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Energy efficiency concepts in sugar house operation – what is old, what is new?

Konzepte für Energieeffizienz im Zuckerhausbetrieb – Was ist alt, was ist neu?

Energy efficiency is a key performance indicator for the efficient operation of a sugar refinery. Water intake via the fine liquor and the amount of wash water used in centrifugals are crucial indicators for steam demand. A range of concepts have for several decades been used in crystallization to automate the process, thus making it to a large extent reproducible. Depending on the crystal size aimed for, one- or two-stage seeding processes are often applied. The water intake by wash water applied in the centrifugals can be significantly reduced by applying syrup washing.

This paper describes the principles of the process steps mentioned and their importance for refinery and recovery in a sugar refinery. It also addresses the reasons why the use of these process steps is time and again called into question. On this basis, the paper presents opportunities that can result from a reassessment of precisely these process steps, taking into account state-of-the-art process monitoring methods.

Key words: crystallization, centrifugation, syrup washing, seeding systems

Die Energieeffizienz ist eine wesentliche Kenngröße für den wirtschaftlichen Betrieb einer Zuckerraffinerie. Entscheidende Größen für den Dampfbedarf sind der Wassereintrag über die Feinkläre und das eingesetzte Deckwasser in den Zentrifugen. In der Kristallgewinnung werden seit mehreren Jahrzehnten verschiedene Konzepte für eine automatisierte und damit in hohem Maße reproduzierbare Kristallisation eingesetzt. Je nach Anforderung an die Kristallgröße werden häufig ein- oder zweistufige Kristallfußprozesse genutzt. Der Wassereintrag durch die eingesetzte Deckwassermenge an den Zentrifugen kann durch die Verwendung einer vorgeschalteten Sirupdecke deutlich gesenkt werden.

Dieser Beitrag beschreibt die Prinzipien der genannten Prozessdetails und deren Bedeutung für die Prozessabschnitte „Refinery“ und „Recovery“ einer Zuckerraffinerie. Er geht auch auf die Gründe ein, warum der Einsatz dieser Prozessmodifikationen immer wieder in Frage gestellt wird. Hieraus werden die Chancen abgeleitet, die sich bei einer Neubewertung genau dieser Prozessschritte unter Berücksichtigung moderner Prozessüberwachung ergeben.

Schlagwörter: Kristallisation, Zentrifugation, Sirupdecke, Kristallfußprozess

1 Introduction

The core task in a sugar refinery is to produce white sugar of a defined quality from raw sugar of an equally defined quality. This is substantially achieved by dissolving the raw sugar and letting it crystallize again. The key requirements to be fulfilled by the process are achieving the specified quality, maximum yield, and minimum energy consumption. Different raw sugar qualities and/or varying qualities of the final product will, from time to time, call for an adjustment of the process parameters. With increasing automation of the process, there is, furthermore, the need to document all knowledge acquired adequately and to use it purposefully.

Thanks to the introduction of suitable process control systems, the automation of sugar processes made great progress in the 1980s. One benefit of automation was that it made processes reproducible – and it still does –, thus helping to advance process optimization along a defined path. This was useful for improving, for instance, the energy efficiency of the processes, significantly reducing energy consumption.

The sections below illustrate some important process steps for crystallization and the significance of their integration into an overall concept.

2 Achieving energy efficiency

The energy demand of a refinery is largely determined by the necessary water evaporation during crystallization, which results from the total amount of water added to the crystallization process:

- Water contained in the added fine liquor (depending on the dry substance content)
- Wash water and screen washing on the centrifugals
- Condensate for dissolving different sugars
- Dilution during crystallization (strike holding, rinsing, etc.)
- Hot water for cleaning, if it is recycled into the process

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The different effects of energy-saving measures can be illustrated using various balances. As a basis, the balance of a traditional refinery according to [1] is shown in Figure 1.

