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PERSPECTIVE ON BAYER PROCESS ENERGY

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Abstract

The three most important cost items in the production of alumina are bauxite, caustic soda, and energy. Alumina energy cost will likely rise more than other costs because of energy price increases and energy related environmental issues. As energy cost increases, refineries will need to adjust the process to maintain or improve their position on the world cost curve.

The Bayer Process is simple, but the various uses of energy in the process are not as simple to understand. The primary uses of energy in the process and their interrelationship will be discussed.

INTRODUCTION

Energy is one of the three largest costs in the production of alumina from bauxite. In the future, energy could be two or three times today's cost (relative to other costs) because of availability and environmental issues. The better the understanding of the factors that make up the usage of energy in the alumina refinery, the better prepared process engineers and designers will be to evaluate refinery design changes in event a significant change in energy supply price or availability.

This paper on energy in the Bayer Process considers only the interrelated areas of Powerhouse (Boilers), Digestion, and Evaporation. Excluded are Calcination and peripheral energy use areas such as Seed Wash, Liquor Purification, Causticization, etc.

As shown in Figure 1, the source of energy to the Bayer Process is the Powerhouse. Energy is distributed essentially as steam to Digestion and Evaporation and electrical power to the refinery. The amount of energy required is determined by four interrelated factors:

1. Type of Bauxite. Trihydrate bauxite (low temperature digest or monohydrate bauxite (high temperature digest).

2. Productivity. The amount of alumina produced for each unit volume of liquor circulating (see Fig. 1)

3. Evaporation. The facility that removes the net amount of water added to the process.

4. Refinery Facilities. The physical design of the refinery with respect to heat recovery facilities.
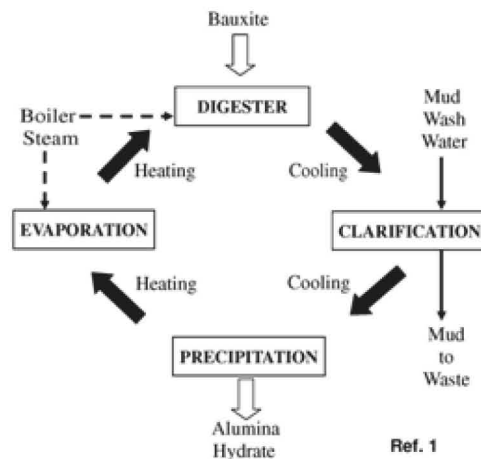

Fig 1. Bayer Process – Circulating Caustic Liquor

DISCUSSION

Type of Bauxite

Digester temperatures range from about 140 °C for trihydrate alumina (THA) bauxite up to about 280 °C for monohydrate alumina (MHA) bauxite. Because of the flash cooling and feed slurry heat recovery system used in Bayer Process refineries, the Digester temperature itself does not have a significant affect on Boiler Steam usage as long as there are no other streams entering or leaving (e.g., Blow Off). That is, theoretically, it will take nearly the same amount of Boiler Steam for a high temperature digest as for a low temperature digest. This is illustrated in Fig. 2.

In practice there are other heat demands that do require additional Boiler Steam to Digestion:

Heat of Reaction (HR). This is the endothermic heat of caustic attacking bauxite alumina hydrate to put the hydrate in solution. The HR is higher for THA than for MHA.

Sensible Heat Loss (SHL). This loss will be higher for MHA digestion because there will be over three times as many flash tanks (FT) and heat exchangers (HX) of which 2/3 will be at significantly higher temperatures than for THA Digestion.

