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THE DESIGN OF PRESSURE SAFETY SYSTEMS IN THE ALUMINA INDUSTRY

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Abstract

The alumina refinery presents the designer with multiple challenges. For a given process flowsheet, the mechanical equipment installed must be routinely inspected and maintained. Piping systems must also be inspected routinely for signs of erosion and/or corrosion. Rapid deposits of chemical species such as lime, silica, and alumina on equipment and piping need special consideration in the mechanical design of the facilities, such that fluid flows are not unduly interrupted. Above and beyond all else, the process plant must be a safe place of work for refinery personnel.

Although much of the alumina plant flowsheet is executed at atmospheric pressure, the digestion facility and boiler plant are two principal areas of the refinery that operate at conditions of up to 400 degrees celcius and 100 Bar pressures. The digestion facility in particular may be comprised of many alternative process designs displaying either inherent mechanical simplicity or complexity. This paper outlines some of the pressure safety considerations to be incorporated into the mechanical design of the digestion facilities for some alternate process flow sheets. Armed with these considerations at the process flowsheet definition stage, optimisation of the process and/or equipment selection is possible preserving the delicate balance of process facility performance, plant operability and maintainability, and personnel safety.

Introduction

Overpressure protection systems may take the form of passive protection or, of more recent consideration in the industry, the use of safety instrumented systems, or some combination of the two. These protection systems guard plant equipment, piping and personnel from process transients that deviate the plant away from normal operating conditions.

Within the digestion facility of the alumina refinery, these transient conditions may include equipment blockages, power failures, hydraulic expansion and pressure transients to name a few.

These transients will manifest themselves with differing characteristics as a function of the mechanical equipment employed and the process flowsheet selected. This paper outlines considerations of particular relevance for accommodating these transients for various equipment selections and digestion facility process designs, to ensure plant reliability and personnel safety. The result of these considerations will be an improved understanding of process and equipment designs that offer the plant designer advantages or disadvantages in intrinsic safety.

Digestion Facility Flow sheets

Three digestion flow sheets employed within the industry are presented schematically in Figures 1, 2 and 3 below.

Figure 1 represents a dual stream or split flow digestion facility whereby caustic liquor is heated separately to bauxite slurry in shell and tube liquor heat exchangers before both are combined in "digester" vessels. These process designs have been employed for "Low Temperature" (digester operating conditions of ~150°C and 7 Bar) and "High Temperature" (digester operating conditions of ~250°C and 37 Bar) digestion facilities. In High Temperature facilities, duplex alloy steels and nickels are often employed in the heater tubes to offer corrosion resistance to caustic at elevated liquor temperatures above 170°C. Final digestion temperature conditions are typically attained with the use of direct steam sparging into the digester vessels.

Figure 1 – Dual Stream Digestion Facility

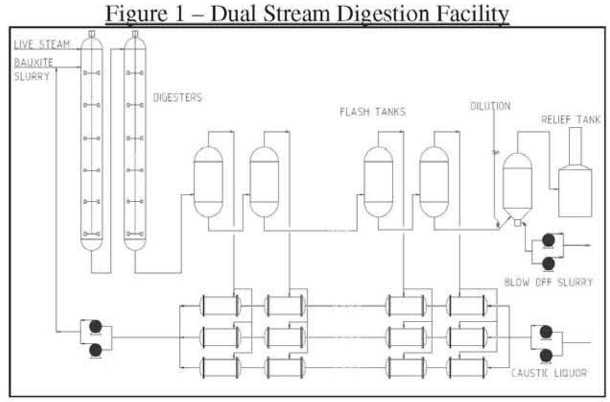


Figure 2 represents a single stream digestion facility incorporating multiple trains of shell and tube heat exchangers. In this process design, the caustic liquor and bauxite slurry are first mixed before being heated in the shell and tube heat exchangers. Following the recuperative stage heating, indirect steam heaters are used to elevate the slurry temperature to the target digestion temperature.

This single stream arrangement has been utilised more typically for Low Temperature digestion facilities (digester operating conditions of ~150°C and 7 Bar) and may incur frequent heater tube replacement as a result of tube blockages and/or tube erosion.