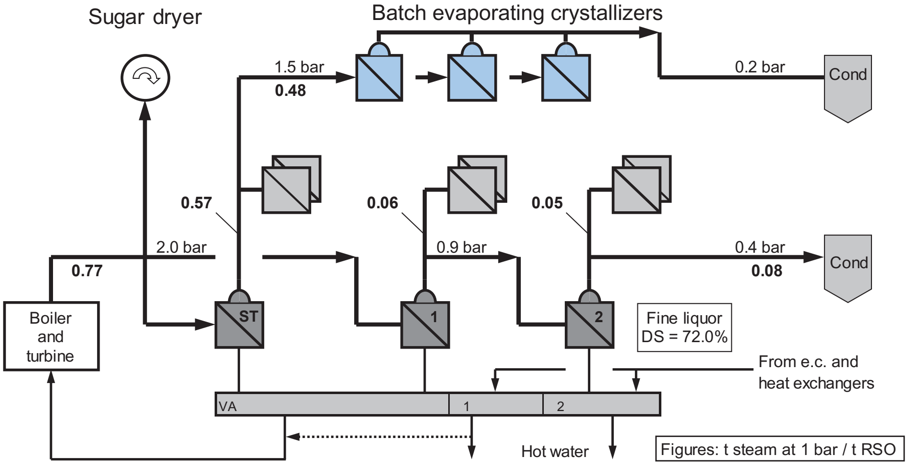

Fig. 1: Balance of a traditional refinery according to [1] (e.c. evaporating crystallizers)

A basic measure to keep the quantity of water added to a minimum is to maximize the concentration of the fine liquor from the evaporator station. Although this variant #1 considerably reduces the steam demand compared to the basic variant, from 0.77 t/t RSO to 0.72 t/t RSO (refined sugar output), a condenser loss remains, which even increases compared to the basic variant, from 0.08 t/t RSO to 0.13 t/t RSO (Fig. 2).

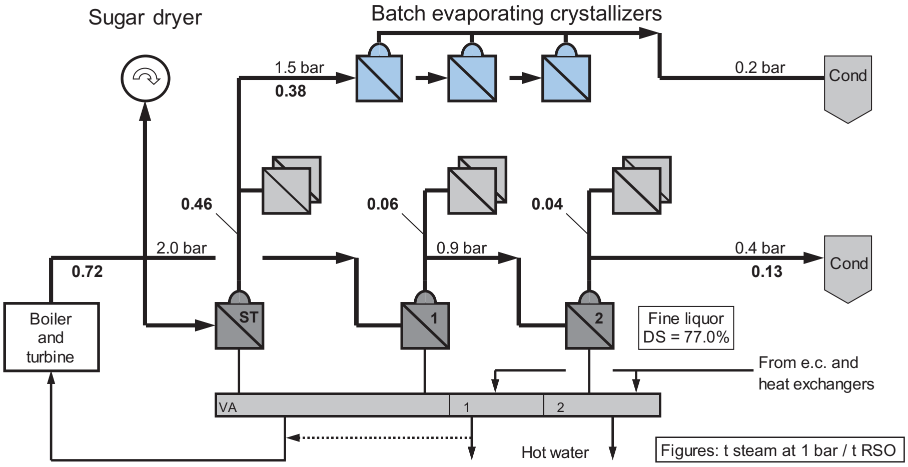

Fig. 2: Optimized variant #1 (concentration increase in the fine liquor from 72% to 77%) (e.c. evaporating crystallizers)

The increased concentration in the fine liquor requires additional measures to improve the crystallization processes; these are described in detail in the next chapter. On this basis, the centrifugals can then be optimized with regard to minimised wash water consumption and screen washing, which is also associated with a higher crystal yield from the centrifugals. As described below, these measures lead to a lower steam demand in refinery operation because less water is added to the centrifugals. This additional effect, which is shown in the optimized variant #2 in Figure 3, results in a steam demand of 0.69 t/t RSO.