Blow Off Vapor. By design, Blow Off vapor is kept small because it is a direct loss of steam (heat) to the atmosphere.
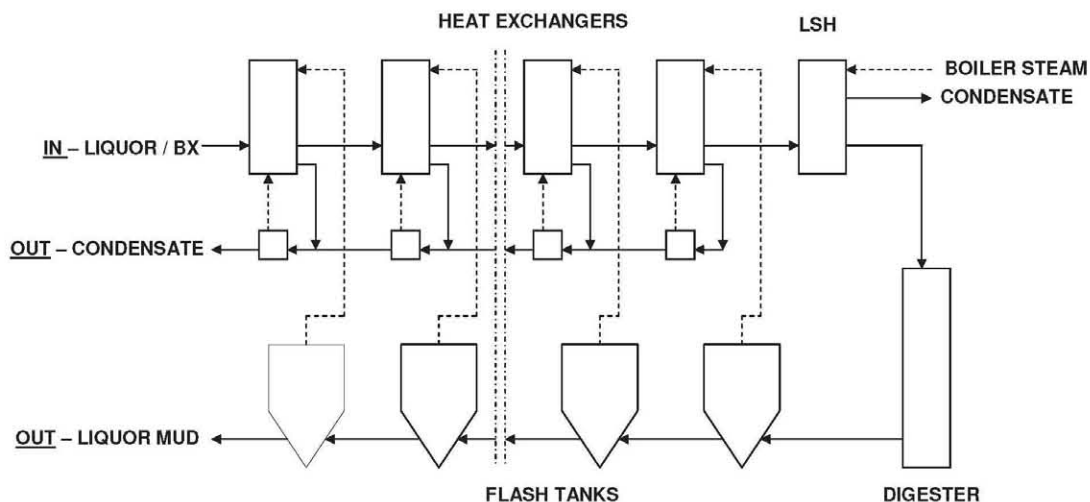
Note: For this illustration no other heat streams enter or leave the system (e.g. Blow Off Vapor, Heat of Reaction, Sensible Heat Loss)

Fig. 2   Heat Out – Heat In = Boiler Steam Heat

Following is an example of the distribution of Boiler Steam in a flash/heater system. The example does not cover the whole range of refinery designs but it does indicate the relative importance of the factors listed above.

	% Distribution
FT/HX (See Fig. 2)	50-60
Heat of Reaction	20
Sensible Heat Loss	5-15
Blow Off	0-?
Washer Overflow	20

THA and MHA Digestion widely differ in the amount of evaporation that occurs. For instance, compare the 275°C digest (MHA) to a 140°C digest (THA) with the same product dissolved solids concentration where the temperature difference is indicative of the amount of evaporation:

MHA digest 275°C - 110°C = 165 °C evaporation
THA digest 140°C - 110°C =   30° C evaporation

For the 275 °C MHA Digestion, the evaporation economy (Fig. 2) is near 7 tons water/ton of steam. To achieve the same total

evaporation (Digestion plus Evaporation) in a THA digest refinery would require a large Evaporator facility at about a 3.6 economy and much more total Boiler Steam than the comparable MHA refinery.

The thermal energy required for Digestion and Evaporation in a MHA Digestion refinery will be less than for a comparable THA digestion refinery because of the higher average efficiency (economy) for total evaporation.

When energy required for refinery electric power is included, the THA Digestion refinery will have a significantly lower electric power requirement. MHA Digestion will require more electric power in order to raise the bauxite slurry to digest pressures as high as 80 atm. (plus the frictional pressure drop in the large number of heat exchangers) compared to THA Digestion pump pressure of less than 10 atm. This will raise the MHA refinery electric power 10% - 15% higher than the THA refinery.

However, that is not the most important difference in providing energy for electric power. With a 60 atm Boiler in the THA

process, electric power can be generated by first passing Boiler Steam through an extraction turbo-generator. In a MHA process an extraction turbo-generator can be used only to a limited extent, and most, if not all, electric power will be generated by condensing turbo-generators at nearly three times the energy required by extraction turbo-generators.

Another factor is the type of fuel supply. The discussion so far has a Powerhouse design of any fuel (coal, oil, natural gas) generating steam to process and generating electric power by extraction or condensing turbo-generators. In the case of natural gas fuel, electric power can be generated by gas turbo-generators with waste heat boilers for steam supply. Natural gas fuel to the Powerhouse can avoid the use of condensing turbo-generators for electric power and provide power at the same low energy usage of extraction turbo-generators.