Figure 2 – Single Stream Digestion Facility

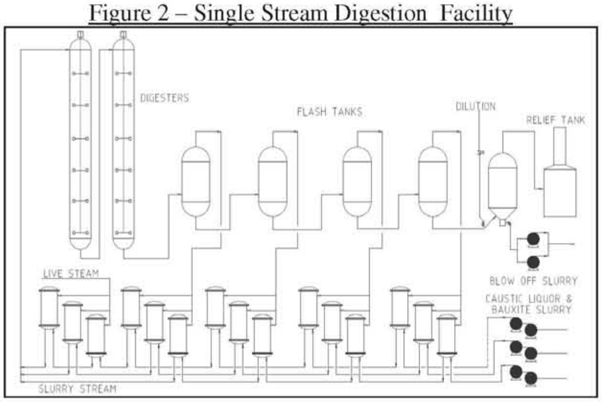
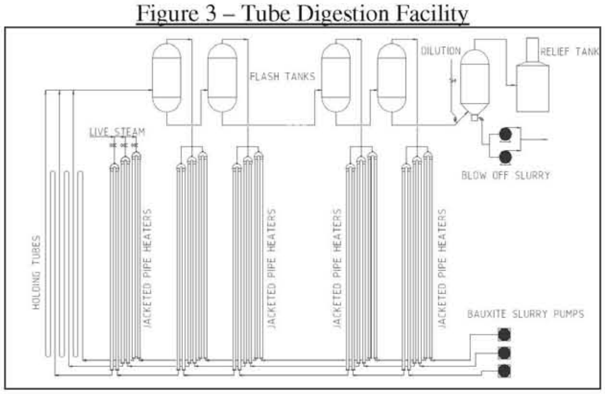


Figure 3 represents the current evolution of single stream technology and is a High Temperature Tube Digestion unit (digestion operating conditions of 280°C and 60 Bar) using technology developed by HATCH Associates Ltd and employed for the KGCC refinery in South Korea and more recently for the Rio Tinto Yarwun Alumina Refinery in Gladstone, Australia. In this design, the shell and tube heaters are replaced by jacketed pipe heaters incorporating tubes of larger bore diameter than utilised in the shell and tube heater. As per the single stream flowsheet of Figure 2, the tube digestion facility mixes the bauxite slurry and caustic liquor prior to the heating circuit. Final digestion temperatures are attained with indirect live steam heating utilising high pressure steam at 310°C and 100 Bar. Alternatively, molten salt at ~400°C may be used.

Figure 3 – Tube Digestion Facility



Transient Considerations and Overpressure Protection Design

The following transients will be reviewed in relation to overpressure protection for the three Figures above.

1. Pump Shut Off Head
2. Vessel Blockages
3. Power Failure
4. Heat Exchanger Burst Tube
5. Heat Exchanger Tube side hydraulic expansion
6. Heat Exchanger Shell side hydraulic expansion
7. Equipment Design for Superheated Steam
8. Pressure Transients / Water hammer

For each transient condition, the application of passive overpressure protection and/or safety instrumented systems will be reviewed.

Passive overpressure protection utilises pressure (and coincident temperature) ratings and the use of safety relief valves and bursting discs to protect mechanical equipment and piping from exceeding code allowable overpressure. Safety Instrumented Systems (SIS) utilise high integrity instrumented controls as part of dedicated and independent safety shutdown systems to achieve overpressure protection.

When utilising a SIS, redundancy is often employed by way of duplicate or triplicate instruments to meet a target "Safety Integrity Level" (SIL). Unlike the use of passive protection, safety instrumented systems must prevent system pressures exceeding the equipment Maximum Allowable Working Pressure.

1. Equipment Design for Pump Shut Off Head

For the tube-side design of the heat exchangers in figures 1-3 above, passive overpressure protection may consider :

(a) a tube-side rating that meets or exceeds the maximum pump discharge pressure.

(b) a tube-side rating that is a suitable margin above normal operating system pressure but below the pump/s maximum shut off head. Bursting discs and/or safety relief valves on the tube-side of the heater or interconnecting piping would then limit tube side pressures to the maximum allowable pressure.