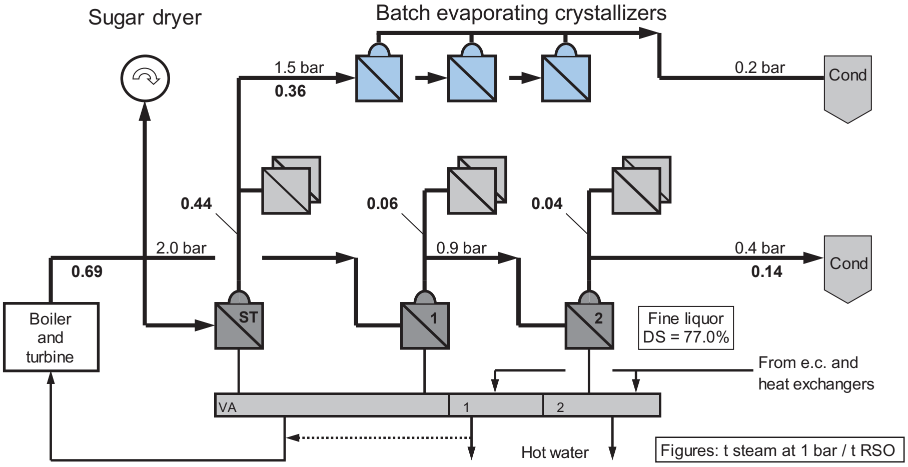

Fig. 3: Optimized variant #2 (reduced wash water consumption) (e.c. evaporating crystallizers)

Continuous crystallization in the R1 product enables the utilisation of vapors from the first evaporation effect as heating steam for crystallization. With a constant steam demand throughout crystallization (0.36 t/t RSO), the condenser loss and thus directly the steam demand can be reduced further to 0.61 t/t RSO (variant #3 – Fig. 4).

As less water is added during crystallization and the vapor utilisation for crystallization changes – as shown in variants #1 to #3 – considerably higher requirements have to be met by crystallization itself:

- An increase in fine liquor concentration induces (a) nearly complete saturation upon entry into the evaporating crystallizer; (b) in case of overheating, quick flash evaporation with supersaturation peaks (formation of fine crystals); (c) absence of the ‘washing effect’ due to the feeding of fine liquor for fine crystal dissolving.
- The wash water consumption can only be reduced, if (a) crystallization is fully controlled and reproducible; (b) there is only a minimum of aggregates in the crystallisate; (c) syrup washing is applied during centrifugation in case of magma with decreasing purity.

Both process design and process control can help with optimization of the crystallization processes. Process design can keep problem areas to a minimum or avoid them altogether, and fully automated process control can keep the ranges of fluctuation below critical limits. In this way, the requested product quality can be achieved even with higher process requirements.

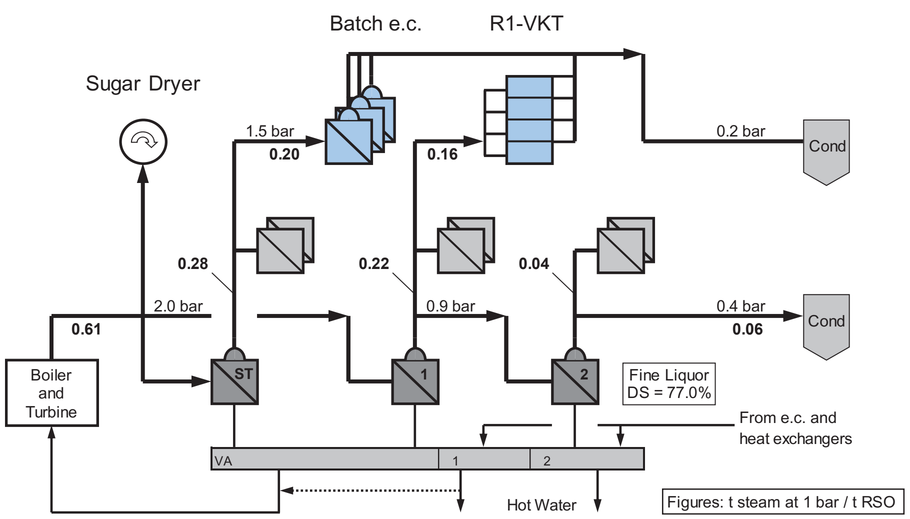

Fig. 4: Optimized variant #3 (continuous crystallization and reduction of the condenser loss) (e.c. evaporating crystallizers)

3 Implementation of seeding systems

The energy saving concepts require selection of a larger heating surface for the batch evaporating crystallizer, to keep the necessary heating steam pressure as low as possible. The disadvantage of this approach, however, is that the volume increases at the beginning of the crystallization process (criterion: heating chamber covered) relative to the final volume.