PRODUCTIVITY

Productivity or Yield is important in design of a refinery. A productivity increase from 75 to 90 gpl alumina is an increase of 20% in production for the same pipes, pumps, and vessels in the refinery. This is very significant in both new and expanding refineries. Higher productivity will give an overall lower Boiler Steam and electric power demand per ton of alumina produced. However, for MHA Digestion, there will be proportionately lower evaporation in Digestion and more mud wash water which will require more evaporation in the Evaporator facility at a lower economy than Digestion. The heat of reaction (HR) per ton of alumina will be the same, but the sensible heat loss in Digestion will be proportionately lower. Overall, perhaps a 10% decrease in Boiler Steam usage per ton alumina for a 20% increase in productivity. The subject requires a more in depth evaluation than can be covered in this paper but it is certainly an important factor in refinery energy usage.

EVAPORATION

An Evaporator facility is usually designed by a contractor with boundary limits provided by the alumina process engineer. This typically results in an evaporation economy of around 3.6 tons water/ton Boiler Steam. As liquor caustic concentrations increase for high yields (productivity), achieving a 3.6 economy becomes more difficult because of a corresponding increase in BPE.

Increasing the Evaporator LSH temperature to over 200 °C and adding more stages (effects) could increase the economy to over 7 tons water/ton steam, a very significant reduction of Boiler Steam. Erosion/corrosion issues would need to be addressed. Silica scale on heater tubes at the higher temperatures can be controlled by chemical additive.

FACILITY DESIGN

An important factor that directly determines the minimum Boiler Steam to Digestion is the nature of the Flash Cooling/ Heat Exchange system itself. This is illustrated in Fig. 3 showing an example of temperatures across the first stage following the digester. Digester discharge at 250 °C immediately flashes to 230 °C, the temperature of both liquor and vapor. The vapor is

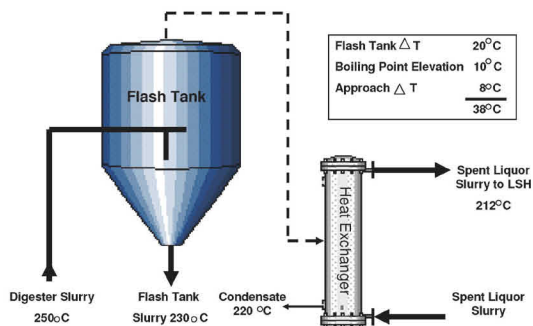| Flash Tank △ T	20°C
Boiling Point Elevation	10°C
Approach △ T	8°C
	38°C

Fig 3. Example - Minimum Temperature Difference Across Digester T °C and LSH Feed T °C

superheated (BPE) 10 °C and must be cooled 10 °C before it can condense in the Heat Exchanger. Cooling of superheated vapor to condensing temperature contributes little to heat transfer (less than 2%) compared to the heat transferred by condensation. In this example, the term 'approach temperature' is the difference of the steam condensing temperature above the slurry heater discharge temperature .

Shown in the table in Fig. 3 is the sum of the three temperature drops (38 °C) which is also the temperature rise across the LSH. If the temperature of the slurry reporting to the last HX is lower than 182 °C, Boiler Steam will increase. In the example of Fig. 3 there will be more Boiler Steam than required by the heat balance described in Fig. 2. This indicates that if there were no other requirement for heat in the system (such as HR, SHL, etc.), there will be Blow Off vapor. Fig. 4 depicts Fig. 3 graphically and at the bottom shows the effect that HX area and number of stages will have on Boiler Steam to the LSH.
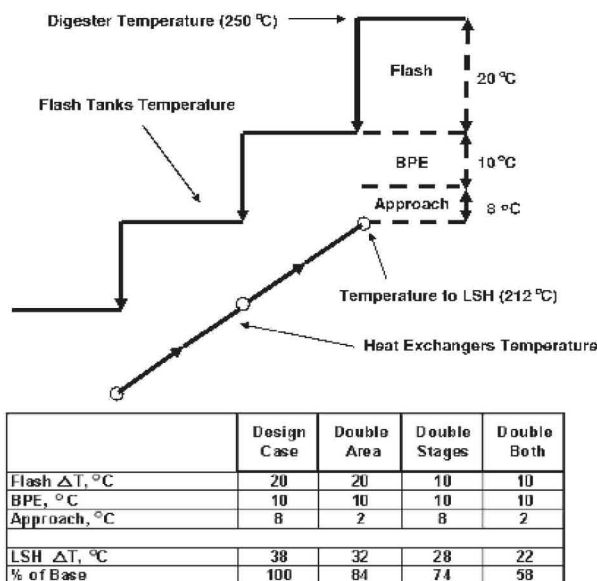
Fig 4. Effect of Heat Transfer Area and Number of Flash Stages

