IMPROVING THE PROCESS STEAM ECONOMY IN A CANE SUGAR FACTORY: A CASE HISTORY OF A SUCCESSFUL COGENERATION PROJECT

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Abstract

The classic procedure to increase cogeneration export from a sugar factory is to maximise the production of high-pressure steam and to minimise the factory requirements of low-pressure process steam, in order to take more advantage of efficient fully condensing turbo alternator systems. Factory configurations that reduce process steam requirements need to be closely examined. This paper reviews the factory configurations that reduce process steam at Ingenio San Antonio, Nicaragua, over recent years, starting from conditions of low steam economy and moving to an operation with high steam economy, appropriate to a full cogeneration factory. The major achievements in this project include the exclusive use of #3 stage evaporator vapour for all raw pan stage operations, using new technology Roberts evaporators, using continuous vacuum pans for all three massecuites, and having the annex refinery operations on #2 stage vapour. Very low process steam consumptions have been consistently achieved. Ideas for further improvements are discussed, and the factors limiting further economies are outlined.

Introduction

To maintain the economic viability of a cane sugar factory, especially in the situation of volatility in the international sugar market, it is seen necessary to seek opportunities for co-products. Cogeneration is particularly appropriate in Nicaragua where power costs are relatively high. Consequently, Ingenio San Antonio (ISA) is undertaking a large project in cogeneration. Factory configurations that reduce process steam requirements have therefore been closely examined in recent years.

The classic procedure to increase cogeneration export from a sugar factory is to maximise the production of high-pressure steam and to minimise the factory requirements of low-pressure process steam, in order to take more advantage of efficient fully condensing turbo alternator systems.

This paper reviews the improvements in factory steam economy at ISA, Nicaragua, over recent years, starting from conditions of low steam economy and moving to an operation with high steam economy, appropriate to a full cogeneration factory. ISA operates with continuous crushing through a 6-month season, and has relatively good cane with a typical cane pol of around 13%, cane fibre of 13% to 14%, and mixed juice purity around 87%.

Process steam economy

It is known that, in beet factories, where fuel (coal or gas) has to be purchased, steam consumption can be reduced to below 30% on beet. To achieve these low steam consumption rates, boiler pressures have to be as high as 60 Bar with a steam temperature up to 500°C. Higher steam process pressures are used with, sometimes, only the last evaporator operating under vacuum. The trend in the beet industry is to have six and even seven effects in the evaporators, with the evaporators taking exhaust steam saturated at 131°C and discharging final vapour at 90°C. This high discharge temperature allows vapour from the last vessel to be suitable for heating or evaporation rather than wasted in a condenser.

With cane sugar, higher temperatures of evaporation are usually avoided as they generate more colour than do lower temperatures. To achieve lower energy use and comparable colour, residence times at
the higher temperature have to be reduced. Detailed knowledge of the effect of operating conditions (temperatures, residence times) on colour formation in cane sugar juice and syrup is needed.

A previous paper by Wright (2000) indicated the need for greater energy efficiency when cogeneration is implemented. It focussed on vapour bleeding as a method to achieve reductions in process steam consumption.

The reduction in process steam consumption would translate directly to an increased potential for cogeneration export or for bagasse saving. In the high-bleed cases, the process steam saving is up to 16.3% on cane, and this can increase power export in factories having co-generation facilities.

Most cane sugar factories produce a variety of sugar types, including direct consumption whites and refined sugars. These require extensive remelting of intermediate sugars and hence extra pan stage heating requirements.

The process steam economies required for cogeneration dictate the universal use of complete vapour bleeding to the pan stage. Some factories operate their pan stations exclusively on #1 vapour in the saturation temperature range of 108°C to 115°C without undue extension of pan cycle times and without pan stirrers.

Some factories operate successfully with juice heating systems configured for extensive vapour bleeding, and with a portion of their pan requirements heated with #2 vapour. One Indian factory has all its high-grade strike batch pans on #2 vapour (nominally 105°C saturation temperature) on a 3.5 h cycle and low-grade batch pans on #1 vapour (nominally 112°C saturation temperature).

Its ‘A’ strikes are slowed by the relatively low concentration of the evaporator syrup, presently required by the syrup sulfitation process. A second factory has #2 vapour bleed (at 102–105°C saturation temperature) to the entire pan stage of six unstirred batch calandria pans and three FCB continuous vacuum pans. Again the batch high-grade pan cycle is 3.5 h, and a moderately low syrup concentration is used.

