

HEAT TRANSFER IN THE BAYER PROCESS

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Abstract

Heat transfer equipment represents a significant portion of Bayer process plant capital and operating costs. Heater operation and maintenance activities can also create potential hazard exposure. Very early flowsheets tended to rely on direct heat transfer, i.e. steam injection heating and flash cooling, and this still persists to some extent today. There has however been an ever increasing utilization of indirect heat exchange over the past 100 years. This has been driven by higher energy efficiency targets and enabled by improvements in heat transfer equipment. In more recent decades there has been a partial shift towards slurry heating and cooling instead of liquor heating and cooling. This paper presents an historical perspective, explores some heater selection scenarios, and looks at future challenges and opportunities.

Introduction

The Bayer process, if treating Gibbsite bauxites, basically converts Gibbsite to Gibbsite. Chemical thermodynamics would therefore suggest zero energy requirement. However, as aptly pointed out by Adamson [1]

“At the extraction end of the cycle we require a high temperature... to dissolve the alumina and at the other end we require a low temperature... to obtain as big a deposition of hydrate as is economically practicable.”

As a result the typical Bayer cycle operates as a thermodynamic “loop” between about 60°C (Precipitation) and about 150°C (for gibbsite digestion) or 240-280°C (for boehmite or diaspore digestion). The energy recovered at 60° during crystallization is around 0.7 kJ per tonne of Alumina [2,3,4] but is generally at too low a temperature to be effectively re-used. So to all intents and purposes the theoretical minimum energy for the Bayer process is about 0.7 GJ per tonne of alumina (plus calcination).

The reality is however that refineries typically consume at least 4 GJ (thermal) per tonne of alumina (excluding calcination) and many run in excess of 10 GJ/t [5]. Energy consumption is typically between 20 and 40% of total alumina production costs.

A significant part of the gap between theoretical and actual consumption is related to the limitations of heat transfer equipment:

- The approach temperature between the slurry entering digestion, and the required digestion temperature, is typically between 30 and 50°C. This results in a sensible heat requirement of between 1.5 and 3 GJ/t.
- Water is added to the process in various locations, mainly for washing of solids streams but also for process/mechanical requirements such as instrument purge and pump seal water. This creates a requirement for evaporation within the Bayer circuit to offset the

water addition. Energy consumption for evaporation typically contributes 0.5 to 1.5 GJ/t.

- Often the lack of suitable heat transfer equipment dictates that direct steam injection be used; this further contributes to evaporation requirement and hence energy consumption.
- The equipment cost to recover the sensible heat from streams exiting the refinery (e.g. bauxite residue and calcined alumina) means that these streams typically exit the refinery at about 80°C, which is a thermal loss of around 0.5 to 1.0 GJ/t

Heat transfer equipment represents a substantial portion of a refinery’s capital and maintenance costs. Figure 1 illustrates just how much of a typical refinery is tied up in heat transfer equipment.

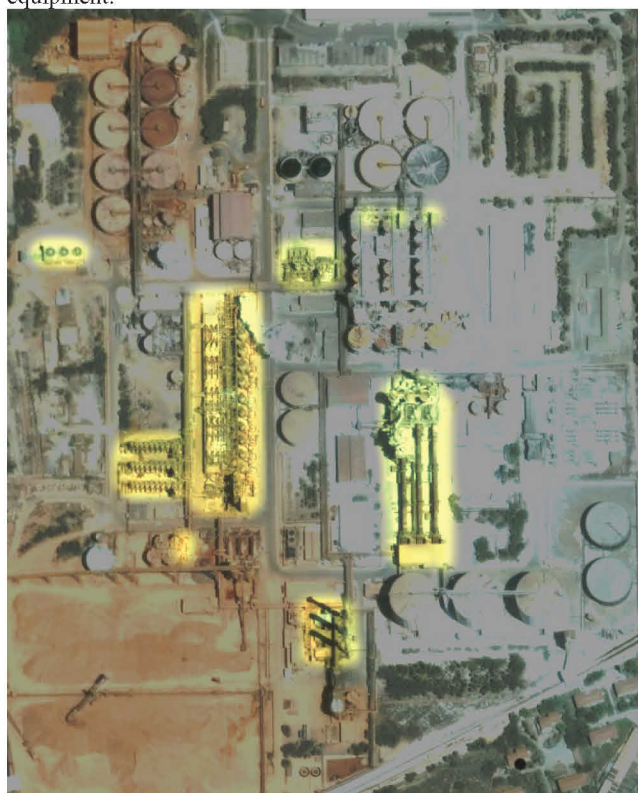


Figure 1 –A typical refinery with heat transfer equipment highlighted.

