Enabling Improved Waste Heat Recovery in Digestion for Energy and Carbon Footprint Reduction

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Abstract

As large industrial process heat users, the CO_2 emissions of alumina refineries are considered “hard to abate”. The digestion area, as the largest single energy consumer in the form of process steam, is critical to both minimising overall refinery energy consumption and delivering net zero carbon in the most economical way. In future, the traditional goal of maximising heat recovery from digestion will be supplemented with a similarly important goal of recovering waste heat from other refinery sources to digestion via mechanical vapour recompression (MVR) as a direct substitute for fossil-fuel based steam. Reviewing the digestion energy balance, single and dual stream heating technologies and incorporating refinery waste heat recovery, this paper explores digestion configurations that lower refinery energy consumption and carbon footprint. It is shown that incorporating waste heat recovery can reduce digestion energy input by more than half with single stream significantly outperforming dual stream. Increasing the number of live steam heating stages further improves this energy benefit.

Keywords: Digestion, Energy, Waste heat, Mechanical vapour recompression (MVR), CO_2 emissions.

1. Introduction

Recent years have seen a global acceleration in the energy transition and the pursuit of efforts to limit climate change temperature increase to 1.5 °C as called for in the 2015 Paris Agreement. There has been widespread adoption amongst nations and corporations of “net zero” CO_2 emissions targets by 2050. The International Aluminium Institute has published its own pathway for the industry in-line with 1.5 degrees of global warming [1]. Like the International Energy Agency’s Net Zero by 2050 roadmap [2], this pathway calls for a pronounced drop in emissions in the decade post 2030. Multiple different decarbonisation initiatives are already underway across the aluminium value chain [3] with 2022 marking the first year total greenhouse gas emissions from the aluminium sector did not grow despite aluminium production growth [4].

Reported uptake of renewable electricity projects as well as future energy use scenarios [5] indicate that there will be an unprecedented change in energy sources globally. Future projections suggest that clean electricity will become the new linchpin of the global energy system. Such a shift is also likely to result in electricity becoming the dominant energy source for alumina refining.

With a vast amount of new energy infrastructure required, the scale of the net zero challenge is huge. Rolling this out at the scale and speed required to meet climate change targets will require a paradigm change to traditional methods of infrastructure delivery [6]. Typically, today’s prices for electricity are substantially higher than those for energy in the form of fossil fuel derived heat. Alumina refineries are very large users of process heat. Thus, despite year-on-year falls in costs as the number of solar photovoltaic (PV) and wind projects accelerates, the alumina industry can anticipate rising energy bills as its energy mix is decarbonised.

To offset energy related alumina production cost increases and limit new energy infrastructure requirements, future step changes in alumina refinery energy efficiency seem likely. As the largest refinery energy consumer, the digestion area will need to be adapted to realise these energy and emission consumption benefits.

2. The Role of Digestion in Bayer Process Energy Consumption

Bayer process alumina refining energy consumption consists of energy for steam generation, energy for alumina calcination and electricity for rotating drives and machinery, with an approximate split of 55 %, 35 %, 10 % respectively (see Figure 1), depending on individual refinery efficiency and equipment technology. The digestion area is the dominant user, typically 75 % or more of the steam energy consumption, with spent liquor evaporation the next largest user. Heating in digestion is needed to meet the desired temperature for dissolution of alumina from the bauxite ore into caustic liquor (145 to 150 °C for “low temperature” gibbsite digestion or 250 to 280 °C for “high temperature” boehmite or diaspore digestion).

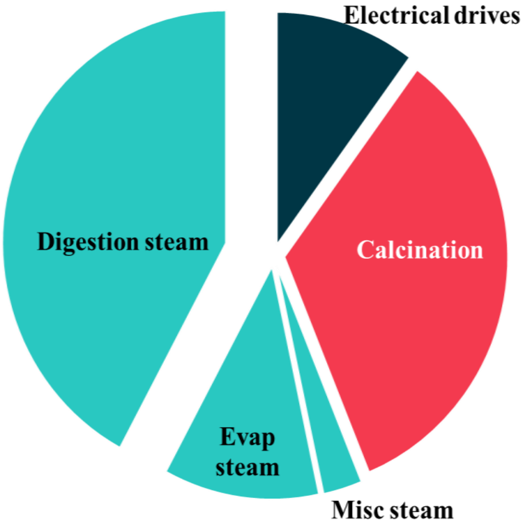

Figure 1. Energy split in the Bayer process by consumer.

