ENERGY BALANCE

The Jet Mixer:

Heat is evolved when nitric and sulphuric acids are mixed together and the temperature rise can be large. In this process mixed acid is recycled to reduce the concentrations of nitric acid and to reduce the rise in temperature.

The heat liberated in mixing may be determined from the chart developed by $M_C C_{URDEY}$ and K_{INLEY} from the experimental data of R_{HODES} and N_{ELSON}. The relative enthalpy of mixtures of nitric acid, sulphuric acid and water is plotted using percent nitric and in anhydrous acid as abscissa and percent by weight of total acid as ordinate. (Ref –1)

Table 5.1

Material In	Relative enthalpy (KJ/kg)	Weight		Heat
		Tons/day	kg/hr	kJ / hr
HNO_3 60%	-202.362	21.05	877.07	-177485.6
H_2SO_4 96%	-81.41	42.76	1781.56	-145036.8
Recycle 75% acid	-290.75	197.88	8244.84	-2397187.2
	Σ =	261.69	10903.5	-2719709.2
Material out				
Mixed acid	-279.12	261.69	10903.5	-3043384.9

By subtraction, the heat evolved on mixing

= -3043384.9 + 2719709.6

= -323675.3 kJ/hr

This is the amount evolved for 10903.5kg of mixed acid.

If each acid is at the same temperature before entering the mixer, the temperature rise will be given by:

$$\frac{\text{Heat evolved in forming 1 lb. Of mixed acid}}{\text{Specific heat of the acid}}$$

Specific heat of mixed acid = 0.46 x 4.186 = 1.9256 kJ/kg k

Heat evolved = 323675.3 kJ / 10903.5 kg acid

= 29.685 kJ/kg

Therefore, temperature rise = $\dfrac{323675.\ 3}{10903.5 \times 0.46 \times 4.186}$ = 15.42 °C

The cycle acid entering the jet will be assumed to be at a temperature of 20°C. If a 15°C datum temperature is assumed throughout, then heat/kg of cycle acid above datum will be given by:

= (20 – 15) x 4.46 x 4.186 = 9.6278 kJ/kg

Therefore sensible heat above the datum in cycle acid.

= 8244.84 x 9.6278

= 79379.67 kJ/kg

The weight of cycle acid entering the jet per kg of mixed acid

$$= \frac{8244.84}{10903.5} = 0.756 \ \frac{\text{kg cycle acid}}{\text{kg mixed acid}}$$

Thus the added sensible heat per kg of mixed acid

= 0.756 x 9.6278

= 7.279 kJ/kg

Now, if the fresh acids enter at 15° C and the cycle acid at 20° C, the rise in temperature in the jet above 15°C will be,

$$= \frac{29.685 + 7.279}{1.926}$$

= 19.2 °C

Thus the temperature of the mixed acid passing to the first nitrator will be 15 + 19.2 = 34.2°C, which is just about the temperature of nitration.

The total heat evolved during the acid mixing is:

323675.3 kJ/hr

Total weight of mixed acid = 10903.5 kg/hr

The sensible heat in the mixed acid above the datum

= (29.685 + 7.279) x 10903.5

= 403037 kJ/hr

Thermal Balance for Acid mixing:

Table 5.2

	Tons/day	Kg/hr	Temp °C	Heat KJ/hr
Heat entering the mixer				
Nitric acid	21.06	877.07	15	0
Sulphuric acid	42.76	1781.56	15	0
Cycle acid	197.88	8244.84	20	79362.7
Heat of dilution				323675.3
			TOTAL IN =	4.3037
Heat leaving the mixer				
Mixed acid	261.69	10903.5	34.2	403037
			TOTAL OUT =	403037

The Nitration Reaction:

Nitration of toluene is carried out with mixed acid in two nitration vessels connected in series. The temperature in the first will be 35°C and in the second it will be raised to 50°C to ensure that the reaction goes to the required degree of completion.

