OPTIMISATION OF CONDENSATE RECOVERY AT HULETT REFINERIES

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Abstract

In past years at the Refinery make-up water to boiler feed has, on average, been used at a rate of 20%. During the past three years much effort has been spent on reducing this make-up rate by providing the means for the recovery of additional condensate from process vapour. The paper describes the modification to the evaporator vapour distribution and hot water heat exchangers which was necessary, as well as the control system which was adopted, in order to minimise the loss of vapour to the evaporator condenser. Other areas of operation where direct steam injection heating has been replaced by indirect contact heat exchangers, is also covered. Levels of boiler water make-up before and after the changes clearly show the advantage to the Refinery of adopting such a scheme.

Introduction

Maximising the recovery of good quality condensate from process for use as boiler feed is of utmost importance, particularly in a central sugar refinery where the starting raw material contains very little natural moisture and therefore requires imported water for processing. Inadequate recovery of condensate during process requires additional imported water to be used for boiler feed make-up. This in turn incurs unnecessary costs by way of energy input and conditioning prior to it being used for steam production.

Make-up levels of up to 20% on total boiler feed have been used in past years at Huletts. During 1986, a replacement evaporator condenser was planned and it was decided that this would be designed for a duty to accommodate both a rearrangement in vapour distribution from the evaporator as well as a change in duty of three hot water heat exchangers. This modification together with a change to the evaporator vapour control logic would provide the opportunity to recover vapour which was previously passed onto the condenser, in the form of usable good quality condensate for boiler feed.

Evaporator Station

Vapour distribution

The evaporator station is used for the concentration of decolourised fine liquor from 66 brix up to 76 brix in preparation for crystallisation. The station consists of two parallel fed single effects of 273 m^2 and 268 m^2 heating surface, each coupled to individual flash vessels operating at a lower absolute pressure. The effects are driven by 120 kPa exhaust steam and are controlled to a vapour space pressure of −40 kPa whilst their flash vessels are controlled to −75 kPa.

The higher pressure of the two vapours generated, has two main uses:

1. The heating of raw sugar melt by injection into a direct contact cascade type heat exchanger.
2. The heating of treated water for process via two shell and tube type heat exchangers connected in series.

Any surplus 1st vapour is let down into the low pressure flash vessel range which together with flashed vapour, is then condensed in a conventional counter current cascade condenser. The vapour distribution as it was prior to the rearrangement is shown in Figure 1, and a vapour balance showing the potential recoverable condensate lost to the condenser, in Table 1.

Table 1

Range of Evaporator vapour quantities (ton/hr)

	Low	High
Generated:		
Evaporation rate	20,5	24,6
Vapour 1	18,4	22,1
Vapour 2	2,1	2,5
Usage:		
Melt heating	5,3	6,1
Water heating	4,0	9,0
Balance:		
Lost to condenser	6,2	14,5

It is evident from Table 1 that there is a wide range in vapour demand by the water heat exchangers. This is due to the fluctuating demand for hot water to process, dictated predominantly by the regeneration if ion exchange resin in the decolourisation station. The potential for recovering up to 14,5 t.p.h. of additional condensate represented some 76% of the total boiler water make-up requirement.

Plant rearrangement

In order to maximise on the recoverable surplus vapour condensate, whilst at the same time ensuring that all other heating requirements were satisfied, the following considerations needed to be addressed before any changes were made:

1. Additional shell and tube condensing surface area was required, using a steady flow cooling medium which is not subject to variable process conditions.
2. The wide variation in vapour demand by both the melter and hot water heat exchangers needed to be met.
3. The new evaporator condenser to be installed, needed to be designed such that it would be able to handle the full vapour condensing duty if required, or the small condensing load as a result of any surplus vapour not able to be condensed by the additional shell and tube condenser.

In order to satisfy these conditions, the following modifications to plant and control system were decided upon:

Shell and tube condenser

A spare shell and tube heater with a surface area of 300 m^2 previously used for process water heating was converted to serve either as a spare hot water heater or a vapour con-