Figure 2 provides another view of the contributors to overall refinery energy consumption. Digestion energy input must ultimately compensate for sensible heating in digestion to overcome the digester approach temperature, irrecoverable low-grade heat as well as discharge stream sensible heat losses [7].

Digestion steam consumption is influenced by a multitude of factors such as liquor yield, bauxite quality etc. but within the digestion area this is practically governed by the digestion heat recovery, i.e., the temperature approach between the last regenerative heating stage and the target digestion temperature. It is important therefore to adequately recover the available heat in digester product discharge to the digester feed with installation of sufficient heat exchange equipment and area [8]. In multi-flash systems this energy recovery is greatly influenced by liquor boiling point elevation and the approach temperature between the condensing vapour and slurry or liquor discharge temperature from each heating stage, with ever increasing amounts of area being required to reduce the temperature approach [9].

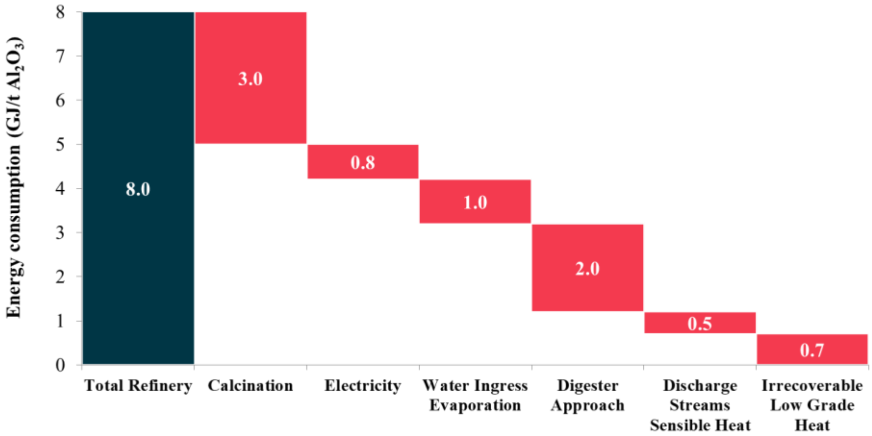

Figure 2. Typical energy split in the Bayer process by consumer.

3. Single versus Dual Stream Digestion

Digestion heat transfer technology has evolved over time. From early batch operation with no heat recovery to continuous multi-flash systems with full indirect heating of digestion slurry. Several digestion flowsheets and equipment technologies are in use today, the most widespread being “single stream” and “dual stream”.

As its name suggests, “single stream” heating refers to the feeding of a combined stream of evaporated spent liquor (test tank liquor) and bauxite slurry fed through flash and live steam heaters (LSH) and into the digester holding vessels. In low temperature digestion circuits, it is carried out using standard shell and tube heat exchangers orientated vertically. Apart from energy considerations, discussed further in this paper, single stream lessens heater silica scaling and free caustic corrosion impacts seen with dual stream heating. Heater tube plugging and / or tube erosion with slurry is a concern to be managed by careful design and operation. Figure 3 illustrates one example configuration employing three flash heat recovery stages with the final stage also acting as digester blow-off. Regenerative flash vapour is exported from the second and third flash stages to vertical shell and tube indirect bauxite slurry heaters and a second washer overflow heater. Other configurations could employ four flash heat recovery stages with less or no export flash steam. The specific configuration is usually selected on a case-by-case basis for each refinery’s circumstances.