The first nitration:

The heat evolved arises from two sources

(a) The nitration of toluene
(b) The further dilution of acid.

The heat reaction of M.N.T. to form 60 percent ortho, 4 percent meta and 36 percent para compound can be calculated as 25.5 Kg cal/mole or 106.77 kJ/mole (1160.54 kJ/kg) of toluene nitrated.

Assume that all the toluene is nitrated.

Rate of feed of toluene

= 18.105 tons/day

= 754.375 kg/hr

$$\text{Heat evolved in the reaction} = 754.375 \times \frac{106.77 \times 10^3}{92}$$

$$= 875485 \text{ kJ/hr.}$$

Table 5.3

	Relative Enthalpy (kJ/kg)	*Weight* (kg/hr)	*Heat* (kJ/hr)
Material In Mixed acid	-279.12	10903.5	-3043384.9
Material Out Spent acid	-290.75	10499.6	-3052758.7

Therefore, Heat evolved = -3052758.7 + -3043384.9

= -9374.7 kJ/hr

Heat evolved by dilution = 9374.7 kJ/hr

Total heat evolved during nitration

= 9374.7 + 875485 = 884859.7 kJ/hr

The temperature of the reactor = 35°C

Inlet temperature of the toluene = 15°C

Inlet temperature of the acid = 34°C

Therefore heat absorbed by toluene

= 754.375 x 0.41 x 4.187 x (35 – 15)

= 25900.26 kJ/hr

Heat absorbed by acid:

= 10903.5 x 0.46 x 4.187 x (35 – 34.2)

= 16800.29 kJ/hr

Assume heat loss by radiation = 13500 kJ/hr

Thus the total heat to be removed by cooling water in order to maintain a temperature of 35°C in the reactor will be

= 884859.7 – 25900.25 – 16800.29 – 13500

= 828659.16 kJ/hr

Cooling water is available at 15°C and 32000 kg/hr

$$\text{Temp rise of water} = \frac{828659.7}{32000 \times 4.187} = 62°C$$

Thus outlet temp of water = 15 + 6.2 = 21.2°C and sensible heat in this water above datum will be 828659.16 kJ/hr

After nitration,

Weight of spent acid = 10499.6 kg/hr

Thus sensible heat above datum

= 10499.6 x 0.46 x 4.187 x (35 – 16) = 404448.8 kJ/hr

Weight of MNT layer,

$$= \frac{27.213 \times 10^3}{24} = 1133.875 \text{ kg/hr}$$

∴Sensible heat above datum

= 1133.875 x 0.43 x 4.187 x (35 – 15)

= 40828.8 kJ/hr

∴Total sensible heat in nitration slurry

= 404448.8 + 40828.8

= 445277.6 kJ/hr

The Second Nitrator:

In this vessel the temperature is raised to 50°C. Since the reaction is almost complete in the first stage, the heat required can be calculated by considering the sensible heat necessary to raise the final products through 15°C.

Heat for spent acid = 10499.6 x 0.46 x 4.187 x (50 – 35) = 303336.6 kJ/hr

Heat for MNT layer = 1133.875 x 0.43 x 4.187 x (50 – 35)

= 30621.6 kJ/hr

The heat loss by radiation will be about twice that from the first vessel, say 27000 kJ/hr.

Therefore the total heat to be supplied = 303336.6 + 30621.6 + 27000

= 360958.2 KJ/hr

The additional sensible heat in the nitration mixture

= 30333.6 + 30621.6

= 333958.2 KJ/hr

Thus total sensible heat above the datum

= 445277.6 + 333958.2

= 779235.8 KJ/hr

The heat required for the second nitrator is best supplied by hot water. If the inlet temperature is 80° C, the outlet temperature should not fall below 70° C in order to maintain a satisfactory temperature difference between the heating medium and the products.

Temperature fall = 80 – 70 = 10° C

Therefore weight flow of water $= \frac{360958.2}{10 \times 4.187} = 8620.9$ Kg/hr

The water can be recycled and reheated to 80° C by steam injection, the details of which are not given here.