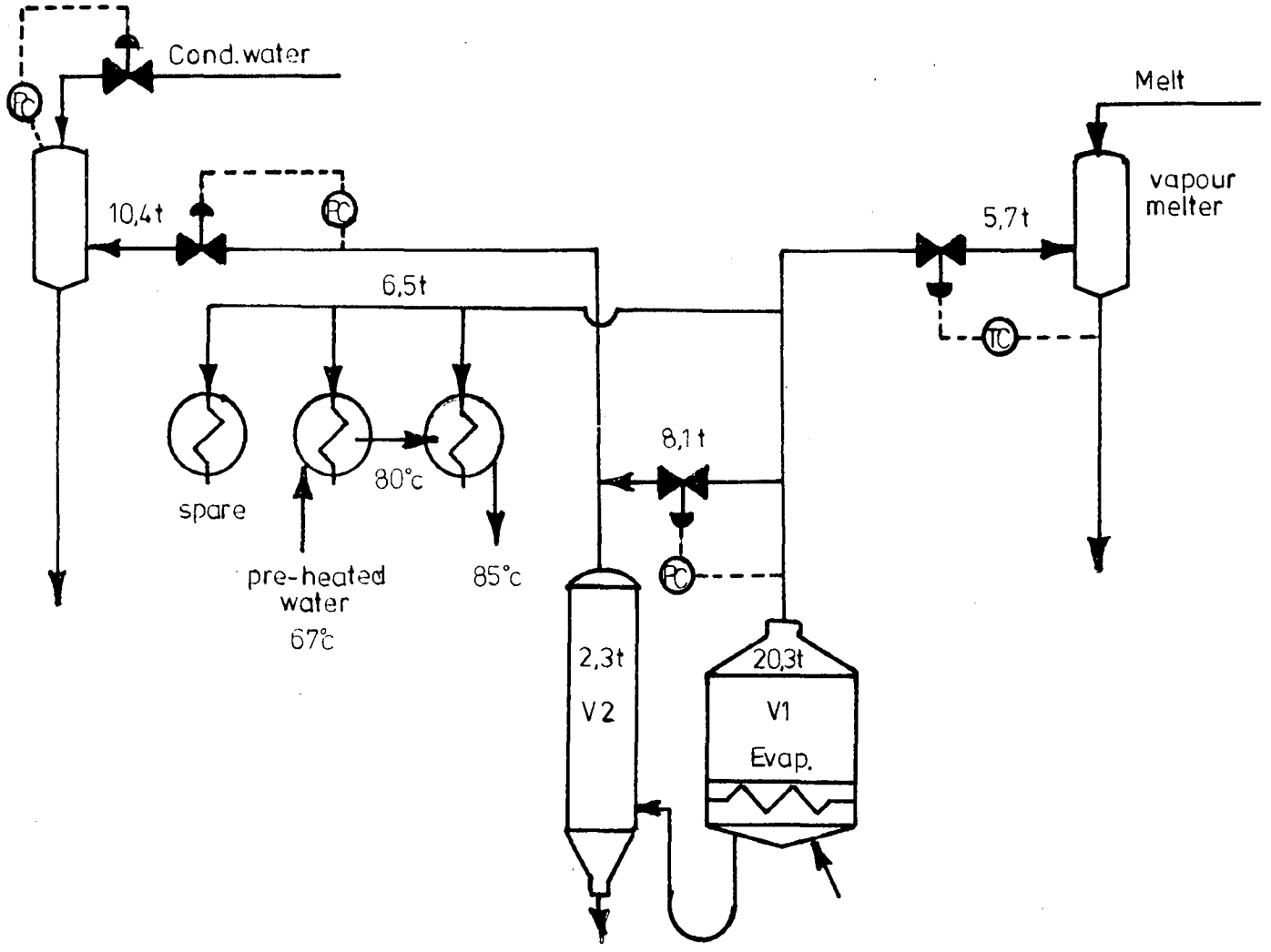

FIGURE 1 Vapour distribution before modification.

denser, using injection water from the closed circuit condenser water system as the cooling medium. This unit would condense vapour 2 and would operate in parallel with the condenser on the vapour side.

New condenser

In order to avoid designing a condenser with a high turn down ratio to cater for the wide variation in condensing load, it was decided that the heated injection cooling water, after passing through the heat exchanger, should be directed through the condenser in series. Additional cold injection water would be taken in by the condenser separately to accommodate any increase in condensing load which might be required. This method also has the advantage of creating the necessary temperature difference to drive the vapour to the shell and tube exchanger first, and only to the condenser once all available condensing surface in the shell and tube unit has been fully used to the extent of its heat transfer rate.

Control system

The modified vapour distribution and control system is shown in Figure 2.

The control system is designed to meet the following requirements in order of priority:

1. Maintain V1 and V2 vapour pressures at −40 kPa and −75 kPa respectively.
2. Meet the V1 demand to the vapour melter.
3. Condense as much surplus V1 as condensing capacity allows in the two series water heat exchangers, before any surplus is let down into the V2 range.
4. Condense as much vapour in the V2 range in the injection water shell and tube exchanger before letting any surplus down to the direct contact condenser.