For the fluids in use in the digestion facility, their tendency to deposit scale on the piping and heater tube walls renders the application of safety relief valves or bursting discs problematic for overpressure protection. In particular, the split flow system (Figure 1) is prone to an almost exponential increase in sodium silicate (sodalite) scaling rates at elevated temperatures beyond 170°C and up to 215°C. Beyond 215°C, liquor scaling rates make the use of shell and tube heaters impractical.

For high temperature digestion split flow flowsheets (Figure 1) employing the use of passive overpressure protection as per (a) above, tube side heater ratings in the vicinity of 50 to 60 Bar are not uncommon.

For a low temperature digestion single stream flowsheet (Figure 2) employing the use of passive overpressure protection as per (a) above, tube side heater ratings in the vicinity of 40 to 50 Bar may be expected.

For the high temperature Tube Digestion facility (HATCH Associates Ltd) incorporating the use of Jacketed Pipe Heaters (Figure 3), optimal use of the transient provisions of the piping code have enabled the use of passive overpressure protection as per (b) above. As positive displacement slurry pumps are utilised, internal pressure relief valves (PRV's) are employed to primarily limit casing pressures developed by the pump in the event of inadvertent downstream valve isolation. These PRV's indirectly protect maximum tubeside heater pressures from exceeding code allowable overpressure. The relief valves are installed on the hydraulic fluid on each pump cylinder and therefore attain the benefit of not being exposed to the scaling nature of the process fluid.

The alternative to passive protection for consideration of the heater tube side design pressure, is the use of a SIS to limit overpressure. Dependent on the required SIL rating that may ensue from a process hazards risk assessment, duplicate or triplicate purged pressure tappings are installed at heater and pump outlet manifolds as part of a dedicated and independent safety shut down system. A tube side rating that provides a reasonable margin over system operating pressures may be specified such that whilst not compromising the mechanical integrity, the result may be an equipment design of lighter construction and transportability, a key benefit for consideration of routine heater maintenance.

This is of particular relevance to shell and tube heat exchangers used for either the split flow (Fig 1) or single stream (Fig 2) flowsheets. A tube side rating in the vicinity of 70% of that used for passive protection would be expected. For comparative purposes for example, the high temperature digestion split flow flowsheet of Figure 1 employing the use of a SIS to protect the heater may employ a tube side rating of between 35 to 40 Bar, relative to the 50 to 60 Bar otherwise required for passive protection.

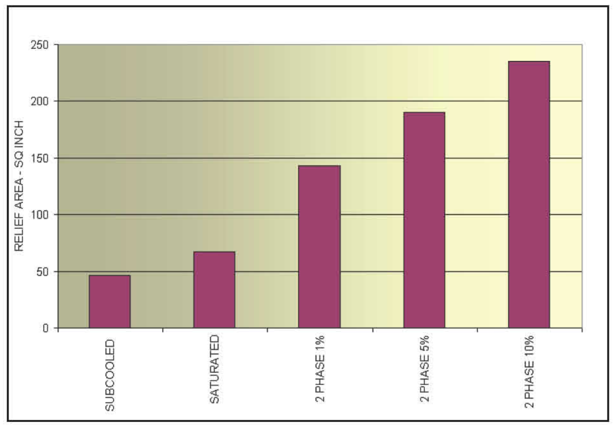
2. Vessel Blockages

Blockages of vessels in the flash tank train may result from internal wall scale dislodgment (particularly during unit startups) or mechanical failure of interposing control valves. The provision of passive overpressure protection via simply rating all vessels to the maximum pump shutoff head (refer (a) above) is generally uneconomic, and more rigorous analyses are required. Independent of the process flowsheet selected, these analyses will be subject to the following variables and assumptions:

1. Nominated vessel design pressures
2. Selected time frame of transient analysis eg 15 mins
3. Normal process flowrates
4. Slurry and liquor supply pump capacities
5. Physical interconnecting piping geometry and componentry used
6. Pump or system supply capacities for miscellaneous injection streams eg. lime and live steam sparging
7. Upstream vessel operating pressures during the downstream vessel blockage i.e. constant or decreasing
8. Fluid state temperatures at the inlet to downstream vessel relief valves following a relief event i.e. subcooled, saturated or two phase mixture
9. Dual phase (slurry) or three phase (flashing slurry) fluid movement between interconnecting pressure vessels following the transient condition.