Heat transfer is governed by the same principles in the Bayer process as it is in other process industries, and it is not the intention of this paper to describe these principles here. However, the elements which distinguish Bayer Process heat transfer from other industries include:

- Process fluid propensity to form scales

- Process fluid abrasiveness and corrosivity
- Process fluid rheology, particularly for high concentration slurries
- Interrelationships between heat transfer options and process water balance.

Brief Historical Perspective

The earliest Bayer process plants employed essentially no heat recovery steps. The total energy consumption must have been of the order of 30 GJ/t. The evolution from this point to current norms of around 10 GJ/t has been made possible by a combination of heat transfer equipment and flowsheet innovations.

The following heat exchange milestones are the earliest examples known to the author:

- 1890 Batch Reverberatory Furnace for Calcination [6]
- 1902 Batch digestion operation at 143°C; no heat recovery; precipitation by carbonation; rotary kiln calcination [6]
- 1910 Batch digestion operation with single flash heat recovery stage [6]
- 1911 Precipitation cooling by “recycle spray tower” [6]
- 1935 Continuous Digestion Single Vessel; Three stage flash heat recovery in digestion; three stage flash cooling of pregnant liquor in Precipitation [6]
- 1938 Rotary Coolers for heat recovery on Alumina kilns [6]
- 1955 Fluidized Coolers for heat recovery on Alumina kilns [6]
- ~1958 Single stream high temperature digestion using steam coils in stirred autoclaves for indirect heating[7]
- 1960 Static Calciners utilizing hot gas cyclones for counter-current heat recovery [6]
- ~1960 Plate Heat Exchangers for pregnant liquor cooling [7]
- 1965 Single-stream digestion with indirect heating to ~200°C using shell & tube heaters [7]
- 1967 Flash Slurry Cooling in Precipitation [8]
- 1968 Predesilication process to reduce fouling rates of digestion heat exchangers [9]
- ~1970 Application of duplex stainless steels for corrosion resistance in high temperature digestion heat exchanger tubes [7]
- ~1971 Tube-in-tube heat exchangers for digestion slurry heating [7]
- 1973 High temperature Tube-in-Tube heat exchangers to 270°C for single-stream digestion (including heat exchange with molten salt at the hot end) [10]
- ~1973 Single-stream digestion heating to 260°C using combination of shell & tube heaters and autoclaves with steam coils [11]
- ~1978 Preheater tower retrofit to existing rotary kiln calciners [7]
- ~1985 Indirect platular slurry coolers in precipitation [7]
- 1988 Indirect heating of thick (mill) slurry in pipe-in-pipe heat exchangers [12]
- 1993 In-draft-tube cooling in precipitation [13]
- 2001? Falling Film evaporator for Bayer spent liquor
- 2001 Double Digestion with Pressure Decantation [14]
- 2004 Chemical inhibitor for suppressing sodalite scale formation [15]

Note that improvements in boiler technology, while important for overall refinery efficiency, have been excluded from the above list.

Heat Transfer Equipment

Choice of heat transfer equipment varies depending on the duty. Typical equipment choices and associated issues are listed below.

Liquor Heating

Spent liquor heating is a relatively straightforward duty. Typically either shell-and-tube or plate heat exchangers are employed. In evaporators, falling-films are also sometimes employed to give simultaneous heat transfer and evaporation.

Sodalite scale is the main liquor-side issue for liquor heating duties, particularly at high temperatures, and this has been well characterized, e.g. [16,17]. Heat transfer decay is rapid [15], and the scale growth is kinetically controlled with an activation energy E_a/R of around 9300K [18] which implies a doubling of scaling rate for every ~16°C temperature increase.