A third Indian factory operates all pans (unstirred batch central calandria and floating calandria pans) on #1 vapour that is at just above atmospheric pressure due to very low process steam pressure. The bleed saturation temperature is only in the 100°C to 103°C range, and there is no sophistication in the pan calandria venting arrangements. Despite this, and despite a low syrup concentration, the ‘A’ strike cycle times are in the range 2.5 h to 3 h, the ‘B’ cycles are 3.5 h, and the C cycles are 6 h to 7 h.

The use of lower temperature bleed vapours free of superheat has a beneficial effect in preventing overheating of the massecuite in the vacuum pan tubes. This improves the sugar quality, as it lowers the formation of colour and crystal inclusions when compared with boiling on process steam, especially where some superheat is present in the steam.

Some other factories successfully operate the vapour line juice heater for the initial heating of the cold mixed juice with final stage vapour. The warm juice then flows in series through another vertical tubular primary heater which is heated by #3 vapour. The secondary heating is carried out in tubular heaters by #2 and then #1 vapour, and clarified juice heating is sometimes carried out using #1 vapour followed by process steam.

The heavy vapour bleed from #1 and #2 stages in ‘high bleeding’ systems results in these stage areas being much larger than the areas of the final three stages. However, as heavy bleed takes off almost as much water as is in the original syrup, there can be control problems with a varying pan stage vapour demand.

The juice concentration after #2 and #3 may sometimes exceed safe values, and the automatic control systems may have to be enhanced with a concentration sensor loop at an intermediate stage. The final vacuum temperature may have to be varied (raised) as a way of controlling the capacity of the set while keeping the intermediate vapour temperatures sufficiently high for effective heating.

Even with high remelt rates for intermediate and reject sugars, low syrup concentration (a requirement of syrup sulfitation operation) and low cane and syrup purities, some Indian factories have lowered their process steam consumption to be effectively below 42% on cane.

**Factory changes at Ingenio San Antonio**

The incentive for reduced steam consumption at ISA had not existed up till 1999, when the cogeneration project was first examined. At that time, ISA, along with many other cane sugar factories, was using the vapour temperature range from 126°C to about 54°C to carry out the evaporation required for process.
This was accomplished generally with single effect evaporation on the pan stage and quintuple effect evaporation in the evaporators. Generally the pan stage was sourced with #1 evaporator vapour. Juice heating was carried out using #1 vapour on the diffuser heaters, #2 and #1 vapour for the first heating stage, #1 vapour and process steam for the second heating stage, and process steam for the clarified juice heaters.

The control of evaporator liquor concentration was poor, and the average liquor brix low. Vapour vent losses were high. The result was relatively high vapour consumption on the pan stage, greater than 56% on cane. The scheme is illustrated in Figure 1.

A large step towards improved steam economy was made in the 1998–1999 season when a 120 m³ Tongatt-Hulett continuous vacuum pan (CVP) was installed and commissioned for processing all the A massecuites. It was operated with #1 vapour at 70 kPag (115.4°C), for the season 1998–1999, but the low head and high heating surface characteristic of this pan enabled a change to #2 evaporator vapour, then at −14 kPag pressure (103.4°C). This change was made in season 1999–2000.

A number of opportunities for minor economies in process steam as listed by Attard (1989), Broadfoot (1999) and Wright (2000) were also considered at this time. Attention was given to minimising evaporator calandria vent losses, and in venting boiler feed tanks to the body of the cane diffuser. The evaporator bleed scheme at this stage is shown in Figure 2.

After May 2000, a large construction program was undertaken in the factory. This included the installation of two additional 140 m³ CVPs, for B and C massecuites, and a re-allocation of the old batch pans to strengthen the refinery pan boiling on #2 vapour, and for seed preparation for the CVPs.

As well, a very large #2 stage evaporator vessel of novel SRI Radial Flow Roberts design (Wright et al., 2002) was installed, and some of the diffuser juice heaters transferred to mainstream juice-heating duties. The improvements allowed the entire pan station to be operated on #2 and #3 stage vapour, and more #2 vapour could be used for juice heating. Better control enabled the evaporator liquor brix to be raised to close to 65 Brix, which then reduced the vapour demand of the pan station. The evaporator bleed scheme at this stage is shown in Figure 3.