	Design Case	Double Area	Double Stages	Double Both
Flash △T, °C	20	20	10	10
BPE, °C	10	10	10	10
Approach, °C	8	2	8	2
LSH △T, °C	38	32	28	22
% of Base	100	84	74	58

The Bayer Process is one large heat exchanger. This is illustrated in Fig. 5. There is practically no temperature drop in liquor across Evaporation and sensible heat loss across Clarification will be small. The temperature drop across Precipitation is composed of induced cooling (cooling water) to lower the process temperature of liquor to the precipitators, sensible and evaporative heat loss during the long holding time in Precipitation, and induced cooling during precipitation to further lower the temperature.
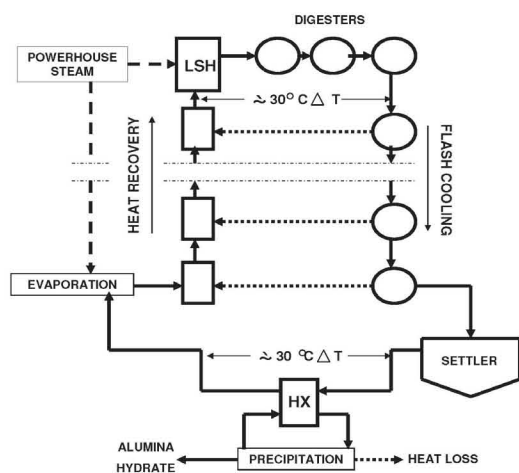

Fig 5. Temperature Difference across Digestion
and Precipitation

The driving force for heat transfer is the temperature difference across Digestion and across Precipitation. As indicated in Fig. 5, in order to reduce Boiler Steam to Digestion, the delta T across Digestion must be reduced. However, this will not be practical or efficient without correspondingly lowering the delta T across Precipitation This will require more recuperative cooling in Precipitation in place of heat loss by cooling water.

Design for MHA bauxite can reach very high temperatures in the Digester (>260°C). The LSH can elevate Boiler pressure above 70 atm. and up to 90 atm depending on Digest temperature. The corresponding temperature of steam will be above the limit to use process condensate for feedwater because of serious corrosion in Boiler tubing. For this reason, Digestion design using Boiler Steam at high pressure (>70 atm), must return LSH steam condensate to the Boilers. For instance, Boiler Steam to LSH at 100 atm and 315°C will be nearly doubled because almost half the steam heat will be returned as condensate to the Boiler. The alternative of routing the condensate to the Digestion Heat Exchangers instead would require using demineralized Boiler feed water at higher cost and, most likely, an increase in Digestion heat exchange area or number of FT stages. Theoretically, the net thermal affect of returning LSH condensate is the same; it's just the issue of physically returning high temperature/pressure condensate to the Boiler. The return of condensate can be avoided by using a different method of heating liquor before the Digester, such as a molten salt HX (2) in place of the LSH.

FINAL COMMENTS

There should be increased interest in investing in heat recovery facilities in the event of a doubling or tripling of energy cost relative to other alumina costs. The possibilities that immediately arise are:

1. Use of natural gas to reduce the heat required to generate electric power in MHA refineries. Refineries that do not have the availability of natural gas may consider using LNG.

2. Increase the number of stages in Digestion Flash Heat Exchange and at the same time increase recuperative cooling in Precipitation to lower the refinery delta T (see Fig. 5).

3. With high temperature MHA Digestion (>260 °C), it should be feasible to eliminate the Evaporation facility.

4. Route Washer Overflow directly to Precipitation. It appears wasteful to heat Washer Overflow for Settler operation and then essentially cool the same stream in Precipitation.

5. Increase the evaporator LSH temperature and the number of stages (effects) to increase the economy in Evaporation.

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