Sensible heat in hot water above datum

= 8620.9 x 4.187 x (80 – 15) = 2346221 KJ/hr

Sensible heat in exit water above datum

= 8620.9 x 4.187 x (70 – 15)

= 1985264 KJ/hr

Thermal Balance for Nitration:

First Nitrator:

Table 5.4

	Tons/day	Kg/hr	Temp ° C	KJ/hr
Heat Entering				
Toluene	18.105	754.375	15	Nil
Mixed acid	261.682	10603.5	34.2	403037.00
Cooling water		32000	15	Nil
Heat of reaction & dilution				884859.70
			Total in =	1287436.70
Heat Leaving				
Nitration mixture	279.789	11657.875	35	445277.60
Cooling water		32000	21.2	828659.16
Radiation loss				13500.00
			Total out =	1287436.76

Second Nitrator:

Table 5.4

	Tons/day	Kg/hr	Temp ° C	KJ/hr
Heat Entering				
Nitration mixture	279.789	11657.875	35	445277.6
Hot water		8620.9	80	2346221.0
			Total in =	2791499.6
Heat Leaving				
Nitration mixture	279.789	11657.875	50	779235.8
Radiation loss				27000.0
Hot water		8620.9	70	1985264.0
			Total out =	2791499.8

Spent Acid Cooling:

Spent acid leaves the separator at a temperature of 50^0C and the greater part is recycled as cycle acid. The temperature of cycle acid, however, is 20^0C and so a cooler is required to remove the excess heat.

The rate of flow of spent acid = 10499.6 kg/hr

Temperature drop = 1.926 KJ/kg ^{0}C

Temperature heat to be removed = 10499.6 x 1.926 x 30

= 606666.89 KJ/hr

This is conveniently done by using 50000 kg/hr of cooling water. Assuming an inlet temperature of 15^0C the temperature rise will be

$$= \frac{606666.89}{50000 \times 4.187} = 2.9^0\text{C}$$

Therefore outlet temperature = 2.9 + 15 = 17.9 $\simeq$ 18^0C

The sensible heat in the outlet cooling water will then be 606666.89 kJ/hr.

The relatively large quantities of water are necessary since the temperature differences between water and spent acid are so small.

The sensible heat in the spent acid from the nitrator above the datum is

= 10499.6 x 1.926 x (50 – 15)

= 707778.04 kJ/hr

Heat to be removed = 606666.89 kJ/hr

Therefore sensible heat in cooled spent acid at 20^0C above the datum

= 707778.04 – 606666.89

= 101111.15 kJ/hr

Only a portion of the cooled acid is recycled and from the thermal balance for this acid mixing it was shown that the sensible heat in cycle acid above the datum was 79362.7 kJ/hr.

Thus heat in excess acid = 101111.15 – 79362.7

= 21748.45 kJ/hr.

Thermal Balance for the Spent Acid Cooling:

Table 5.5

	Tons/day	Kg/hr	Temp^{0}C	Kj/hr
Heat entering				
Spent acid	252.576	10524	50	707778.04
Cooling water			15	Nil
			Total In =	707778.04
Heat leaving				
Cycle acid	197.88	8245	20	79362.7
Excess acid	54.696	2279	20	21748.45
Cooling water			18	606666.89
			Total out	707778.04

Washing Process

After separation the mixture of mononitrotoluenes is washed with water and dilute alkali. The total wash liquor amounts to18.75 tons /day of water and 9.59 tons /day of alkali. The washing is carried out continuously with countercurrent flow in a Holley-Mott washer.

Any heat of reaction can be neglected. The wash liquors enter at 15^0C and the mononitrotoluenes , in passing through the washer will be cooled to about 15^0C.