In the traditional one-stage process with slurry seeding, there is the basic dilemma that the heating surface is too large at the beginning of the crystallization process, and too small at the end. At the beginning, water evaporation is so high that the concentration increase in the mother liquor cannot be balanced by crystallization and more undesirable secondary nucleation (formation of fine crystals) takes place. At the end of the process, it is just the opposite: water evaporation is limited and, owing to the higher massecuite level in the batch evaporating crystallizer, can only be raised further by an increased heating steam pressure. The need for a larger heating surface as described above aggravates the situation.

As regards crystallization, this relationship can be expressed by the ratio of the crystal surface available for crystallization to the mass of crystallizable molecules [2, 3, 4]. For this ratio, the coefficient σ has been introduced. Figure 5 shows the relationship according to [2]. The course of the classical process (blue line) clearly shows the effect described above:

- At the beginning of the process after seeding, coefficient σ is very low (0.1 m^2/kg).
- At the end of the process, coefficient σ is very high (1000 m^2/kg).

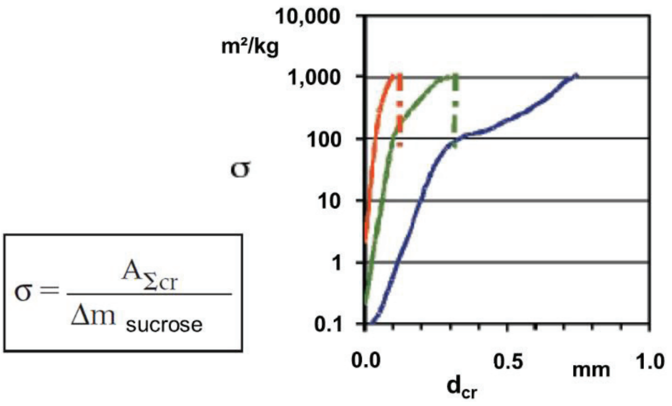

Fig. 5: Development of coefficient σ during evaporating crystallization [2]

Various seeding systems were therefore developed in the 1980s to shift the critical crystallization process after seeding with slurry to a separate small system [3]. The process most frequently used was the one developed in the so-called Sugar Institute at the Braunschweig University.

At the core of the process is the defined adjustment of a supersaturation of approx. 1.1 at the seeding point by concentrating the feed solution used and setting a controlled seeding temperature. After cooling crystallization to a final temperature of approx. 30 °C, a crystal content of about 20–25% is achieved, with a typical crystal size of approx. 0.1 mm.

At the beginning of the 1990s, BMA developed a universal variant of this process (Fig. 6), which was applied in different plant types, depending on the individual requirements. If the 1st seed produced by cooling crystallization is directly used as seed material in a batch evaporating crystallizer to produce the final product (one-stage seeding system), this can be recommended for crystal sizes up to approx. 0.5 mm. To obtain a larger product particle size, the cooling crystallisate is processed further in an intermediate stage by evaporation crystallization to produce 2nd seed (approx. 0.3–0.5 mm). This seed massecuite can then be used for seeding to produce the final product (two-stage seeding system).

The red line in Figure 5 shows coefficient σ for the process to produce 1st seed, and the green line the coefficient for 2nd seed. Practical experience would suggest that it is sufficient for coefficient σ to be in a range of 40–80 m^2/kg to largely avoid secondary nucleation.

Figure 6 shows the crystal size ranges for the one-stage and two-stage seeding system variants. For the threshold ranges, it is recommended to determine what process control would best suit the individual plant configuration.

