that of the LCO lower cut point but can still be used in a valuable way. In the specific case if it would be possible to increase the LCO ASTM 95 specification from 355 to 365°C, the LCO flash point would be improved by 3.5°C (see Figure 7). Besides the 85 t/d of LCO yield improvement due to the shift from HCO to LCO an additional 35 t/d of gasoline can be shifted to the LCO before meeting the LCO flash point specification again (see Figure 8). The total LCO yield gain for the above increase of LCO ASTM 95 is therefore  $85 + 35 = 120$  t/d (=5.2% wof).

By using the LCO reboiler, without the application of any stripping steam, a dry LCO product can be obtained, avoiding a drier and consequent slugs make. Also usually reboiled LCO strippers are equipped with somewhat more stages than steam strippers.

## **Appendix 1**

### **Process data for Figure 1**

For all Figures, except stated otherwise, the base run from which the perturbations have been made is defined as follows:

Fresh feed intake	= 2300 t/d
HCO recycle	= 460 t/d (CFR = 1.2)
Feed to fractionator	= 2753 t/d (including 98 t/d steam)
Feed temperature	= 480 °C
Bottom temperature	= 360 °C
Yield LCO	= 770 t/d
Yield gasoline-minus	= 1280 t/d
Wash oil flow	= 124 t/d
Slurry recycle flow	= 4250 t/d
TCR flow	= 6565 t/d
Auxiliary reflux	= 0 t/d
Bypass reflux	= 0 t/d
Extra HY gasoline	= 0 t/d
HCO stripping steam	= 2.0 % wob
LCO stripping steam	= 2.6 % wob
Slurry stripping steam	= 0 % wob
LCO reboiler duty	= 0 tcal/day

## Appendix 2

### Ammonium Salts Sublimation

The following data give information on the sublimation of ammonium salts from the vapour phase. Particularly the deposition of ammonium chloride ( $\text{NH}_4\text{Cl}$ ) can give fouling problems in the top of CCU fractionators.

The following reactions may turn up in the top of the fractionator:



Both reactions take place at the sublimation temperature. The presence of free water is *not* required. When the fractionator top temperature is in the vicinity of the  $\text{NH}_4\text{Cl}$  sublimation temperature, the possibility of direct  $\text{NH}_4\text{Cl}$  deposition has to be reckoned with. For reaction (2) the  $\text{NH}_4\text{HS}$  sublimation temperature is much lower than that of the fractionator top temperature.  $\text{NH}_4\text{HS}$  deposits will therefore not be formed.

The sublimation temperature of ammonium salts is given by the following equation:

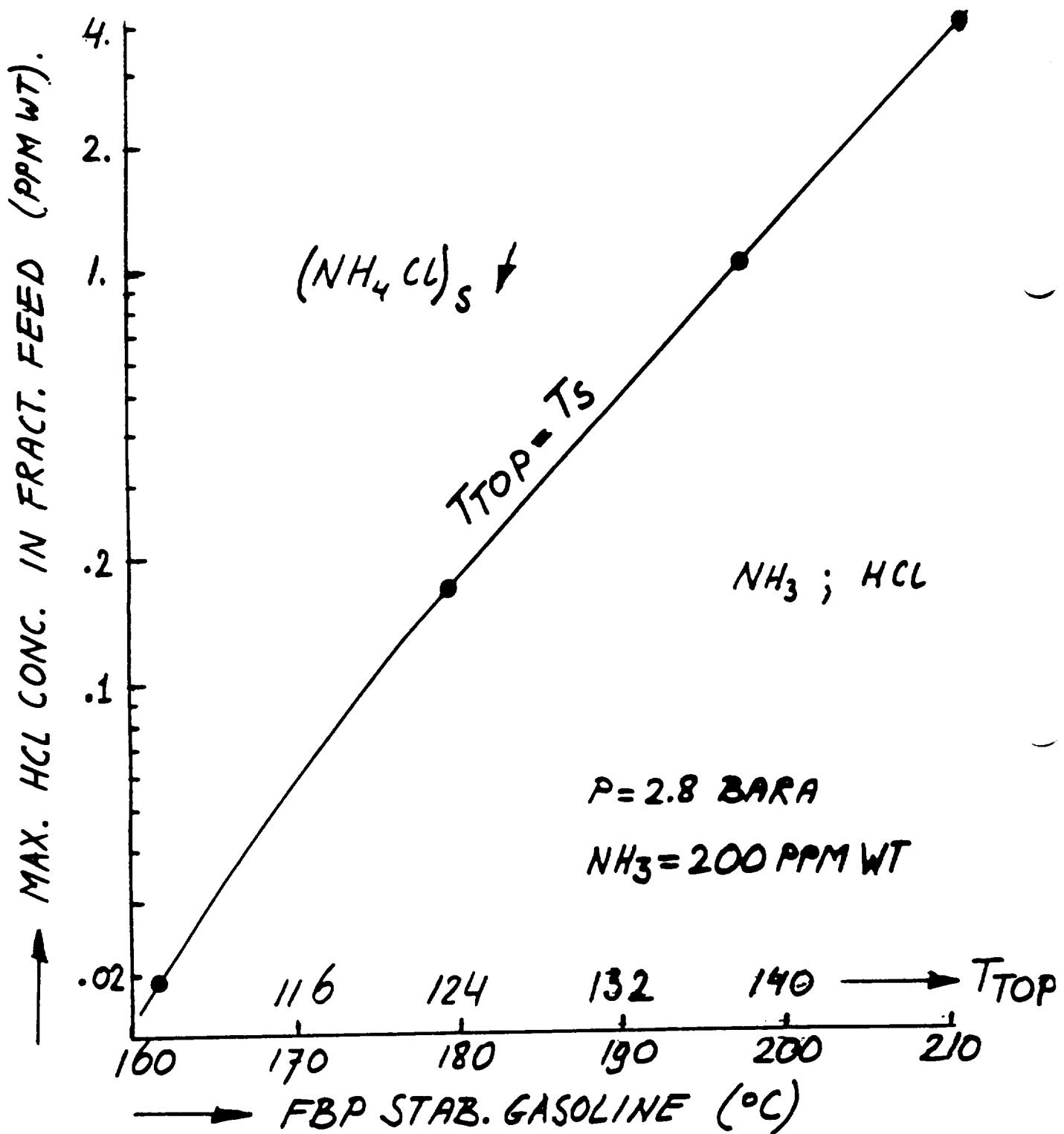
$$\ln(K_p) = \frac{-H^\circ_{\text{sub}}}{R * T_s} + C$$

where:  $K_p$  = Partial press ( $\text{H}_2\text{S}$ ) \* Partial press ( $\text{NH}_3$ ) (bar<sup>2</sup>)  
 $K_p$  = Partial press ( $\text{HCl}$ ) \* Partial press ( $\text{NH}_3$ )  
 $T_s$  = sublimation temperature (°K)  
 $R$  = 8.315 (J/° Kg mole)  
 $H^\circ_{\text{sub}}$  = heat of sublimation (J/g mole)  
 $C$  = characteristic constant of the system

Component	$H^\circ_{\text{sub}}$ (J/g mole)	C
$\text{NH}_4\text{HS}$	$90.23 * 10^3$	36.12
$\text{NH}_4\text{Cl}$	$176.1 * 10^3$	34.17

From the above formula one can now calculate the maximum allowable HCl concentration in the fractionator feed, so that the fractionator top temperature is precisely at the point at which no  $\text{NH}_4\text{Cl}$  deposits occur. In the attached Figure the maximum allowable HCl concentration in the fractionator feed is given as a function of the final boiling point of the gasoline, for a given system (given pressure,  $\text{NH}_3$  conc., amount of steam etc.). From this Figure it can be seen that, at a gasoline FBP of 210°C, 4 ppm wt HCl is allowed, while at an FBP of only 160°C the maximum allowable concentration drops to 0.02 ppm only. In the case where the fractionator feed contains 1 ppm wt HCl too much, the result will be a deposition of 105 kg  $\text{NH}_4\text{Cl}$  in the top circulating reflux section per 1000 t feed processed.

MAX HCl CONC. FRACT. FEED VERSUS FBP GASOLINE



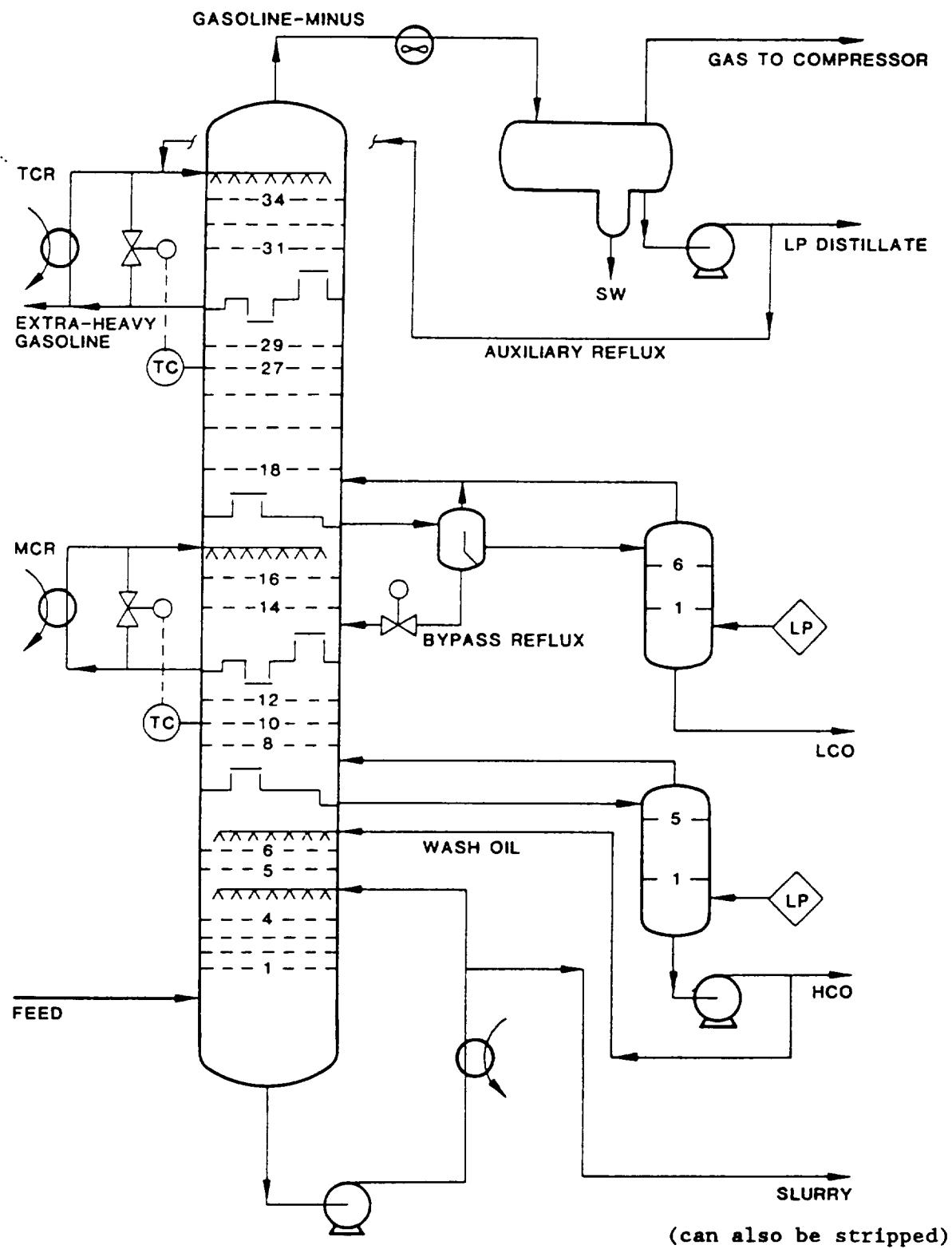
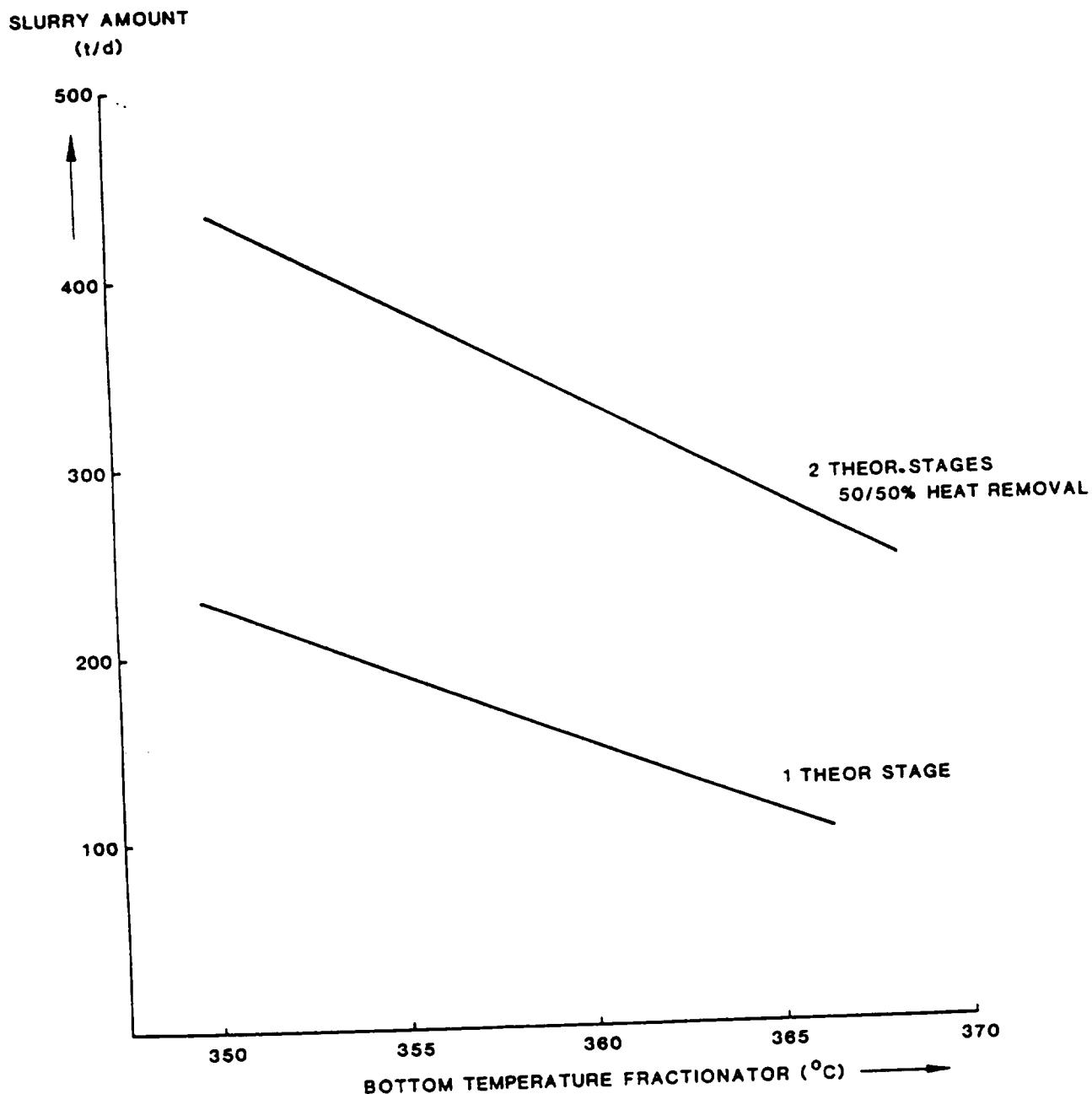
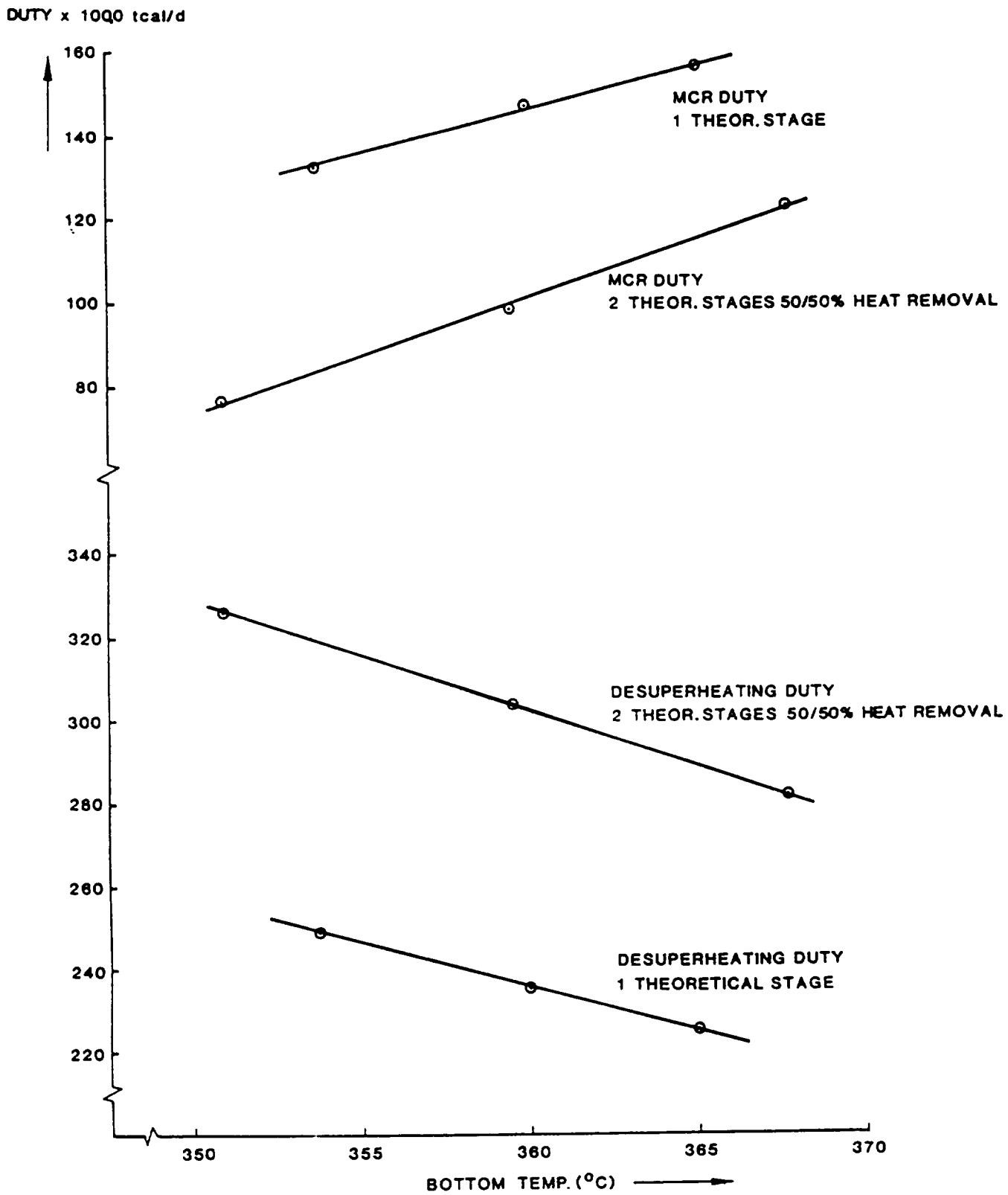


FIGURE 1 : SIMPLIFIED PROCESS FLOWSCHMIE FOR A TYPICAL  
CAT CRACKER MAIN FRACTIONATOR



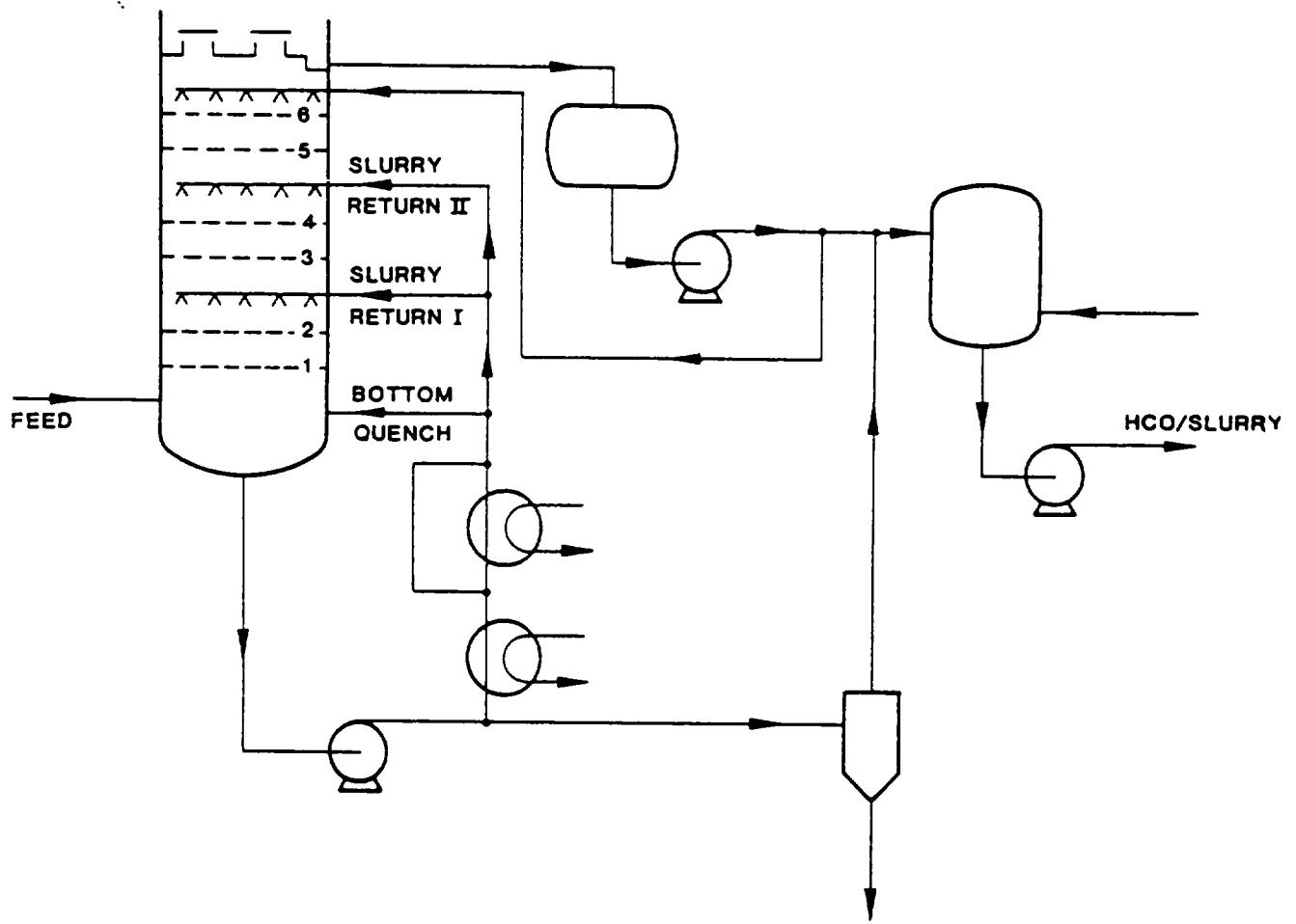
SLURRY AMOUNT AS A FUNCTION OF BOTTOM TEMPERATURE  
 REACTOR FRESH FEED 2300 t/d : FEED TO FRACTIONATOR 2756 t/d  
 (480 t/d HCO RECYCLE) WASH OIL 124 t/d.