Rate of flow of wash liquors = $\frac{(18.75 + 9.59)}{24} \times 10^3$

= 1180.83 kg/hr

Heat in the mononitrotoluenes above the datum will be

= 1133.875 x 0.43 x 4.187 x (50 – 15)

= 71450.4 kJ/hr

All this heat is transferred to the wash liquors and so the temperature rise of the latter will be:

$$= \frac{71450.4}{1180.83 \times 4.187} = 14.5^0C$$

Therefore outlet wash liquor temperature = 15 +14.5 = 29.5^0C

Thermal Balance for mononitrotoluenes washing:

Table 5.6

	Tons/day	Kg/hr	Temp^{0}C	KJ/hr
Heat into washer				
Hot mononitrotoluenes	27.213	1133.875	50	71450.4
Wash liquors	28.338	1180.75	15	Nil
			Total in =	71450.4
Heat leaving washer				
Washed mononitrotoluenes	26.536	1150.67	15	Nil
Wash liquors	29.005	1208.54	29.5	71450.4
			Total out =	71450.4

The Topping Still:

The washed mononitrotoluenes contain toluene, paraffins and water as impurities and heat sensitive by-products. These are removed in a batch still under high vacuum. The most suitable method of treatment was found to be the distillation of one-quarter of a week's production on each of the first four days. Forerunning are removed first and 80 per cent of the crude mononitrotoluenes. The residues from these distillations are collected and redistilled on the fifth day. In this way the impurities are removed without building up excessive concentrations of explosive poly-nitro compounds.

Daily output of mononitrotoluenes = 26.536 tons

Therefore production per week = 5 x 26.536 = 132.68 tons.

Thus quantity distilled on each of the first four days = $\frac{132.68}{4}$ = 33.17 tons

The amount to be distilled on the fifth day can be shown to be 26.05 tons and for the purpose of this energy balance only the primary distillation of 33.17 tons will be considered.

The still pressure is such that the boiling point of the mixture is 105^0C. If the charge enters at 15^0C the heat required will be

$= 33.17 \times 10^3 \times 0.34 \times 4.187 \times (105 - 15)$

$= 4249813.37$ KJ

This quantity of heat can be supplied in half an hour.

The charge to this still amounts to 33.17 tons per day and it is distilled into three fractions-forerunnings, distillate consisting of 80 per cent of the mononitrotoluenes and residue.

Table 5.7

Component	**Feed**	**Forerunnings**	**Distillate**	**Residue**
Mononitrotoluenes	32.562	-	26.05	6.512
Toluene	0.229	0.229	-	-
Paraffins	0.229	0.229	-	-
Water	0.160	0.160	-	-
	33.180	0.618	26.05	6.512

Total mononitrotoluenes distilled

$$= 26.05 \text{ tons}$$

$$= \frac{26.05 \times 10^3}{137}$$

$$= 190.146 \text{ kgmol}$$

The latent heat of vaporization $= 11900 \times 4.187$

$$= 49825.3 \text{ KJ/kg mol.}$$

Since no reflux is used in this distillation, the heat to be supplied is given by

$$= 190.146 \times 49825.3 = \underline{9474081.5 \text{ KJ}}$$

The whole distillation may be carried out in 12h.

Therefore total time including heating up =12 + 0.5

$$= 12.5 \text{ hrs}$$

Total heat required = 4249813.37 + 9474081.5

$$= 13723894.87 \text{ KJ}$$

Thus hourly rate of supply $= \frac{13723894.87}{12.5}$

$$= 1097911.59 \text{ KJ/hr.}$$

This heat is supplied by steam at 2.81 kg/cm^2

The distillation rates for the three fractions are given as follows:

Mononitrotoluenes $= \frac{26.05 \times 10^3}{12} = 2170.83$ kg/hr

Forerunnings $= \frac{0.618 \times 10^3}{12} = 51.5$ kg/hr

Residues $= \frac{6.512 \times 10^3}{12} = 542.67$ kg/hr

Therefore feed rate = 2765 kg/hr

The sensible heat in the residues at 120^{0}C above the datum will be

$$= 542.67 \times 0.34 \times 4.187 \times (120 - 15) = 81116.09 \text{ KJ/hr}$$

Therefore total heat in the mononitrotoluenes vapors

= 1097911.59 – 81116.09

= 1016795.5 KJ/hr

Thermal Balance for Topping Still:

This balance is shown on an hourly basis to bring it into line with the others, although in fact it is a batch process. The total heat in the forerunnings is not considered and the sensible heat in steam condensate is neglected. This procedure is satisfactory in avoiding unnecessary calculation, and the only heat in the steam to be considered will be latent heat.