Cleaning frequencies of every 5 to 20 days are not uncommon [15,19]. However, the sodalite scale is readily removed by acid cleaning, so management of these heat exchangers relatively straightforward.

Mild steel is a suitable low-cost material of construction, although corrosion issues must be addressed, including caustic-related corrosion issues and acid corrosion during acid cleaning [20,21,22]. Above about 200°C, management of both sodalite scale and caustic corrosion issues tends to favor direct steam addition type heaters.

The predictability of spent liquor heater performance, and the ready availability of proven heater designs, has been a contributor to the ongoing popularity of the dual-stream digestion process, despite the emergence of alternative single-stream technologies in the late 1960's / early 1970's.

“Thin” Bauxite Slurry Heating

In “single-stream” digestion, the bauxite is mixed with the spent liquor and then heated to digestion temperature. This requires the use of slurry heaters. Three distinct heat exchanger preferences types emerged from the early development phase:

- Shell-and-tube [23]
- Pipe-in-pipe [24]
- Stirred autoclave with internal steam coils [11]

All of these heaters must deal with the issues of erosion, scale growth and scale removal. Heat transfer coefficients are comparable to those in liquor heaters [15,25] and scaling rates are reported to be lower [24].

The relative benefits of the different digestion heater types are a subject of ongoing debate in the industry. Table 1 is a brief attempt to summarize the main advantage and drawback of each type. The fact that all four technologies continue to be installed in new refineries is a strong indication that the debate is far from resolved.

Table 1 - Digestion Heater Benefits and Drawbacks

Heater Type	Main Advantage	Main Drawback
Shell&Tube Liquor Heater	Compact and hence low cost per unit area	Only heats the liquor stream; high scaling rates
Pipe-in-Pipe Slurry Heater	Long service life	High capital cost; difficult to access for repair or shell-side cleaning
Shell&Tube Slurry Heater	Compact, off-the shelf items mean low cost per unit area	Smaller pipes and more inlet-outlet transitions has potential to impact on erosion, blockage risks and pressure drop
Autoclave Slurry Heater	Main advantages relate to pressure rating and low slurry-side pressure drop	High pressure agitator seals; relatively high scaling rates

A fifth potential candidate warrants mention, namely slurry-slurry heat transfer. All of the options above use steam (flash vapor or boiler steam) as the heat source. Slurry-slurry transfer offers some significant potential approach temperature advantages. To date, this has achieved only limited commercial application [26]. Another interesting variation on slurry-slurry heat transfer has been proposed by Fulford and Chinloy [27].

Thick Slurry Heating

Bauxite slurries in Bayer liquor exhibit significant changes in rheology as solids concentrations increase. Above about 300 g/L solids, the slurries begin to exhibit yield-stress or power-law type behavior (e.g. Zirnsak *et al* [28]). This presents a significant challenge for heat exchanger design, and warrants a separate category for “thick” slurry heating.

The main application for thick slurry heating is for predesilication. Typical temperature requirement is 100°C or a little higher. The slurry has high scaling potential due to the rapid dissolution of kaolin and associated high silica concentrations. These heat transfer challenges have typically been addressed (or perhaps more correctly, avoided) by using direct contact heaters [6].

The particular challenges associated with indirect heating in this service are addressed in the Case Study below.

Heating of Wash Water and Other Dilute Streams

For heating of dilute streams such as wash water, direct steam type heaters are almost always employed. This is because the dilution penalty in this part of the flowsheet is negligible. The only advantage of indirect heat transfer in this case is the production of a high purity condensate stream as a byproduct.

Bauxite Slurry Cooling

With the exception of the pulp-pulp technology mentioned earlier [26], bauxite slurry cooling is almost universally accomplished by flash cooling. This carries a significant approach temperature penalty, typically around 20°C, due to the combined effect of

boiling point elevation and the stepped temperature drop. However this is generally outweighed by the benefits in terms of water balance and reduced scaling issues on the hot fluid side of the heat exchangers.

Flash vessel design and slurry transfer systems need to address the issues of hydraulic transfer, three-phase flow management, slurry entrainment during flashing, slurry disentrainment in the vapor space, scaling and erosion.