4 Implementation of syrup washing

Syrup "washing" of the sugar in a centrifugal is a very old method. As far back as 1985, Reinefeld [4] pointed out that the method is not new as such, but had previously not managed to gain lasting acceptance – presumably because of the deficiencies connected with practical application. However, with the automation of crystallization and introduction of the seeding technique, syrup washing was more and more commonly used before water washing to reduce the amount of wash water to a minimum. In a first process step in the centrifugal, a "syrup" of adequate quality replaces the low-quality mother liquor in the massecuite (syrup washing). Afterwards, the remaining adhering good "syrup" is washed off using only

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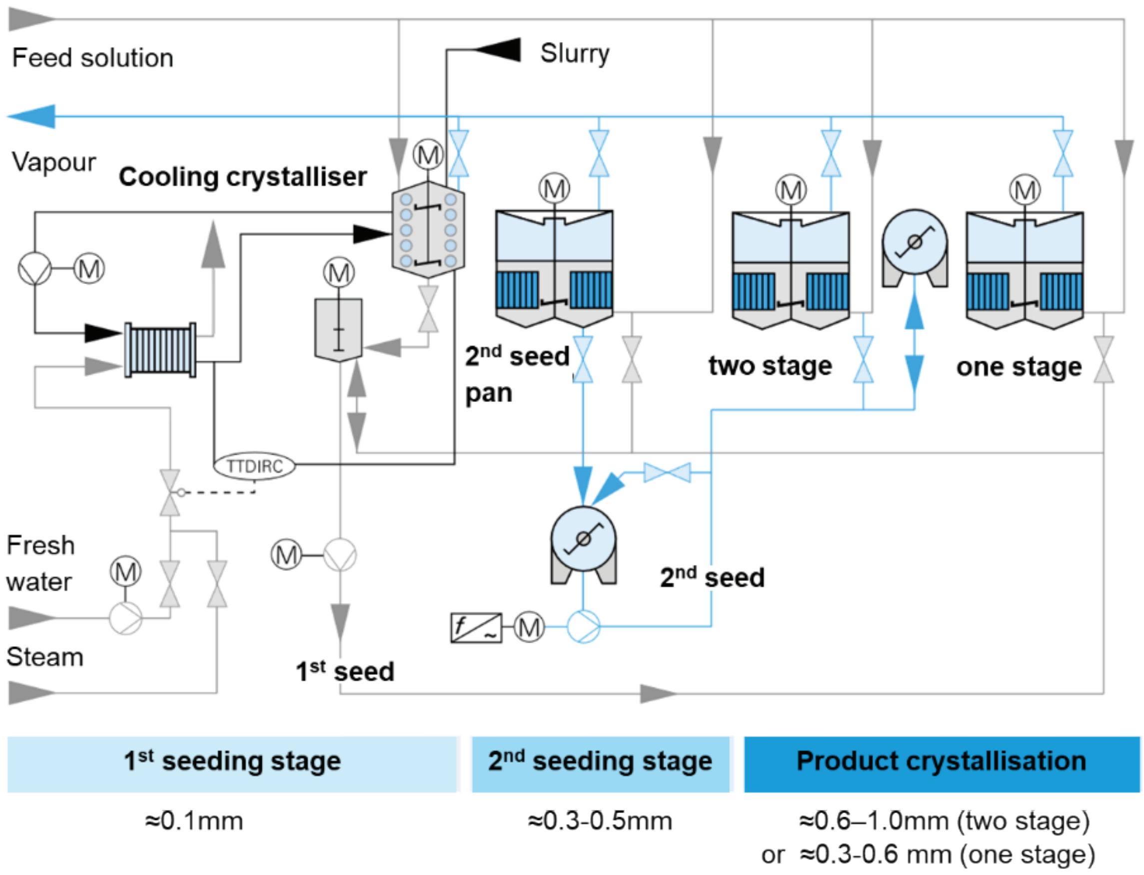

Fig. 6: Seed preparation as a one-stage or two-stage process

a small amount of water (water washing). Two media are generally used for syrup washing, which have different effects on the process.

Variant #A (Fig. 7) uses the wash syrup from the centrifugal as the medium. Bypassing the centrifugals, it is pumped in a cycle via a heat exchanger (with the target temperature adjusted). Finally, a pressure retention valve maintains a constant feed pressure at the centrifugals. The excess wash syrup is discharged into the green run-off. Recycling the wash syrup into the feed tank of the same crystallization stage has proved ineffective.

