

MORE FLOW!! What HAZOPs need to know about Flow Assurance and Process Control Dynamics

John Evans, Laurent Gazeaud and John Gill, g3baxi Partnership

The Problem

Surely there cannot be anything new to say about Hazard and Operability Studies (HAZOP). In 2013 the process celebrated its 50th birthday. The format is tried and trusted. Like the Rolling Stones, who started in the same year, it has become part of the establishment.

But can HAZOP say that 50 is the new 40? Can the old dog teach us new tricks?

The origin of this paper was a HAZOP for a debottlenecking study. Perhaps naïvely, we assumed the plant operated at a steady flow rate X and all that was necessary was to check that all the process parameters would cover the case for X + 10%. Unfortunately, the real situation was that the flow fluctuated between X +/- 15%, over a timescale of less than an hour. There were both apparent random and quasi-periodic variations of flow. In these circumstances what certainty or assurance could the HAZOP confidently add? How could we use it to identify potential hazards in the debottlenecking? And how typical is this of real plants everywhere?

This paper therefore seeks to bring to wider attention the occurrence, causes, consequences and assessment of flow instability, based on the experience of both the safety and flow assurance groups of g3baxi partnership. We wish to communicate a process we have developed by which plants can be confidently reviewed despite flow instabilities. We even dare to add some new guidewords for HAZOPs to those in the IChemE's HAZOP: Guide to Best Practice, ref. 1.

Fluctuating Flows are Widespread

The simple step change in constant flow we considered for our plant would indeed be an exception if it happened in reality. Fluctuating flows are actually very common, even when things are working normally.

The puzzle is more why this type of variability is not always recognised by HAZOP studies as an inherent feature of flow. Variation is the focus of much attention when it creates serious issues and may still be a nuisance if it is not. Consequences can include increased wear and tear on equipment, a requirement for greater operator attention, decreasing confidence in alarms and problems recognising incipient dangerous conditions.

In upstream operations, significant flow instability and change in composition are commonplace (see ref. 2). Oil with associated gas will have inherently variable flow. The amount of instability may depend on whether the flow is near the limits of the design (either higher or lower). In downstream operations, feedback loops and instrument response times can create either regular or irregular oscillations. In one case at least, flow instability in a low flow regime resulted in a major insurance loss, but its role as a contributing factor was only discovered after the loss. Records of oscillations with a period of several hours were discovered. These showed that the initial fluctuations started nearly 36 hours before the loss.

Arguably the worst incident caused by unstable flows was, of course, the Chernobyl accident. Although this example seems far away from hydrocarbon processing, the multi-layered complexities will be familiar to many operators of downstream processes involving reactions, even endothermic ones. Chernobyl also illustrates the difference between flow instability and runaway reactions, and how the former can cause the latter. It will be discussed in more detail to show how each of the steps exacerbated the problem.

It is important, then, to be aware of the type of upsets that are implicit in the design, so that the HAZOP is informed about the problems that it may be asked to audit. One would expect information exchange between the HAZOP and the process flow modelling as part of the HAZOP preparation on a project. Equally, detailed discussion of operating histories on an existing plant is essential.

The Causes of Unstable Flow

Unstable flow can arise from many causes. We examine two common areas below, but it is important to recognise that these are only samples to demonstrate the need for links between process flow analysis and HAZOP auditing.

Upstream Operations

Upstream operations have long been recognised as needing to cope with changing flows, arising from:-

- Multiphase flow and slugging;
- Depletion of wells; and
- Change in composition.

A HAZOP should recognise that:-

- depleted wells can give highly unstable flows, which have an impact on control valve lifetimes;
- the response time of valves and the process needs to be matched to limit the effect of flow changes in production lines;
- closed loop control systems can cause upsets if not designed correctly; and
- there are flow conditions along production systems that are inherently unstable (e.g. hydrodynamic or terrain-induced slugging).

The HAZOP needs to cover these issues in addition to the design addressing them because HAZOP is an audit of the design. Thus, we should record the flow regime in which the system is operating, its characteristics and the implications for the process and control systems. By doing so, the HAZOP makes instability visible and a recognised part of the plant dossier.

Despite the best efforts of design and modelling, some process variations will only become apparent once the plant is built. This makes it particularly important that plant revalidation HAZOPs cover the flow stability issue.

Example – Simulation of Pipeline Flows with the Software OLGA

The standard flow regime map for horizontal pipes is presented in Figure 1, the types of flow being:

- 'Stratified' flow: liquid will flow at the bottom of the pipe with gas above it; the liquid surface is smooth at low gas velocities. At higher velocities, the liquid surface is wavy.
- 'Annular' flow: at high gas velocities, most of the liquid forms a film around the circumference of the pipe while the gas and entrained liquid droplets are in the centre.
- 'Bubble' flow: at low gas content the flow becomes a frothy mixture of liquid and gas.
- 'Slug' flow: at moderate gas and liquid velocities, waves appear at the gas/ liquid interface which can completely bridge the pipe and then lead to slug flow.