This balance is not strictly accurate but it provides sufficient information for design estimates.

Table 5.8

	Kg/hr	Temp ^{0}C	KJ/hr
Heat entering	2765	15	Nil
Steam 40psi	510	130	1097911.59
		Total in =	1097911.59
Heat leaving			
Mononitrotoluenes vapour	2170.83	92	1016795.5
Mononitrotoluenes residue	542.67	120	81116.09
Forerunnigs	51.5	92	Nil
Condensate	510	130	Nil
		Total out =	1097911.59

The Batch Still Condenser and Product Cooler:

The vapor rising to the condenser amounts to

$$= \frac{2170.83}{137} = 15.845 \text{ kg mol/hr}$$

The latent heat = 49825.3 KJ/kg mol

Therefore condenser loading = 15.845 x 49825.3

= 789481.88 KJ/hr

The vapor condenses to a liquid at a temperature of 92^0C in the condenser and so the above figure represents the latent heat in the material.

The total heat above the datum of the vapors

= 1016795.5 KJ/hr

Thus sensible heat in condensed liquid

= 1016795.5 – 789481.88

= 227313.62 KJ/hr

The amount of cooling water required must be such that the outlet temperature is not excessive. The most suitable method of cooling and condensing consists of passing all the water through the cooler and condenser respectively. This enables high water velocities to be maintained and improves heat transfer, especially in the product cooler.

The temperature of the condensing vapor 92^0C and after leaving the condenser the product must be cooled to a suitable storage temperature say 25^0C.

Heat load on the cooler = 2170.83 x 0.34 x 4.187 (92 – 25)

= 207053.46 KJ/hr

Sensible heat in the cooled liquid = 227313.62 – 207053.46

= 20260.16 KJ/ hr

The total load on the both units = 789481.88 + 207053.46

= 996535.34 KJ/hr

If water at the rate of 16500 kg/hr is used at 15^0C then the rise in temperature

$$= \frac{996535.34}{16500 \times 4.187} = 14.4^0\text{C}$$

Therefore outlet water temperature = 15 + 14.4 = 29.4^0C

which is not excessive.

The outlet temperature of water from the cooler and hence the inlet temperature to the condenser is = 15 + 207053.46/(16500x4.187) = 18^0C

Sensible heat in cooling water above the datum :

Leaving cooler = 207053.46 KJ/hr

Leaving Condenser = 996535.34 KJ/hr

Thermal Balance Over Condenser and Cooler:

The Condenser:

Table 5.9

	Kg/hr	Temp ^0C	KJ/hr
Heat entering			
Mononitrotoluenes vapour	2170.83	92	1016795.5
Cooling water	16500	18	207053.46
Forerunnings	51.5	92	Nil
		Total in =	1223848.96
Heat leaving			
Mononitrotoluenes liquid	2170.83	92	227313.62
Cooling water	16500	29.4	996535.34
Forerunnigs	51.5	92	Nil
		Total out =	1223848.96

Product Cooler:

Table 5.10

	Kg/hr	Temp ^0C	KJ/hr
Heat entering			
Mononitrotoluenes liquor	2170.83	92	227313.62
Cooling water	16500	15	Nil
		Total In =	227313.62
Heat leaving			
Mononitrotoluenes liquor	2170.83	25	20260.16
Cooling water	16500	18	207053.46
		Total out =	227313.62

The Residues Cooler:

The residues, which are returned for redistillation, must be cooled before storage. Since this is a batch operation the product cooler may be used for this purpose.