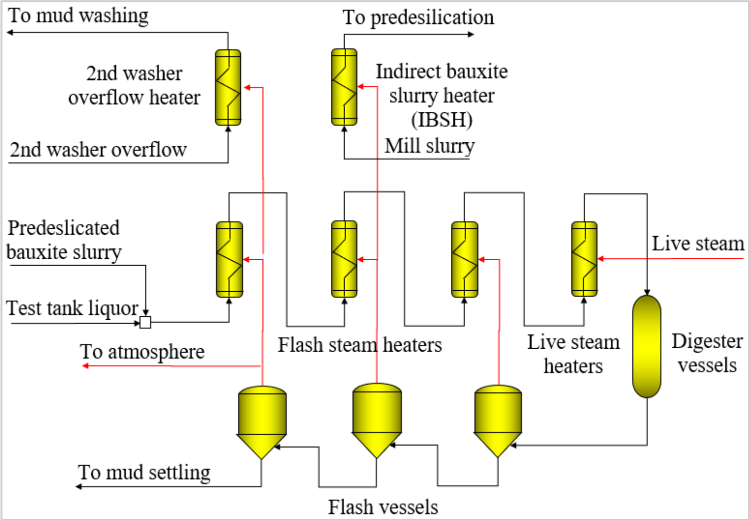

Figure 3. Example single stream low temperature digestion flowsheet.

"Dual stream" heating refers to the separate feeding of test tank liquor and bauxite slurry into the digester holding vessels. Only the liquor stream is fed through the flash and live steam shell and tube heat exchangers which can be orientated horizontally or vertically. With no slurry that could plug tubes the heater tube velocity can be set lower than in single stream. Figure 4 shows a dual stream version of the single stream flowsheet given in Figure 3. Regenerative flash vapour destinations are the same although, in the absence of indirect slurry heat exchangers, bauxite slurry heating occurs via direct injection contact heating.

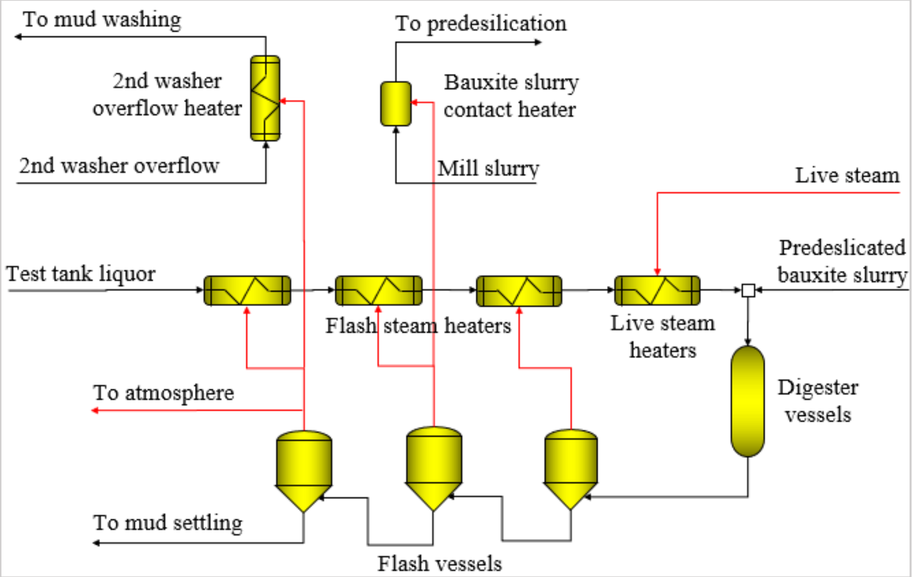

Figure 4. Example dual stream low temperature digestion flowsheet.

One significant difference between single and dual stream flowsheets is the required live steam pressure which is approximately double in the case of dual stream. Based on typical installed heater areas and allowing for adequate heater approach temperature, the minimum required steam pressure for single stream is 6 to 6.5 bar(g). For dual stream the equivalent value is 12 to 12.5 bar(g), although 15 bar(g) pressure was required in this study because of the bauxite charge ratio and high yield. In single stream, slurry discharges from the live steam indirect heaters at ~145 °C, whereas in dual stream liquor must discharge at ~180 °C to achieve the same target 145 °C digester temperature after bauxite slurry is separately added downstream. There is an additional impact of the higher live steam pressure for dual stream where steam supply is via a combined heat and power (cogeneration) system. In this instance the higher pressure reduces the power generation opportunity and cogeneration efficiency.