FIG. 2



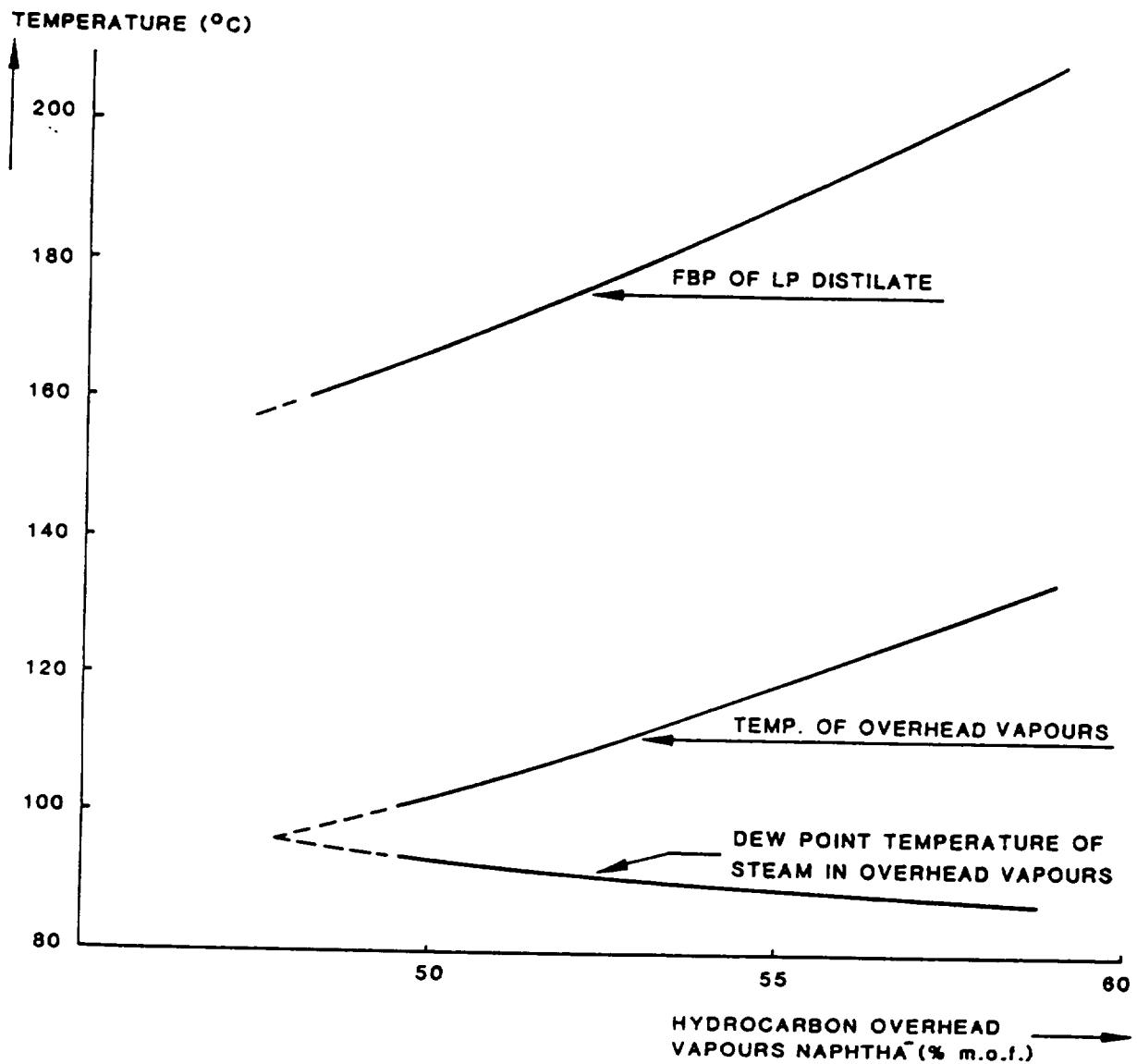
HEAT DUTIES OF SLURRY LOOP AND MCR LOOP  
 REACTOR FRESH FEED 2300 t/d : FEED TO FRACTIONATOR 2753 t/d  
 (480 t/d HCO RECYCLE) WASH OIL 124 t/d LCO<sup>-</sup> = 2049 t/d HC  
 FBP LCO 370  $^{\circ}$ C

FIG. 3



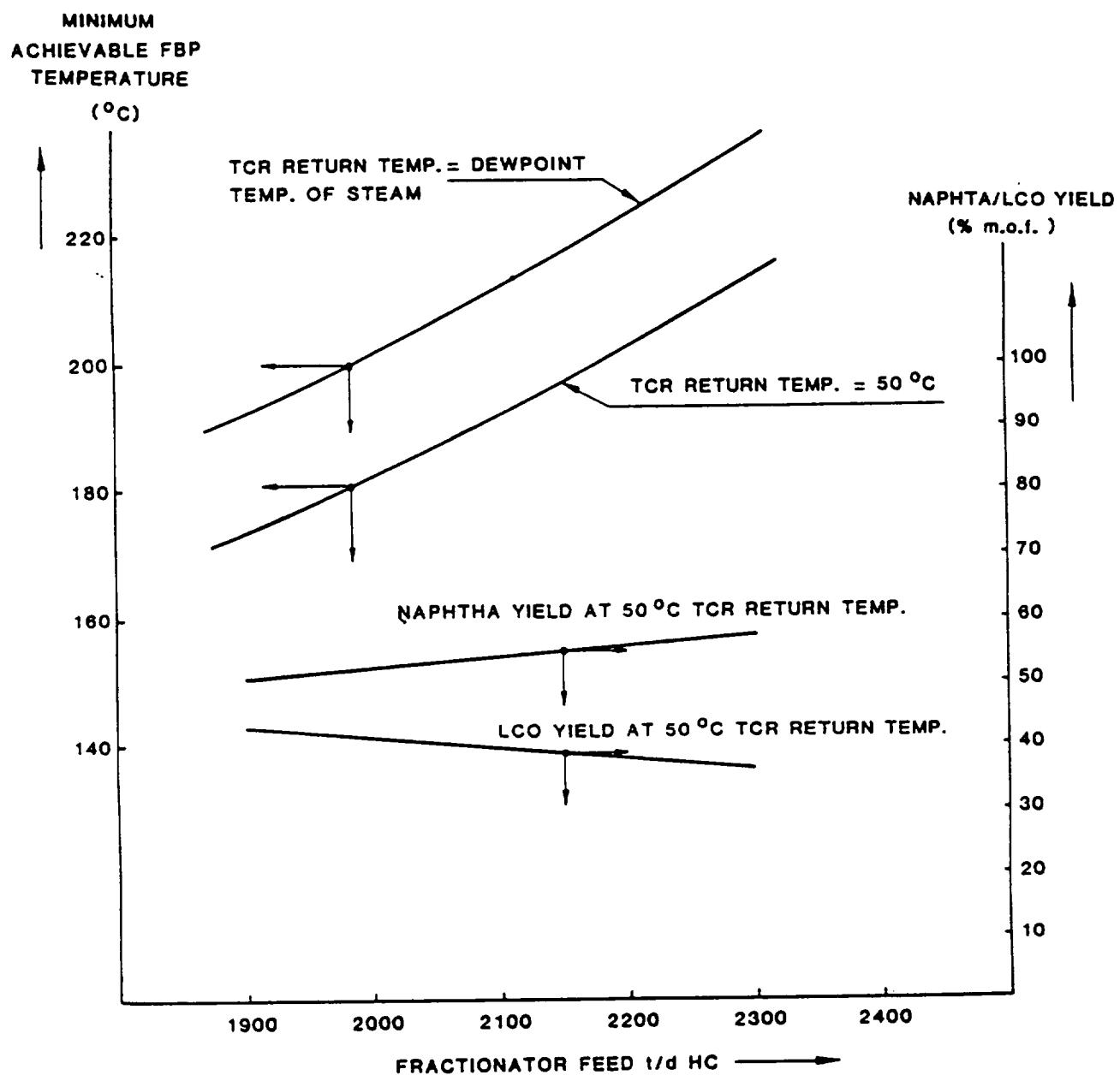
CAT. CRACKER FRACTIONATOR DESUPERHEATING SECTION  
WITH TWO SLURRY RETURNS AND A BOTTOM QUENCH

FIG. 4



TEMPERATURE OVERHEAD VAPOURS AND DEW POINT TEMPERATURE  
OF STEAM AS A FUNCTION OF FBP OF LP DISTILLATE .  
NO CONSTRAINT FROM COLUMN LOADING IN THE TOP CIRCULATION  
REFLUX SECTION .

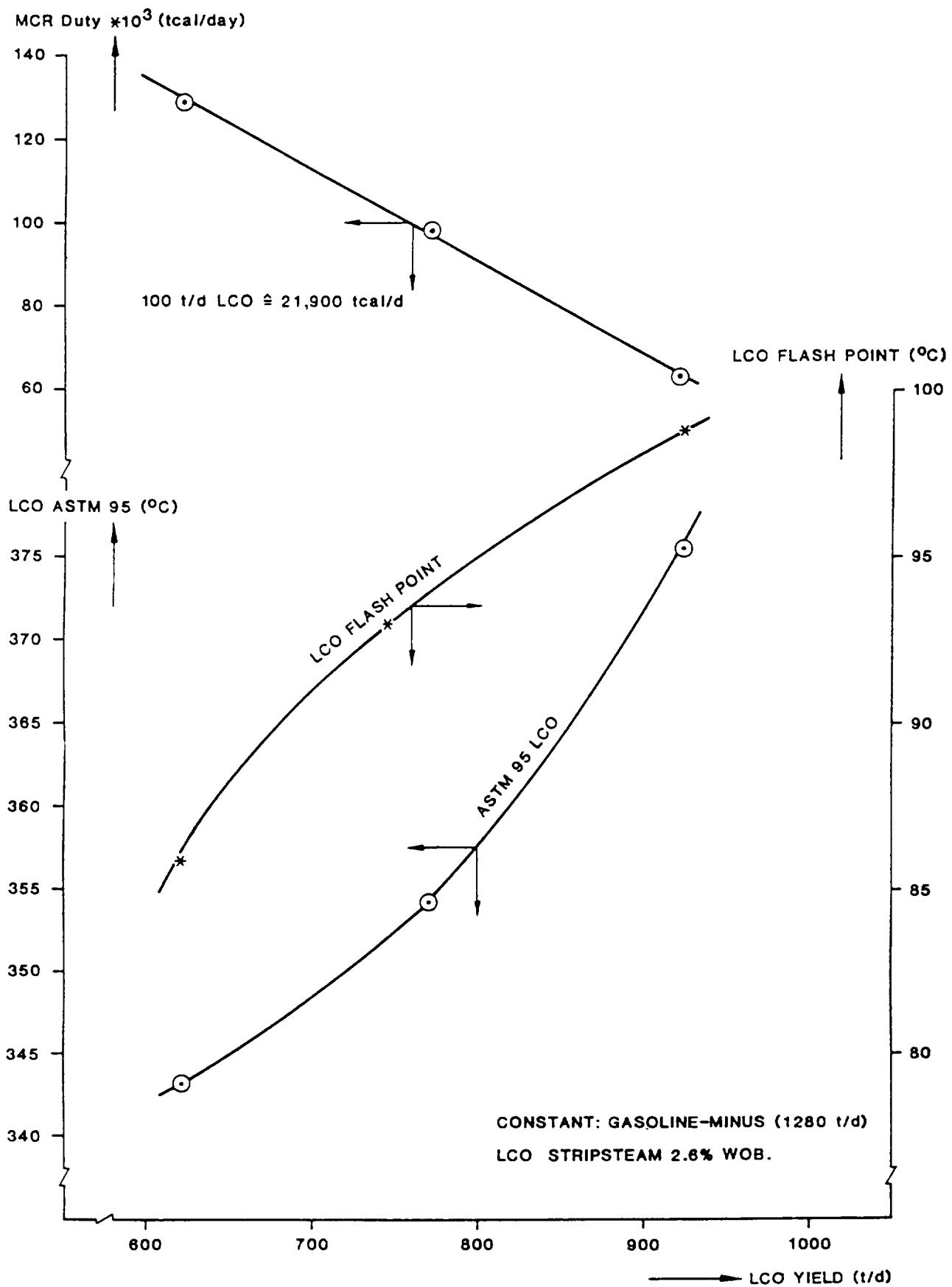
FIG. 5



Note : With a TCR return temperature of 50°C water condensation problems may occur. See section 4.

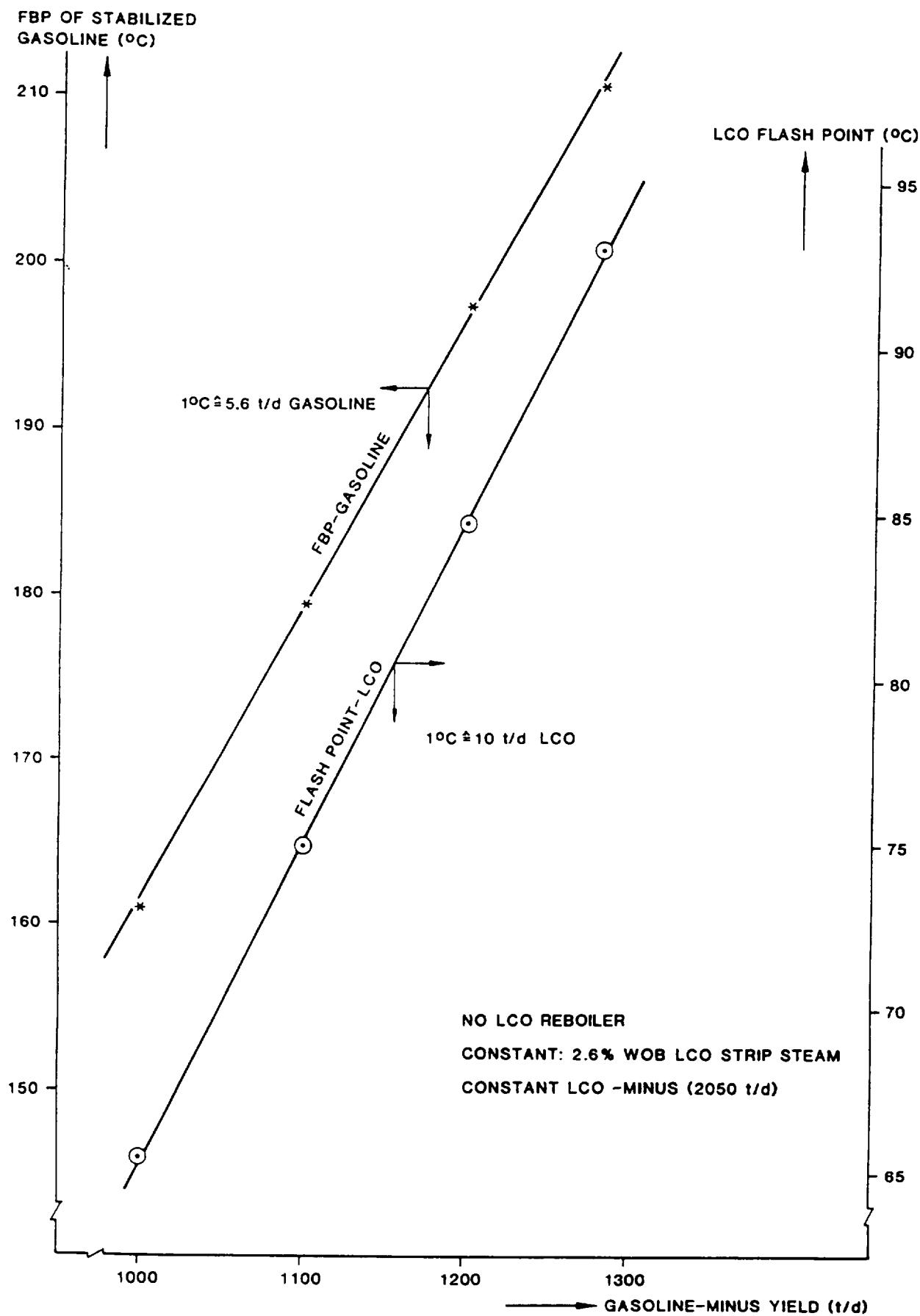
MINIMUM ACHIEVABLE FBP OF NAPHTHA , DEPENDING ON THROUGHPUT WHEN COLUMN IS CONSTRAINED BY LOADING OF TCR SECTION FOR TCR RETURN TEMPERATURE SET AT 50 °C AND AT THE DEW POINT TEMPERATURE OF STEAM IN THE OVERHEAD VAPOUR PRODUCT RESPECTIVELY

FIGURE 6



LCO ASTM 95, LCO FLASH POINT AND MCR DUTY vs. LCO YIELD

FIGURE 7



FBP OF STABILIZED GASOLINE AND LCO FLASHPOINT vs.  
THE GASOLINE-MINUS YIELD

FIGURE 8

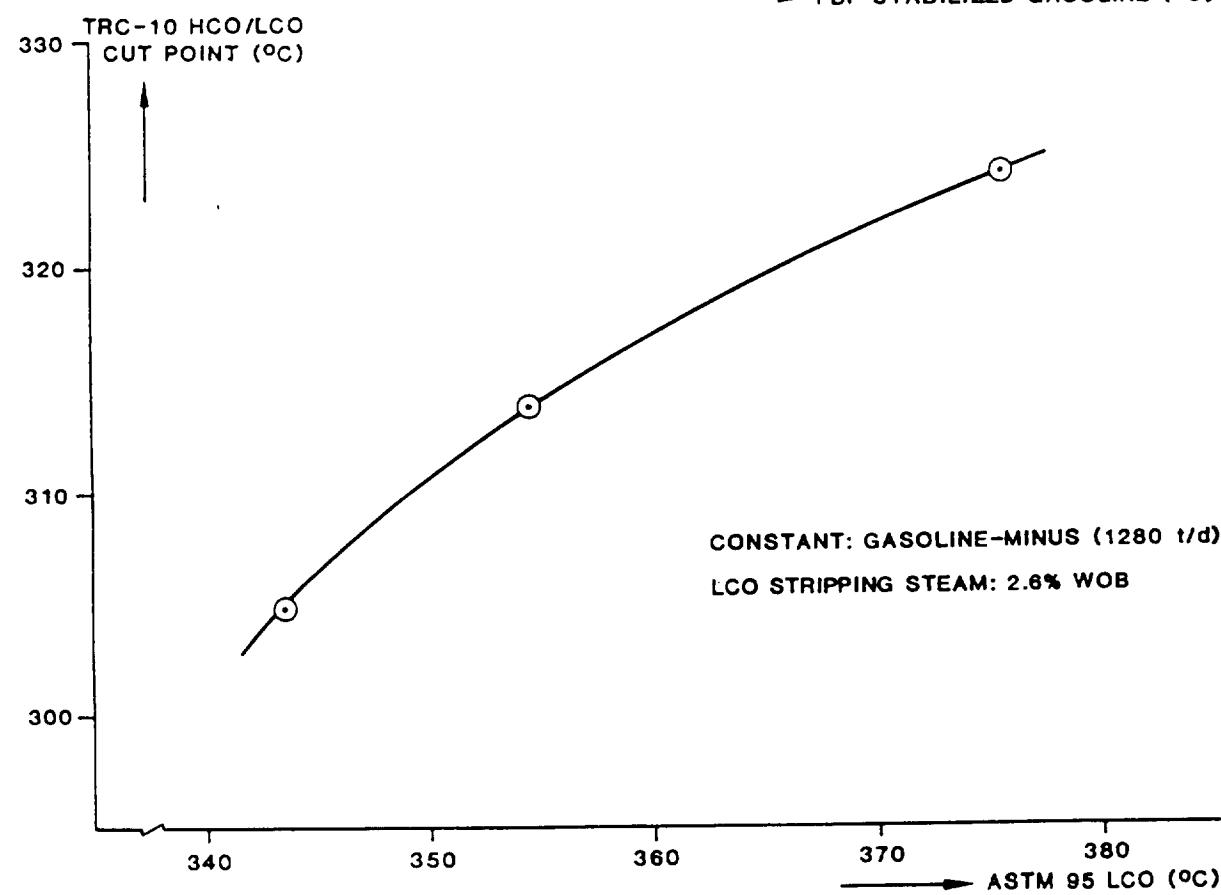
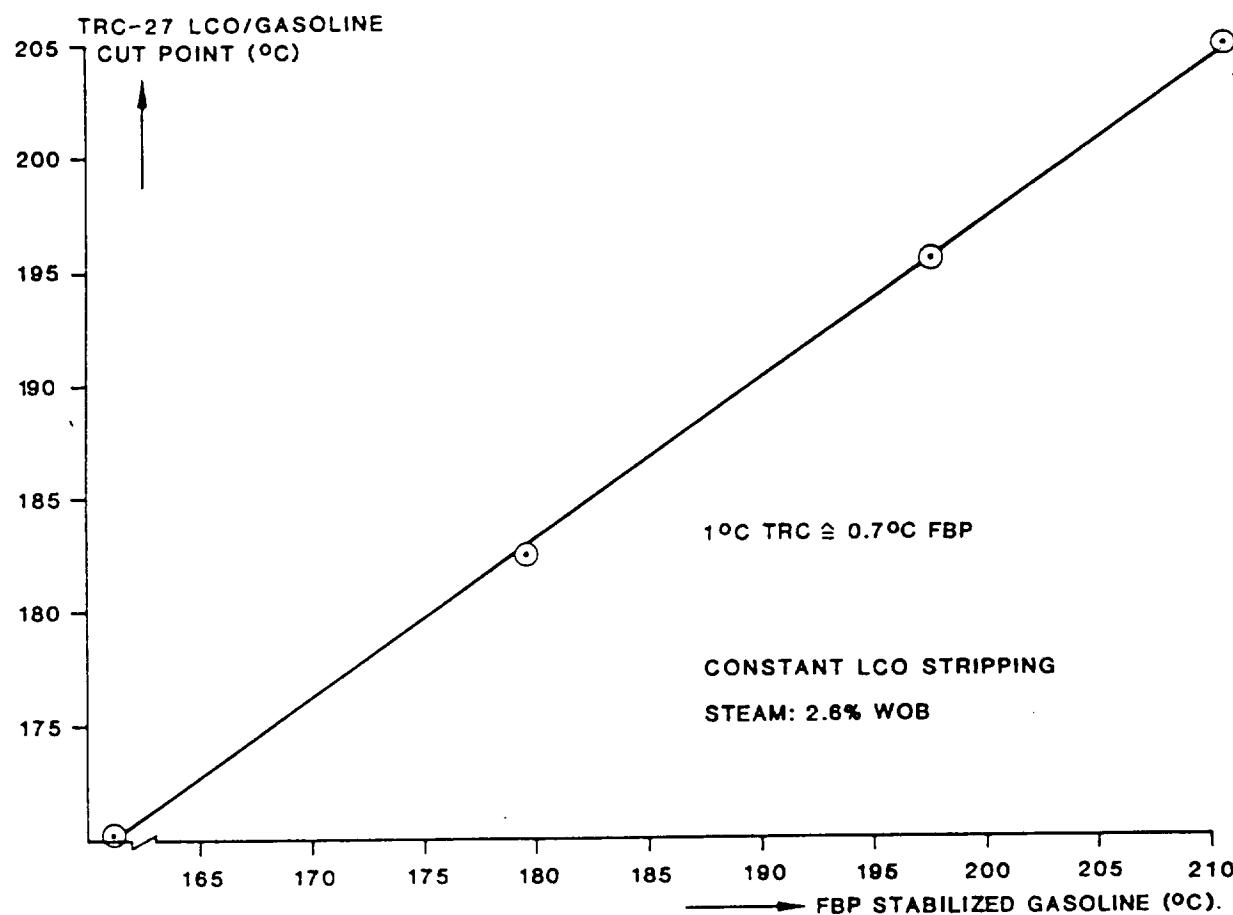
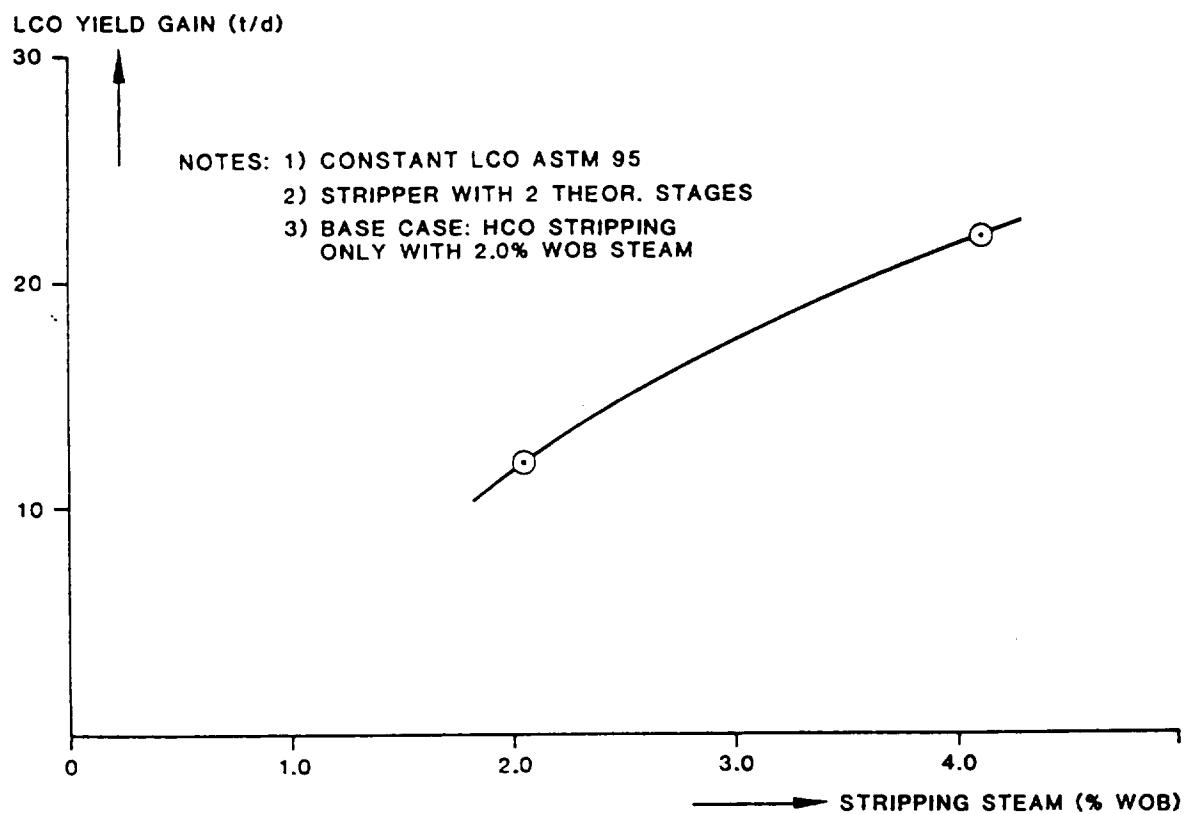