The weight of residue = 6.512×10^3 kg

The liquor may be passed out in two hours and so the rate of flow becomes

$$= \frac{6512}{2} = 3256 \text{ kg/hr}$$

The residue temperature is 120^0C and is cooled to 40^0C

The heat load = 3256 x 0.34 x 4.187 x (120 – 40)

= 392626.71 KJ/hr

The sensible heat of the hot residues above the datum

= 3256 x 0.34 x 4.187 x (120 – 15)

= 486693.53 KJ/hr

Therefore sensible heat in cooled residue = 4866393.53 – 392626.71

= 94066.82 KJ/hr

Using 16500 kg/hr of water as before

$$\text{Temperature rise} = \frac{392626.71}{16500 \times 4.187} = 5.7^0C$$

Thus outlet temp = 20.7^0C

Thermal Balance over Residues Cooler:

Table 5.11

	Kg/hr	Temp ^0C	KJ/hr
Heat entering			
Mononitrotoluenes residue	3256	120	486693.53
Cooling water	16500	15	Nil
		Total In =	486693.53
Heat leaving			
Mononitrotoluenes residue	3256	40	94066.82
Cooling water	16500	20.7	392626.71
		Total out =	486693.53

The Continuous Distillation Unit:

The continuous still separates the purified mononitrotoluenes into two fractions, an almost pure ortho product and a residue containing a mixture of all three isomers, mostly the para compound.

Several heat exchange units are required, namely:

(a) Feed pre heater to raise the feed to 133^0C

(b) Total condenser for condensing the whole of the vapor at 117^0C

(c) Cooler for cooling the ortho product to 25^0C

(d) Bottoms cooler for cooling residue to 80^0C

(e) Reboiler at the column base.

The Feed Preheater:

The rate of flow of feed = 25 tons/day

$$= \frac{25 \times 10^3}{24}$$

$$= 1041.67 \text{ kg/hr}$$

Thus feed preheat required = 1041.67 x 0.34 x 4.187 x (133 – 15)

= 174982.27 KJ/hr

Steam at 90psi will be used with a latent heat of 2093.5 KJ/kg

Therefore steam required = $\frac{174982.27}{2093.5}$ = 83.58 kg/hr

Thermal Balance over Feed Pre heater:

Table 5.12

	Kg/hr	Temp ^{0}C	KJ/hr
Heat entering			
Mononitrotoluenes feed	1041.67	15	Nil
Steam at 90psi	83.58	160	174982.27
		Total In =	174982.27
Heat leaving			
Hot mononitrotoluenes	1041.67	133	174982.27
Condensate	83.58	160	Nil
		Total out =	174982.27

The Product Cooler and Total Condenser :

The water cooling arrangements on this still will be similar to those on the batch still, i.e, all the water is passé though the product cooler and then the condenser.

Product rate = 15 tons/day

$= \frac{15 \times 10^3}{24} = 625$ kg/hr

The product is cooled from 117^0 to 25^{0}C and so the temperature drop is 92^{0}C.

Therefore heat load on cooler = 625 x 0.34 x 4.187 x 92

= 81855.85 KJ/hr.

and the sensible heat of the hot product above the datum

= 625 x 0.34 x 4.187 x (117-15)

= 90753.225 KJ/hr

Thus the sensible heat in the cooled product

$$= 90753.225 - 81855.85$$

$$= 8897.375 \text{ KJ/hr}$$

Using 16500 kg/hr water as before

Temperature rise of water = $\frac{81855.85}{16500 x 4.187} = 1.2^0C$

and the inlet water to the condenser will be approximately 16^0C

since the cooling water enters at 15^0C the sensible heat in the cooling water leaving the cooler will be 812855.85 kg/hr

The vapor load on the condenser depends on the reflux ratio used. For a reflux ratio of 5:1 the vapor load will be

$$= \frac{625 \text{ x } 6}{137} = 27.372 \text{ kgmole/hr} = 3750 \text{ kg/hr}$$