3.1 Heat and Mass Balance Modelling

Heat and mass balance modelling using SysCAD software was carried out to explore the overall energy efficiency of the single stream and dual stream digestion flowsheets shown in Figures 3 and 4. Further scenarios were then examined to evaluate the energy efficiency impacts of importing recompressed vapour arising from low-grade waste heat streams currently lost to atmosphere into digestion, as described in section 4.1 below.

3.1.1 Modelling Basis and Assumptions

The modelling used a base case single stream digestion flowsheet designed to produce 1.4 Mtpa of alumina in a single digestion unit, i.e., single flash tank train, with a refinery yield of approximately 90 g/L (i.e., 90 grams of alumina produced per litre of pregnant liquor fed to the precipitation area).

For simplicity modelling was restricted to the digestion area but inclusive of flash tank vapour export duties (to bauxite slurry and second washer overflow heating in this instance). The same vapour export duties and heat transfer areas were retained for each scenario. Any unrecovered vapour from the flash train was reflected as blow-off vapour to atmosphere.

For each scenario the mass flows and temperatures of predesilicated bauxite slurry and test tank liquor to digestion were kept constant.

145 °C digester temperature was targeted in all scenarios. In dual stream the model produces an elevated liquor temperature from the live steam heaters that then delivers a 145 °C digester temperature once the bauxite slurry is mixed further downstream. For dual stream, all gibbsite dissolution, which is endothermic, was modelled to occur in the digester vessels; for single stream a portion of the gibbsite dissolution was assumed to occur during flash and live steam heating.

Different flash heater area (A) and overall heat transfer coefficients (HTC) were applied for single and dual stream. Single stream HTC values are calibrated from an operating refinery using a four month hydroblast cleaning cycle. Dual stream HTC is an average from Worley's internal database of operating refineries with the higher HTC reflecting much shorter acid cleaning cycles despite lower tube velocities. Single stream flash heat transfer area basis is a real installed project designed by Worley using the Figure 3 flowsheet. Dual stream area was adjusted in the SysCAD model runs to achieve the same blow-off vapour exhaust tonnage as single stream.

3.1.2 Modelling Results

SysCAD model results are provided in Table 1. Some key input values are provided in the shaded cells.

For dual stream, the digestion energy consumption shown excludes any additional consumption resulting from direct injection heating of bauxite slurry. Direct injection dilutes process liquor requiring additional spent liquor evaporation. In this example, 39 t/h of directly injected flash vapour is required. At an assumed evaporator steam economy of 4.0, this equates to nearly 10 t/h of additional evaporation steam or 5.5 MW thermal energy input.

Table 1. Digestion energy efficiency - single versus dual stream heating.

Scenario	Single stream	Dual stream
Relative flash heater area (A)	1	1.390
Relative heat transfer coefficient (U)	1	1.105
Relative U.A value	1	1.546
Live steam heater discharge temperature (˚C)	146.1	182.1
Blow-off vapour (t/h)	5.2	5.3
Digestion steam energy consumption (MW thermal)	85.4	85.5

4. Mechanical Vapour Recompression

Mechanical Vapour Recompression (MVR) is a technology that recovers waste heat otherwise lost through low-grade vapour streams to the atmosphere or cooling water systems. MVR can also significantly reduce water consumption by recycling the water currently lost in this way. MVR is already widely used in a variety of other industries such as water treatment, dairy, chemical and pulp and paper.