The blockage analysis can multiply in complexity for the high temperature digestion facility as a result of the multiple number of flash tanks employed and sound engineering judgement is required to bound the analysis within practicable limits. For the flowsheets depicted in Figures 1, 2 and 3 above, a vessel blockage downstream of the digester or 1st stage flash tank will require a relieving capacity from PRV's as dictated by the capacity of the delivery liquor, bauxite slurry and lime supply pumps, and the direct steam injection supply capacity where direct steam sparging is used (Fig 1). Items 1, 4, 5 and 6 above are of particular relevance for this analysis. For a blockage of a flash tank further downstream in the flash tank train, the specification of relieving capacity will be dictated by Items 1 to 9 inclusive, and the analyses for determination of relief valve area become particularly sensitive to Item 8 above. This point is illustrated in Figure 4. below illustrating the required relief valve area for differing fluid states to the relief valve inlet.

Figure 4. Relief Valve Area as a function of Fluid State



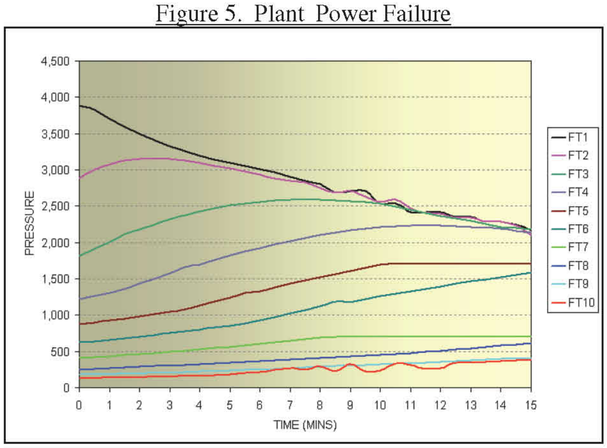
The use of a SIS for vessel overpressure protection is well proven for low temperature digestion applications (Fig's 1 or 2). Multiple (purged) pressure tappings are incorporated in the various pump discharge manifolds and pressure vessel head spaces or vapour lines, where the risk of contamination from the process fluid can be mitigated. Pressure trips are then utilized to sever the energy source from the downstream equipment. The reliance solely on a SIS for a high temperature digestion unit application (Fig 3) however is not recommended, as interconnecting vessels will not be responsive to upstream interlocks and safety trips on motive power supplied by liquor or slurry pumpsets. The process thermodynamics and piping system will then dictate response characteristics and required overpressure relief capacities.

3. Power Failure

As with the vessel blockage transient above, passive overpressure protection for a power failure requires quantification of both the fluid physical properties and required relieving capacity for specification of the number and size of pressure relief valves and/or bursting discs. To quantify these parameters, an understanding of the mechanics of the power failure transient is required. Analytical techniques to understand the mechanical response characteristics of the digestion unit to a power failure were developed by HATCH Associates Ltd in March 1999 and fundamentally require an understanding of energy flows. These analytical methods employ the thermo-hydraulic response characteristics of the equipment and piping following the power failure, to assess relieving rates, such that vessel pressures during the transient do not exceed code allowable pressures.

Figure 5 below is the result of a dynamic simulation and typically illustrates the changing and increasing vessel operating pressures experienced in a flash tank train during the power failure, reflecting the increasing energy states of the vessels over time.

Figure 5. Plant Power Failure



These simulations allow the quantification of relieving rates required to contain energy states within the pressure vessels to allowable overpressures. They are also critical tools in establishing the capacity of passive relief to protect against the transient condition for alternate process flowsheets.

For all three digestion flowsheets above (Fig's 1-3), passive overpressure protection of downstream flash tanks will usually require sizing of PRV's for vapour relief only. There are however some salient differences between these flowsheets that illustrate aspects of inherent safety in the flowsheet design.