The point at which separation between green and wash syrup run-off occurs in the centrifugal has a major impact on the quality of the wash syrup: if it is switched too early, too much green run-off gets into the wash syrup, affecting the quality; if it is switched too late, there is not sufficient high-quality wash syrup for syrup washing.

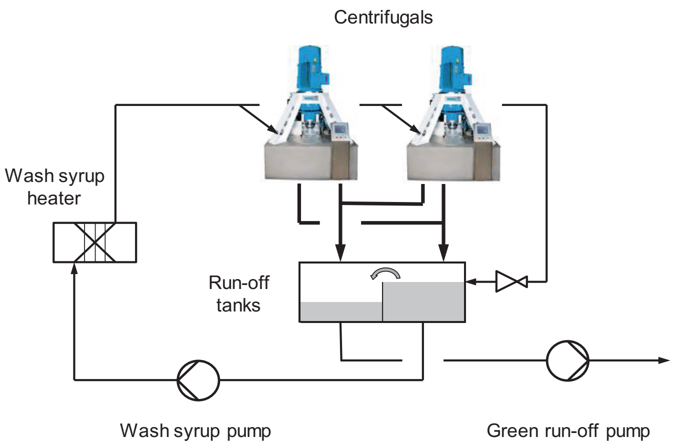

Fig. 7: Variant #A for the centrifugal process with syrup washing according to 4

Variant #B uses the feed solution of the crystallization stage as wash syrup, in beet sugar factories, also the thick juice directly from the evaporator station. Depending on the situation, a separate pump is needed to produce the necessary feed pressure and to ensure a constant pressure for a continuous flow. The benefit of this variant is the simpler process. Its disadvantages are usually a lower dry substance content and the lack of a separate temperature control, which can result in increased dissolving of sugar in the centrifugal. This variant therefore generally uses less wash syrup.

Where a specific washing effect is to be achieved, variant #A generally yields more wash syrup and has a lower dissolving effect than variant #B. On the other hand, variant #A needs more time for the task, which limits the possible acceleration of the centrifugal. Variant #B has the disadvantage that there is no crystallization yield from the amount of wash syrup bypassing the respective crystallisation step. With both variants, results can vary due to fluctuations in the concentration of the feed solution or temperature variations.

A further impact on the outcome when applying syrup washing comes from the separation process in the centrifugal. Figure 8 shows the typical time curve of the different phases of a centrifugal batch. Syrup washing starts after the green run-off has been separated off. This means that the pores between the crystals must be mostly open. After the end of syrup washing then, the wash syrup must be separated off before water washing can start. If syrup washing or water washing are started too early, there is the risk that the dammed liquid may cause vibrations. However, if water washing is started too late for safety reasons, the sugar layer will have become so dense, as

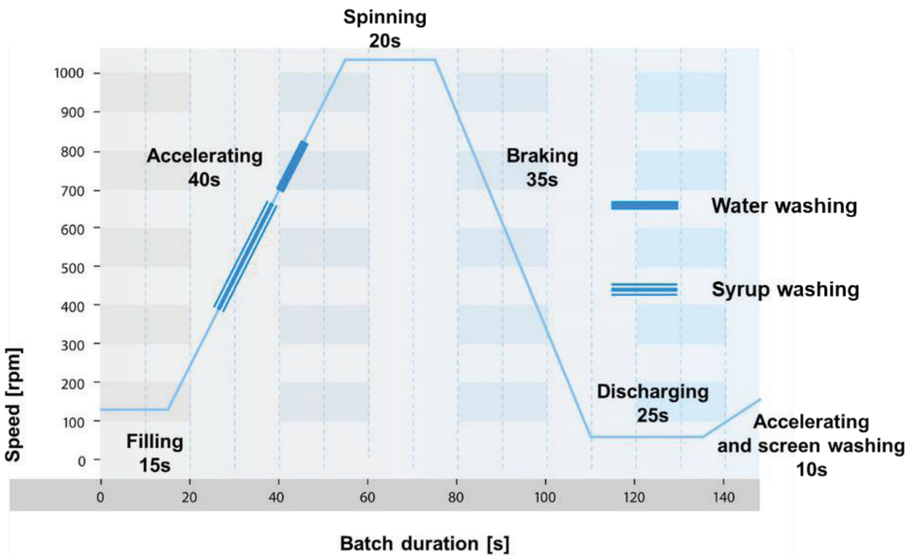

Fig. 8: Typical centrifugal cycle

the centrifugal force increases along with acceleration, that a sufficient washing effect can no longer be achieved.