MVR is a very energy efficient technology, offering a significantly higher Coefficient of Performance (COP) versus other electrification technologies such as electric boilers. COP refers to the ratio of steam energy output to electrical power input. With MVR, a modern greenfield alumina refinery recovering waste heat from several sources across redside and whiteside could reduce the process steam energy consumption by ~ 60 % [10]. MVR efficiency (COP) depends on the vapour compression ratio. Vapour compression is most applicable to low temperature digestion refineries and using carbon-free electricity has the potential to eliminate all CO_2 emissions from refinery steam generation. The COP for high temperature digestion is significantly reduced compared to low temperature, eroding its economic advantages as its steam pressure requirement of ~ 50 to 110 bar(g) vastly exceeds the ~ 6 to 15 bar(g) needs of low temperature digestion.

MVR has advantages over other electrification-based decarbonisation pathways with much lower operating cost due to its energy efficiency, as well as lower capital cost for electrical infrastructure due to the lower electricity demand. Therefore, MVR is likely to play an important role in future refinery decarbonisation, although this may be restricted to refineries that operate low temperature digestion circuits.

4.1 Waste Heat Recovery into Digestion

Worley estimates that all normal digestion steam demand could be met using Bayer process waste heat coupled to MVR. Recovering this waste heat as useful steam would reduce digestion energy consumption because MVR systems always have COPs higher than 1. Recycling waste heat vapour does not change digestion steam demand, but the thermal energy input of current steam generation technology is replaced with a lesser amount of compressor electrical energy input.

MVR system COP is related to the Carnot cycle. The Carnot cycle is an ideal thermodynamic cycle that represents the theoretical maximum efficiency that can be achieved by a heat engine during the conversion of heat into work, or conversely work into heat, when operating between two temperature reservoirs. Lower temperature differences between these reservoirs result in higher COP values, per equation (1). In practice the maximum efficiency is not achieved as the Carnot cycle assumes no friction, no heat losses, reversible processes, isothermal compression, and a variety of other inefficiencies exist such as limits on the rate of heat transfer between the hot and cold reservoirs. For this work, calculated COPs were downrated to 65 % of the theoretical maximum to approximate real-world industrial performance, whilst the temperature reservoirs were the low-grade waste heat source (cold) and the final compressed steam (hot).

$$COP = \frac{T_h}{T_h - T_c} \tag{1}$$

where:

COP Coefficient of Performance
T_h Hot reservoir temperature, K
T_c Cold reservoir temperature, K

Various scenarios were modelled where recompressed waste heat vapour entirely displaces boiler generated steam for digestion live steam heating. Both single and dual stream heating flowsheets were analysed. These included introducing a modification to live steam heating to further reduce energy consumption. This modification involves splitting live steam heating into two or more stages, each stage operating with recompressed vapour at a different pressure. Steam pressure (and temperature) is lowest in the first live steam heating stage and increases from stage to stage. Compared to operating with only one stage of heating this allows a higher COP on the portion of steam to the first stage heaters with a higher overall average COP achieved. An example with two live steam heating stages is given in Figure 5. Here, 50 % of the total steam mass flow is sent to the first stage using recompressed vapour (steam) at 3.5 bar(g) pressure, with the remaining 50 % performed in the second stage at 6.5 bar(g) steam pressure. This compares with a single stage of live steam heating (Figure 3) where all heating is performed using 6.5 bar(g) steam. The MVR waste heat recovery scenarios evaluated were:

- Single stream flowsheet: 1 LSH stage fed with MVR steam at 6.5 bar(g).
- Single stream flowsheet: 2 LSH stages each fed with MVR steam at different pressures of 3.5 bar(g) and 6.5 bar(g). Refer Figure 5.
- Dual stream flowsheet: 2 LSH stages each fed with MVR steam at different pressures of 7.5 bar(g) and 15 bar(g).
- Single and dual stream flowsheets: 2 LSH stages each fed with MVR steam at different pressures per above plus available blow-off vapour recompressed to the 1st stage. Figure 6 shows this configuration for dual stream.

For calculation of MVR system COP, a cold reservoir temperature of 41 °C was used for all scenarios except 80 °C when recompressing blow-off vapour. The latter assumes blow-off vapour is used to indirectly raise clean steam as a feed into MVR fans and results in a more conservative COP versus compressing the 100 °C saturated blow-off vapor directly.