The latent heat of vaporization = 11300 x 4.187 = 47313.1 KJ/kgmol

Therefore total heat load = 27.372 x 47313.1 = 1295054.173 KJ/hr

The sensible heat in the hot condensate

$$= 3750 \text{ x } 0.34 \text{x} 4.87 \text{ x } (117 - 15)$$

$$= 544519.35 \text{ KJ/ hr}$$

Total heat in the vapour above the datum

$$= 544519.35 + 1295054.173$$

$$= 1839573.523 \text{ KJ/hr}$$

16500 kg/hr of water are used and the temperature rise will be

$$= \frac{12950540.173}{16500 \text{x} 4.187} = 18.7^0C$$

Since the inlet temperature of the water is 16^0C the outlet temperature will be 34.7^0C.

The sensible heat in the hot cooling water

$$= 1295054.173 + 81855.85$$

$$=1376910.023 \text{ KJ/hr}$$

Thermal Balance over Condenser and Cooler

The product cooler:

Table 5.13

	Kg/hr	Temp ^0C	KJ/hr
Heat entering			
Mononitrotoluenes liquor	625	117	90753.225
Cooling water	16500	15	Nil
		Total In =	90753.225
Heat leaving			
Mononitrotoluenes liquor	625	25	8897.375
Cooling water	16500	16.2	81855.85
		Total Out =	90753.225

The vapor condenser:

Table 5.14

Heat entering			
Mononitrotoluenes vapour	3750	117	1839573.523
Cooling water	16500	16	81855.85
		Total In =	1921429.373
Heat leaving			
Mononitrotoluenes Liquor	3750	117	544519.35
Cooling water	16500	34.7	1376910.023
		Total out =	1921429.373

The Bottoms Cooler:

Para-mononitrotoluenes is obtained from the residue by a crystallization process and so it is convenient to store the residue at a temperature above the crystallizing point of 46^0C a value of 80^0C will be chosen.

Rate of flow of residues = 10tons/day $= \frac{10 \times 10^3}{24}$ == 416.67 kg/hr

The residues leave the still at 150°C and are cooled to 80°C, a temperature drop of 70°C.

∴ Heat to be removed = 416.67 x 0.34 x 4.187 x 70

= 41521.416 kJ/hr

The sensible heat in the hot liquor

= 416.67 x 0.34 x 4.187 x (150 – 15)

= 80077.016 kJ/hr

Thus sensible heat in cooled liquor.

= 80077.016 – 41521.416

= 38555.6 kJ/hr

Since this is a small quantity and since the bottoms temperature required is relatively high. It is possible to utilize the water leaving the condenser for cooling.

Although 16500 kg/hr of water are available, the entire quantity is not necessary, only one-half need be used.

The temperature rise in the cooling water = $\frac{41521.416}{8250 \times 4.187}$ = 1.2°C

∴ The final temperature of water = 34.7 + 1.2 = 35.9°C

The sensible heat in cooling water leaving is

bottom cooler = $\frac{1376910.023}{2}$ = 688455.012 kJ/hr

Thus sensible heat in the cooling water leaving

= 688455.012 + 41521.416

= 729976.428 kJ/hr

Thermal Balance over Bottom Cooler:

Table 5.15

	Kg/hr	Temp°C	KJ/hr
Heat entering			
MNT liquor	416.67	150	80077.016
Cooling Water	8250	34.7	688455.012
		Total In =	768532.028
Heat leaving			
MNT residue	416.67	.80	38555.6
Cooling Water	8250	35.9	729976.428
		Total Out =	768532.028

The Still Reboiler:

The heat required for reboiling can be determined by a heat balance over the still. Heat leaves the unit in the vapors and the residues and the quantities must all be expressed as heat above the datum.