The new evaporator was commissioned in January 2001 and was able to maintain the #2 vapour pressures at above 20 kPag (105°C) even when all the CVPs were on #3 vapour and all batch vacuum pans were being operated on #2 vapour. This achieved an improved steam economy performance for the factory, even at the peak of the season when the liquor solids to be handled were at their highest.
Fig. 2—Scheme for process heating at ISA after the installation of the CVP for A massecuite.

Fig. 3—Scheme for process heating at ISA after the installation of the CVP for all massecuites, a new No. 2 stage SRI evaporator, and juice heating changes.
In the 2002–2003 season, ISA operated with the two big SRI evaporators (5300 m²), and two large Alfa Laval plate heaters for heating juice from 44°C to 95°C. The A CVP was expanded to 180 m³ nominal capacity. In the 2003–2004 season, ISA added two additional Alfa Laval plate heaters, so these were used for heating juice from 44°C to 95°C and then from 93°C to 105°C, using #3, #2 and #1 vapour in sequence. All the CVPs used #3 vapour at 7 kPa, while the refinery pans use #2 vapour. The A CVP was expanded still further, to have a total of 240 m³ nominal capacity, with twin units in the same body. The nominal residence time in the strike CVPs (at 14 065 t/day cane rate) is currently 4.10 h, 3.43 h and 7.35 h for A, B and C respectively. There is presently an obvious overcapacity in the A-strike capacity, but this was provided to minimise disruption during scheduled A-pan cleaning operations.

A survey of vent losses from factory vessels indicated relatively high losses associated with steam ejectors used to maintain vacuum in the sulfitation towers, and in vessel off-line cleaning operations. Ideas for minimising these losses were investigated. An additional item that was considered was the flash loss from condensates collected at the boiler feed water tanks. The scheme documented in a recent paper (Wright and Joyce, 2001) was favoured. In this scheme, the flash loss can be largely eliminated if the ‘primary’ condensate, produced by condensation of the process steam at a temperature of ~125°C is held in a small pressure tank, and used preferentially as the major boiler feed water. It was estimated at ISA that this would result in the production of approximately 2.8% extra high-pressure steam from the same bagasse.

A chart summarising the improvements in steam economy over recent years is given in Figure 4. Here, the values for each season are shown subdivided into early (a), middle (b) and late season (c). One would expect the late season values to be higher because of higher sugar yield from the cane (and hence higher vapour use for crystallisation), but this is offset in some cases by the higher brix in clarified juice late in the season. Values for individual weekly periods in the most recent season have been below 41% on cane, which is considered excellent for a factory that re-processes about one third of its product sugar to refined sugar. The latter would impose an extra evaporation load of approximately 4% on cane on the system.

When the daily variations are examined, as shown for the most recent season in Figure 5, it is seen that good steam economy values are achieved only when the cane rate is above ~85% of the budgeted rate. This (as well as experience in Indian factories) indicates that any slowing of the factory crushing rate results in a significant jump in the process steam requirement. This is due to the fact that some process items, such as vacuum pans, have increased vapour consumption relative to their throughput during slowdowns or idle periods.
Modelling and simulation

The mass/energy balances for the factory were modelled through the changes. Estimates for the various configurations and operating conditions were compared and matched with the actual values, after allowance for estimated vapour vent losses.

A series of predictions was then run to examine various improvement strategies. The process steam pressure assumed in these predictions was 138 kPag (126°C), planned for in the future cogeneration boiler arrangements. The predictions showed that there is promise in the strategies of:

- Using a vapour line juice heater (VLJH) to use some final vapour (from the evaporator set or from the continuous vacuum pans) to preheat the coldest juice;
- Adding extra heating surface to the #3 stage in order to keep the #3 pressure up to around atmospheric pressure so that #3 vapour can be used on almost all of the vacuum pans;
- Eliminating unnecessary venting of vapours to atmosphere, or at least use these to preheat cold juice using a direct contactor (perhaps after and in series with the VLJH).

In an exploration of schemes, the most promising were those that had the characteristic that 83% of the vacuum pans would be run from #3 vapour, leaving some batch pans for seed preparation on #2 vapour.

The #3 vapour would be maintained at or above atmospheric pressure by installation of a new large #3 evaporator vessel, and by using appropriate pressure control.

A continuous pan (or modified batch pans) for the refined sugar strikes would operate on the #3 vapour supply. Under these conditions the juice concentration leaves the #3 stage at 62 brix.