For the split flow flowsheet (Fig 1), a pertinent aspect of relevance to the power failure analysis is the use of live steam sparging into the digester. Following the power failure, a condition can arise whereby the slurry level in the digester falls following the loss of feed liquor and bauxite slurry streams from electrically driven pumps. The steam pressure may then act as a hydraulic ram displacing slurry from the digester vessel through the back pressure station and into the 1st stage flash vessel at a rate faster than the slurry can flow from the 1st flash vessel to the 2nd flash vessel. No practical amount of relief valve area installed on the 1st stage flash vessel will mitigate the overpressure condition. In addition to the driving pressure differential between the steam pressure in the digester and the initial vapour pressure in the 1st flash vessel, the digester may have up to a 20m static head rise over the initial slurry level in the 1st flash tank, further increasing the instantaneous driving force for the transient flow. The back pressure control station is the critical conduit for energy flow between the digester and flash tank train, and therefore its isolation is one avenue to mitigate the relieving requirement. The second avenue is isolation of the live steam supply. Both mitigation measures lie outside the realm of passive overpressure protection, and lend themselves ideally to the utilisation of SIS to protect against this condition. Typically, such isolations have required SIL 3 ratings to satisfy the risk assessment.

The single stream flowsheet in Fig 2 is one layer of inherent safety ahead of its counterpart in Fig 1 as there is no direct steam sparging into the digester vessels. All heating post the recuperative flash stages is via indirect steam heaters. One layer of inherent process safety over and above that in Fig 2 is the Tube digestion flowsheet of Fig 3, whereby both the digester vessel (with its static head rise) and live steam sparging have been eliminated from the flowsheet design.

4. Heater Burst Tube Considerations

The vessel codes ASME VIII and AS1210, require pressure relief for heat exchangers to protect against internal tube failure. This is particularly relevant where the shell side design pressure is significantly less than the tubeside design pressure and the shell side may be fully isolated. These requirements are applicable to both the split flow and single stream flowsheets in Fig's 1 and 2 above.

Passive overpressure protection may be achieved via :

(a) provision of individual PRV's or bursting discs on each heater shell where individual heaters shells may be isolated,

(b) use of the provisions of API521 in selection of the appropriate shell side design pressure to mitigate the provision of PRV's or bursting discs.

Strict adherence to sizing relief valve or bursting disc capacities in accordance with the requirements of (a) above may prove impracticable particularly where there is a high differential between the tube side rating and the shell side rating, as the burst tube analysis (and consequential relief area determination) requires consideration of relieving rates from both tube orifices at the tube break.

This can result in significant relief area to be installed on individual heater shells. Of particular relevance to the recuperative shell and tube heaters are issues of shell side fouling which would require frequent inspections to ensure operability of the installed relief protection and safe discharge of the effluent.

Unlike the use of shell and tube heat exchangers in Fig's 1 and 2, the Jacketed Pipe Heaters of Figure 3 are of more robust construction and do not require the same burst tube provisions. As a result of the use of schedule piping for the inner tubes, this ‘tubing’ is considered no more likely to fail or rupture than any other refinery area piping. This as always, is subject to sound engineering judgment.

5. Heat Exchanger Tube side hydraulic expansion

Hydraulic expansion of the heater tube side fluid will occur when the heater train is isolated at the inlet and outlet manifolds with fluid at an initial temperature of the inlet to the 1st stage heaters. Alternately, fluid in the heaters may have been ‘bottled’ up as a result of a short term operational disturbance and allowed to cool. Subsequent opening of the indirect steam supply and/or recuperative flash vessel vapour will impart sensible heat to thermally expand the isolated tubeside fluid. Rapid increases in system pressures may eventuate from relatively small rises in fluid temperature.

The determination of required passive relief may be simplified by neglecting volumetric expansion of the system piping. Liquid relieving rates will then be a function of :

1. The heat input rate, Q (J/h)
2. The fluid heat capacity, Cp (J/kg.°C), and
3. The fluid coefficient of volumetric expansion at constant pressure, β (°C^{-1}).