Precipitation Slurry Cooling

Early precipitation circuits tended to be self-cooling due to a combination of:

- Batch precipitation in relatively small precipitators (smaller size means greater temperature drop from convection and radiation losses)
- Long holding times allowing further natural cooling
- Use of airlift style agitators which provide significant evaporative cooling

The shift towards continuous precipitation circuits, larger tanks, and mechanical agitation created a need for cooling processes.

Flash cooling provided a good solution, having the advantage of avoiding cooling surface in contact with scale-forming and abrasive slurry. However the normal water balance benefit of evaporative cooling is offset in this case by unfavorable impacts on gibbsite equilibrium solubility, to the point that it may even be attractive to re-inject the flash condensate.

More recently, indirect slurry coolers have found increasing use. Options for indirect cooling include wide-gap platular heat exchangers, shell-and-tube heat exchangers, and in-draft-tube double plate coolers.

Liquor Cooling

Liquor cooling is required for pregnant liquor feeding precipitation and for spent liquor feeding oxalate removal facilities. Both flash cooling and indirect cooling can be employed, with plate heat exchangers finding preference in recent years due to compact design and good approach temperatures.

Calcination

Calcination is primarily a heat-transfer operation. Traditional rotary kilns have made way to more compact and energy-efficient stationary calciners which utilize a cascade of gas-solid separators (gas cyclones) and contactors (venture mixers, fluid beds, etc) to provide high efficiency counter-current heat exchange.

Heat Exchanger Design

Design of heat exchange systems in the Bayer process is an iterative process encompassing the steps of:

- Flowsheet development
- Pinch analysis (and optionally water pinch) for optimization
- Heat exchanger technology selection
- Heat exchanger sizing and thermal design
- Sensitivity analysis to assess robustness of selected design across a range of operating scenarios

→ Detailed design of heat exchanger

The core step of heat exchanger technology selection and design requires a detailed understanding of operational issues including achievable heat transfer coefficients, scale growth rates and scale removal options. Attention to detail at the design stage can mean the difference between a trouble-free system and an ongoing operations and maintenance headache.

Case Study – Thick Slurry Heating

Some of the challenges of thick slurry heating are described above. This is arguably the most difficult heat transfer challenge in the Bayer process. The four key challenges are:

- High viscosity non-Newtonian slurry
- High scaling potential
- Erosion and corrosion potential
- Unsteady flow rates (because heat exchangers are typically coupled to individual mills)

Option 1: Don't Do It

Some problems are best avoided altogether. It is possible to forego pre-desilication altogether, and manage silica via extended digestion holding time or a post-desilication step.

Alternatively the pre-desilication solids concentration may be reduced until the slurry viscosity moves out of the “thick” slurry region and the heat transfer problem becomes easier to solve. This approach however is effectively limited to single-stream digestion units processing high-silica bauxites (due to a combination of heat balance and desilication chemistry issues).

Option 2: Direct Steam Injection

This is the proven approach, but it typically adds around 50 t/h of dilution to the process, which must be offset by additional evaporation plant capacity. Given the difficulties guaranteeing robust performance of any of the options presented below, installation of a direct injection “backup” stage should always be considered.

Option 3 – Pipe-in-Pipe

This approach was pioneered by Kumar [12].

The most important criterion for good heat transfer in thick slurries is to achieve turbulent flow. This requires Reynolds number of nominally >2500. If the pipe-in-pipe heater is to be coupled to an individual mill discharge flow of, say, 300 m³/h, and the target slurry velocity is 10-15 ft/s as suggested by Kumar, then the corresponding tube diameter is 160mm. A Reynolds number greater than 2500 will be achieved provided the slurry viscosity is less than about 400 mPa.s.

The main problem with this approach is the very long lengths of pipe-in-pipe exchanger in order to achieve the required heat transfer area. This can be partially offset by using multi-pipe variations with, say, 3 internal pipes per shell. But even with triple-pipe exchangers the installation requires a significant footprint and steel quantities, as illustrated in Figure 2. The issue is compounded by requirements for insulation, vapor distribution and condensate collection.