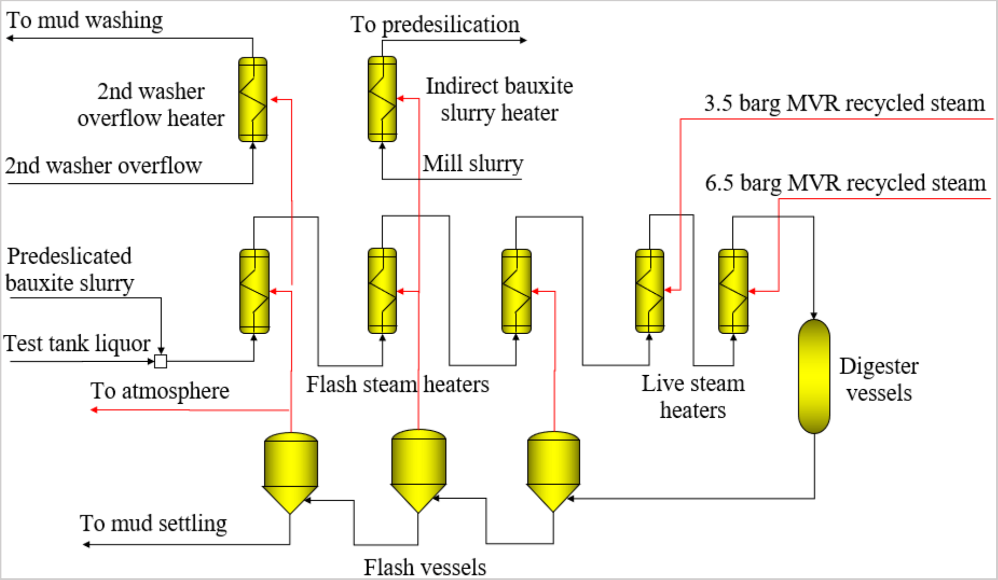

Figure 5. Single stream with waste heat recovery to split LSH stages.

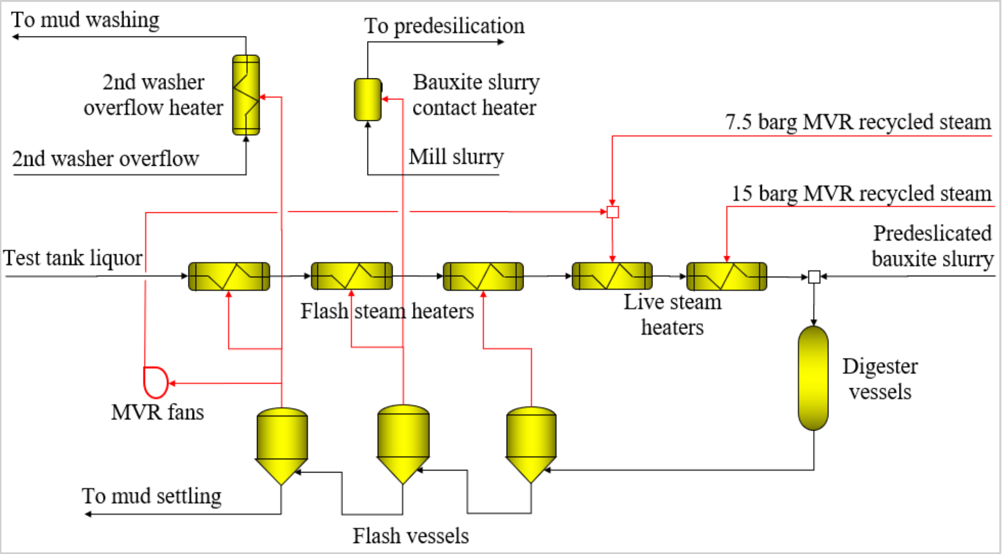

Figure 6. Dual stream with blow-off waste heat recovery and split LSH stages.

4.1.1 Modelling Results

SysCAD model results are provided in Table 2. Again, for clarity, some key input values are provided in the shaded cells.