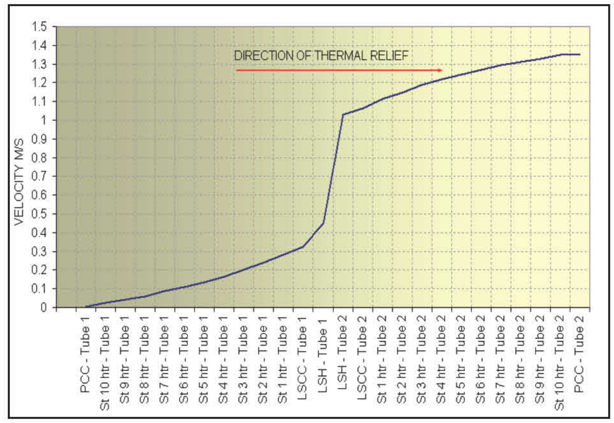
The provision of passive relief for this transient on heater piping associated with flowsheets represented in Fig’s 1, 2 or 3 will suffer from the same limitations as discussed previously in relation to the scaling nature of the process fluid, and its potential to render ineffective, passive overpressure protection without frequent inspection of the devices installed.

The SIS is a viable alternative to passive relief for this condition, particularly if already implemented as part of a broader application for overpressure protection. The same purged pressure sensors utilised at heater outlet manifold piping for protection of heater tubeside ratings may be utilised to isolate live steam energy sources. The SIS will not however restrict inadvertent operation of manual valves which are often employed in the recuperative flash vessel vapour lines as a result of their ability to fully isolate.

The determination of passive relief for the Jacketed Pipe Heaters of Figure 3 is a more complex analysis than its equivalent for the shell and tube heaters as a result of the extended piping runs involved for each heater tube.

To limit transient tube pressures to code allowable, the instantaneous inlet frictional losses (from the expanding fluid) to the relief valve must be added to the relieving pressure. These inlet losses will be dictated by both the number and location of thermal relief valves selected, as well as the thermal relief scenario. Where the fluid thermal expansion results from an accumulation of multiple stage heat inputs, the system frictional losses must be effectively integrated over the piping system and must take into consideration the relative contribution of each piping segment to the overall fluid volume expansion. An integrated velocity profile for one thermal relief scenario is illustrated in Figure 6 below.

Figure 6. Integrated Velocity Profile for Jacketed Pipe Heater Hydraulic Expansion



6. Heat Exchanger Shell side hydraulic expansion

Hydraulic expansion of the heater shell side fluid will occur when the heater train is isolated at the steam or vapour inlet and condensate outlet with cold (ambient) condensate. Alternately the shell side of the heaters may be filled with cold process water during acid cleaning to limit tube side temperatures where isolation of vapour valves is limited. The passage of hot tubeside fluid through heater will then impart sensible heat to thermally expand the isolated shell side fluid. As per the case for tubeside hydraulic expansion, rapid increases in system pressures may eventuate from relatively small rises in fluid temperature.

The determination for passive overpressure protection is generally as described above for the tubeside hydraulic expansion. The configuration of the jacketed pipe heaters of Figure 3 present no special requirements for these analyses, relative to the shell and tube heaters in Fig’s 1 and 2 although the mechanical layout and arrangement of each heater (individual unit or banks) are pertinent considerations for any installation.

For the shell and tube heaters, the heater burst tube provisions (Section 4. above) or requirements of ASME VIII (UG-128) for liquid only relief (minimum ½ inch inlet relief valve) may dictate the shell side passive protection installed, relative to the considerations for shell side hydraulic expansion.

7. Equipment Design for Superheated Steam

The indirect live steam heaters of Fig’s 2 and 3 are generally sized to transfer the latent heat of vapourisation of the supplied steam ($\Delta\hat{H}_v$) to the tubeside fluid. They rely on desuperheating stations either at the upstream boiler station battery limit or locally at the heat exchanger, to provide ideally not more than10-25°C of superheat to ensure optimal use of heat transfer area is made. These desuperheating stations may be required to desuperheat the steam supply by up to 140 degrees celcius. Under such conditions, failure of the desuperheater on equipment not designed for the temperature transient may impose excessive thermal stresses on the heater shell, tubes and tubesheet through significant reduction of material allowable stresses at the elevated steam temperature.

Designing the heat transfer equipment however, for such extremes of pressure and coincident uncontrolled temperature would prove unduly onerous. Here, the heater design may employ the provisions of API Standard 521 concerning double jeopardy. Double jeopardy is the simultaneous occurrence of two or more unrelated causes of overpressure which need not be used as a basis for design.