The time curve in Figure 8 shows clearly that an optimal application of syrup washing is only possible if the periods for syrup washing and water washing are well-timed and consistent. The necessary conditions for a consistent crystal quality must be established earlier, at the crystallization stage.

5 New control cycles

The increasing requirements to minimise energy demand represent new challenges, also for process automation. Processes must be controlled with increasing precision and the lowest feasible ranges of variation to meet the requirements for optimal process parameters and the requested product quality. It will no longer be sufficient to have optimal process control for single equipment items and/or process stations. Instead, several stations or even whole factories will have to be controlled intelligently and with as much foresight as possible. This can be illustrated by two examples:

- Raw sugar intake: The quantity and quality of raw sugar fed into the process should be as consistent as possible. Any change in these parameters should be slow and controlled. The process parameters in downstream stations can then be anticipated and adapted based on proven knowledge.
- Batch crystallization cycle: The individual cycles of the batch evaporating crystallizers should be coordinated in such a way that the average heating steam consumption of all crystallizers is as constant as possible. This requires an intelligent control of the filling levels in the different buffers like mixers and tanks, and certainly at least a rough estimate of the volumes of “floating products” in each equipment item. A higher-level control system for the centrifugals must be integrated to avoid disturbance variables.

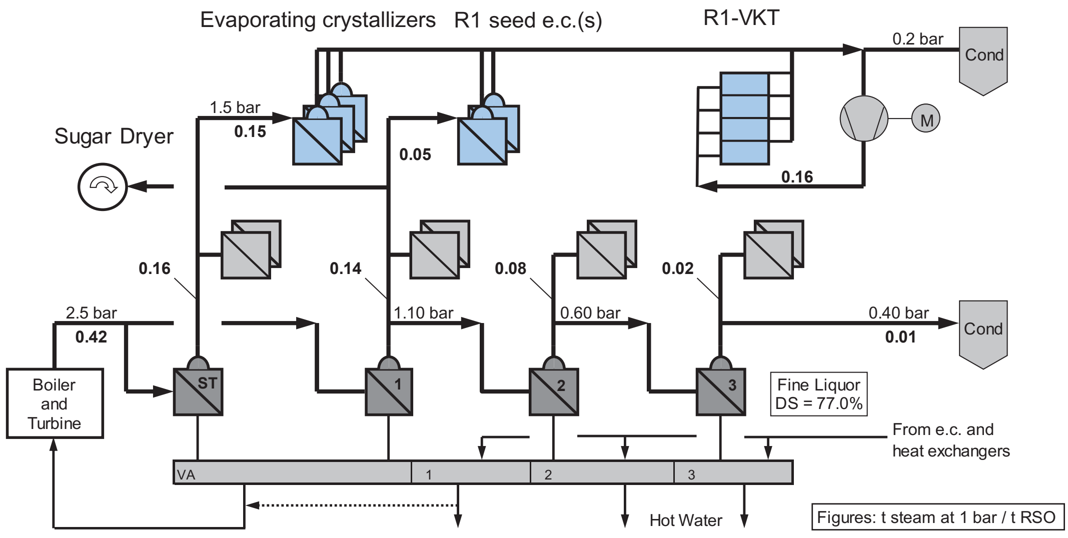

Fig. 9: Mechanical vapor recompression for R1-VKT (variant #4) (e.c. evaporating crystallizers)

Higher process control requirements also affect equipment design and must therefore be taken into account. Considering the maximum capacity of each item in isolation leads to a capacity deficit in optimized operation, unless reserves are planned elsewhere.