The two vapour ranges are each controlled by single loop programmable controllers, which each take their respective inputs from the pressure in the range, and provide a split range output to two control valves. The vapour let down valve in each case has the lower priority of the split output in both systems. In the case of the V1 pressure control, the second valve introduces cold water to the pre-heated water which enters the heater supply in order that these heat exchangers are able to condense to their maximum before the let down valve relieves pressure into the V2 range. Likewise, in the V2 pressure control system, the valve on the heat exchanger discharge is required to be fully open, allowing maximum injection water flow through the unit for optimum condensing, before the let down valve relieves any excess vapour to the direct contact condenser. The absolute pressure controller on the condenser and temperature controller on the vapour melter both operate independently of the rest of the system.

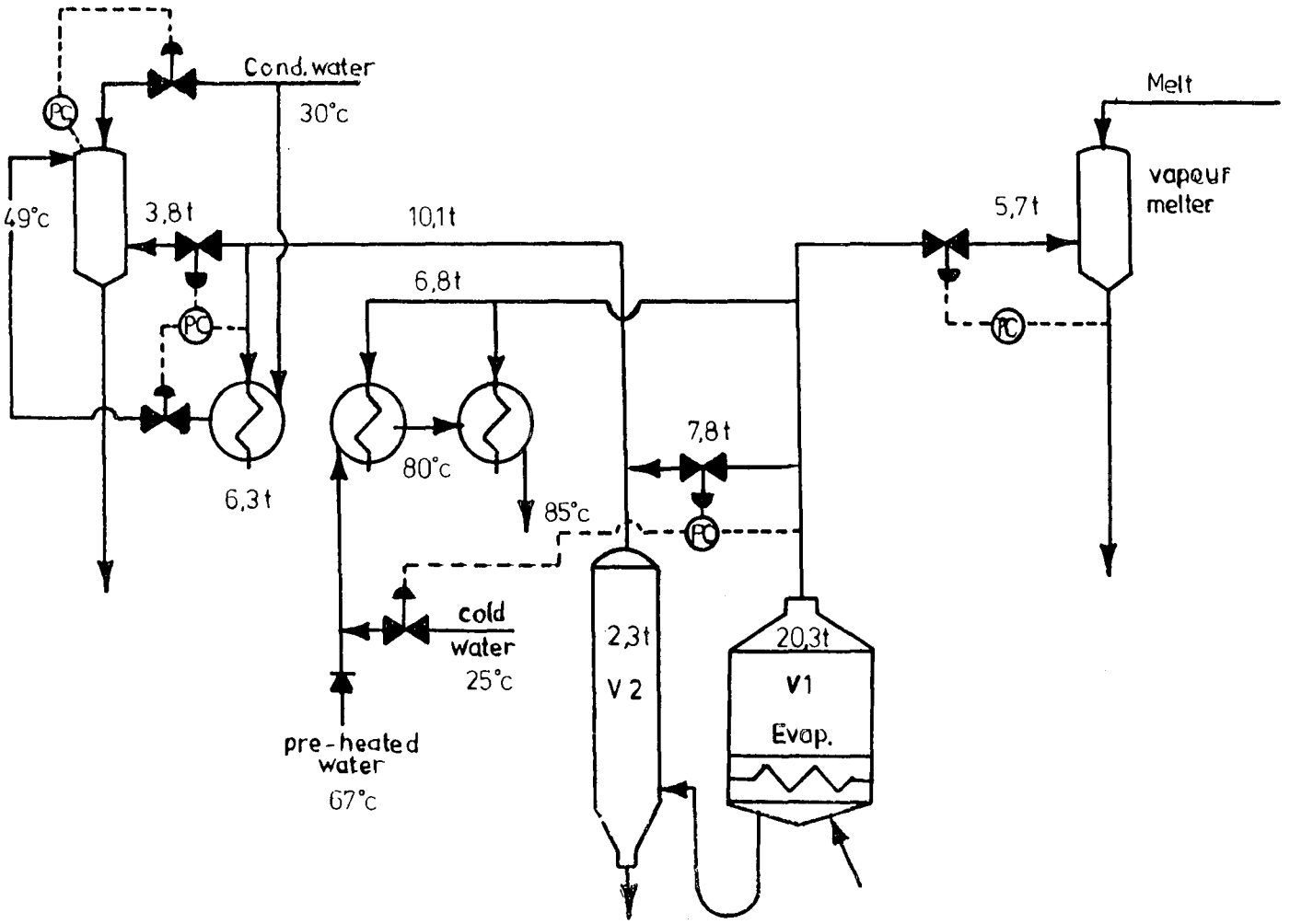

FIGURE 2 Modified vapour distribution and control system.

Direct steam injection

Wherever possible in the factory, direct injection heating with steam or vapour is being replaced by installing an indirect contact heat exchanger which enables condensate to be recovered. Where this is not possible, flashed vapour from some of the high temperature condensates is used for direct injection. Such a scheme has been introduced in the white pan station.