Figure 1 – Standard Flow Regimes in Pipe Flow

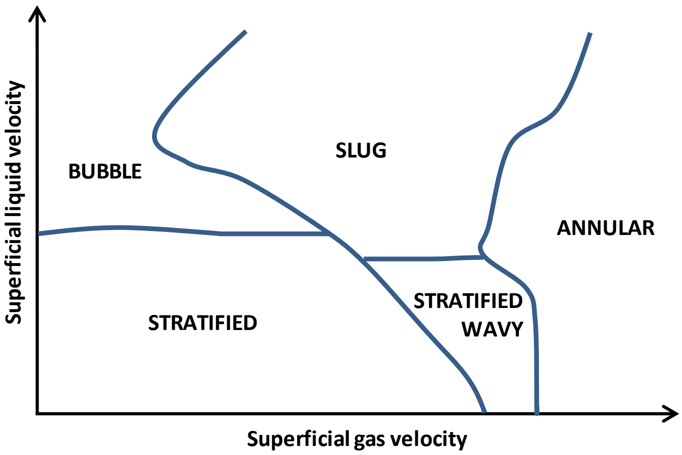


Figure 2 and 3 show slugging for a 5 km subsea tie-back to a fixed offshore platform. The results were generated using the transient thermo-hydraulic software OLGA (ref. 3). There were flow instabilities at normal production rates from day one (hydrodynamic slugging during early field life and potentially severe slugging later in field life at higher water-cuts).

Figures 2 and 3 show the instantaneous gas and total liquid production rates as a function of time at the 1st stage separator during early field life at 100% flow:

Figure 2 – Sample Flow in a Subsea Pipeline

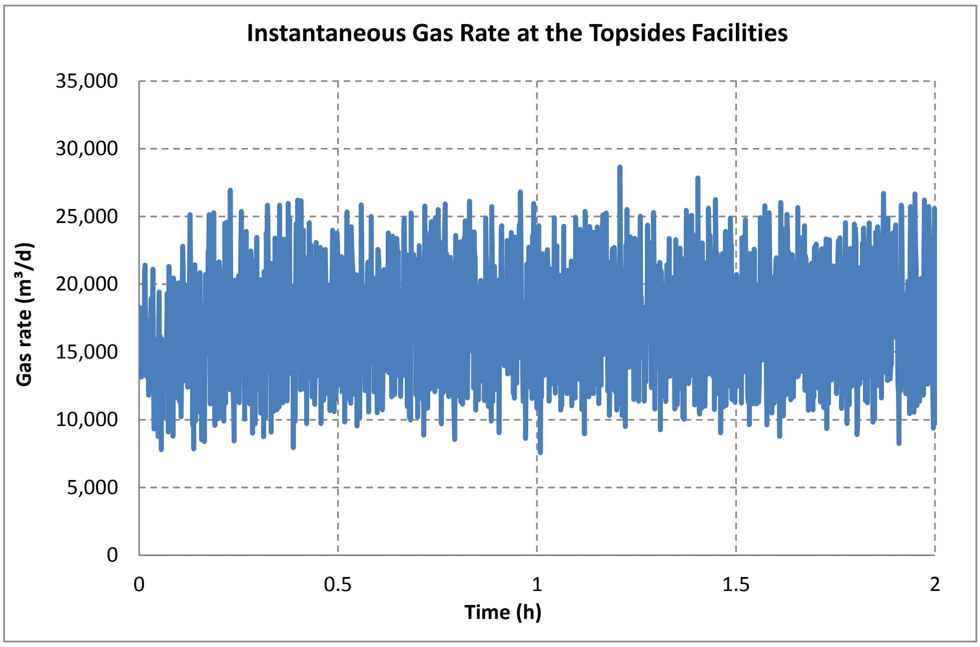
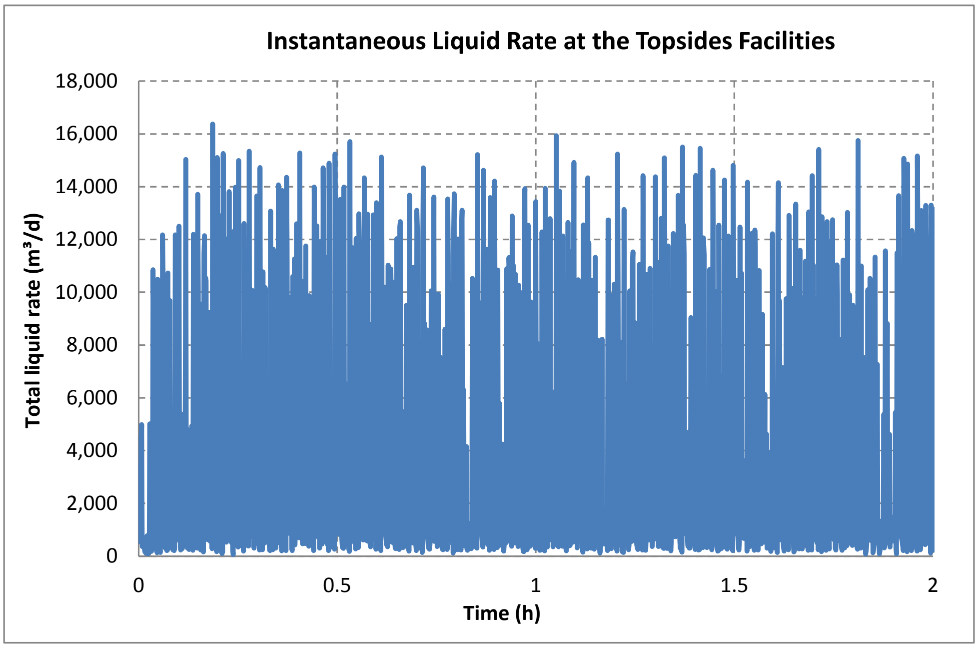