The failure of the desuperheating station above (resulting in a peak temperature excursion of up to 140˚C) at conditions coincident with the shell side design pressure (typically the set pressure of a pressure relief valve at the desuperheating station) need not be considered if the basis of both events show they are unrelated. In this context, the heater design may be examined for the two alternatives below :

(a) Tube side design pressure/coincident design temperature, Shell side design pressure/coincident design temperature

(b) Tube side design pressure/coincident design temperature, Shell side normal operating pressure/coincident peak (un-desuperheated) temperature.

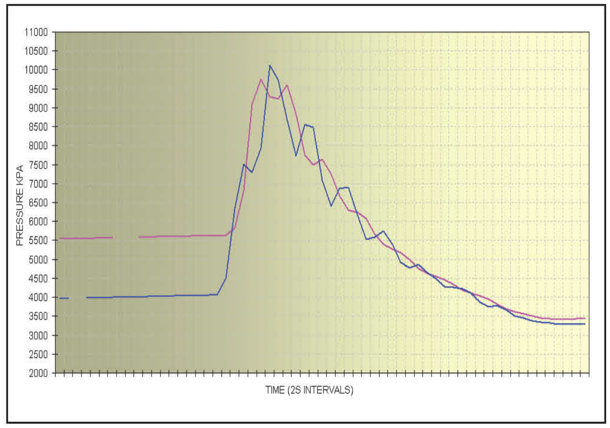
Alternatively, the heater design may be based solely on (a) above and incorporate a SIS to isolate the live steam supply in the event of failure of the desuperheating station/s. In either case, to mitigate the temperature transient, there is no impediment to implementing such isolation as part of the plant control system.

8. Pressure Transients / Water hammer

Pressure transients may take the form of water hammer (hydraulic shock waves), steam hammer or condensation induced hammer (collapsing of trapped steam pockets). All types can manifest themselves with an elevation of piping system internal pressures by multiples of the normal operating system pressure. Where closure of valves is concerned, the general method employed to limit the water hammer pressure wave is to restrict the closure time of control valves to not less than 30s or more. This control however, is not intrinsically failsafe nor self limiting where manual angle valves may be inadvertently closed against online process streams. The flow characteristic of this valve type is in fact well disposed to an almost instantaneous flow stoppage.

Figure 7 below (plotted at 2 second intervals) indicates an example of a digestion piping system transient where the pressure wave invoked as a result of sudden control valve closure peaked at over three times the piping system normal operating pressure. The single stream flowsheets employing dedicated pump station and heater trains (Fig's 2 and 3) may be particularly susceptible to this transient following inadvertent operator actions, as the pressure wave is readily propagated between the pump discharge valving and final downstream isolator.

Figure 7. Pressure Transient in Digestion Unit



The wave reflections between inlet and outlet valve stations are clearly evident in Figure 7, with the time response between pressure peaks (6-7 seconds) being characteristic of the celerity of the wave front of approximately 1,300m/s. The alternating peaks of Figure 7 represent pressure measurements at the pump discharge location (i.e. the start of the hydraulic system), and the back pressure station (i.e. the end of the hydraulic system). A complementary approach to overpressure protection employing the layers outlined below (i.e. neither purely passive or safety instrumented systems), is deemed to offer a satisfactory level of risk mitigation to this transient condition. Note that the response time of PRV's and bursting discs are generally below that required to mitigate the rate of rise of the pressure wave.

1. Administrative controls incorporating locks on all manual isolators, along with strictly supervised operating procedures.
2. Surge arrestors. Where positive displacement pumps are employed the dampener must be sized for both the normal acoustic attenuation and the pressure transient.
3. Control interlocks on manual isolators to initiate pump shutdown during impending incorrect valve operation.
4. Modification of angle valve isolators to incorporate an intrinsic Cv characteristic with a limiting low speed gear ratio to limit the valve closure period.

Conclusions.

This paper has outlined considerations to be incorporated into the overpressure protection design of process plants for various transient conditions. As indicated, the process and mechanical equipment types and flowsheet configurations have intrinsic features that change the response characteristic of the overpressure event. Some equipment and flowsheet configurations demonstrate inherent safety for certain transients, whilst demonstrate susceptibility to others.

An understanding of these considerations makes possible early optimisation of the process flowsheet at the definition stage, to retain the coincident requirements of process facility performance, plant operability and maintainability, and personnel safety.