FIG. 9 TRC-27 (LCO/GASOLINE CUT POINT) vs. FBP STABILIZED GASOLINE

FIG. 10 TRC-10 (HCO/LCO CUT POINT) vs. ASTM 95 LCO

FIGURE 9,10

SLURRY STRIPPED TOGETHER WITH THE HCO IN HCO STRIPPER



HCO STRIPPING ONLY

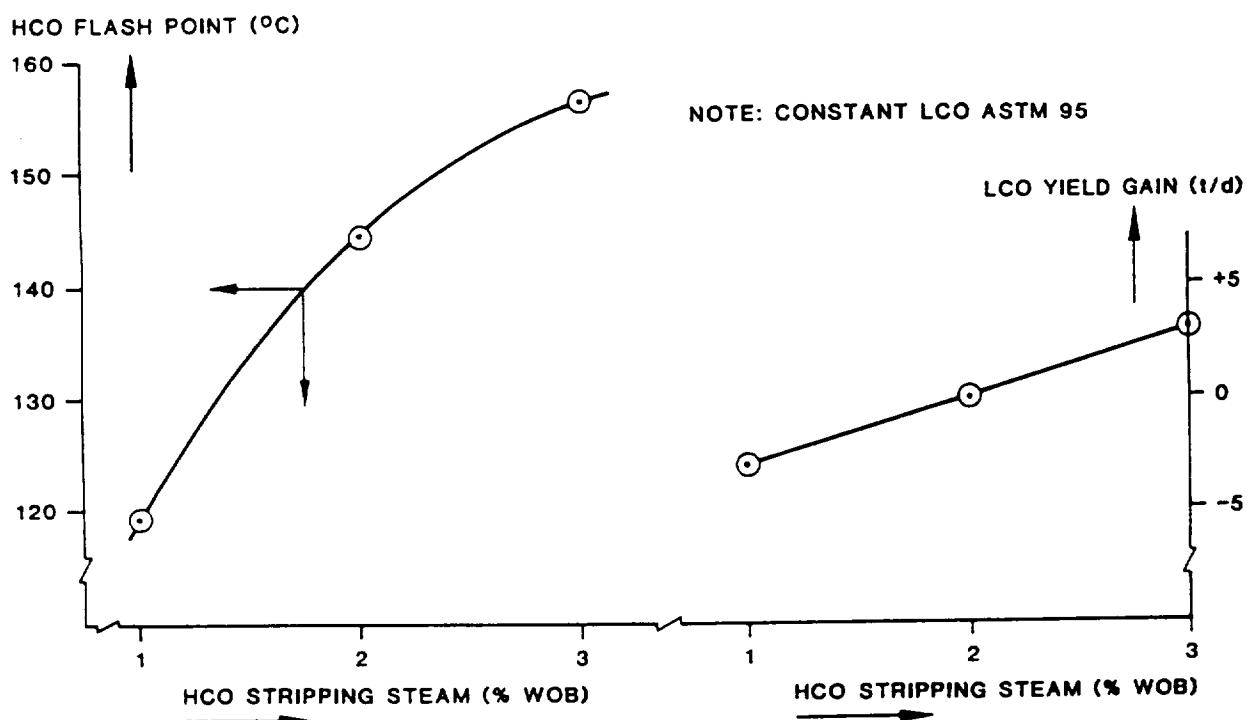
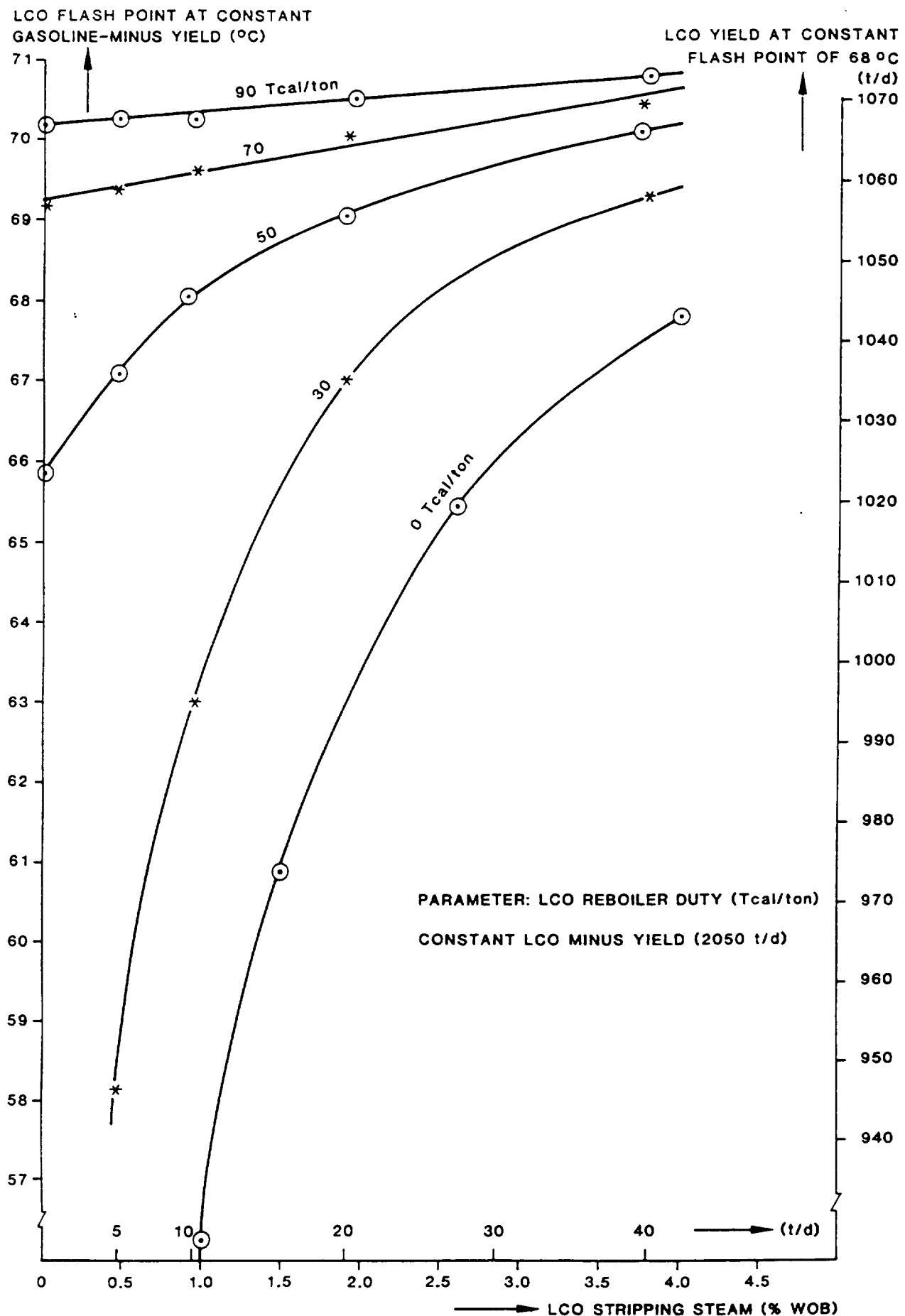


FIG. 11 LCO YIELD GAIN vs.  
STRIPPING STEAM FOR COMBINED HCO/SLURRY STRIPPING  
HCO FLASH POINT, AND LCO YIELD GAIN RESPECTIVELY vs.  
FIG. 12 HCO STRIPPING STEAM

FIGURE 11,12



LEFT AXIS: LCO FLASH POINT AT CONSTANT GASOLINE MINUS  
(1000 t/d) YIELD vs. LCO STRIPPING STEAM

RIGHT AXIS: LCO YIELD AT CONSTANT FLASH POINT OF 68°C vs.  
THE LCO STRIPPING STEAM

FIGURE 13  
MAX. LCO CASE

# **Developments of Cat. Cracker Fractionator**

## **1. Introduction**

The CCU continues to be one of the key refinery upgrading units not only in respect of new designs but also in view of the continuing revamp studies and proposals for the Group's various units. Most revamps focus on increased throughput at a higher rate of conversion, thereby stretching the existing equipment to its limits. The revamp philosophy is, therefore, logically to re-utilise the existing equipment to the maximum. Hence, only occasionally can radically new ideas be incorporated into such units.

In this paper we will look at some new design aspects with the intention to highlight key process and equipment aspects as a guide to staff involved with the design and operation of CCU fractionation facilities. Before we turn to these new designs, let us briefly describe the characteristics of the reactor effluent which forms the fractionation section feed. This effluent contains:

- Highly unsaturated hydrocarbons. The molecules are unstable and prone to polymerisation reactions.
- Substantial amounts of Hydrogen Sulphide.
- Large volumes of non-condensable gases such as Hydrogen, Nitrogen and CO<sub>2</sub>.
- Considerable amounts of steam.

The superheated vapours are at a high temperature. To stop after-cracking the stream is desuperheated below the cracking temperature range as soon as possible when it enters the column. However, the temperature should not be reduced too much because we do not want the condensation of the light components. The liquid is cooled further in the bottom compartment to prevent coke formation.

## **2. New Design Applications**

Figure 1 shows the line-up for a typical modern unit design. In this particular facility the main fractionator consists of a two stage desuperheating/slurry section, a two stage MCR section and TCR section. The fractionation sections for LCO/HCO and LCO/gasoline separation are in between the various condensation sections.

The aim of the actual design will always be to apply existing and new hardware in such a manner that the plant can be kept on stream with a satisfactory separation performance for at least the period between statutory inspections. In many cases several solutions will be possible and the designer will select the solution which at that particular moment is considered to be the optimum for the particular service. It should, however, be borne in mind that the performance often depends on such "intangibles" as coking tendency, fouling tendency and unforeseen feedstocks. Similarly, it should be borne in mind that the steady development of new catalysts will have certain impacts on the product distribution, hence, a certain built-in flexibility will be appreciated.

### **2.1 Desuperheating Section**

The hot reactor section effluent enters the lowest section of the column in a superheated state. This superheat is removed by means of slurry circulation in which

heat is exchanged against for example, splitter reboilers, CCU feed, crude unit feed and MP steam generation. The heat removal should be fast in order to prevent after-cracking and potential coke formation, thereby taking into account the presence of some (minor) entrainment of catalyst particles, depending on the efficiency of the upstream reactor cyclones. Notwithstanding this, only a minimum part of the feed should be condensed at this stage, i.e. only minimum slurry product, just enough to cater for the removal of the catalyst fines and to obtain an acceptable bottom temperature in view of possible coke formation. As shown in Figure 3, this is done by applying a section consisting of two sets of a double layer of grid trays with vertical baffles (Figure 5) and a double set of tapered slurry spiders plus a simple quench slurry inlet device in the liquid bottom phase. The two slurry circulation spiders are both of the tapered design (Figure 6) in order to maintain constant velocities and back pressure so as to ensure equal liquid distribution for all the holes and hence, prevent the creation of dead zones prone to coking-up.

A low heat and mass transfer efficiency is a common feature of grid trays. High efficiencies are not needed because ideally the desuperheating section should represent only about one theoretical stage. A higher efficiency than that assumed in the design can result in considerably more than one theoretical stage and give rise to an increased amount of slurry at the expense of HCO, together with an increased heat duty for the slurry loop. If this duty shift happens than a consequence will be a reduction in the flow of vapours rising over trays 8, 9, 10 and 11 - the HCO/LCO fractionation section (see Figure 2), being condensed and returned as internal reflux. This results in less traffic over these trays which means a reduced HCO/LCO separation. This may manifest itself as a more heavy tail on the LCO product ASTM distillation. Figure 4 shows the amount of slurry produced as a function of the bottom temperature, simulating the desuperheating section with one theoretical stage and two theoretical stages with 50/50% heat withdrawal. As can be seen, the amount of slurry at a certain bottom temperature almost doubles.

The bottom quench enables an independent reduction of the liquid bottom compartment temperature. There is often a requirement to do this set by the mechanical components of the slurry loop. It should be mentioned that the installation of a coke catcher at the bottom outlet nozzle must be considered a part of a sound design for all modern columns with semi-elliptical bottoms (as compared to some older designs with steep conical bottoms).

## 2.2 Wash Oil Section

The function of this section is to prevent entrained catalyst from penetrating into the upper parts of the column, where it could contaminate (particularly) the HCO product. Older units used trays in this section, latest thinking is to apply packings as used in High Vacuum Units. A "vane pack" device should provide sufficient catching efficiency while being robust enough for this service. The packing should be kept wet via a wash oil spray distributor system.

## 2.3 Mid Circulating Reflux Section

In the unit shown in Figure 1 the decision to install two MCR sections was based on heat integration considerations. In Table 1 we see a comparison of a conventional and double MCR unit. Also we see the heat requirements for the associated gasplant. In the conventional line up some of the debutaniser heat requirements must have come from the (high temperature level) slurry loop. In the case as given in the table at least 8.7 MW must come from the slurry loop. However, in the double MCR case the absorber

heat can come fully from the MCR II, the debutaniser heat fully from the MCR I and no heat is required from the slurry loop. Therefor there is an additional 8.7 MW of high grade slurry heat that can potentially be used to save direct fuel. Moving to a double MCR system in an existing unit does result in a reduction of LCO/gasoline separation (at constant number of trays in the fractionation section). However, in a new design a few more trays can be added to bring the separation within specification.

## 2.4 The Top Circulating Reflux Section

To avoid free water in the top of the column two requirements must be fulfilled:

- 1) The temperature of the overhead vapours should be above the dew point temperature of steam in the hydrocarbon vapour leaving the column. This requirement sets the ultimate minimum naphtha FBP which can be achieved in the fractionator.
- 2) The reflux return temperature should be above the dewpoint temperature of the steam in the hydrocarbon vapours leaving the column.

Figure 7 shows the relationship between the FBP of LP distillate, the temperature of overhead vapours and the dew point temperature of steam in the overhead vapours for a typical case. As can be seen, the minimum FBP is around 165°C. In the cases where this FBP is too high for the required naphtha production, the possibility must be considered either of installing adequate accepting condensation (see section 4 of paper 1 in this section). The latter means upgrading of tray materials in the TCR section (in order, at least, to obtain an acceptable lifetime) and the installation of facilities to cater for water condensation, salt deposition etc.

In view of the need to maintain a certain minimum TCR return temperature, the circulating reflux flow has to be such that the required heat duty is withdrawn and the return temperature is acceptable. This sets a certain TCR flow, given this flow the trays installed will have to cope with, what is often relatively high liquid volumes.

## 3. Gasplant

### 3.1 Recontacting System

The recontacting system typically applies two-stage condensation, and compression with the wet gas compressor compressing the gases to the absorber pressure level. Good recontacting between the stages is used to reduce first-stage condensate, which is relatively lean in light ends, as absorber lean oil. The condensate is send via a chiller to be used as lean oil in the absorber, thereby avoiding the far costlier application of a debutanised gasoline recycle to the absorber.

### 3.2 Rectifying Absorber

The operating pressure for the absorber is selected at such a level that the absorption factor is sufficiently high to produce an overhead product of dry gas which is lean in (recoverable) LPG components. In order to reduce this pressure level, chilling of the lean oil/overhead reflux system is applied. The application of the chilling system can result in a saving of typical 15 - 20% of the energy otherwise consumed by the wet gas compressor due to the lower absorber operating pressure. The energy for the chilling system proper is low level heat delivered from the main fractionator TCR in the form of hot water or LLP steam. In order to optimise the heat integration, heat input to the

bottom part of the absorber can be at different levels. Water is removed in a conventional large liquid-filled settler.

### **3.3 Debutaniser, Depropaniser**

Both columns typically apply well known technology in the form of calming-section Hifi trays, thermosyphon reboilers and low-pressure operation commensurate with the recovery requirements for light products, without applying chilling at this stage (this may be introduced later as an increased recovery/debottlenecking exercise).

## **4 Conclusion**

The benefits of a new approach of CCU main fractionator can be considerable. For example Table 2 shows the overall heat balance for 10,000 t/sd feed intake. The heat supply of an “old” and “new type” of unit is similar but there is a considerable reduction on waste heat cooling. The difference of 41 MW is equivalent to some 90 t/sd SRF direct fuel saving. In addition there are savings on wet gas compressor power requirements due to the absorber chiller.

Here we see some of the advantages of a modern approach the unit design and integration which is based on constant development of large and small items of technology.

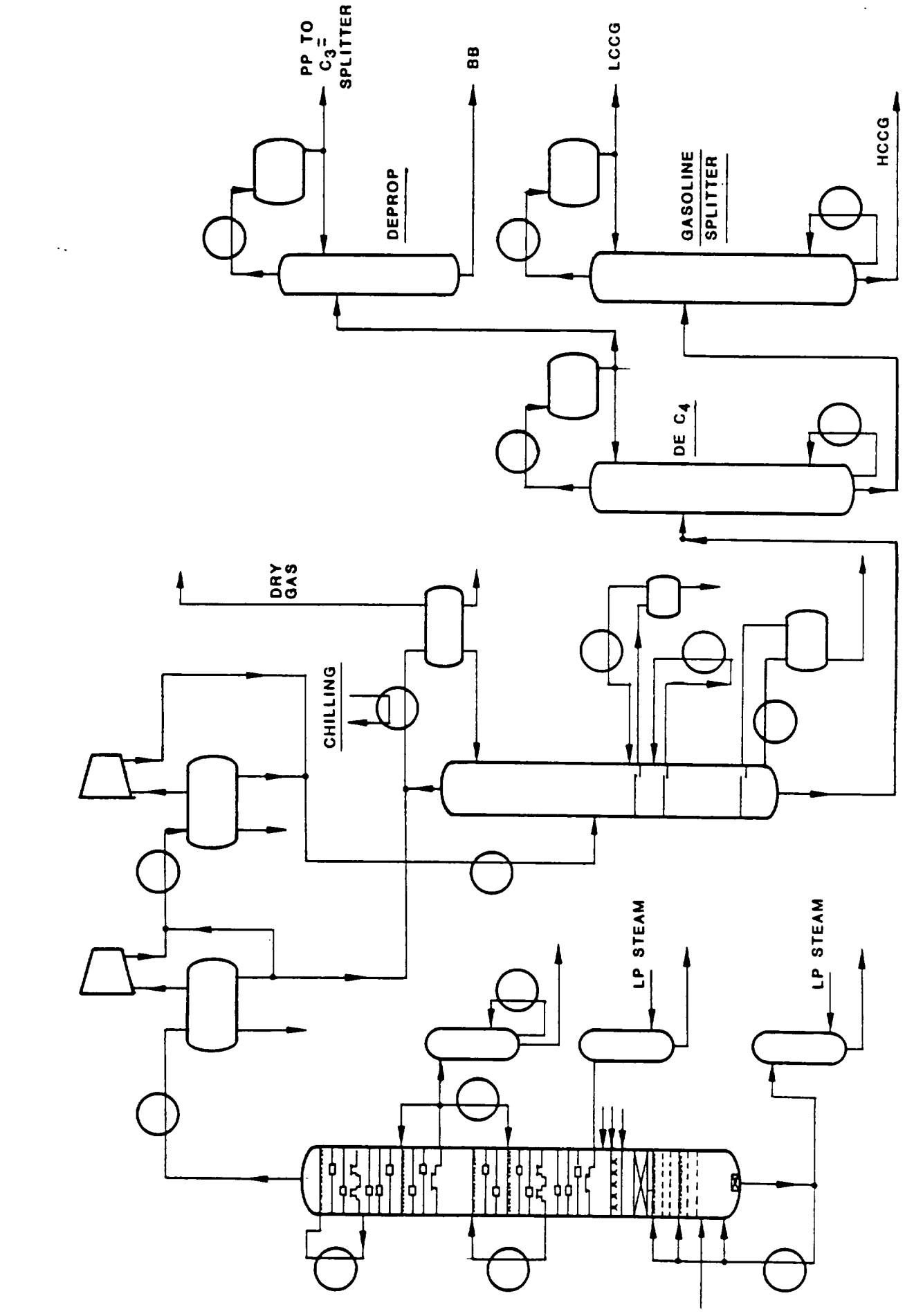
**Table 1 - Heat duties**

Conventional			Double MCR		
	MW	Temp. (°C)		MW	Temp. (°C)
TCR	34.5	110 - 150	TCR	15.1	120 - 150
			MCR II	19.4	160 - 220
MCR	22.0	240 - 290	MCR I	22.0	240 - 290
Slurry	53.2	300 - 360	Slurry	53.2	300 - 360

Gas Plant heat requirements		
	MW	Temp. (°C)
Depropaniser	5.3	90 - 91
Absorber	11.5	120 - 150
Debutaniser	19.2	165 - 195
Chiller	4.5	90 - 100

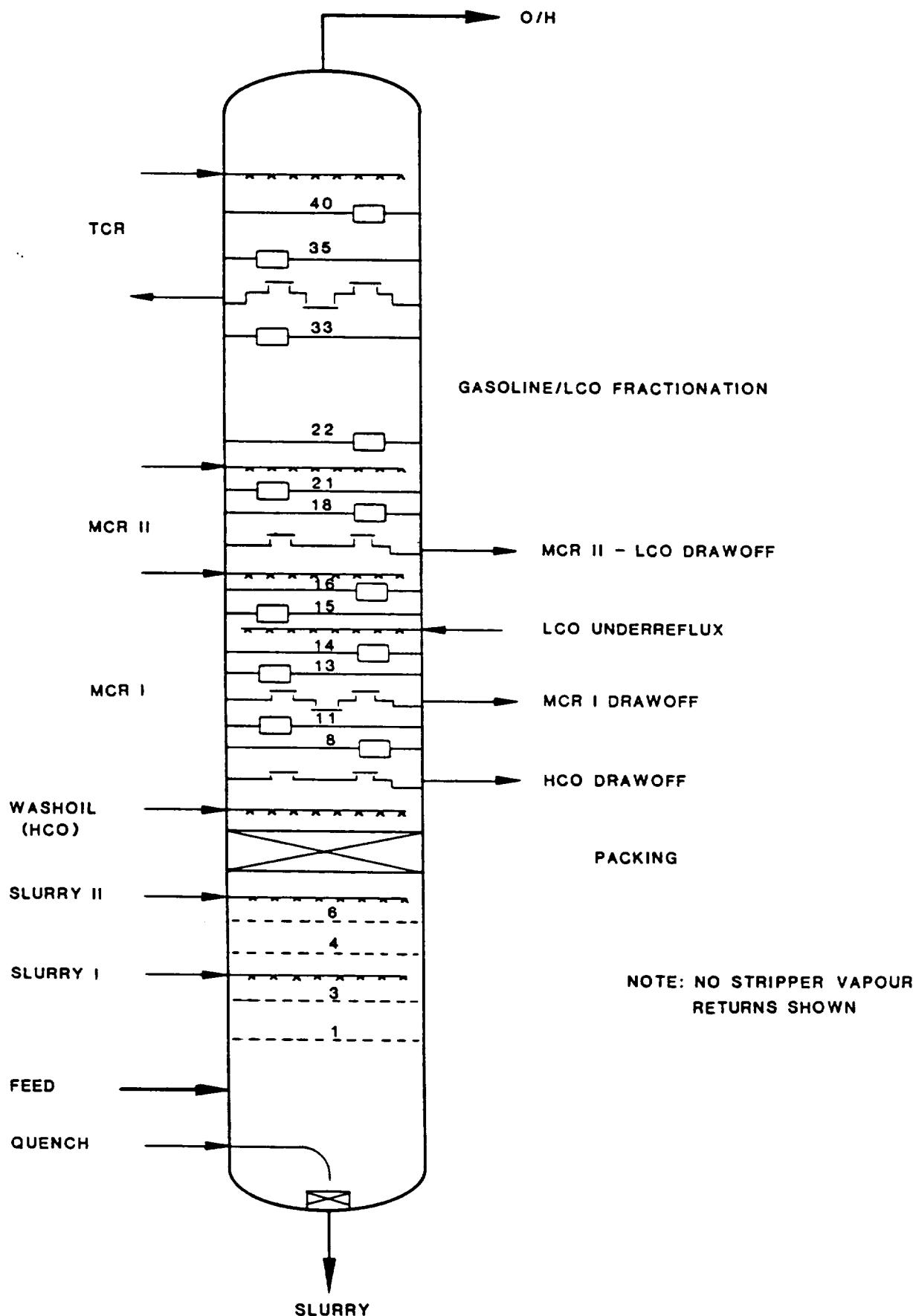
**Table 2 - CCU Fractionator heat balance**