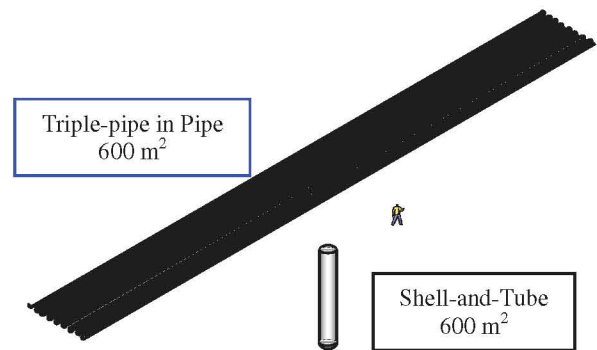


Figure 2 - Graphical illustration of quantity differences between pipe-in-pipe and shell-and-tube heat exchangers.

Option 4 – Compact Heat Exchangers

More compact heat exchanger designs such as shell-and-tube, spiral or wide-gapped plate/platular all offer cost-effective heat transfer area in a compact unit. For the sake of simplicity the calculations below assume a shell-and tube arrangement although the same arguments apply equally to the other exchanger types.

The narrow flow channels involved make it more difficult to achieve turbulent flow with high viscosity slurries. For example if the tube diameter is 30mm and the velocity is 3 m/s then achieving Re>2500 requires viscosity < 60 mPa.s. This places a significant constraint on pre-desilication slurry density, which may have downstream consequences, particularly in dual-stream digestion processes.

Also if the flow rate is still tied to mill discharge flow rate then this places a limit on the number of tubes that can be installed per heat exchanger pass, while still achieving the target slurry viscosity.

The heat exchanger described by deBoer et al [29] addresses these issues by using slightly larger-diameter tubes (to increase Reynolds number) and increasing the flow per heater to ~600 m³/h (to allow a reasonable number of tubes per pass), but still requires a reasonably large number of modules in series.

The biggest issue with thick slurries in compact heat exchangers relates to the fluid dynamics of parallel flows of non-Newtonian fluids. This is unique to power-law and yield-stress fluids and goes well beyond the simple need for a good flow distributor to present slurry equally to all tubes. Figure 3 shows how pressure drop varies as flow is increased in a 50mm tube for a power-law fluid ($\tau = K\dot{\gamma}^n$) with K=0.011 Pa.s and n=0.1. In a heat exchanger, all parallel tubes see the same pressure drop, since they connect the same feed and return chambers. Unfortunately in this case this does not guarantee an equal flow distribution. There is quite a broad dead-band at around 20 kPa pressure drop, wherein it is possible for one tube to operate in laminar flow at about 0.1 m/s while its neighbors operate in turbulent flow at around 2 m/s. This is analogous to two other maldistribution phenomena reported in the literature when heating gases or

boiling fluids (tubes with low flow heat up more which leads to expansion which generates backpressure and hence low flow) and cooling liquids (tubes with low flow cool more and become more viscous and hence run at low flow)³⁰. However, shear-thinning induced maldistribution (tubes with low flow tend to thicken and hence run at low flows) appears to have received relatively little attention in the literature.

This maldistribution places the tubes with laminar flow at high risk of stagnation, scaling and ultimately complete blockage. To an external observer, there is little or no evidence of what is going on inside the heater until it is too late.

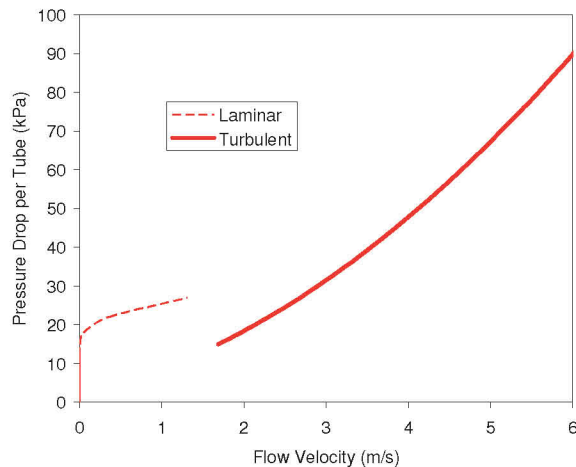


Figure 3 - Pressure Drop vs Flow for Power-Law Fluid in a heater tube (D=0.032, L=9, K=0.011, n=0.1).