Total heat in the vapors = 1839573.523 kJ/hr

Sensible heat in the residues = 80077.016 kJ/hr

∴Total heat leaving = 1919650.539 kJ/hr

Wight of reflux returning to still = 5 x 625

= 3125 kg/hr

Sensible heat returning to still = 5 x 90753.225

= 453766.125 kJ/hr

Sensible heat entering in feed = 174982.27 kJ/hr

∴Total heat entering = 453766.125 + 174982.27

= 628748.395 kJ/hr

The deficit in heat = 1919650.539 – 62874.395

= 1290902.144 kJ/hr

This must be supplied by steam in the reboiler at a temperature of 160°C.

The latent heat of steam at 160°C = 2210.736 kJ/kg

$$\therefore \text{Weight of steam required} = \frac{1290902.144}{2210.736}$$

= 583.924 kg/hr

Thermal Balance over Still Reboiler:

Table 5.16

	Kg/hr	Temp °C	KJ/hr
Heat entering			
MNT feed	1041.67	133	174982.27
Reflex	3125	117	453766.125
Steam	613.84	160	1290902.144
		TOTAL IN	= 1919650.539
Heat leaving			
MNT vapor	3750	117	1839573.523
MNT residue	416.67	150	80077.016
Condensate	613.84	160	NIL
		TOTAL OUT	= 1919650.539

The Crystallizing Unit:

Most of the para compound can be crystallized from the residues off the continuous still in the pure state. Using cooling water at 15°C, and allowing for 1°C temperature rise the mixture can be cooled to 20°C. The crystallizing point is 46°C and most of the para compound deposits between these temperatures.

Since the bottoms are stored at 80°C it is advisable to cool them to 50°C prior to their entering the crystallizer. A simple cooler can be used and water from the continuous till condenser at 34.5°C may be utilized.

The heat to be removed = 416.67 x 0.34 x 4.187 x (80 – 50)

= 17794.892 KJ/hr

Thus the sensible heat in the cooled liquor

= 38555.6 – 17794.892 = 20760.708 KJ/hr

4250 kg of water at 34.7°C, the temperature raise will be

$$= \frac{17794.892}{4250 \times 4.187} = 1°C$$

Thus outlet temperature water = 35.7°C

Sensible heat in cooling water entering = $\frac{1376910.023}{4}$ = 344227.506 kJ/hr

∴Sensible heat in cooling water leaving

= 344227.506 + 17794.892

= 362022.398 kJ/hr

In crystallizer itself the mixture is cooled to 20°C and crystals are deposited. The heat to be removed is the sensible heat and the heat of crystallization.

Sensible heat = 416.67 x 0.34 x 4.187 x (50-20)

= 17794.892 kJ/hr

The heat of fusion = 27 x 4.187 = 113.049 KJ/kg

The weight of crystals = 7.5 tons/day =312.5 kg/hr

Thus the heat of Crystallization = 312.5 x 113.049

= 35327.813 KJ/hr

Therefore total heat to be removed = 35327.813 + 17794.892

= 53122.705 KJ/hr

Since the cooling water temperature rise is only 1°C, the water required will be 53100 kg/hr .

Thermal balance for Crystallizing Unit:

Cooler:

Table 5.17

	Kg/hr	Temp °C	KJ/hr
Heat entering			
Mononitrotoluenes isomers	416.67	80	38555.6
Cooling water	4250	34.7	344227.506
		Total In =	382783.106
Heat leaving			
Mononitrotoluenes isomers	416.67	50	20760.708
Cooling water	4250	35.7	362022.398
		Total Out=	382783.106

Crystallizer:

Table 5.18

Heat entering			
Mononitrotoluenes liquor	416.67	50	20760.708
Cooling water	53122.705	15	Nil
Heat of fusion			35327.813
		Total In =	56088.521
Heat leaving			
Mononitrotoluenes liquor	104.17	20	2965.816
P- mononitrotoluenes crystals	312.5	20	Nil
Cooling water	53122.705	16	53122.705
		Total Out =	56088.521