Feed intake 10,000 t/sd	New (MW)	Old (MW)
Heat supply	190	181
High grade heat export	-50	-35
Medium grade heat export	-37	-2
Cooling	-103	-144



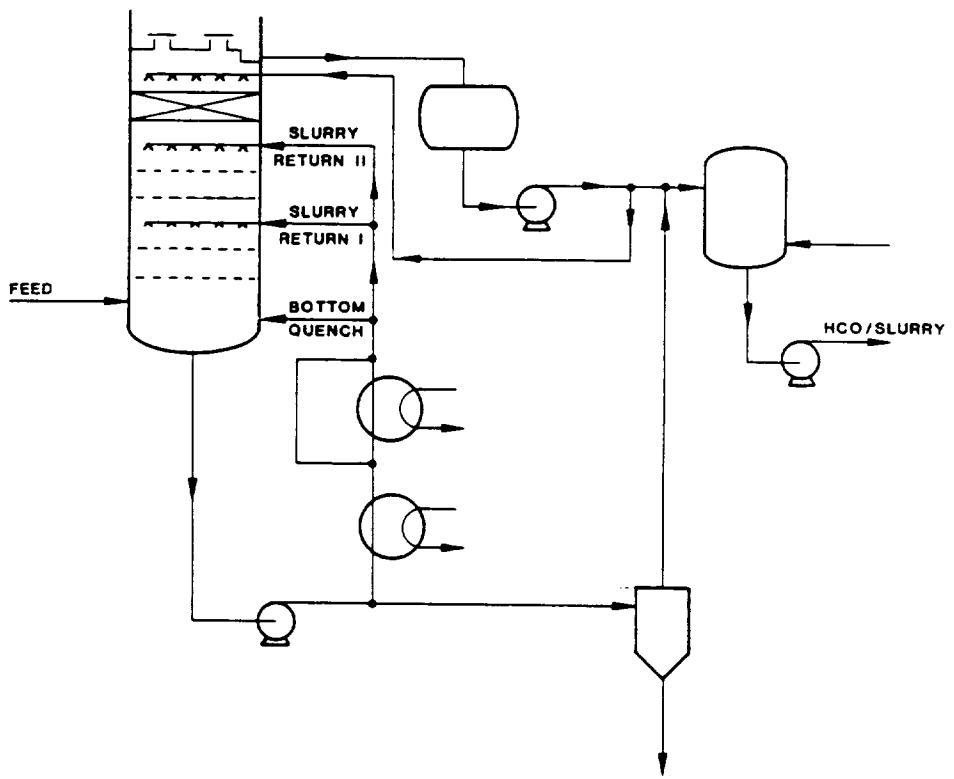
SIMPLIFIED PFS OF CC FRACTIONATION PLANT

FIG. 1



CAT. CRACKER MAIN FRACTIONATOR

FIG. 2



CAT. CRACKER FRACTIONATOR DESUPERHEATING SECTION  
WITH TWO SLURRY RETURNS AND A BOTTOM QUENCH

FIG. 3

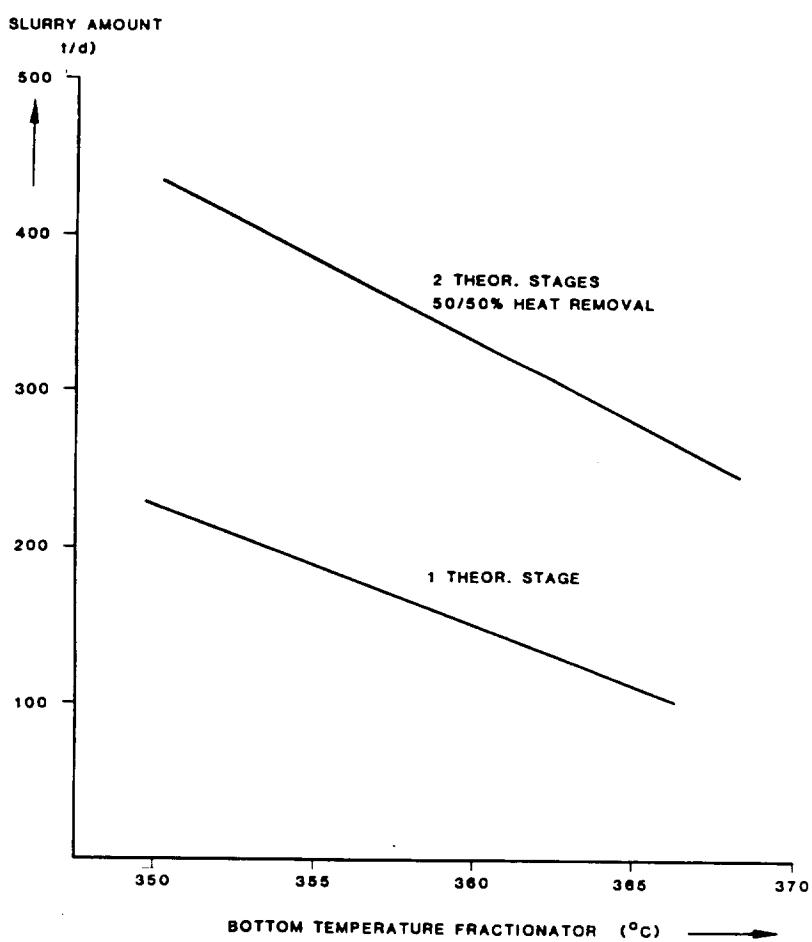


FIG. 4

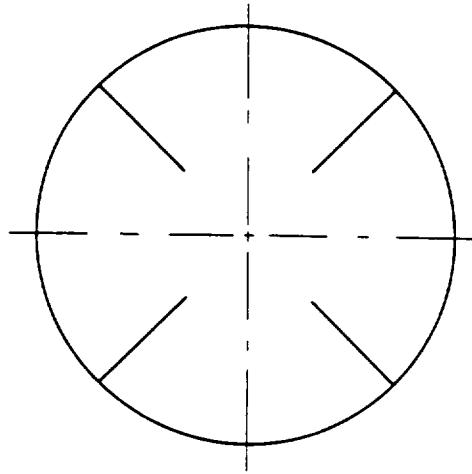


FIGURE 5 : ROUND BAR GRID TRAY WITH VERTICAL BAFFLES

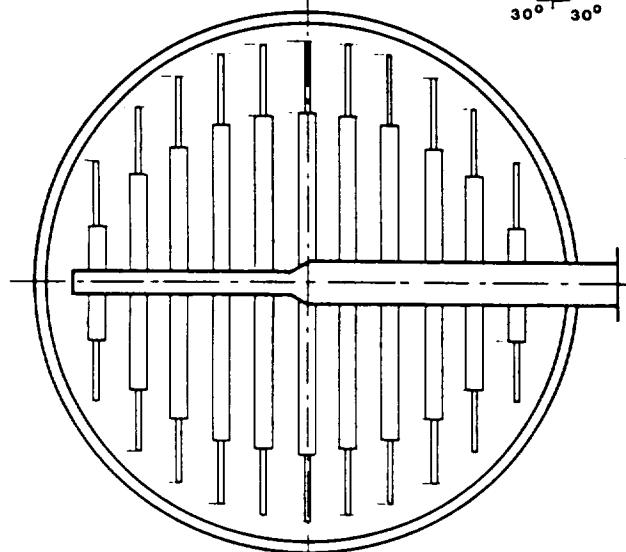
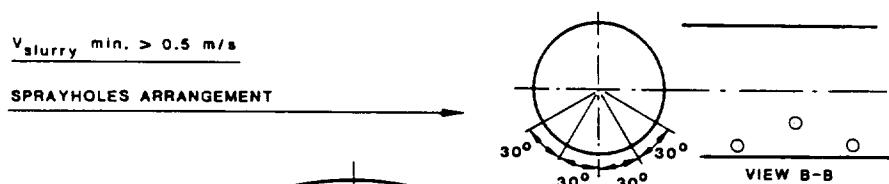
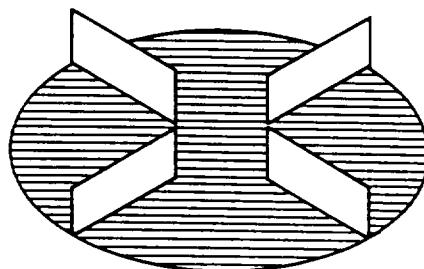
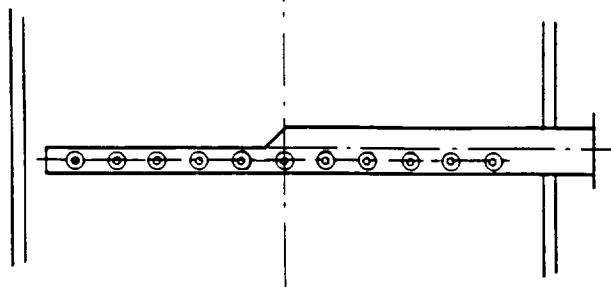
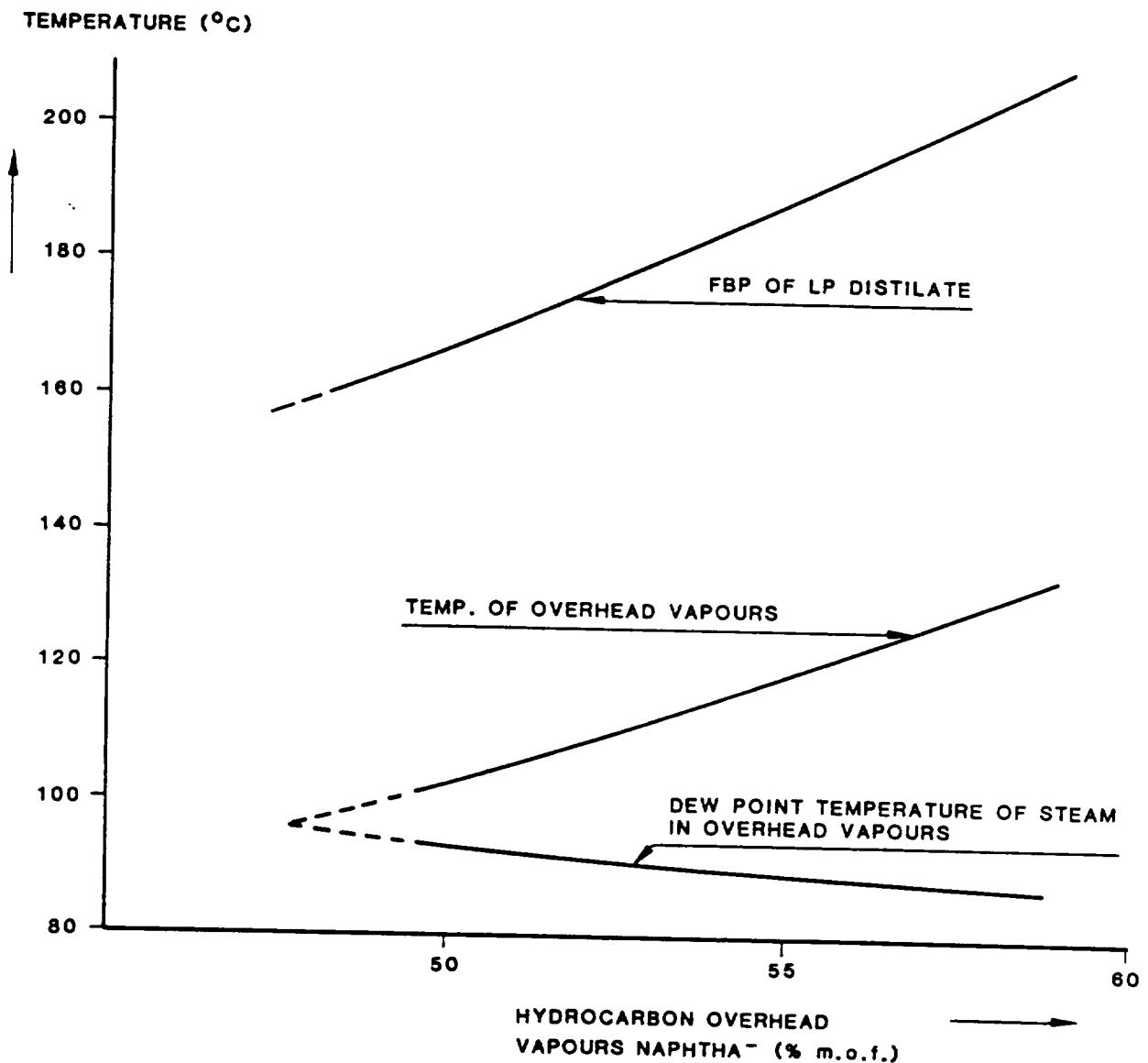


FIGURE 6 : TAPERED SLURRY SPIDER FOR DESUPERHEATING SECTIONS





TEMPERATURE OVERHEAD VAPOURS AND DEW POINT  
TEMPERATURE OF STEAM AS A FUNCTION OF FBP  
OF LP DISTILLATE

FIG. 7

# **Rectifying Absorber**

## **1 Introduction**

When the overhead stream from a column contains a large amount of light and volatile components such as C1, C2 etc., condensation (via cooling water or air) to obtain reflux is not usually practical. Increasing the column operating pressure (and hence temperature) in order to enable the top vapours to be partly condensed against the cooling medium tends to result in very low relative volatilities between components resulting in very high numbers of stages and reflux ratios required to effect the desired separation. Application of a cold enough utility to condense the overheads at moderate pressures often leads to the requirement for expensive refrigeration systems. Moreover “deep cold” cannot be applied for wet gases without the problem of water freezing out on the inside of heat exchangers etc.

In these circumstances a better alternative is to use an absorption system. Basically an external (cooled) absorption oil stream is introduced as extraneous top reflux to the column. This absorption oil is chosen to have a higher boiling range than that of the overhead vapours. The extraneous reflux enters the top of the column as “lean oil”, absorbs some of the light hydrocarbons that would otherwise go overhead and leaves the column bottom with the absorbed material as “fat oil”. By reboiling in the bottom of the column the required fractionation and light component content of the bottom product can be achieved.

## **2 The Cat Cracker Gasplant Absorber**

In principle absorbers can be used for any system that is required to be split into light-end components as top product (which cannot be condensed against the cooling medium) and a remaining part as bottom product. However in the refinery environment the Cat Cracker Gasplant Absorber is the most typical example (see Figure 1) and we will discuss this unit now in some more detail.

The effluent from the Cat Cracker reactor section is first fed to a main fractionator column (similar to a CDU main column) where heavy fractions (slurry, cycle oils) are separated. The overheads stream from the main fractionator contains a large percentage of light components (propane, propylene, ethane, ethylene, methane, H<sub>2</sub>, H<sub>2</sub>S and some inert such as N<sub>2</sub>, CO, CO<sub>2</sub>) as well as the gasoline (C5 - 220°C) fraction. This stream is partially condensed, the condensate being pumped to the absorber column, the remaining vapours being routed via a (usually 2 stage) compression system to the absorber which typically operates with a pressure of around 20 bar.

### **2.1 Description of the Process**

The absorption process involves counter current contacting of rising vapours with the descending lean oil in the absorber column. The heavier components in the rising vapours are preferentially absorbed into the lean oil and leave the bottom of the column with the lean oil. The remaining light components leave the top of the absorber as a vapour stream, usually referred to as “dry gas”. From a process flow view point lower temperatures, higher pressures and high lean oil flow rates will contribute to

reduced loss of valuable components from the top of the column. However for each particular unit there will be various constraints on these process conditions such as:

Temperature	-	Heat exchanger capacities
Pressure	-	Design pressures, compressor capacities
Lean oil rates	-	Column, pump capacities

In the typical conventional line up as shown in Figure 1 the cooled lean oil is introduced to the top tray of the absorber. In this system the lean oil supply is taken from the bottom of the downstream debutaniser and is therefore stabilised Cat Cracked full range gasoline. Although this lean oil is relatively heavy hydrocarbon fraction it is inevitable that some of the lean oil is lost to the dry gas leaving the top of the absorber. The vapour leaving the top of the column will be at dew point and it will take some components of the lean oil with it in order to fulfil the dewpoint condition, at the prevailing temperature and pressure conditions. To reduce these losses, in the conventional line up, the dry gas is routed via a small "sponge oil" section. This is a small separate absorption column containing a few trays. The dry gas is counter currently contacted with a cooled heavy oil fraction (such as light cycle oil from the main fractionator system). The enriched sponge oil containing lean oil and heavier dry gas components is returned to the main fractionator. The ultimate effect of the sponge oil section is to increase somewhat the overall recovery of valuable hydrocarbons.

The bottom section of the column (the stripping section) is usually reboiled. The light ends content of the bottom product may be controlled by adjusting the column reboil ratio. It should be noted that any C2 components that slip through the bottom of the absorber will be found back in the top of the debutaniser and depropaniser columns downstream where they may put products off spec. or upset column operation.

The bottom product from the absorber is routed to the debutaniser. This column operates at a lower pressure (typically 10 bar) and as a result part of the hydrocarbons flash off in the feed line. In the debutaniser fractional distillation is achieved by splitting the feed into a C3/C4 top product and a stabilised gasoline as bottom product. Part of the bottom product is recycled as lean oil to the absorber after being cooled down, the balance is sent to storage or to further gasoline splitting facilities.

## 2.2 Absorber and Gasplant Products

Only the line up of a typical conventional Cat Cracker Gasplant Absorber system is shown in Figure 1. The absorber will be accompanied in the gasplant by a debutaniser (to separate the C3/C4 fraction from the gasoline fraction), a depropaniser (to separate the C3's from the C4's) and often propane/propylene and gasoline splitting columns.

## 3 Miscellaneous

Wash water may be injected upstream of the absorber feed coolers, and in the top of the column to protect against equipment fouling by salts deposition, corrosion and also to reduce the cyanide concentration in the dry gas to prevent problems in the downstream gas treating facilities.

Water draw off facilities are provided in the column to draw off the water/hydrocarbon mixture, separate off the water and return the hydrocarbon phase to the column again.

The H<sub>2</sub>S present in the hydrocarbon feed to the absorber will distribute between the dry gas and the bottom product where in downstream processing it will go with the C3/C4 fraction. The dry gas is usually treated in an amine type system, the C3/C4 is usually treated in a caustic type treater.

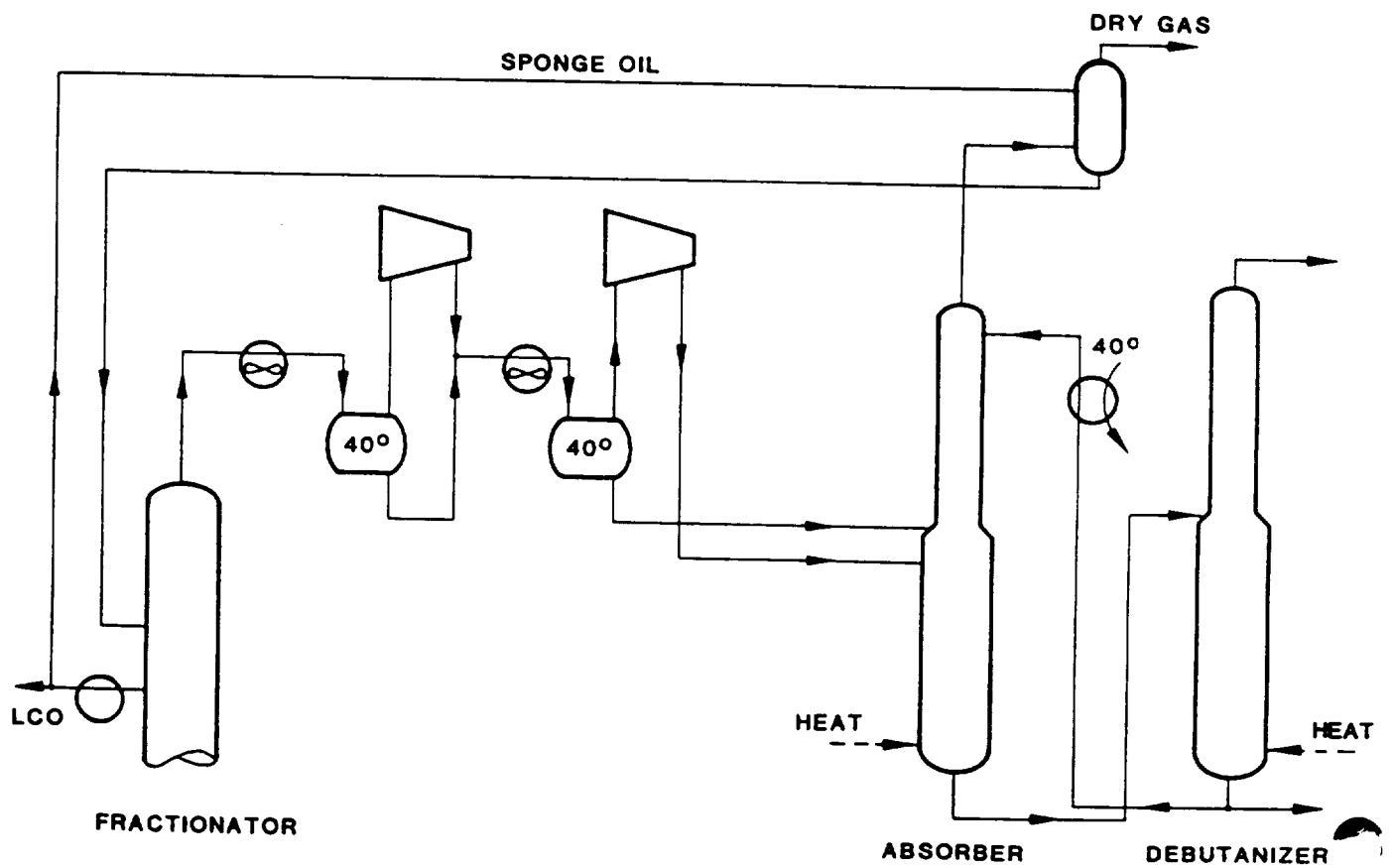


FIGURE 1 TYPICAL CONVENTIONAL LINE - UP

# Cat Cracker Absorber Operation

## 0 Summary

The use of debutaniser bottom product as lean oil and light cycle oil product or mid-circulating reflux stream as sponge oil for the absorber of a typical cat. cracking unit presents several disadvantages. Considering options to improve the operation of existing units reveals that the application of chilling in conjunction with using low-pressure distillate from the fractionator overheads to replace debutaniser bottom product as lean oil obviates the use of sponge oil and can offer substantial savings in absorber reboiler heat and compression energy for the wet gas compressor. It can also result in a significant reduction in both absorber and debutaniser column loadings, at the same time, enhancing C<sub>3</sub> recovery. There are several schemes for the application of chilling the selection of which should be determined by an economic evaluation exercise.

## 1 Introduction

The typical line-up of a cat. cracker fractionator overhead system and the absorber consists of the following features (Figure 1):

- a) The low-pressure distillate (LPD) condensed by air cooling or water cooling in the fractionator overhead condensers is sent to recontact with the compressed O/H vapour. The mixture is cooled by air or water and enters a vessel where the gas and the enriched liquid (HP Dist.) are separated. The gas is further compressed and sent to the absorber together with the HPD as absorber feed. The typical temperature after air/water cooling is 40°C.
- b) Part of the debutaniser bottom product (debutanised gasoline) is recycled as lean oil to the top of the absorber to absorb the C<sub>3+</sub> components that would otherwise leave with the absorber top gas stream.
- c) Part of the LCO product or the mid-circulating reflux stream (MCR) is used as sponge oil to further absorb some of the C<sub>3+</sub> components in the absorber top gas before it is routed to the Adip. treater. The fat sponge oil is sent back to the fractionator for stripping.

This line-up has several drawbacks. The recycling of debutanised gasoline inevitably requires higher absorber reboiler duties and (to a lesser extent) debutaniser reboiler duties. Furthermore, loadings in the absorber and debutaniser columns are increased. This may pose a constraint to any increase in throughput or any increase in gasoline yield. Further the C<sub>3+</sub> components absorbed by the sponge oil are recycled back to the fractionator, thereby wasting compression energy and taking up compressor capacity which could otherwise be gainfully used processing outside gases to improve the refinery overall C<sub>3+</sub> recovery.

It is evident that the elimination of the use of debutanised gasoline as lean oil and LCO as sponge oil is an attractive route to pursue. LPD which is essentially unstabilised gasoline can be used to replace the debutanised gasoline as lean oil and chilling of the lean oil or the absorber overheads can be applied to offset the somewhat inferior lean oil quality and to replace the sponge oil. As an example of what can be done to improve this type of system the results of a typical revamp study are presented below.