The function of the #4 and #5 stages is to flash this hot juice down to a final vapour temperature of 60°C, and, in so doing, achieve a final concentration of 68 brix. Using a clarified juice Brix of 16, the theoretical steam % cane value was estimated at around 38.2 units.

A small additional loss has then to be made for vessel cleaning and venting losses. The estimated steam economy from this arrangement is 38.7% on cane. The arrangement is shown in Figure 6.
With such a high bleed load taken from the #3 stage, the vapour loading on the following stages is very low, and they act as flashing vessels rather than heat transfer vessels. The control of the concentration would be more complicated, and probably would require brix sensors on the #3 syrup as well as on the final evaporator syrup.

Disadvantages of the scheme would be an increase in the generation of colour in the evaporator set (estimated as an additional 3% over the present scheme) and an increased sucrose loss by inversion (estimated, for the syrup pH conditions used at ISA, as an increase of the loss from 0.17% to 0.38% of the sucrose in the cane).

These factors would have to be considered when considering the costs and benefits of the proposed schemes.

**Conclusions**

The ISA factory has made extraordinary progress in reducing the factory requirement of process steam in recent years. The recent values with steam consumption around 40% to 41% on cane must be considered excellent for a factory that produces roughly equal quantities of raw sugar, direct consumption sulfited sugar, and recrystallised refined sugar.

The major achievements in this project include the use of #3 stage evaporator vapour exclusively for all raw pan stage operations, new technology Roberts evaporators to maintain a low pressure drop, continuous vacuum pans for all three massecuites, and #2 stage vapour for annex refinery operations.

Very low process steam consumptions have been consistently recorded, as long as the cane crushing rate is maintained at near-target values.

Ideas for further improvements have been considered, and the factors limiting further economies have been taken into account.

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REFERENCES


ECONOMIE DE VAPEUR DE FABRICATION DANS UNE SUCRERIE
DE CANNE–UN PROJET DE COGENERATION

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Résumé
L’APPROCHE classique pour augmenter l’exportation de la vapeur produite dans une sucrerie est de maximiser la production de vapeur à haute pression et de minimiser la demande de vapeur à basse pression pour la fabrication. Cela permet d’optimiser l’utilisation de turbogénérateurs efficaces. Il faut donc revoir les techniques pour réduire la demande de vapeur à la fabrication. Dans ce papier on discute le travail fait à la sucrerie Ingenio San Antonio, au Nicaragua, pour passer d’une situation très pauvre en économie de vapeur, à une opération permettant une forte cogénération. Les points les plus performants comprennent l’utilisation de la vapeur d’évaporateur #3 pour les cuites, l’utilisation d’évaporateurs Roberts employant une nouvelle technologie, des cuites continues pour les trois massecuites, et l’utilisation de la vapeur #2 dans la raffinerie. Avec cette approche la demande de vapeur a toujours été basse. On donne aussi des idées et les limites pour des améliorations futures.

MEJORA EN LA ECONOMÍA DEL PROCESO DEL VAPOR EN LOS INGENIOS AZUCAREROS—UNA HISTORIA DE CASO DE UN PROYECTO EXISTOSO DE COGENERACIÓN

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PALABRAS CLAVE: Economía de Vapor, Proceso, Extracción de Vapor, Cogeneración.

Resumen
EL PROCEDIMIENTO clásico para incrementar la exportación de cogeneración de un ingenio azucarero es mediante la maximización de la producción de vapor de alta presión y la minimización de requerimientos de vapor de proceso de baja presión, con el fin de tomar ventaja de los eficientes sistemas de turbo alternadores de condensado total. Las configuraciones de fábrica que reducen los requerimientos de vapor de proceso necesitan examinarse de cerca. Este artículo revisa las configuraciones de fábrica que en años recientes han permitido reducir el vapor de proceso en el Ingenio San Antonio, en Nicaragua, iniciando desde una condición de baja economía de vapor y avanzando hacia una operación con alta economía de vapor, adecuada para una fábrica de cogeneración total. Los principales logros de este proyecto incluyen el uso exclusivo de vapor de etapa #3 del evaporador para todas las operaciones en crudo de la etapa de los tachos, usando los evaporadores Roberts de nueva tecnología y usando tachos continuos al vacío para las tres massecuadas y teniendo las operaciones anexas a la refinería en la etapa #2 del vapor. De esta manera se han conseguido consistentemente consumos muy bajos de vapor de proceso. Se tratan aquí ideas para mejoras ulteriores, y se delinean los factores que limitan dichas mejoras adicionales.