6 Prospects for the future

In future, the challenge will be to decarbonise the refinery process as much as possible. Some technical solutions can already be considered the state of the art, even though they have so far only been applied in individual cases [1, 5]. With the use of mechanical vapor recompression (MVR), the exhaust steam produced in the boiler from fossil fuels can be replaced with green electric energy.

Based on variant #4 shown below, the steam demand can be significantly reduced by applying mechanical vapor recompression in conjunction with the R1-VKT (Fig. 9).

A three-effect evaporator station is included to avoid considerably higher condenser losses. The associated increased pres-

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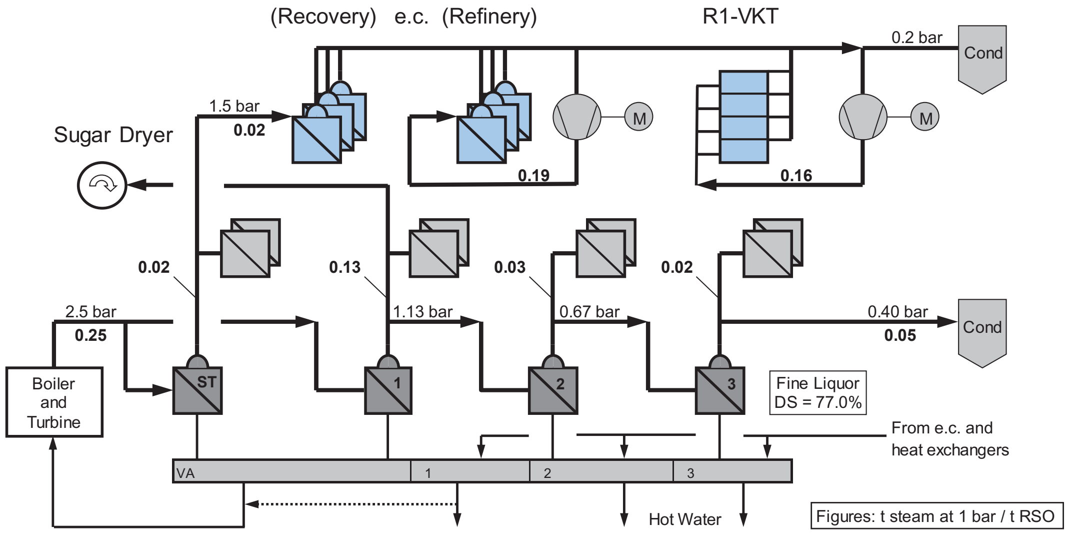

Fig. 10: Mechanical vapor recompression for all refinery crystallizers (variant #5) (e.c. evaporating crystallizers)

sure in the first effect also permits the use of 1st vapors for R1 seed massecuite production as well as for nearly all heat exchangers.

Steam demand can be further reduced if all evaporating crystallizers of the refinery are integrated into mechanical vapor recompression (Fig. 10). At 0.25 t/t RSO, it will then amount to only 1/3 of the steam demand of the basic variant in Figure 1.

An important parameter for the energy demand of the mechanical vapor compressors is the pressure ratio of delivery pressure to suction pressure. This is clearly lower for the R1-VKT ($\pi \approx 3$–3.5) than for the batch crystallizers in the refinery ($\pi \approx 5$–6).

In a refinery with a capacity of 3000 t/d, the vapor compressor requires an electric capacity of approx. 1.3 MW for the R1-VKT, and even approx. 3.1 MW for the batch crystallizers, with a similar vapor quantity. The key factor for the electric energy demand of crystallization is that the batch crystallizers require about twice as much electric energy as continuous evaporating crystallizers such as the VKT.

7 Conclusions

The variants shown above describe different theoretical concepts for process optimization that would result in potential savings. Essential prerequisites are a constantly well operated crystallization process and a higher-level automation concept, to compensate for disturbance variables already right from the start. The individual measures represent the actual state of the art. The challenge for the future will be consistent and complete application to always achieve the best possible result in daily operation.

The examples shown can be used to derive further ways of reducing energy demand, especially with regard to the design of the evaporating crystallizers. However, while no "green" fuel is available as an alternative, mechanical vapor recompression clearly offers the greatest potential for decarbonising refinery operation.

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