In this department the use of 120 kPa steam make-up to assist temperature control of the centrifugal wash water, sugar drier air heating and reject white sugar melting, has been eliminated by making more efficient use of flash vapour from high temperature pan condensate. The white pans operate on 410 kPa steam and this condensate had previously not flashed completely. This resulted in lost vapour being flashed in the boiler feed tank. A system has now been installed whereby additional flashed vapour at 35 kPa gauge pressure is used for these duties without the need for exhaust make-up. Centrifugal hot water tank temperature control has also been modified by replacing the direct injection system with a tubular heat exchanger from which flash condensate is now recovered and sent to boiler feed.

Results

The reduction in make-up rate achieved from these modifications is shown graphically in Figure 3. A reduction in feed water make-up of about 40% is shown from an average level of 20,4% during years 1977 to 1985, down to around 12% for the last two years.

The lack of sufficient indirect condensing surface area means that not all the surplus vapour generated by the evaporator at the higher evaporation rates can be condensed. The vapour balance under the modified system is shown in Table 2. A maximum rate of vapour lost to the condenser is now shown to be 7,6 t.p.h. which is a saving of 6,9 t.p.h. from the maximum recoverable surplus shown in Table 1.

Table 2

Range of Vapour quantities since modification (ton/hr)

	Low	High
Generated:		
Evaporation rate	20,5	24,6
Vapour 1	18,4	22,1
Vapour 2	2,1	2,5
Usage:		
Melt heating	5,3	6,1
Water heating	4,3	9,2
Additional exchanger	6,0	6,6
Balance:		
Lost to condenser	0	7,6

Substituting direct injection steam heating for an indirect contact heat exchanger in the white pan station hot water tank, as well as more efficient use of the flashed vapour from the white pan condensate, have provided a further 0,9 t.p.h. contribution towards boiler feed as well as an exhaust steam saving of 1,8 t.p.h. Table 3 provides an overall picture of the situation, showing the previous make-up rate together with the areas of additional recovery which are responsible for the make-up reduction.

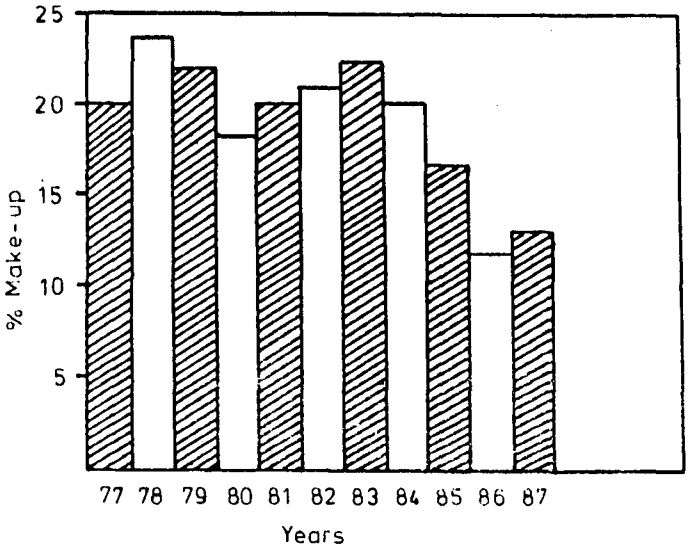

FIGURE 3 Feed water make-up 1977–1987.

Table 3

Overall make-up reduction

	Ton/hr	%
Previous make-up rate	19,0	20,4
Less: Additional evap. vapour recovery	6,6	7,1
Additional pan station recovery	0,9	1,0
Present make-up rate	11,5	12,4

Estimated net cost savings in energy and boiler chemicals as a result of the additional recovery of condensate, is shown in Table 4. The energy saving is based on the net effect of the coal equivalent of reduced energy required to raise the total boiler feed water temperature to 120°C, versus the cost of additional power required to be imported to compensate for the reduced power output by the Refinery power station due to the steam saving.

Table 4

Estimated cost savings due to additional condensate recovery

	Saving p.a.	% Total cost
Coal equiv. of net energy saving	R53 100	0,8
Boiler chemicals	R16 800	45
Water softener regenerant	R 1 700	35
Total annual saving	R71 600	1,10

Conclusion

Making more efficient use of existing equipment and rearranging the vapour distribution at the evaporator and white pan stations, have resulted in a 40% reduction in boiler feed water make-up.

Insufficient indirect condensing surface for the recovery of surplus evaporator vapour indicates a potential for recovering up to a further 7,6 t.p.h. condensate. The installation of additional condensing surface in the future would result in boiler water make-up being reduced further by 4% in this area of operation alone.

Progress thus far on reducing make-up water has saved the refinery an estimated R71 600 p.a. in net energy and boiler chemical cost. Potential for a further saving of R110 000 p.a. exists in recovering the additional 11,5 t.p.h. of condensate needed to eliminate boiler make-up altogether.

Acknowledgements

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