Table 2. Digestion energy efficiency with recompressed waste heat vapour as steam source.

Scenario	**Single stream**			**Dual stream**	
	One LSH stage	**Two LSH stages**	**Two LSH stages + blow-off recovery**	**Two LSH stages**	**Two LSH stages + blow-off recovery**
Relative flash heater area (A)	1	1	1	1.390	1.390
Relative heat transfer coefficient (U)	1	1	1	1.105	1.105
Relative U.A value	1	1	1	1.546	1.546
Last LSH exit temperature (°C)	146.1	146.1	146.1	182.1	182.1
Blow-off vapour (t/h)	5.2	5.3	5.3	5.3	5.3
1^{st} LSH steam energy (MW thermal)	85.4	43.3	43.3	43.9	43.9
2^{nd} LSH steam energy (MW thermal)	N/A	42.1	41.6	41.0	41.0
Compressor COP for steam to 1^{st} LSH	2.26	2.56	2.56	2.20	2.20
Compressor COP for steam to 1^{st} LSH (source blow-off vapour)	N/A	N/A	4.03	N/A	3.12
Compressor COP for steam to 2^{nd} LSH	N/A	2.26	2.26	1.92	1.92
Weighted average compressor COP	2.26	2.40	2.42	2.06	2.07
Total compressor electrical energy (MW electrical)	37.8	35.3	35.1	41.3	41.1

5. Discussion

Table 1 illustrates that for a similar digestion thermal energy consumption and blow-off vapour exhaust to atmosphere dual stream requires approximately 39 % more installed flash heater area than single stream. Single stream's better heat recovery arises because the flowsheet is more thermally balanced with similar mass flows up and down the heater and flash trains. This advantage can further increase if considering the additional evaporation needed in dual stream to compensate for direct injection heating of bauxite slurry and the lower cogeneration plant efficiency arising from dual stream's higher live steam exhaust pressure.

Table 2 shows that displacing digestion live steam with recompressed low grade waste heat vapour dramatically lowers overall energy consumption. For single stream with one stage of live steam heating, the electrical energy input to the MVR system compressors is only 44 % (the reciprocal of the COP) of the thermal input energy from boiler steam. This electrical energy can itself be lowered by ~ 7 %, and overall system COP improved, if two stages of live steam heating are used each with its own dedicated steam pressure. Given that energy prices are projected to play a larger role in future alumina production costs this could be a valuable improvement.

For dual stream there are similar significant energy consumption benefits of recovering low-grade heat into digestion in lieu of boiler steam, although the numbers are not quite as attractive. Table 2 shows that with two live steam heaters the MVR system electrical energy is ~ 17 % higher than the equivalent single-stream scenario, driven by the higher required steam pressures.

Due to the efficient design, blow-off vapour quantities were minor in the scenarios studied and therefore recovery of this stream had only a minor impact on energy consumption. However, much higher COPs can be achieved when recovering this stream compared to other refinery low grade heat sources because of the hotter blow-off vapour temperature. Therefore, existing refineries with large amounts of blow-off vapour could find blow-off MVR projects as attractive energy efficiency and decarbonisation pathways.

6. Conclusions

Electrification will be the key global pathway to meet net zero climate change targets. This is likely to extend to the alumina industry. There are several electrification technologies but refinery waste heat recovery via mechanical vapour recompression will play an important role due to its efficiency advantages, especially for low temperature digestion refineries. Such waste heat recovery can fully electrify digestion steam production and lower its energy input by more than half.

Of existing digestion heating technologies, single stream requires significantly less flash heater area versus dual stream for similar digestion energy consumption and blow-off vapour loss. Single stream is expected to be the preferred technology in the future MVR world where compressed waste heat vapour is imported into digestion with its much lower steam pressure requirement driving COP efficiency gains. This will be important given that energy prices are projected to play a larger role in future alumina production costs. COP can be further improved by splitting the indirect live steam heating system into two or more heating stages with different steam pressures set for each stage.

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