## **2 Basis of the Study**

The operation of a typical Absorber were simulated with a flowsheet. The Absorber operating pressure is 26.0 bara.

The stream data is:

	<u>t/d</u>	<u>°C</u>
HP gas (including some outside gases)	340	90
HP distillate	475	40
LP distillate (lean oil)	500	40
C3 in the combined feeds	129	-
Debutanised gasoline as lean oil	500	40
Sponge Oil (MCR)	150	40

Properties and composition of the streams are given in the Appendix.

## **3 Optional Cases**

For the purposes of this study the target C2 content in the absorber bottom product is set at 1.0% wt. on C3.

The following optional cases were considered:

- Case 1 - Base case. Line-up as per description above.
- Case 2 - LPD as lean oil, no sponge oil.
- Case 3 - Chilled LPD as lean oil, no sponge oil.
- Case 4 - LPD + pre-saturation at 15°C, no sponge oil.
- Case 5 - Lower-pressure operation of Case 4.

### **Case 1**

The typical conventional line-up as shown in Figure 1 is considered as the base case. The C3 recovery with this line-up is represented by line 1 in Figure 4. The higher the reboiler heat, the lower the C2 content and the lower the C3 recovery. The dotted part of line 1 represents the region where the C2 content in the absorber bottom product is above 1.0% wt on C3 and is therefore (for the purposes of this study) an unacceptable operating region.

The optimum operation of the absorber with regard to maximum C3 recovery at minimum reboiler heat is therefore indicated by point 1, where the C2 content is just 1.0%. The maximum achievable C3 recovery is 95%, which is equivalent to about 6.5 t/d (out of 129 t/d) of C3 loss. The reboiler load for this operation is 114 ktcal/d.

### **Case 2**

An equal amount of LPD from the fractionator overhead is used to replace the debutanised gasoline as lean oil and no sponge oil is used. The line-up is shown in Figure 2 (note however that this case is without chilling of the LPD and it is used at 40°C), the operation of the absorber is represented by line 2 in Figure 4.

Owing to the somewhat inferior lean oil, which contains some light ends, and the lack of sponge oil, the C3 recovery deteriorates significantly. In fact, there is a large increase in the C3+ content in the absorber off-gas. The best operation of this line-up is indicated by point 2, where the maximum attainable C3-recovery is only 89%,

which is equivalent to 14.2% t/d C3 loss. However as there is now no circulation of material over the absorber and debutaniser the absorber reboiler heat requirement is drastically reduced to some 65.5 ktcal/d and the column loading in the lower half of the absorber is reduced by an average of 20%. New LPD pumps however would most likely be required for this operation.

### **Case 3**

This case is the same as Case 2, except that the lean oil (LPD) is cooled to 15°C by using chilled water from a chilling unit. The operation for this case is represented by line 3 in Figure 4. Effectively, the chilled lean oil lowers the temperature at the top part of the absorber column and brings the C3 recovery back to about the same level of the base case. Point 3 represents the best achievable C3 recovery of 93.5% at the minimum reboiler heat of 67 ktcal/d and specified bottoms stream C2 content. A 6 ktcal/d duty chilling unit, cooler and (probably) new LPD pumps are required.

### **Case 4**

This line-up is shown in Figure 3. Instead of introducing the lean oil (LPD) at the top of the absorber, it is first mixed with the absorber overhead gas and the mixture is cooled by chilled water to 15°C. The chilled mixture enters a vessel, where gas is separated and sent to the Adip. treater and the liquid, which is pre-saturated with light ends, is sent to the top of the absorber. What is essentially achieved by applying the pre-saturation step is that the last equilibrium stage of the absorber (consisting of the cooler and the vessel) is no 15°C (lower than the 39°C in Case 3). Consequently, the temperature profile along the column is further lowered as compared with Case 3. The end result is that the C3 recovery is improved significantly, as is shown by line 4 in Figure 4. Point 4 represents the maximum obtainable C3 recovery of 99% (1.3 t/d C3 loss) at a reboiler heat of 38.5 ktcal/d and a refrigeration duty of 10.7 ktcal/d.

New equipment required here would be the chiller unit, cooler, vessel with pumps and probably LPD pumps.

### **Case 5**

In the line-up without pre-saturation step, lowering the absorber pressure will result in a substantial drop in C3 recovery. A check on the operation of Case 4 at a lower pressure of 23.0 bara shows that C3 recovery would not suffer much deterioration in such a case as compared with Case 3 (2% versus 12%). The benefits that can be obtained from a lower-pressure operation of the absorber are a lower reboiler heat requirement for the absorber and a lower compression ration, i.e. a lower compression energy requirement for the wet gas compressor. The latter may offer some gain in compressor capacity for outside gases if the compressors are power-limited.

## 4

## Discussion

The result of the five cases are summarised below.

Case	1 base case	2 Lean oil LPD 40 °C	3 Lean oil LPD 15 °C	4 Pre-sat. 15 °C	5 Pre-sat. 15°C low P
C2 in bottoms (% wt on C3)	1.0	1.0	1.0	1.0	1.0
Op. Press (bara)	26	26	26	26	23
C3 recovery (% wt)	95.0	89.0	93.5	99.0	97.0
Reboiler heat (ktcal/d)	114.0	65.5	67.0	68.5	61.0
Refrigeration duty (ktcal/d)	-	-	6.0	10.7	10.9
New equipment required:					
LPD pump		x	x	x	x
Chilling unit			x	x	x
Cooler, vessel, pump				x	x

The C3 recovery for Case 2 is relatively low. Comparing Case 3 with the Base Case about the same C3 recovery level can be maintained without a sponge oil section by replacing the debutanised gasoline with chilled LPD as lean.

By installing 2 x 500 t/d LPD pumps, a chiller unit of 6.0 ktcal/d and a cooler of about 40 m<sup>2</sup> for the lean oil, the following benefits can be achieved;

- a saving of 47 ktcal/d in absorber reboiler heat;
- an average reduction of 20% of the column loading in the lower half of the absorber;
- elimination of the recycling of some 13 t/d of light ends (average MW = 60). (This is equivalent to a saving of some 33 KW of compression energy or some 6 - 8 t/d of additional intake of outside gas to the compressor).

A further improvement can be achieved with Case 4. By installing new LPD pumps, pre-saturation facilities with chilling consisting of 2 x 500 t/d pumps for the pre-saturated lean oil, a larger chilling unit of 10.7 ktcal/d duty, a vessel and a cooler of about 80 m<sup>2</sup>, a 99% C3 recovery can be enjoyed at the minor expense of an additional 1.5 ktcal/d reboiler heat.

With virtually the same equipment as is required for Case 4 the following extra benefits can be obtained with a column pressure of 23.0 bara. at the expense of some 2% (2.6 t/d C3) reduction in C3 recovery:

- a further saving of 7.5 ktcal/d of reboiler heat;
- a further saving of about 9% of the compression energy used in the second stage of the wet gas compressor.

## **5 Chilling of Fractionator Overheads**

One drawback of using LP distillate (LPD) as lean oil is that the loading of the second-stage wet gas compressor is increases due to the reduced amount of LPD available for re-contacting before the second stage suction. This may in return result in a higher energy consumption for the second-stage compressor, which may nullify the saving mentioned above and necessitate the backing-out of some outside gases if the compressor capacity is limiting. This problem could be overcome by installing a chilled water condenser downstream of the fractionator overhead condenser and bringing the condensing temperature down to, say 25°C. The resultant reduced vapour loads to the first and second stages of the wet gas compressor would amount to a total saving of some 20% in compression energy or the equivalent of the spare compressor capacity for outside gasses. The possibility could even be considered of extending the chilling to the recontacting cooler downstream of the first stage of the compressor so as to enjoy a further 10% reduction in compression energy.

## **6 Chiller Unit**

Of the conventional refrigeration methods which produce chilled water of (lowest) 5°C by mechanical refrigeration, absorption refrigeration and steam-jet refrigeration, the absorption refrigeration system (Figure 5) is the one used most in distillation processes and it is briefly described below.

The system uses LIBR as absorbent and water as refrigerant. The refrigerant vapour is absorbed by the absorbent at a low pressure (typically 0.01 bara) and the “fat” solution is then pumped to the generator, where the refrigerant is condensed and passes through an expansion device to the low-pressure side, thus creating cooling. The refrigerant vapour enters the absorber and is re-absorbed by the “lean” absorbent from the generator, thus completing the absorption/desorption cycle. The process stream is cooled by a unit of chilled cooling water between the process and the refrigeration unit. The generator uses low temperature level heat obtainable from low -pressure steam (typically 1.8 bara) or hot water. This is an attractive feature, as heat is potentially available from refinery waste heat or there may be a surplus of heating media.

As a rough estimate the heat required to drive the chilling unit is about 1.4 times the chilling duty required. Note that therefore 2.4 times the chilling duty is required in cooling water.

At the temperatures (10°C) and pressures (25 bar) typical in some systems problems with hydrates have occurred and checks on the hydrate formation conditions for the prevailing gas conditions should be made if this is a concern. Some units are provided with hydrate inhibitor injection facilities.

## **7 Conclusions**

The above study demonstrates that much improvement of the conventional process line-up of the absorber cat. cracker can be achieved by the application of chilling. The discontinued use of sponge oil and replacement of the debutanised gasoline by chilled low-pressure distillate as lean oil will result in significant saving in reboiler heat, a reduction in the absorber (as well as the debutaniser) column loading and some savings in compression energy while maintaining about the same level of C3+ recovery.

The application of a pre-saturation step with chilling can lead to a significant increase in C3+ recovery. Pre-saturation with chilling provide the flexibility to operate the absorber at a lower pressure level and at a somewhat lower C3+ recovery and, hence, offers an opportunity for reducing the operating costs (reboiler heat and compression energy) in situations where a saving in costs out weighs a marginal gain in C3+ recovery.

Part of the chilling of lean oil can be carried out upstream at the fractionator overhead condensers to obtain a substantial saving in compression energy or to free some compressor capacity which can then be gainfully used for outside gases to improve the refinery overall C3+ recovery.

How far one should go in applying chilling to improve the efficiency of the absorber/fractionator of a cat. cracker is a matter of economic evaluation of capex versus savings and gains for each step of the application. For example, start from lean oil (LP dist.), chilling and proceed to pre-saturation plus fractionator overhead chilling. Each incremental step should be justified on its own incremental merits.

## Appendix

### Stream Properties and Composition used in this study

Component	HP gas
H <sub>2</sub>	22.9
N <sub>2</sub>	3.0
CO	0.9
H <sub>2</sub> S	3.5
C1	15.6
C2	12.1
C3	11.1
C3=	13.2
C4	2.5
iC4	6.9
Butene-1	2.7
T-Butene-2	1.5
C-Butene-2	0.9
iC5	3.2

	HP Dist.	LP Dist.	Debutanised gasoline	Sponge oil
d 15/4	0.718	0.744	0.747	0.997
MW	93	102	103	210
UOPK	12.0	11.8	11.8	10.1
ASTM IBP	-	35	40	176
5%	4	48	49	215
10%	27	60	62	246
20%	47	74	77	267
30%	65	87	89	283
40%	81	98	101	297
50%	95	108	112	310
60%	109	120	124	324
70%	124	132	136	338
80%	143	148	152	355
90%	160	164	170	378
95%	169	172	181	398
FBP	189	192	200	441