Good heater design must seek to avoid this dead-band zone by setting a minimum velocity limit and a maximum viscosity threshold. However even with good design under normal operating conditions, any temporary turndown in flow rate has potential to cause tube stalls.

Option 5 – Compact Heat Exchangers with Recirculation

By breaking the coupling between mill slurry flow and heat exchanger flow, it is possible to overcome two of the key limitations above:

- Avoids exposure to flow downturns during mill or refinery upsets
- Allows high flow rate which permits a larger number of tubes per pass and hence more economy of scale in the heat exchangers.

Figure 4 shows one such arrangement with two-stage heating plus a contact heater as backup. In reality typically only one indirect heating stage is required. The sketch shows two different ways of providing the recirculating flow.

The duty of the mill discharge pump is essentially just to transfer slurry to the predesilication tanks. The energy to overcome friction and provide turbulent flow in the heat exchangers comes from the booster/recirc pumps.

The alternative configuration shown in Figure 5 totally decouples the heating operation from the mill capacity.

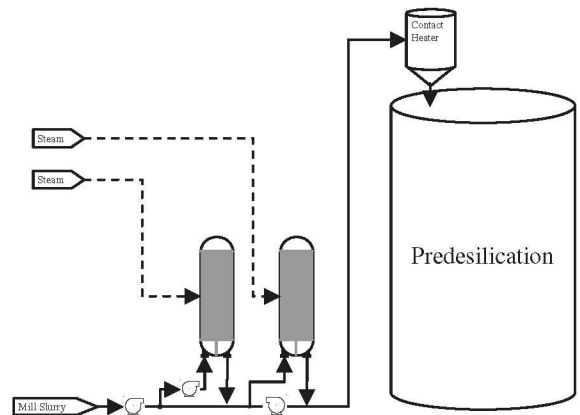


Figure 4 - Thick Slurry Heating with Recirculation

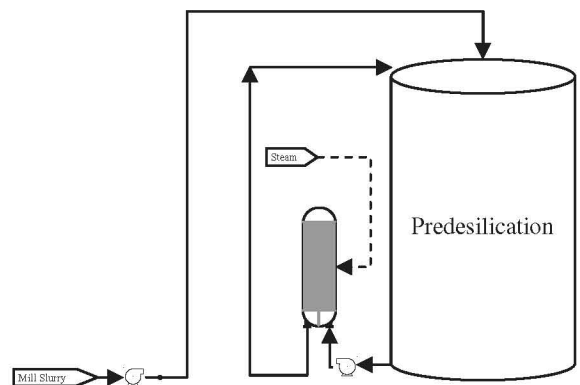


Figure 5 - Alternative Configuration

There is a potential process risk with this second configuration depending on the extent of the desilication reaction taking place within the lead desilicator vessel. There are a range of chemical and particle size related processes taking place in this vessel [31], all of which have potential impacts (both positive and negative) on both scaling rates and slurry viscosity.

The slurry recirculation approach makes it possible to:

- Achieve economy of scale by using large heat exchangers (1000m² or higher)
- Achieve good flow distribution in the heater heads by using single-pass or two-pass configurations.
- Re-optimize tube velocities after installation or in response to future process changes.
- Achieve a high degree of mill turndown without compromising heater operation

A drawback of these configurations is a slight loss of driving force in the heat exchanger due to recirculation of heated slurry. This is more than compensated for by the above efficiency gains.

Future Challenges and Directions

Review of the literature suggests a peak in R&D activity in the 1990's. Despite apparently promising results from novel approaches such as fluid-bed liquor heating, boiling films, fancy

tube geometries and various scale suppression strategies, there has been very little pull-through into commercial applications.

The challenge remains to deliver significant reductions in energy consumption, water usage and environmental footprint. The required improvements will not come without significant breakthroughs in heat transfer. Ongoing R&D, plus a commitment to transfer the results to the real world, should be a priority.

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