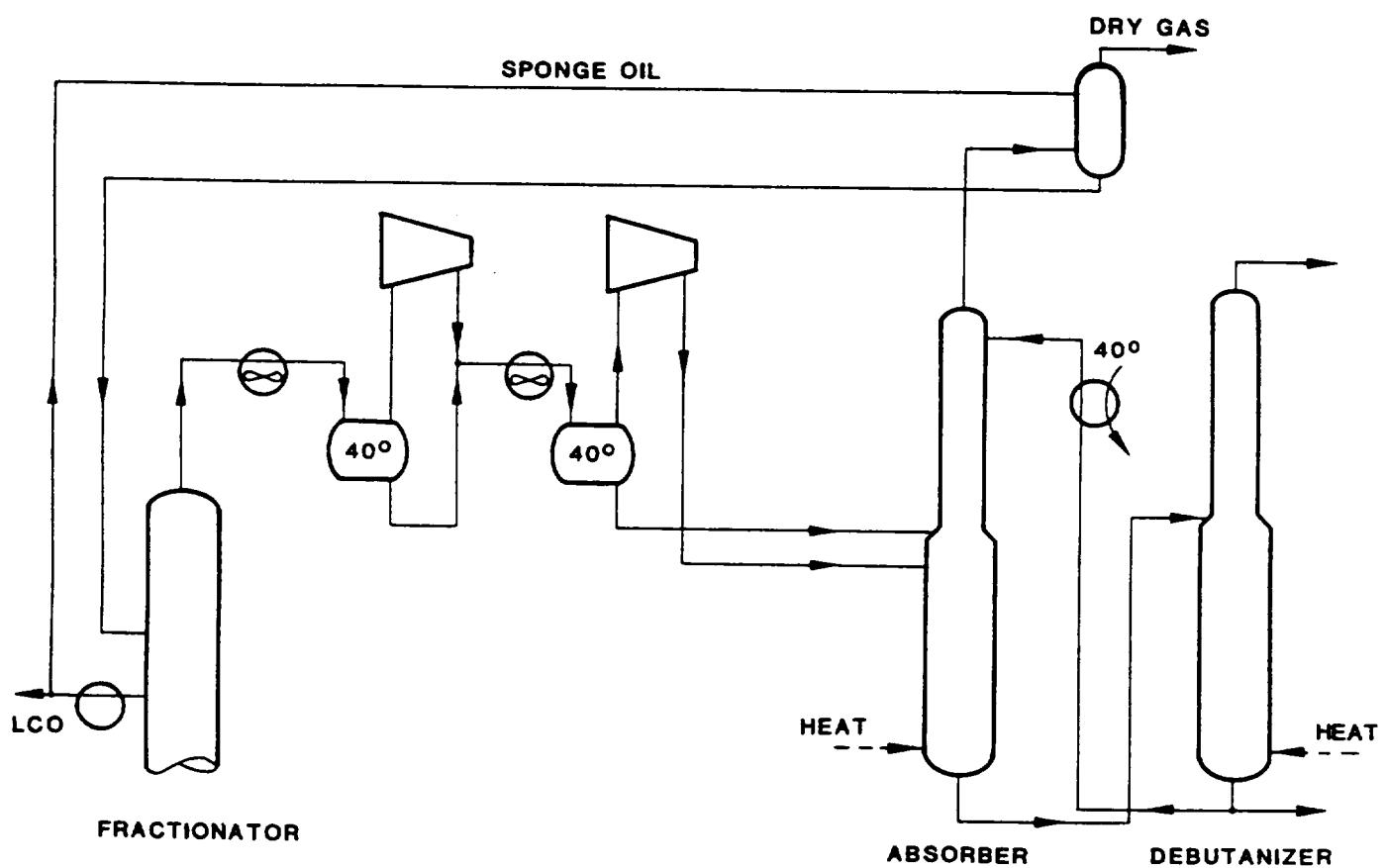


FIGURE 1 TYPICAL CONVENTIONAL LINE - UP

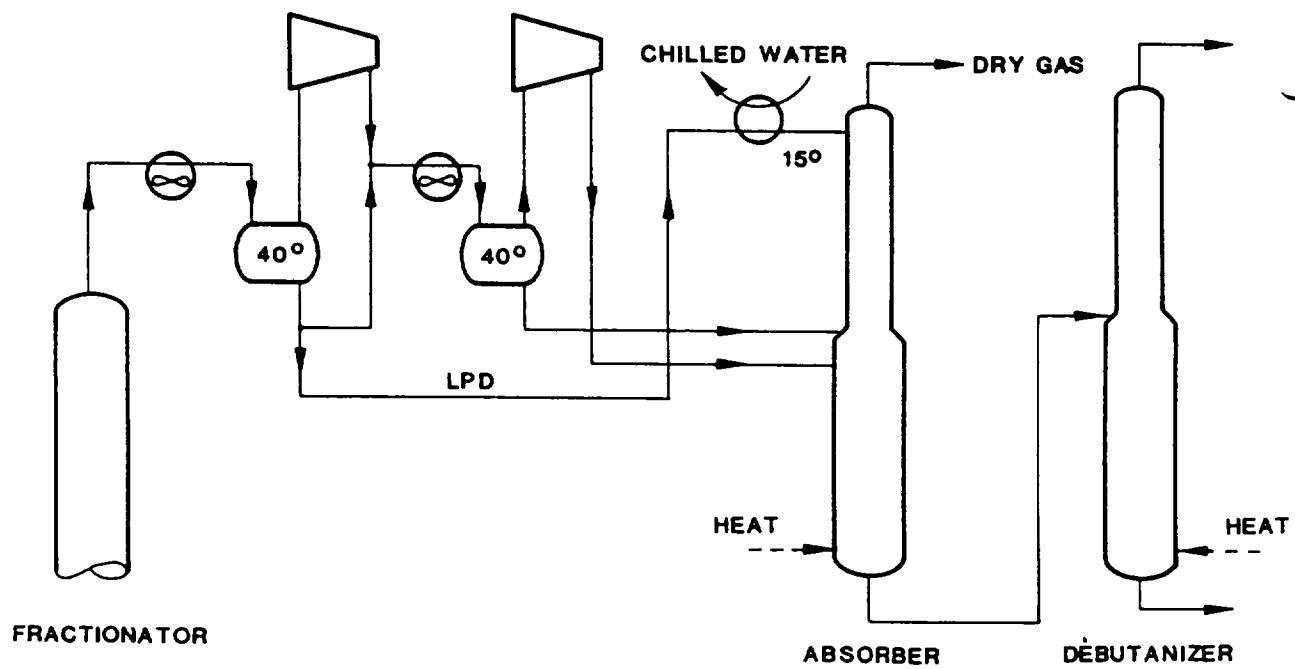


FIGURE 2 CHILLED LPD AS LEAN OIL

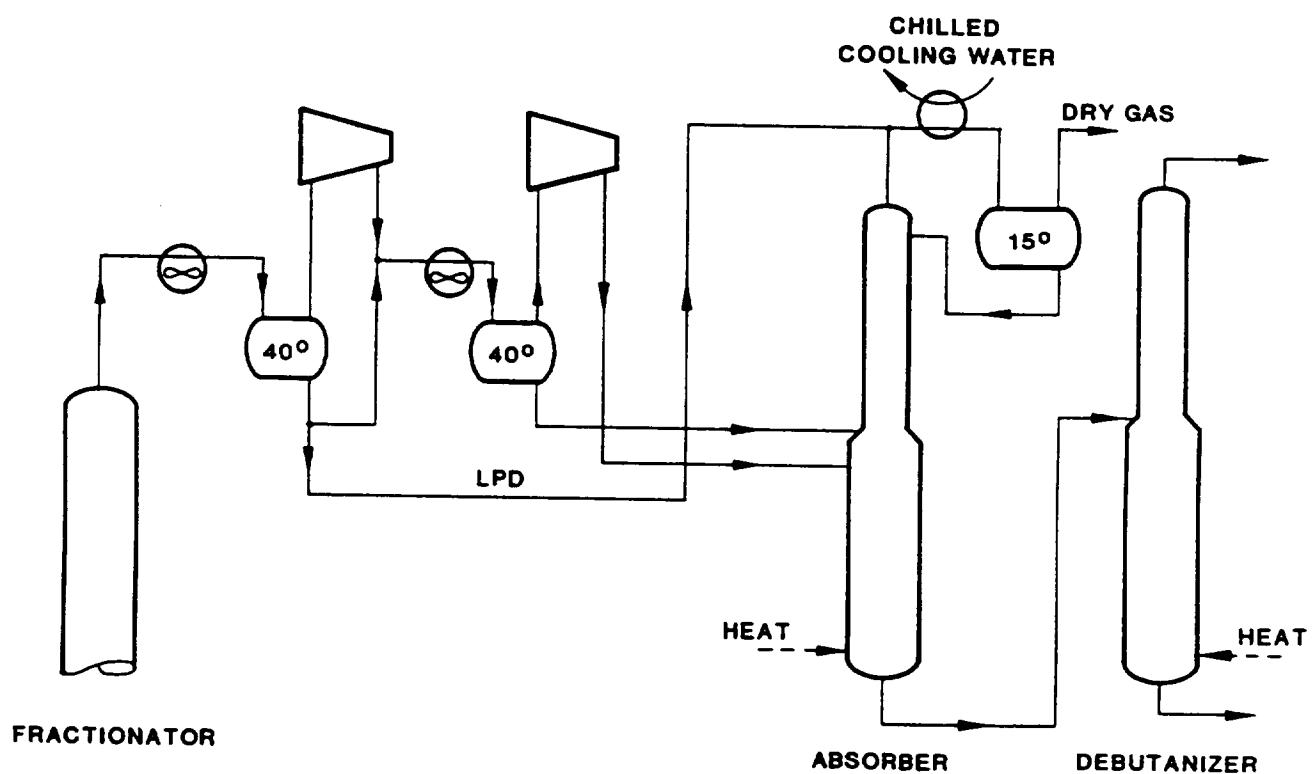


FIGURE 3 LPD PRE-SATURATION WITH CHILLING

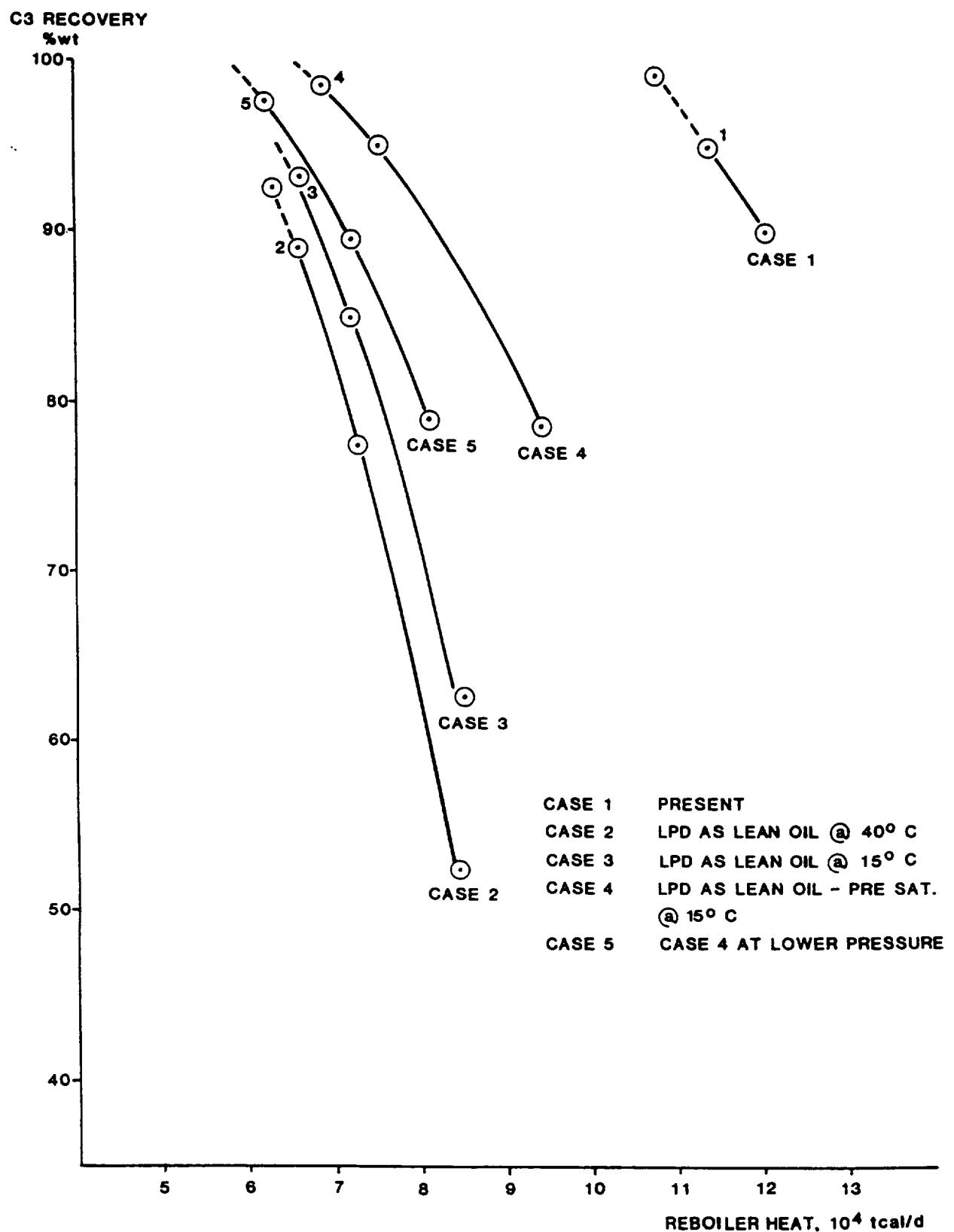


FIGURE 4 TYPICAL OPERATING LINES OF AN ABSORBER

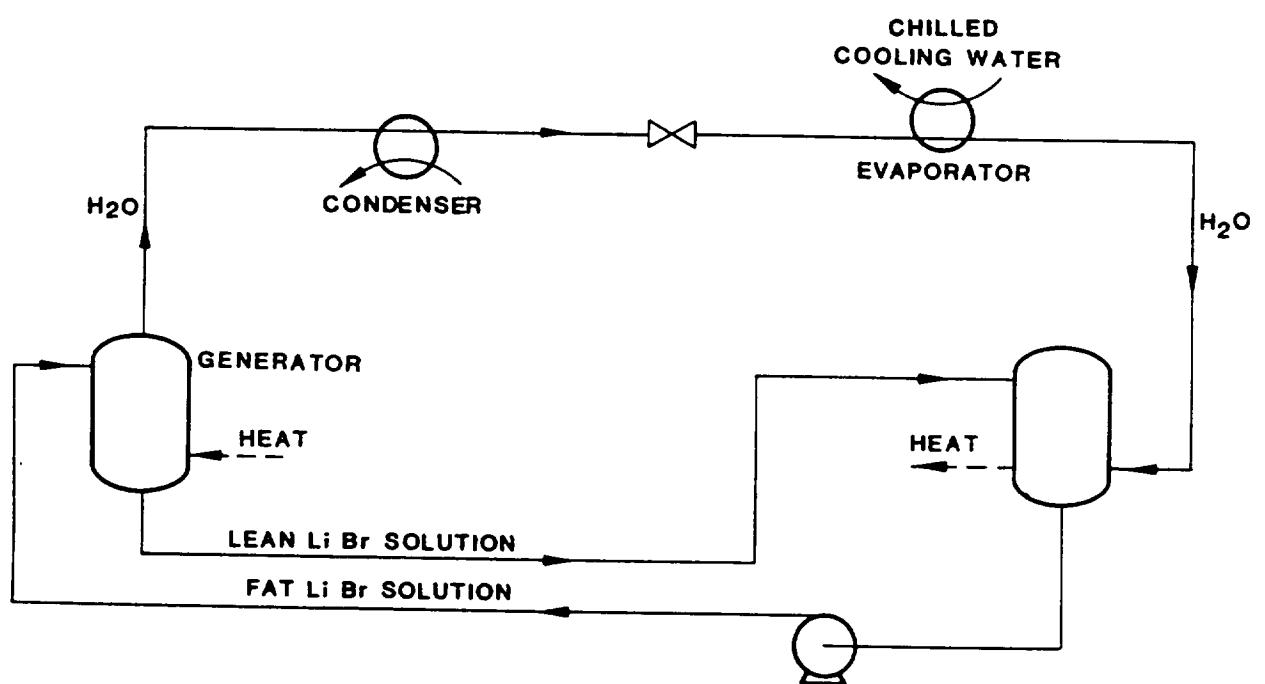


FIGURE 5 ABSORPTION REFRIGERATION



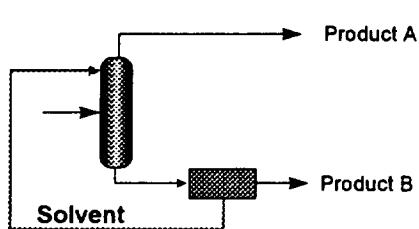
**18**

## **Modified distillation**

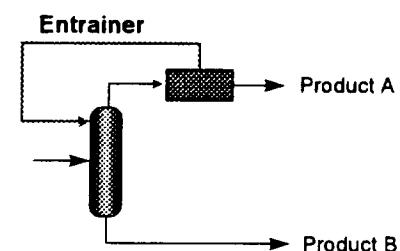
## Modified Distillation

- Adding an extraneous liquid that shifts the apparent boiling points

Heavier than feed

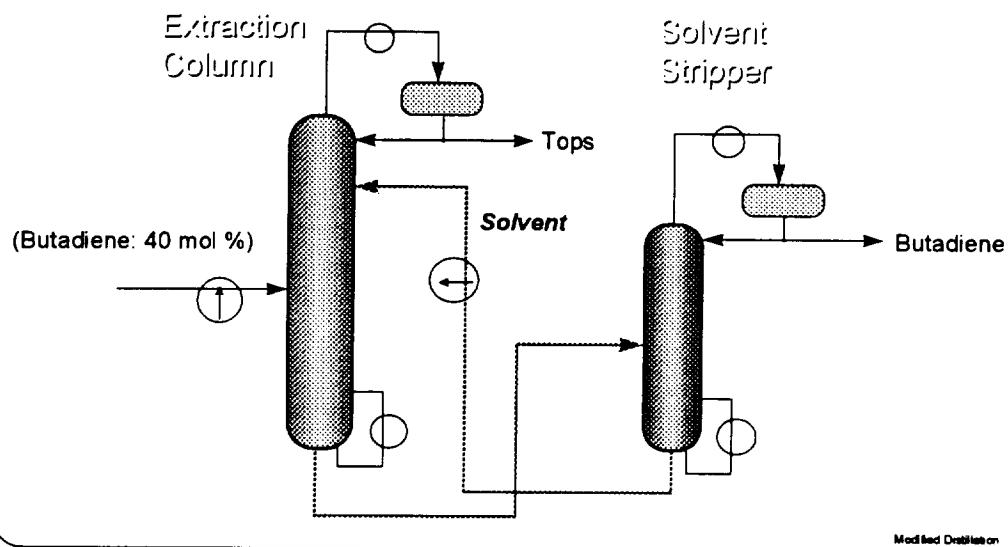


More volatile than feed



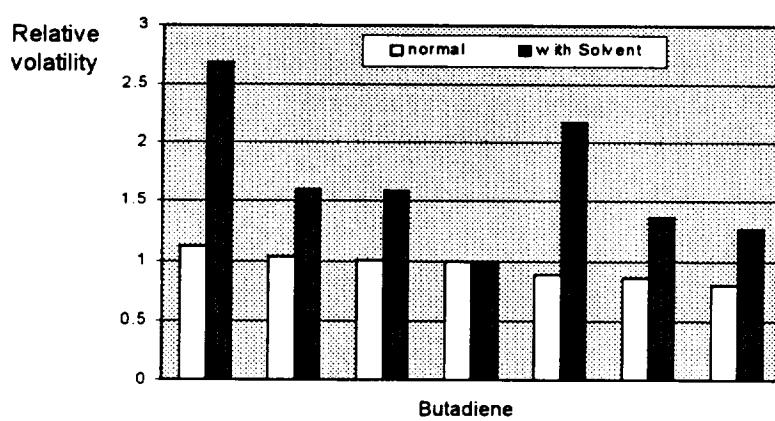
Modified Distillation

## Extractive Distillation



Modified Distillation

## Acetonitrile Solvent



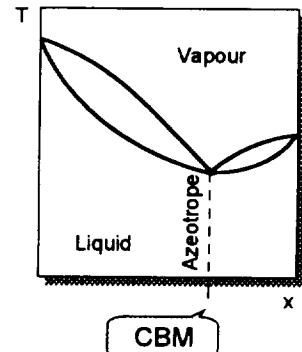
Modified Distillation

## Azeotrope

### Constant Boiling Mixture

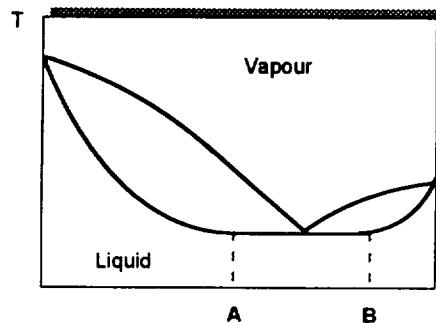
- Binary Azeotropes  
*Water - Alcohol*  
*Methanol - Benzene*
- Ternary Azeotropes  
*Water - Ethanol - Benzene*

### Minimum-boiling Azeotrope



Modified Distillation

## Heterogeneous Azeotrope

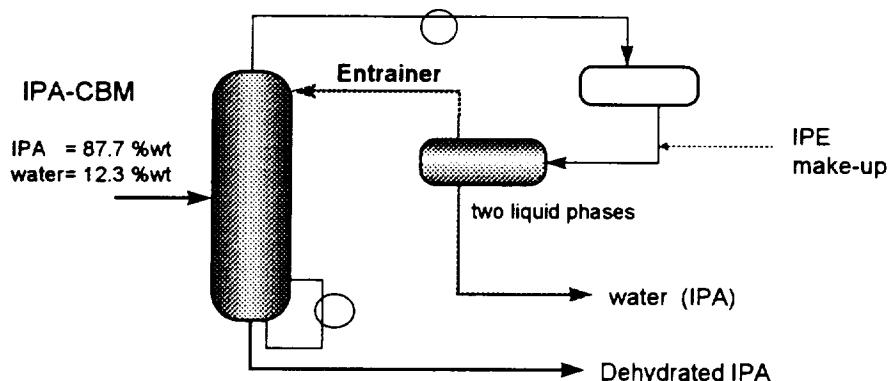


Two immiscible liquid phases

Modified Distillation

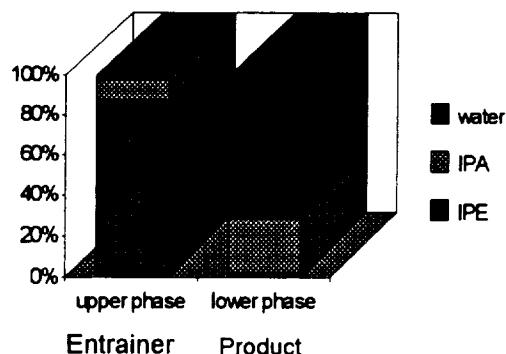
## Azeotropic Distillation

- Dehydration of Iso-Propyl-Alcohol (CBM)
- Entrainer: Iso-Propyl-Ether



Modified Distillation

## Liquid phase separation



Modified Distillation

## Modified Distillation

### 1. Introduction

Separation by distillation can be difficult when the boiling points of the components are very close. For these cases ordinary distillation is often found not economic because of expensive equipment (many stages) and high operating cost (reboiler heat). Other examples where ordinary distillation falls short are when the components do form an azeotrope. Separation is than even physically not possible because azeotropic mixtures have a boiling point and not a boiling range, essential for the distillation process.

Modified distillation uses an added component; the extraneous liquid. Inside the column it acts as an agent to enhance the separation. The extraneous liquid is called a **solvent** when its volatility is less than the feed, or it is called an **entrainer** when it is more volatile than the feed. So solvents are always discharged together with the bottom product while entrainers always leave the column with the overhead vapour.

Modified distillation requires an additional process step to separate the extraneous components from the final product. The recovered extraneous liquid is than recycled to the column and there is basically no consumption of extraneous material.

There are two different types of modified distillation processes:

- **Extractive Distillation**

The extraneous liquid is typically a solvent that is added at the top of the column. The solvent shifts the apparent boiling point of the components such that relative volatility's are increased. This promotes the separation such that less reboiler heat and fewer stages are required.

The process is called extractive distillation because the function of the extraneous material is superficially similar to a solvent in a chemical extraction process.

The solvent is recovered with an auxiliary column; a conventional two-cut Splitter. Since the solvent has a lower volatility than any of the other components the separation can be done with little difficulty.

- **Azeotropic Distillation**

The extraneous liquid is an entrainer that may either be added at the top of the column or it is mixed with the feed before it enters the column. The characteristic of the entrainer is that it forms a minimum-boiling azeotrope with one or more of the overhead products. Hence this operation is called azeotropic distillation.

Recovery of the entrainer by distillation is not possible because the entrainer forms an azeotrope with one or more of the overhead components. A partial separation however is possible when the condensed overhead vapour has two different liquid phases.

The entrainer-rich liquid phase is recycled to the column.

## 2. Extractive Distillation

An example of extractive distillation is the production of 1,3-butadiene. This product, used as feedstock for the manufacture of poly-butadiene rubber, is recovered from a narrow boiling range hydrocarbon feed as given in Table 1.

**Table 1, Feed composition**

Component	Atm. b.p.	mol %
propene	-47.7	0.26
isobutane	-11.7	0.29
isobutene	-6.9	26.3
butene-1	-6.3	21.2
1,3-butadiene	-4.4	40.2
n-butane	-0.5	0.6
trans-butene-2	+0.9	6.9
cis-butene-2	+3.7	3.8
vinyllacetylene	+5.6	0.45

As can been seen from the component atmospheric boiling points, separation with the normal distillation route would be very difficult. However, addition of aqueous acetonitrile ( $\text{CH}_3\text{CN}$ ) favourably affects the relative volatility's with respect to 1,3-butadiene as shown in Table 2.

**Table 2, Relative volatility shift by the presence of the solvent**

Component	Relative volatility (at 7 bara)	
	solvent free	with aqueous $\text{CH}_3\text{CN}$
isobutane	1.12	2.68
isobutene	1.03	1.60
butene-1	1.01	1.59
1,3-butadiene	1.00	1.00
n-butane	0.88	2.18
trans-butene-2	0.86	1.36
cis-butene-2	0.80	1.27

By presence of the solvent, the volatility's of the paraffin's and olefins are enhanced and can now easily be separated from the butadiene. The solvent has also an effect on the n-butane, trans- and cis-butene-2. These components, originally heavier than butadiene become, by the influence of the solvent, more volatile than butadiene and are leaving the column with the overhead product.

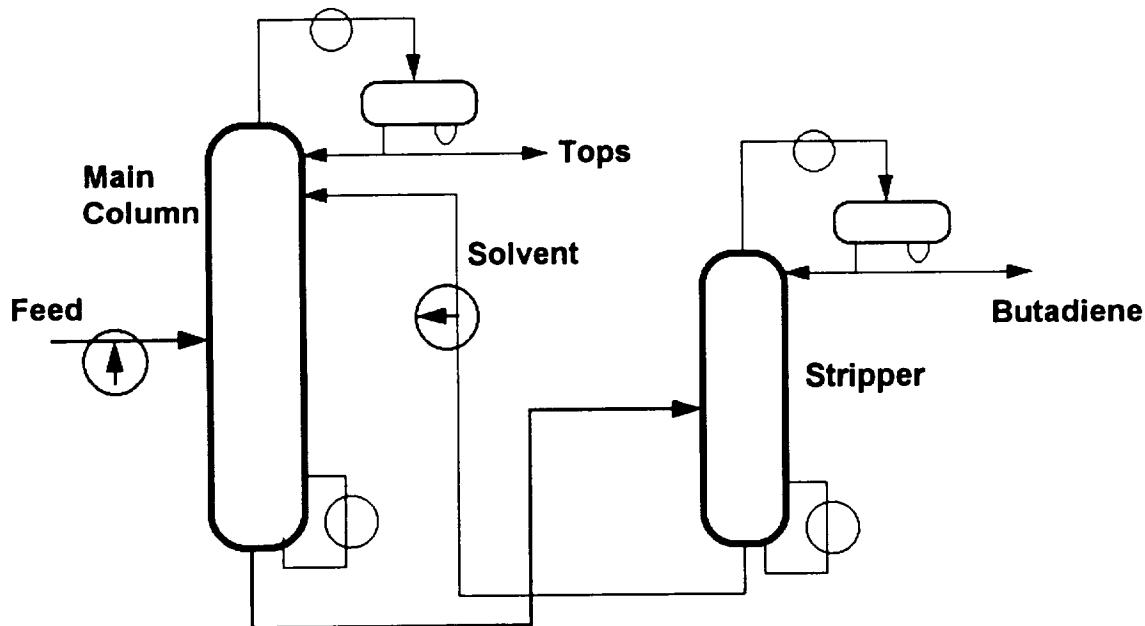
The line-up of the extractive distillation process is shown in Figure 1.

Feed is preheated and charged to the extractive distillation column as a vapour stream in order to stay below the limit of maximum solubility of hydrocarbons in solvent. This is important because

possible formation of two liquid phases in the column will have a negative effect on the tray efficiency.

The solvent is charged 3 trays below the top of the column to reduce loss of solvent by possible entrainment in the column overhead vapour stream. The total overhead vapour is condensed and partly used as reflux to the column.

The 1,3-butadiene and the acetylene are leaving the extractive distillation column as bottoms together with the solvent. This bottom stream is routed to a stripper column. This is a normal two-cut splitter column where the volatile hydrocarbons are separated from the solvent. The top product is sent to a tailing section for further purification of 1,3-butadiene. Finally, the solvent is recovered as bottom product and recycled to the top of the extractive distillation column.



**Figure 1, Extractive Distillation of 1,3-butadiene**

### 3. Azeotropic Distillation

An example of azeotropic distillation is the dehydration of iso-propyl-alcohol (IPA) to produce pure, water free IPA, a commercial product for the pharmaceutical, cosmetic and coating industry.

Because IPA and water do form a homogeneous binary azeotrope at a concentration of 87.9 %wt IPA, no higher purity can be obtained by normal distillation. However, with iso-propyl-ether (IPE) as an extraneous component an IPA purity close to 100% is reached. The azeotropic distillation process scheme is shown in Figure 2.

The feed to the unit is a Constant-Boiling-Mixture of IPA and water (IPA-CBM). This feedstock is obtained by distillation of an IPA / water mixture which yields IPA-CBM as top product. It consists of 87.7 %wt IPA, close to the binary azeotropic composition.

The extraneous component IPE is charge to the top of the azeotropic distillation column as reflux. In the top of the column the IPA, water and IPE forms a ternary azeotrope with a composition given in Table 3. Because the boiling point of this ternary azeotrope is substantially lower than the boiling points of the pure components and the IPA / water azeotrope the IPE is an entrainer and leaves the column with the ternary azeotrope overhead vapours.

Table 3, Composition of overhead vapours

Component	Composition %wt.
IPE	83.9
IPA	10.4
water	5.7

The IPA is dehydrated since the water leaves the column via the overhead vapours. This is accomplished by the entrainer because the ratio of water to IPA is higher in ternary azeotrope of the overhead stream than in the binary water-IPA azeotrope of the feed.

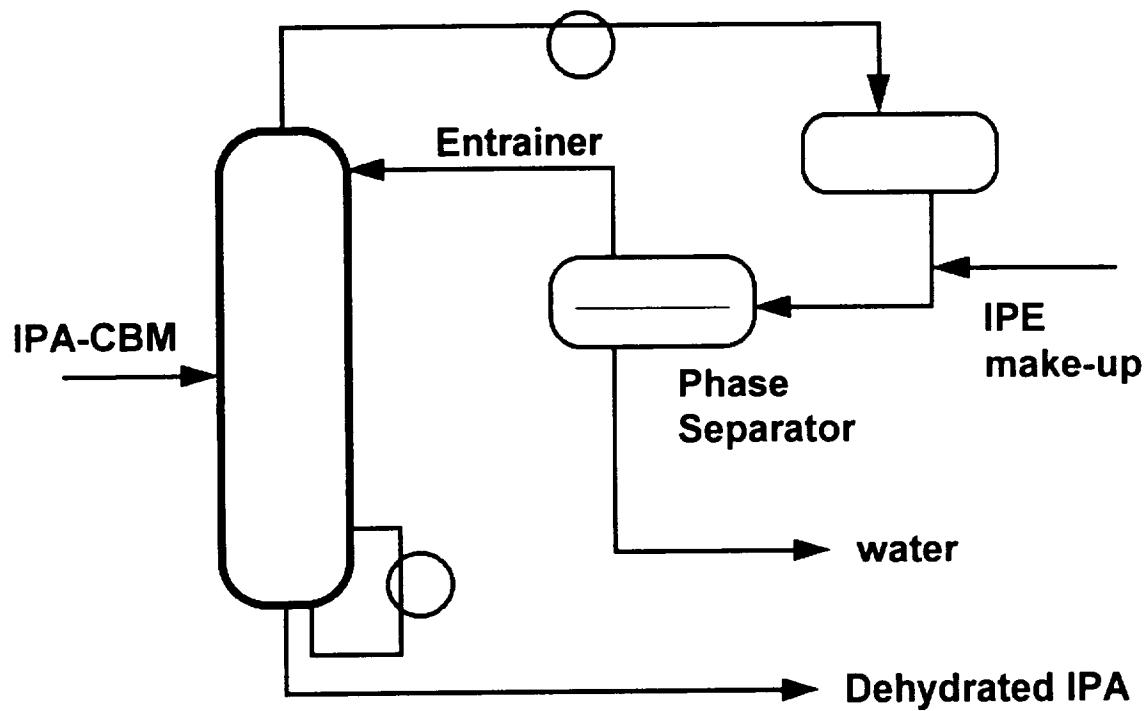
The total overhead vapour is condensed and the condensate is pumped to the phase separator. As the ternary IPE / IPA / water mixture is a heterogeneous azeotrope, two liquid phases with different densities are formed in the phase separator. The miscibility of the components in the two liquid phases are shown in Table 4

Table 4, Composition of two liquid phases

Component	Composition %wt	
	upper phase	lower phase
IPE	88	2
IPA	10	26
water	2	72

The decanted upper phase liquid, containing the major part of the IPE entrainer is recycled to the top of the column as reflux. The aqueous lower phase is the net top product which is rundown to the IPA storage tank. It contains water and IPA and some of the IPE entrainer. To compensate the loss of the IPE in the an intermittent make-up of entrainer is required.

The anhydrous IPA bottom product is nearly 100 % pure. It contains only traces of water but no IPE. The IPA recovery of the azeotropic distillation process is 94 %.



**Figure 2, Dehydration IPA-CBM by Azeotropic Distillation**

# Extractive distillation saves energy

Adding a solvent often enhances the separation of close-boiling compounds, cutting the amount of energy needed for distillation.

The key tradeoff is the investment in a solvent stripper.

Ian Sucksmith, Arco Chemical Co.

Components of nearly equal volatility are difficult to separate by conventional distillation, requiring many stages and high reflux rates. In such a case, adding a solvent that increases the relative volatility of one component will make the separation easier. This is known as extractive distillation. In this article, we will show how to choose a solvent, and how to evaluate the feasibility of extractive distillation.

Until recent years, extractive distillation was rarely used unless conventional distillation was impractical or impossible—in separating an azeotropic mixture, for example. But the situation is different now, and extractive distillation is worth considering even when the conventional approach is feasible.

The key reason for the change is the increase in energy costs. Extractive distillation uses less energy than conventional distillation does, and when energy costs are high the saving will pay for the solvent-separation equipment in many cases.

The other reason is that new methods for choosing solvents simplify evaluation of extractive distillation. In the past, predicting the effect of any proposed solvent generally required expensive laboratory work. Today, group-contribution methods provide an inexpensive method of evaluating solvents; only the most promising need to be checked in the lab.

## How the technique works

Extractive distillation uses an added component, the solvent, to increase the relative volatility of the components to be separated. For example: Suppose that we have a mixture of two components ( $X$  and  $Y$ ) that are to be separated by distillation. The conventional approach illustrated in Fig. 1 can separate them, but this will be very costly if the volatility of  $X$  relative to  $Y$  is close to 1.0.

Fig. 2 illustrates the same separation performed by extractive distillation. A solvent is added at the top of the first column, the extraction column. Since the solvent is much less volatile than either  $X$  or  $Y$ , it flows down the column to leave with the bottoms product. While in the column, the solvent increases the volatility

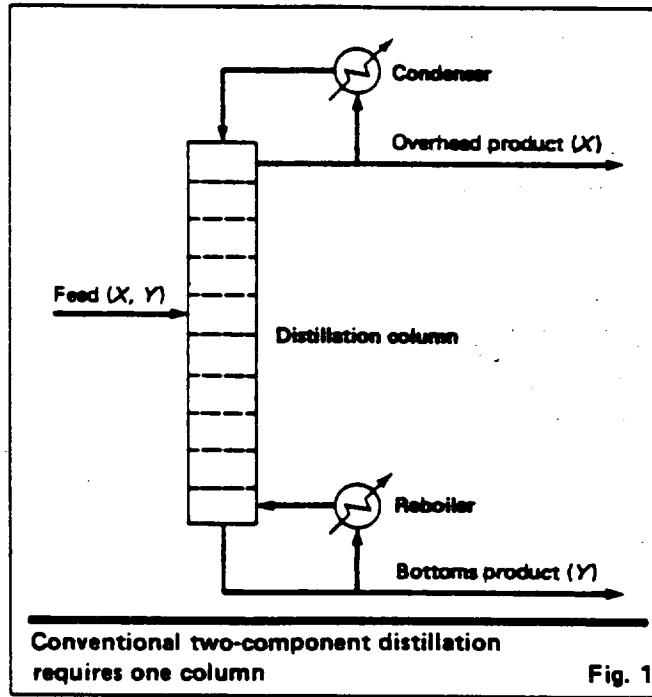


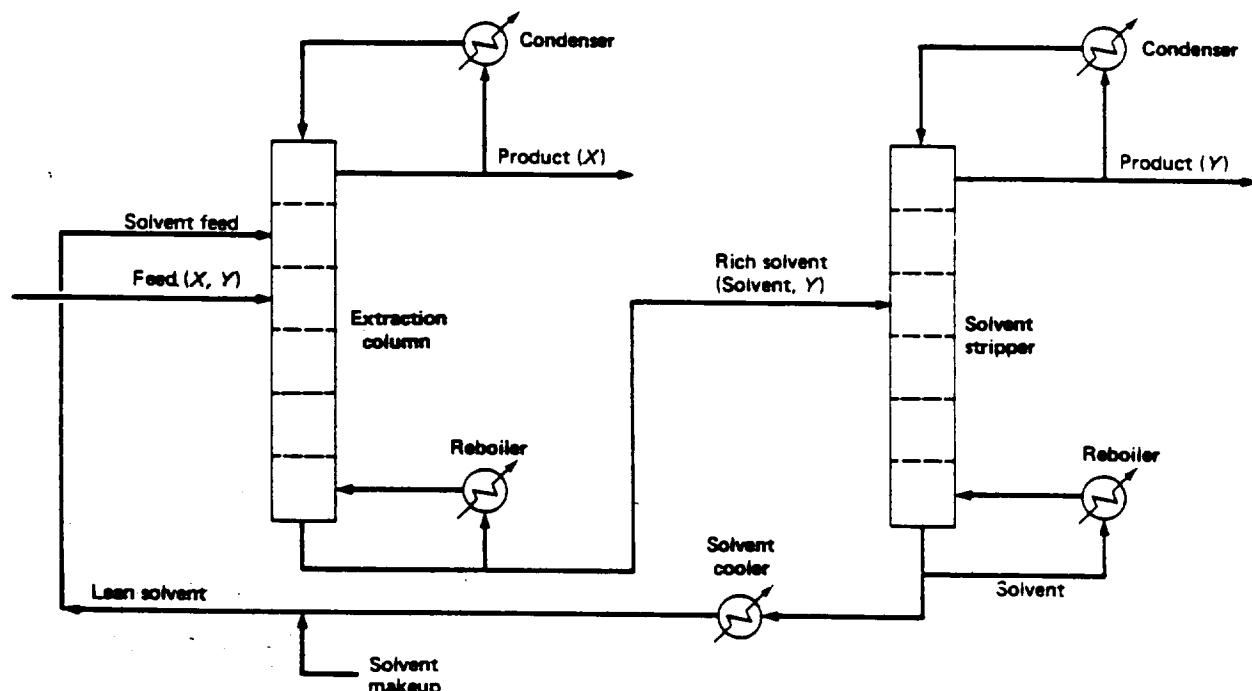
Fig. 1

of  $X$  with respect to  $Y$  and thus makes the separation easier.

The second column in Fig. 2 is the solvent stripper, which removes  $Y$  from the solvent. This is an easy separation because the solvent is much less volatile than  $Y$ . The lean solvent is then cooled and recycled back to the extraction column.

Net consumption of solvent is very low. Heat is lost in the solvent cooler but can often be recovered elsewhere in the process because it is at a high temperature. Even though the solvent is less volatile than either  $X$  or  $Y$ , a few trays are needed above its feedpoint in the extraction column to assure that little is lost in the overhead stream.

Fig. 3 shows a typical temperature profile for an extraction column, which may help to illustrate what happens inside. (This profile is derived from the exam-



Extractive distillation requires an extra column to recover the solvent

Fig. 2

ple to be given later.) There are three step changes in temperature:

1. Temperature increases by 18°F at the top of the column, just above the solvent addition point.
2. Temperature decreases by 3°F at the primary feedpoint, because of dilution by the liquid feed. This

would not occur if the feed were a vapor at its dewpoint.

3. Temperature rises sharply at the bottom stage, where a large quantity of the heavier component is boiled out of the solvent. This is described in detail in [1], and may provide a suitable point for controlling the separation.

The relative volatility of the components has a profile similar to that of the temperature: Above the solvent feed, it is close to 1; below the solvent feed, it is much greater than 1 and fairly constant; at the bottom stage, it becomes still greater because the solvent concentration is increased.

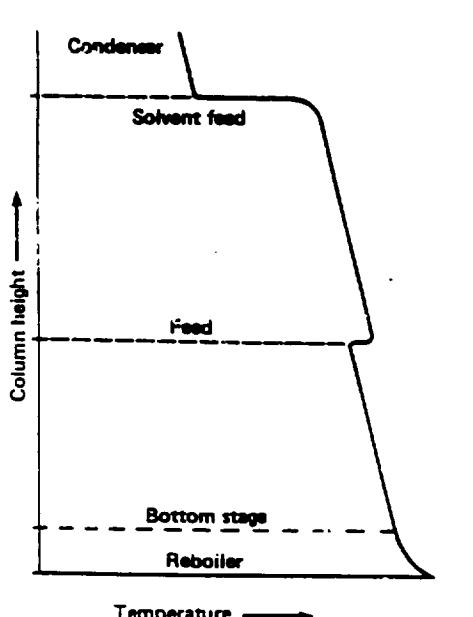
#### Relative volatility

The ratio of component volatilities determines how difficult the separation is. In azeotropes, the relative volatility of the components is 1, and conventional distillation cannot separate them. Extractive distillation has generally been used for components with relative volatility between 1 and 1.05, but now it can pay for itself in separating components above 1.05.

The relative volatility of component *X* with respect to *Y* ( $\alpha_{XY}$ ) is the ratio of their vapor-liquid equilibrium ratios. At equilibrium, the fugacity of any component *i* is the same in the vapor and liquid phases. This is expressed as:

$$\phi_i \gamma_i P = \gamma_i x_i f_i^0$$

where  $\phi_i$  is the vapor-phase fugacity coefficient,  $\gamma_i$  is the vapor-phase mole fraction,  $P$  is the total system pressure,  $\gamma_i$  is the liquid-phase activity coefficient,  $x_i$  is the liquid-phase mole fraction, and  $f_i^0$  is the standard-state fugacity. At low to moderate pressures and tempera-



Typical temperature profile in extractive distillation column

Fig. 3

tures.  $f_i^0$  can be approximated by the pure-component vapor pressure ( $p_i^0$ ).

Assuming that we can use the pure-component vapor pressure, the equilibrium ratio for component  $i$  is:

$$K_i = y_i/x_i = \gamma_i p_i^0/\phi_i P$$

The relative volatility of two components,  $X$  and  $Y$ , is then:

$$\alpha_{XY} = K_X/K_Y = (\gamma_X/\gamma_Y)(p_X^0/p_Y^0)(\phi_Y/\phi_X)$$

From this last equation, we can see that there are three ways to change the relative volatility:

1. Alter the ratio of pure-component vapor pressures ( $p_i^0$ ). This ratio increases slightly as temperatures are reduced, but not usually enough to enhance separation to a significant degree.

2. Alter the ratio of vapor-phase fugacity coefficients ( $\phi_i^0$ ). These are measures of the nonideality of the vapor-phase mixture. At moderate pressures, these coefficients are usually close to 1 and do not provide a practical means of changing relative volatility.

3. Alter the ratio of liquid-phase activity coefficients ( $\gamma_i$ ). Many liquid mixtures are highly nonideal, and therefore these coefficients can be much greater than 1. The ratio of the two coefficients can be changed substantially by adding a solvent that is chemically more similar to one component than to the other. This last approach is the basis of extractive distillation.

### Choosing a solvent

The two components generally differ significantly in molecular structure in systems where relative volatility is near 1. The extracting solvent is chosen to take advantage of this.

In general, the solvent should be similar in structure

to the heavier (less volatile) of the two components. It will then form a near-ideal mixture with the heavy component, and a highly nonideal mixture with the light one. This will increase the liquid activity coefficient of the light component, and therefore its relative volatility.

For example: Suppose that we have a mixture of acetone and methanol ( $\text{CH}_3\text{COCH}_3$  and  $\text{CH}_3\text{OH}$ ). Acetone is the more volatile of the two, with a relative volatility of 1.4 in a 50:50 bubble-point mixture. If a polar solvent were added to this mixture, we would expect to see the relative volatility of acetone increase because acetone is less polar. Water is a suitable polar solvent because its boiling point is 64°F above methanol's and therefore it is far less volatile than either component of interest. When water is added so that it makes up 80 mol% of the mixture, the acetone:methanol relative volatility is increased to 3.6.

In choosing a solvent for extractive distillation, many compounds appear to be candidates because of their structural similarity to the heavy component of interest. Therefore, several other criteria are used to pick the best solvent.

First, any good solvent should satisfy the following conditions [7]:

1. The solvent should be substantially less volatile (higher boiling point) than either component of interest. This assures that it can be stripped readily.
2. Within this volatility constraint, the boiling point should be as low as possible to minimize the column temperatures.
3. The solvent should be completely miscible with both components over the full concentration range.
4. It should have no unacceptable corrosive, toxic or flammability properties.

### Group contribution methods predict liquid activity coefficients

Predicting the effects of a solvent on a mixture of components requires liquid activity coefficients. These are costly to obtain experimentally, and are generally not to be found in the literature because there are so many possible mixtures.

Group contribution methods are derived from experimental data, and often provide a reasonable estimate of activity coefficients. The methods consider all components as mixtures of their constituent groups, and calculate activity coefficients based on the behavior of the groups and their interactions with each other.

For example: *n*-nonane is treated as a mixture of two  $\text{CH}_3$  groups and seven  $\text{CH}_2$  groups; toluene as five aromatic ACH groups and one aromatic ACC $\text{H}_3$  group; and acetaldehyde as one  $\text{CH}_3$  group and one CHO group. To predict the liquid activity coefficients for a mixture of these three compounds, one need only know the behavior and interactions of five groups.

### Available methods

There are now two methods in widespread use: UNIFAC [3] and ASOG [4]. Both are nonproprietary, and programs for their use are available in the open litera-

ture. UNIFAC has also been incorporated into some commercial simulation packages.

Both methods base their predictions on properties of individual groups and interactions between groups, but they differ somewhat in their approach. Practical use of either technique is limited by the following:

1. Not all group parameters or interaction parameters have been determined.
2. All components must be condensable.
3. Temperatures are restricted to about 30–150°C.
4. Real components will usually behave differently than the sum of their constituent groups.
5. Neither method is likely to predict the formation of two liquid phases.
6. These methods do not work in the case of electrolytes or polymers.

Because they are limited, group contribution methods should be employed only as a first step in choosing a solvent. If reliable experimental data are available, those should always be used. And the final choice of solvent should always be checked in the laboratory. This can sometimes be done quite rapidly, using gas-liquid chromatography [2].

## EXTRACTIVE DISTILLATION

Effect of solvents on a mixture of *n*-hexane and benzene

Table I

Solvent	$T_H^{\circ}$	$T_B^{\circ}$	$S_{H/B}^{\circ}$	$a_{H/B}$	Solvent boiling point, °F
Cyclohexane	1.038	1.398	0.743	1.099	177.3
<i>n</i> -Heptane	0.991	1.367	0.725	1.039	200.2
<i>n</i> -Octane	0.971	1.258	0.772	1.060	258.2
<i>n</i> -Nonane	0.899	1.108	0.811	1.076	303.4
(No solvent)	—	—	(1.000)	(1.370)	—
Ethylbenzene	1.388	0.968	1.414	1.800	277.1
Toluene	1.493	0.972	1.635	1.947	231.1
Phenol	13.184	10.454	1.585	2.090	369.3

\*Liquid activity coefficients for *n*-hexane (*H*) and benzene (*B*) are based on infinite dilution.

Example: Conventional distillation requires  
42 million Btu/h    *n*-Hept.-Toluene Table II

Design parameter	Design estimate
Number of stages	55
Feed location (stage number)	46
Assumed tray efficiency (percent)	65
Number of trays	82
Reflux ratio	20
Reboiler duty (million Btu/h)	42

Example: Selectivity of potential solvents  
for *n*-Heptane-Toluene Table III

Solvent	$T_H^{\circ}$	$T_T^{\circ}$	$S_{H/T}^{\circ}$	Solvent boiling point, °F
Cyclohexane	1.024	1.425	0.719	177.3
<i>n</i> -Octane	0.965	1.316	0.728	258.2
(No solvent)	—	—	(1.000)	—
Ethylbenzene	1.311	0.968	1.357	277.1
Phenol	2.612	1.877	1.584	369.3
Styrene	2.862	1.133	2.452	293.4

\*Liquid activity coefficients for *n*-heptane (*H*) and toluene (*T*) are based on infinite dilution.

Example: Extractive distillation requires  
only 18 million Btu/h    *n*-Hept.-Toluene Table IV

Design parameter	Design estimate	
	Extraction column	Solvent stripper
Number of stages	12	20
Feed locations (stage number)	4 (feed) 3 (solvent)	10
Assumed tray efficiency (percent)	25	55
Number of trays	40	33
Reboiler duty (million Btu/h)	29.9	9.3
Heat recovered (million Btu/h)	11.2	—
Net heat consumption (million Btu/h)	9.7	8.1

5. It must not react with either of the components being separated.

The solvents that satisfy these conditions should then be compared for differences in selectivity. This is a measure of the effect of the solvent on relative volatility, and thus ease of separation. Selectivity is defined as the ratio of the liquid-phase activity coefficients for the components of interest when each alone is present in the solvent at infinite dilution [2]. For components *X* and *Y*:

$$S^{\circ} = \gamma_X^{\circ}/\gamma_Y^{\circ}$$

The greater the selectivity value, the greater is the ease of separation.

As an example of selectivity, consider the system *n*-hexane and benzene. The relative volatility of *n*-hexane to benzene is 1.37 in a 50:50 molar mixture at the bubble point. We can expect that adding a solvent similar to *n*-hexane will reduce the relative volatility, and that adding an aromatic solvent similar to benzene will increase it. This is illustrated in Table I. The aromatic solvents have high selectivities, and increase the relative volatility of *n*-hexane to benzene, while the aliphatic solvents have poor selectivities. If this were a projected operating system, toluene and phenol would be considered promising candidates.

The liquid activity coefficients needed to determine selectivity had to be gathered experimentally in the past; this made choosing a solvent very costly. Today, analytical methods can be used to predict these coefficients, saving considerable time and expense. These prediction techniques are called group contribution methods because they consider compounds to be mixtures of their constituent groups—for example, *n*-nonane as two CH<sub>3</sub> groups and seven CH<sub>2</sub> groups. The box on the previous page describes the methods in detail.

## Design considerations

Extractive distillation systems require more design analyses than do conventional systems because there is an extra component and an extra tower. Here we will look at four factors that would not arise in conventional distillation, and then at an example that illustrates design in detail:

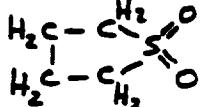
**Solvent flowrate.** A greater solvent flowrate generally yields a better separation, but there are two tradeoffs to consider. The first is that solvent must be heated in both the extraction and stripping columns. Therefore, more heat is required at greater flowrates. The other is that higher solvent concentrations require higher distillation temperatures. This leads to a slight reduction in relative volatility, which works against separation.

**Reflux ratio.** A higher reflux ratio leads to better separation, as in a conventional distillation, but it also causes the solvent to be more diluted and therefore have less effect on relative volatility. Deciding on a reflux ratio requires an optimization. Another concern here is that two liquid phases must not be permitted to form.

**Tray efficiencies.** Low tray efficiencies are often encountered in the extraction column. This can partially offset the reduction in the number of stages needed for a given separation.

**Crossover in relative volatility.** Up to now, we have as-

shall use Sulfolane for extraction/extractive distillation of aromatics.



sumed that the solvent will always increase the volatility of the lighter component more than that of the heavier. It is possible to increase the volatility of the heavy component so that it becomes more volatile than the light, but this is generally not desirable.

For example (as described in [1]): The relative volatility of *n*-butane to 1-butene is typically about 0.85. If furfural is added to a mixture of these components, the relative volatility becomes greater than 1 because this polar solvent preferentially increases the activity of the "heavy" and nonpolar *n*-butane. The problem with such a separation occurs at the top of the extraction column. Here, the two components assume their normal relative volatility in the absence of the solvent. The distillation that takes place at the top of the column therefore works against the desired separation, because of the crossover in relative volatility.

### Design example

Determining whether extractive distillation is more attractive than the conventional approach requires a preliminary design and economic evaluation. If it appears more economical after such an analysis, then laboratory solvent-testing and more-precise design work will be needed to assure a payoff. One should:

1. Define the problem.
2. Design a conventional distillation system.
3. Choose potential solvents and evaluate their selectivity by group contribution methods, or by using laboratory data if any are available.
4. Design an extractive distillation system that uses the best of the solvents evaluated.
5. Compare the economics of each approach.

The rest of this article consists of an example that illustrates the evaluation procedure. We will assume that there are no experimental vapor-liquid equilibrium data available for the components of interest, or any potential solvents. Instead of such data, we will use the UNIFAC group-contribution method to calculate liquid activity coefficients directly. Vapor-phase fugacities will be derived from the Redlich-Kwong expression. (In fact, experimental data for the system to be discussed are available [5].)

**Problem definition.** The feed is a mixture of *n*-heptane and toluene, containing 50 mol% of each. Flowrate is 30,000 lb/h, pressure is 10 atm and temperature is 100°F. It is required to recover heptane at 94 mol% or better purity, and 83% or better recovery. There are no equipment constraints on the system. All vapor pressure data are available in [6], and UNIFAC constants are listed in [3].

**Conventional distillation.** Details of the column that will meet the design target are listed in Table II.

**Evaluating potential solvents.** Two aliphatics and three aromatics will be considered potential solvents. These are: cyclohexane, *n*-octane, ethylbenzene, styrene and phenol. We can expect that the aromatics will increase the relative volatility of *n*-heptane to the aromatic toluene, and as such will be preferred.

Table III lists the selectivities for each of the potential solvents, based on UNIFAC calculations. Styrene and phenol offer the best selectivity, and generally meet the other criteria for choosing a solvent. However, styrene is

But Phenol is quite Toxic!

likely to polymerize at the temperatures required, which is not acceptable, and therefore phenol appears to be the best of the solvents.

**Designing the extractive-distillation system.** Table IV lists details of the extraction column and stripper needed to meet the design objectives, based on using phenol as the solvent. Note that tray efficiency in the extraction column is assumed to be only 25%, vs. 65% for the conventional case. Assuming that most of the heat lost in the solvent cooler can be recovered, the net heat consumption for the extractive approach is only 17.8 million Btu/h—57% less than is needed in the conventional approach.

**Comparison and followup.** If heat is valued at \$5 per million Btu, the energy cost for the conventional system will be \$1.7 million for an 8,000-h operating year. The net energy cost of the extractive system will be only \$710,000 per year. The net saving, therefore, is about \$1 million/yr—or about 1¢ per lb of *n*-heptane product. \*

Against this energy saving must be traded off the cost of the additional column, reboiler, condenser, heat-recovery exchanger, pump and solvent-storage system. The cost difference between the extraction column and the conventional distillation column should also be considered—the extraction column usually costs less.

Such an analysis shows that extractive distillation should be considered as an alternative to conventional distillation for this application. If extractive distillation appears more economical, then optimization of the design as well as laboratory evaluation of the solvent (if not already done) are justified.

Mark Lipowicz, Editor

### \* Variable cost savings.

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#### The author

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**19**

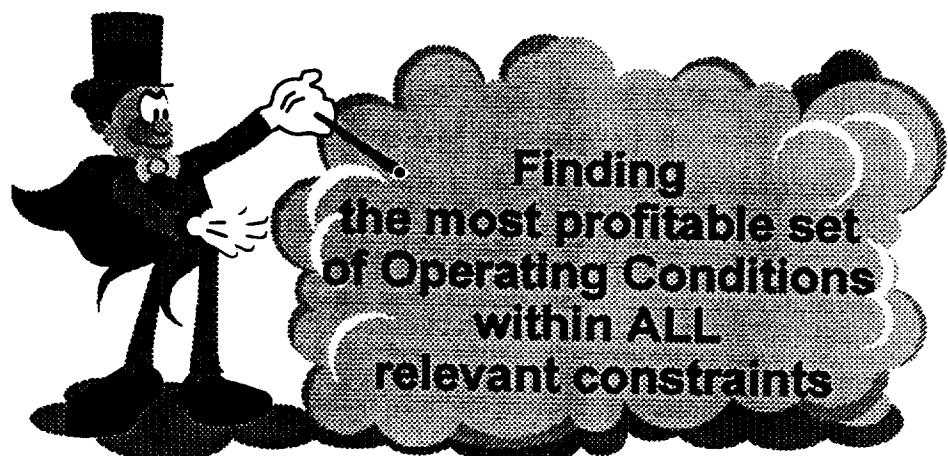
## **On-line optimisation**

# **ON-LINE CLOSED LOOP UNIT OPTIMISATION MINI COURSE**

M111: Distillation and Control

ORTTL/46 10/11/97

## **A Definition of Unit Optimisation**



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## Off-line Optimisation

- Uses **testrun data** as input
- Optimisation is done **on request**
- Data reconciliation is **recommended**
- Used by technologist / engineer for **case studies**

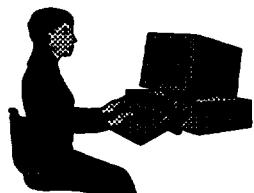


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## On-Line Open Loop Optimisation

- Uses **on-line**, hourly averaged **data** as input
- The optimiser calculates advisory targets, which can be **implemented manually**
- Data reconciliation is highly recommended
- Used by technologist / engineer for **case studies**



Operation



Optimiser

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## Closed Loop Optimisation

- Uses on-line, hourly averaged data as input
- The optimiser calculates targets, and **passes them to the control layer**
- Data reconciliation is highly recommended
- Used by technologist / engineer for case studies



Operation

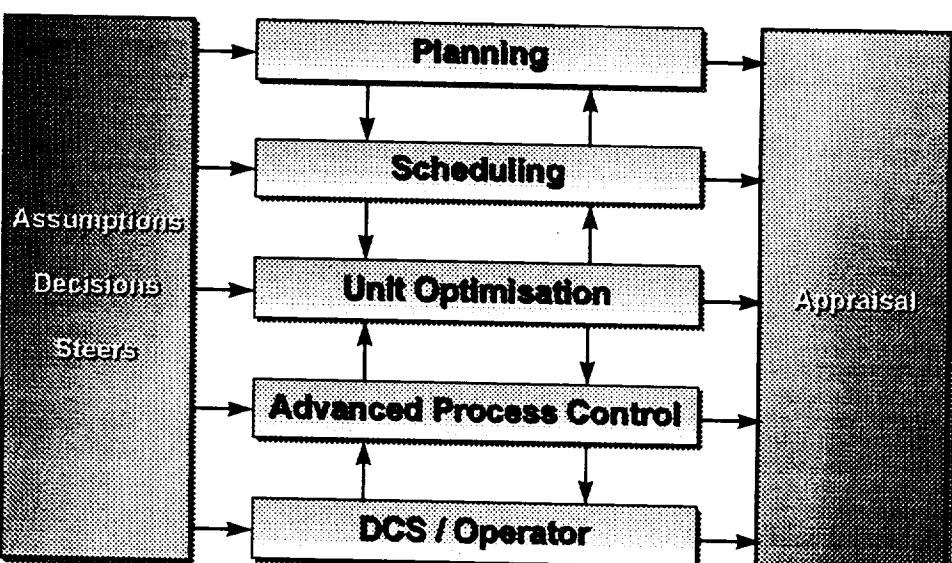


Optimiser

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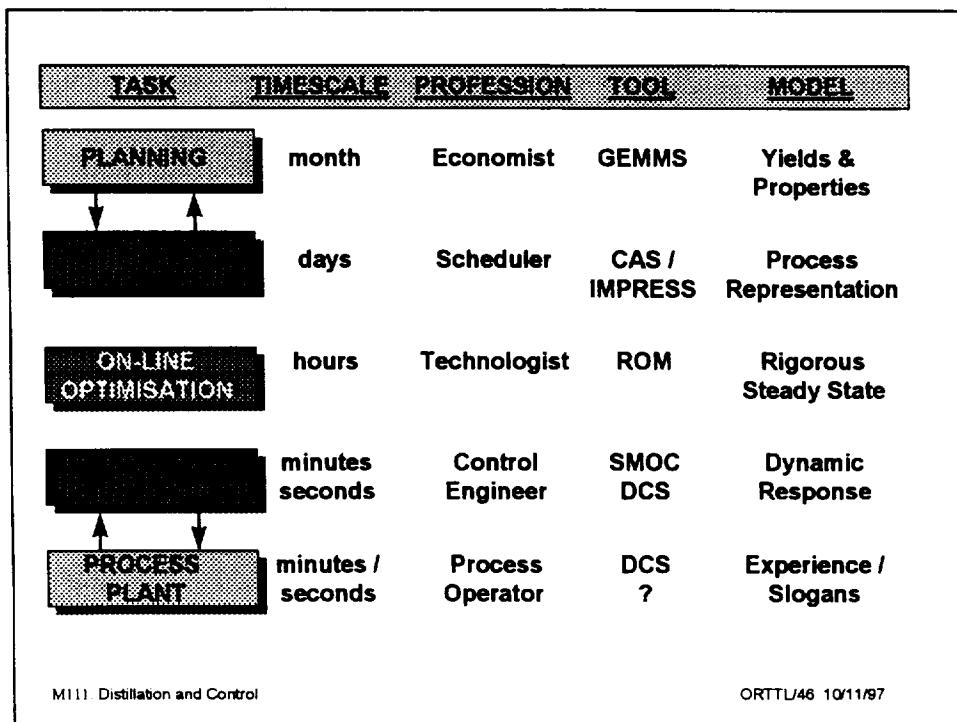
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## Where does Optimisation fit in?



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F1 1006 TPD  
F2 510 TPD  
E3 502 TPD

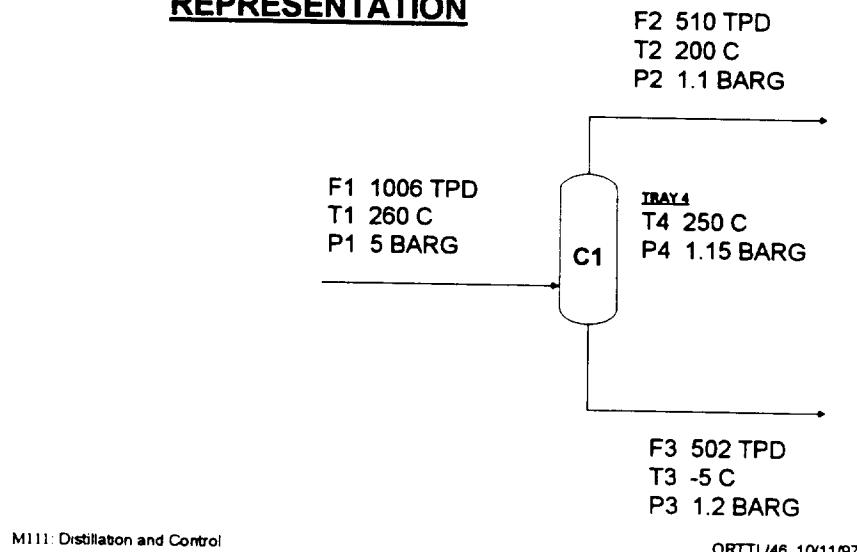
P1 5 BARG  
P2 1.1 BARG  
P3 1.2 BARG  
P4 1.15 BARG

T1 260 C  
T2 200 C  
T3 -5 C  
T4 250 C

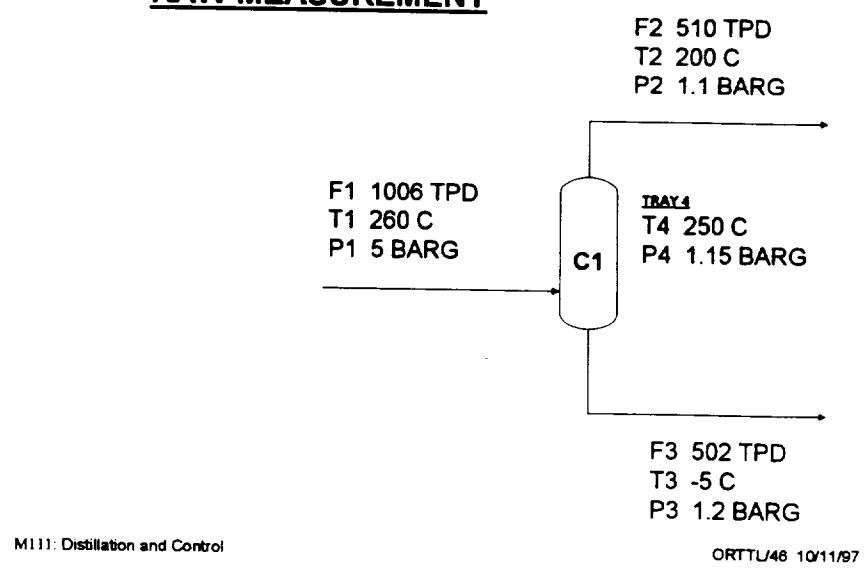
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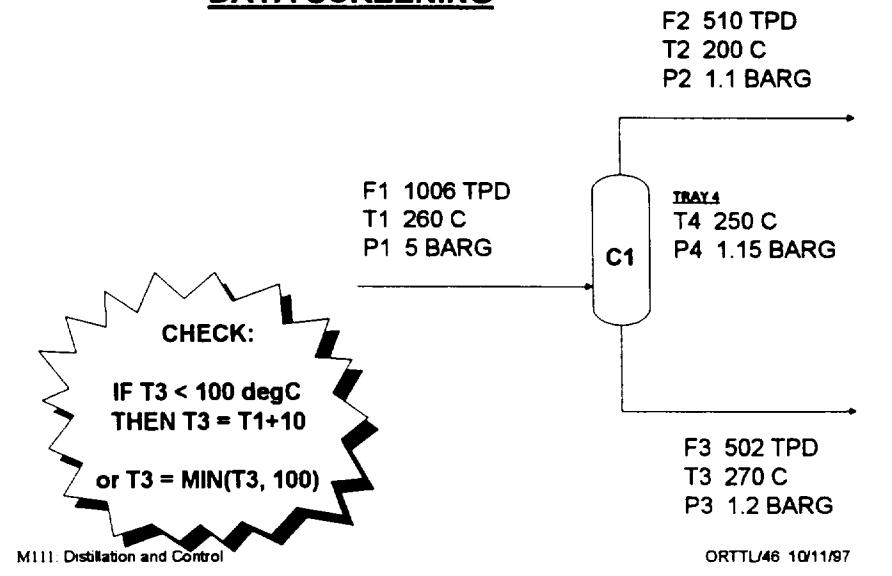
**GRAPHICAL  
REPRESENTATION**



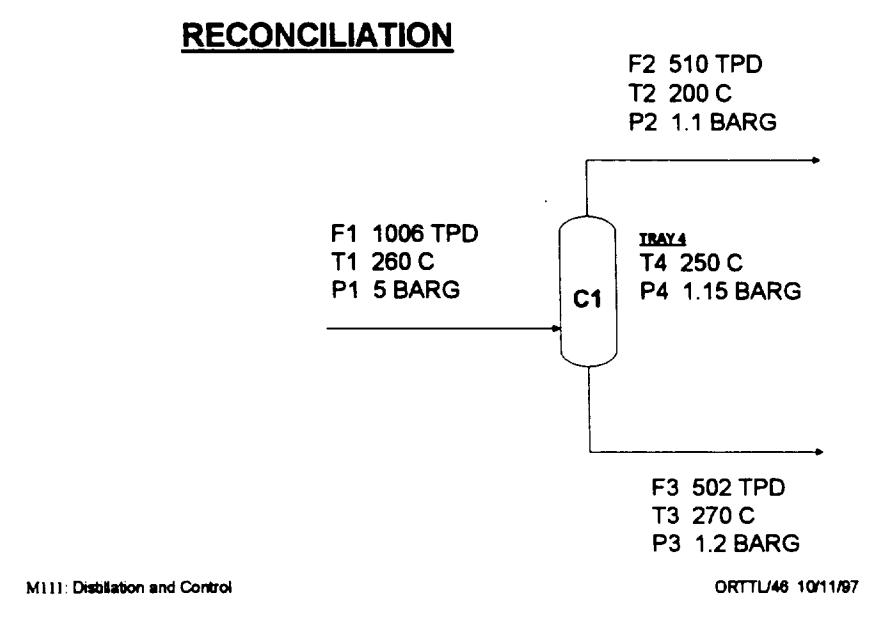
**RAW MEASUREMENT**



## DATA SCREENING

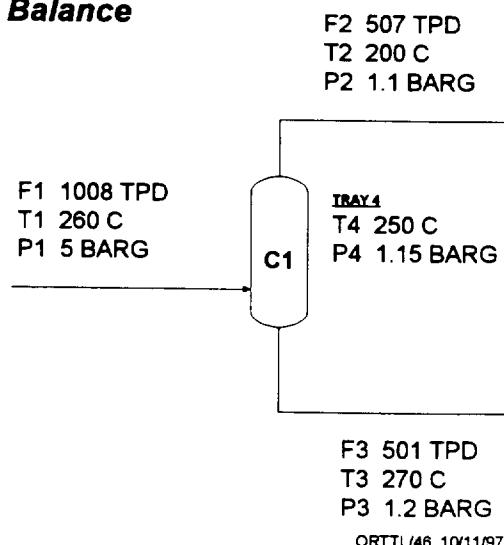


## RECONCILIATION



## RECONCILIATION

### **1<sup>st</sup> Deliverable: Unit Balance**

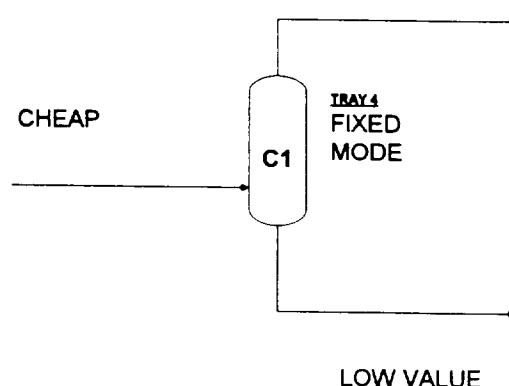


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## SLOGAN OPERATION

HIGH VALUE

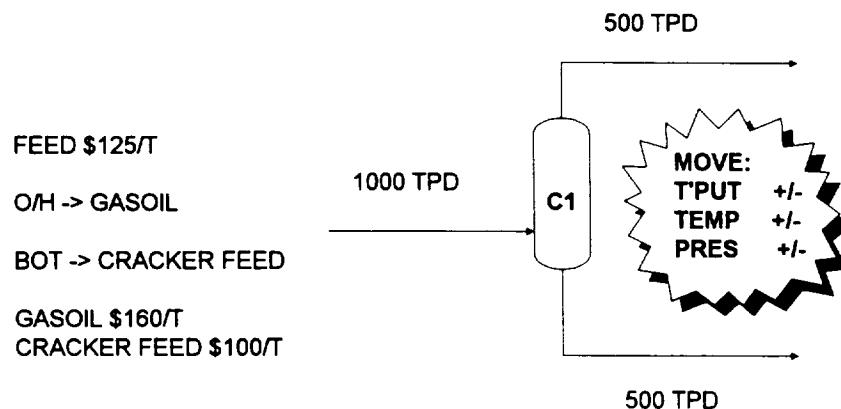


LOW VALUE

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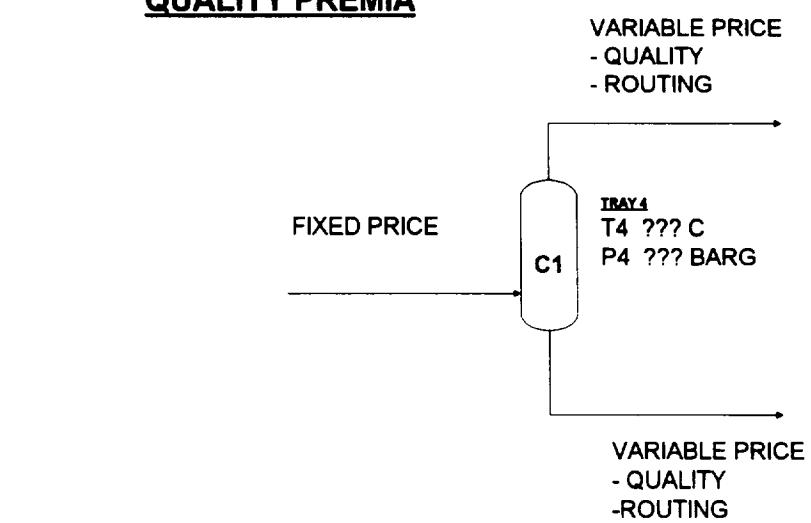
## SIMPLE ECONOMICS



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## QUALITY PREMIA



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## MORE ECONOMICS

FEED \$125/T

O/H -> GASOIL

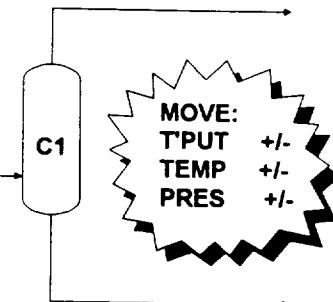
BOT -> CRACKER FEED

GASOIL	\$160/T
Ref. Cloud	-6 degC
Cost of Cloud	\$2/T/degC

CRACKER FEED \$100/T

500 TPD  
Cloud +2 degC

1000 TPD



### MODEL RESULTS:-

Increasing overhead yield by 10 tpd  
will increase cloud by 0.5 degC

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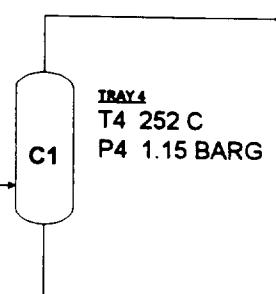
500 TPD

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## RATING CASE

F2 507 TPD  
T2 203 C  
P2 1.1 BARG

F1 1008 TPD  
T1 260 C  
P1 5 BARG

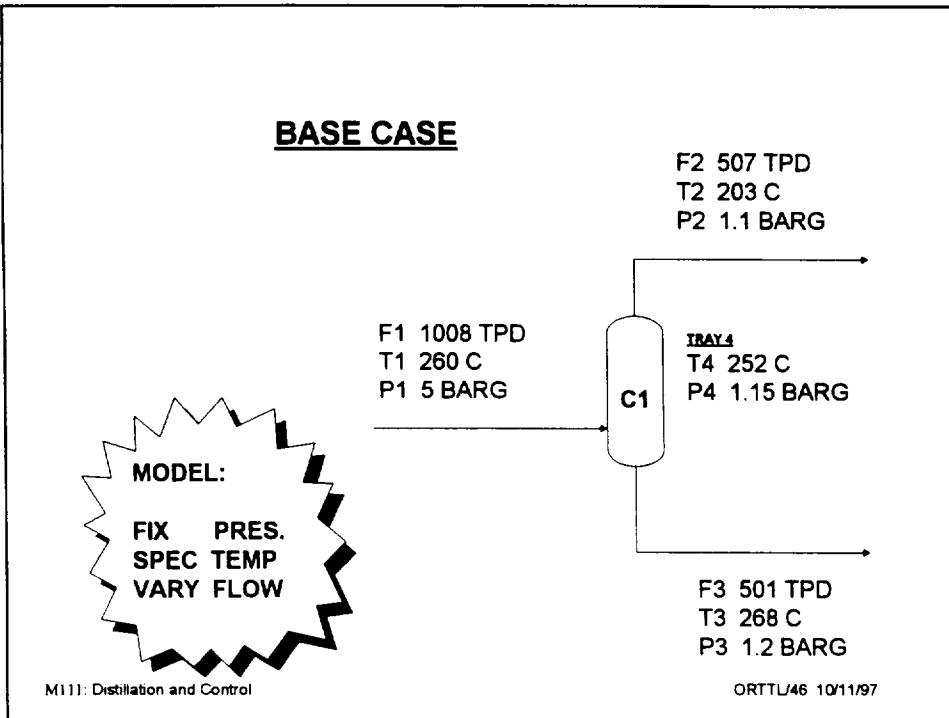


F3 501 TPD  
T3 268 C  
P3 1.2 BARG

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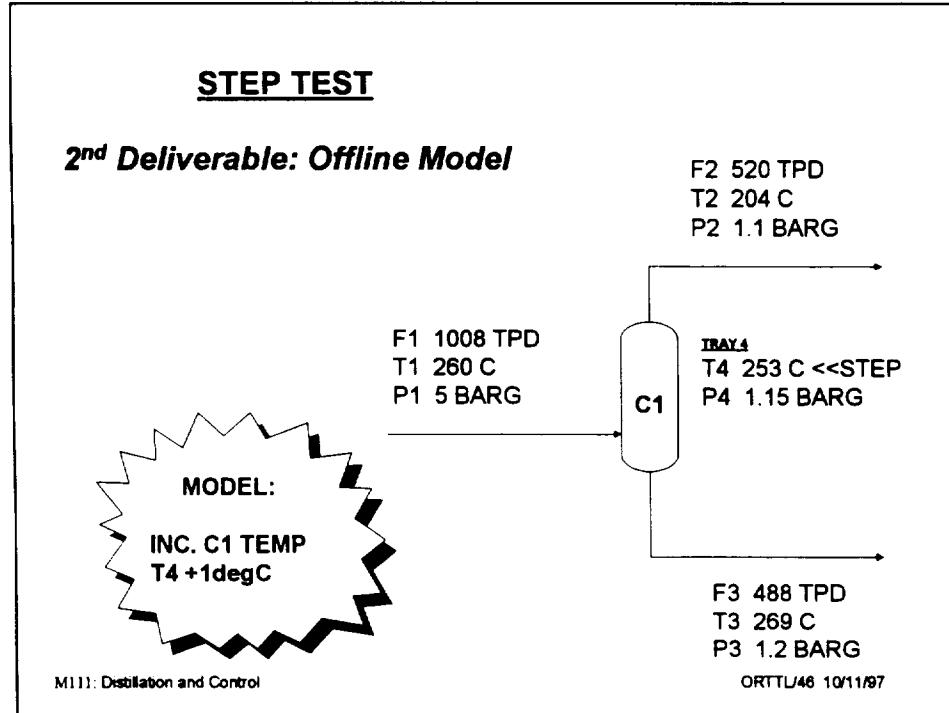
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## BASE CASE



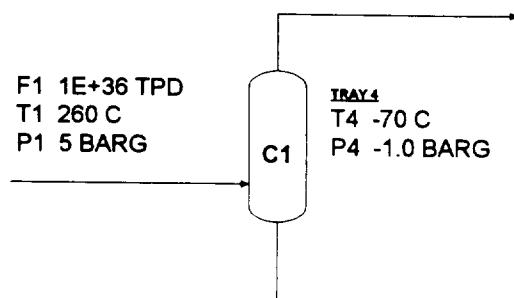
## STEP TEST

### **2<sup>nd</sup> Deliverable: Offline Model**



OPTIMISATION CASE  
UNCONSTRAINED

F2 1E+36 TPD  
T2 -70 C  
P2 -1.0 BARG

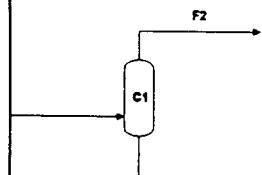


F3 1E+29 TPD  
T3 -70 C  
P3 -1.0 BARG

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CONSTRAINT CALCULATION



F2, HIGH LIMIT

Raw Measurement:	510 TPD
Base Case Value:	507 TPD

PROCESS HIGH ALARM 1000 TPD

ECONOMIC HIGH LIMIT 750 TPD

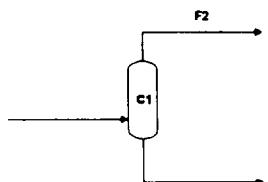
MAX MOVE (NORMAL STEP) +/- 50 TPD

VALVE MAX OPENNING	100%
Valve Position	94%
Valve Gain	6 TPD/%

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## CONSTRAINT CALCULATION



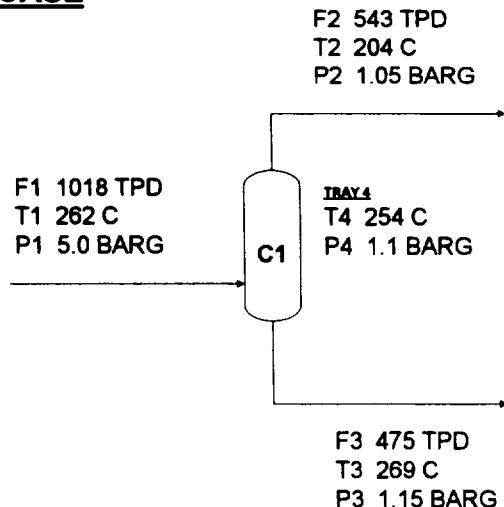
**F2, HIGH LIMIT**

Raw Measurement:	510 TPD
PROCESS HIGH	1000 TPD
ECONOMIC HIGH	750 TPD
NORMAL STEP	510 + 50 TPD =560 TPD
VALVE MAX OPENNING	510 + 6*(100-94) =546 TPD
MIN of MAX	MIN(1000,750,560,546) =546 TPD
Base Case Value	507 TPD
BIAS	507-510 TPD =-3 TPD
Model Constraint	546 + (-3) TPD = 543 TPD

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## OPT CASE



F2 543 TPD  
T2 204 C  
P2 1.05 BARG

F1 1018 TPD  
T1 262 C  
P1 5.0 BARG

F3 475 TPD  
T3 269 C  
P3 1.15 BARG

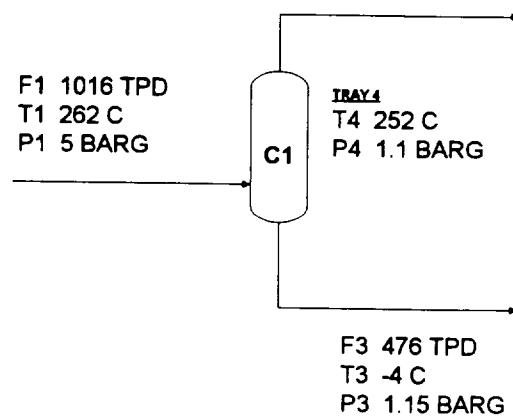
M111: Distillation and Control

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## OPT CASE, UNBIASED

### **3<sup>rd</sup> Deliverable: On-line, Open Loop**

F2 546 TPD  
T2 201 C  
P2 1.05 BARG



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## IMPLEMENT

### **4<sup>th</sup> Deliverable: On-line, Closed Loop**

#### **BASE LAYER SETPOINTS**

Temp  
Pressure

(Reflux Rate, Feed Rate)

#### **SMOC SETPOINTS**

Any (inc. Product Rate)

**!!OPT AND APC MODELS MUST BE (REASONABLY) CONSISTENT!!**

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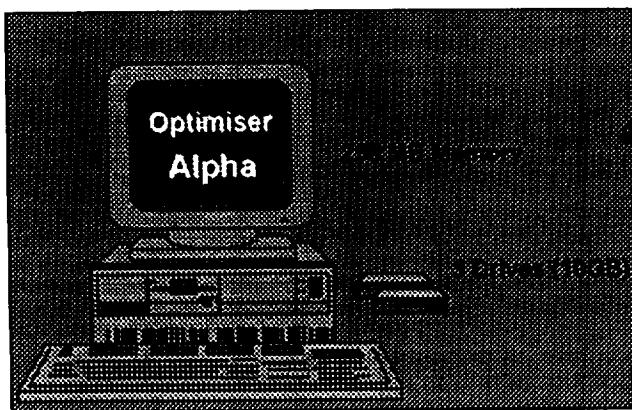
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			RAW	SCREENED	RECOND	RATE	BASE	(STEP)	OPT	OPT
F1	FEED	TPD	1008	1008	1008	1008	1008	1008	1018	1018
F2	OVERHEAD	TPD	510	510	507	507	507	520	543	546
F3	BOTTOMS	TPD	502	502	501	501	501	488	475	476
T1	FEED	degC	260	260	260	260	260	260	262	262
P1	FEED	berg	5	5	5	5	5	5	5	5
T4	TRAY 4	degC	250	250	250	252	252	253	254	252
P4	TRAY 4	berg	1.15	1.15	1.15	1.15	1.15	1.15	1.1	1.1
T2	OVERHEAD	degC	200	200	200	203	203	204	204	201
T3	BOTTOMS	degC	-5	270	270	268	268	269	269	264
			SCN-RAW	REC-RAW	RATE-RAW	BASE-RATE	(STEP -BASE)	OPT-BASE	OPT-RAW	OPT UNBIASED
F1	FEED	TPD	0	2	2	0	0	10	10	10
F2	OVERHEAD	TPD	0	-3	-3	0	13	36	36	36
F3	BOTTOMS	TPD	0	-1	-1	0	-13	-26	-26	-26
T1	FEED	degC	0	0	0	0	0	2	2	2
P1	FEED	berg	0	0	0	0	0	0	0	0
T4	TRAY 4	degC	0	0	2	0	1	2	2	2
P4	TRAY 4	berg	0	0	0	0	0	-0.05	-0.05	-0.05
T2	OVERHEAD	degC	0	0	3	0	1	1	1	1
T3	BOTTOMS	degC	275	275	273	0	1	1	1	1

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# **SPOT** Hardware



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# Optimisation and Control

## On-line Optimisation

### Advanced Process Control

Requires  
Generates  
Rigorous  
Steady State  
Total  
Global Plant

Provides  
Implements  
Dynamic  
Response  
Section(s) of  
Local

### Characteristics

Stable Operation  
Set Points  
Models  
Plant Operation  
Optimisation