Figure 3 – Liquid Flow in a Subsea Pipeline



Thus, rapid fluctuations in gas and liquid rates were predicted at the topsides facilities, which is typical for hydrodynamic slugging. The usual strategies for reducing instability did not minimise the fluctuations, while any significant increase in backpressure would decrease production and could not be sustained. Such considerations need to be highlighted to the HAZOP team, so that they can audit how the process can cater for such inherent variation.

Matching OLGA predictions with real pipeline data is a very time-consuming activity and discrepancies are always observed, especially in terms of pressure fluctuations along the production system. On the other hand, OLGA gives a good indication of the slugging tendency of a particular production system, so that appropriate mitigation measures can be implemented.

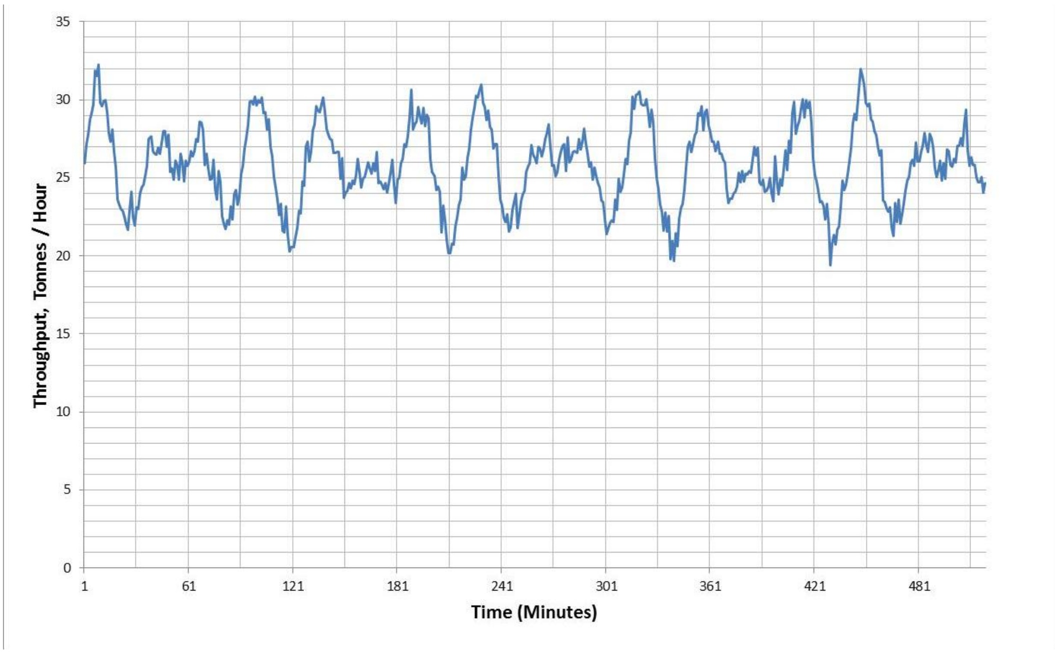
Process Plant Flow Variations in Downstream Assets

Process plant output can also be complex as control systems and thermodynamics act in response to each other. Tuning a process plant is harder than might be thought, as the thermal inertia of events can be on the order of tens of minutes or even hours, whilst fluctuations in pressure and flow and the response times of control valves are usually much quicker. We discuss this in two stages, first focussing on analysing the flow and secondly on attempting to reproduce the response of process control systems.

Assessing Oscillating Flows

The chart below shows the simulated throughput of a section of process plant, based on the experience of a specific process plant that g3 have worked on. The throughput is given as 25 tonnes per hour, however, the range varies from 19.3 to 32.2 over a period of several hours.

Figure 4 – Simulated Output from a Process Plant

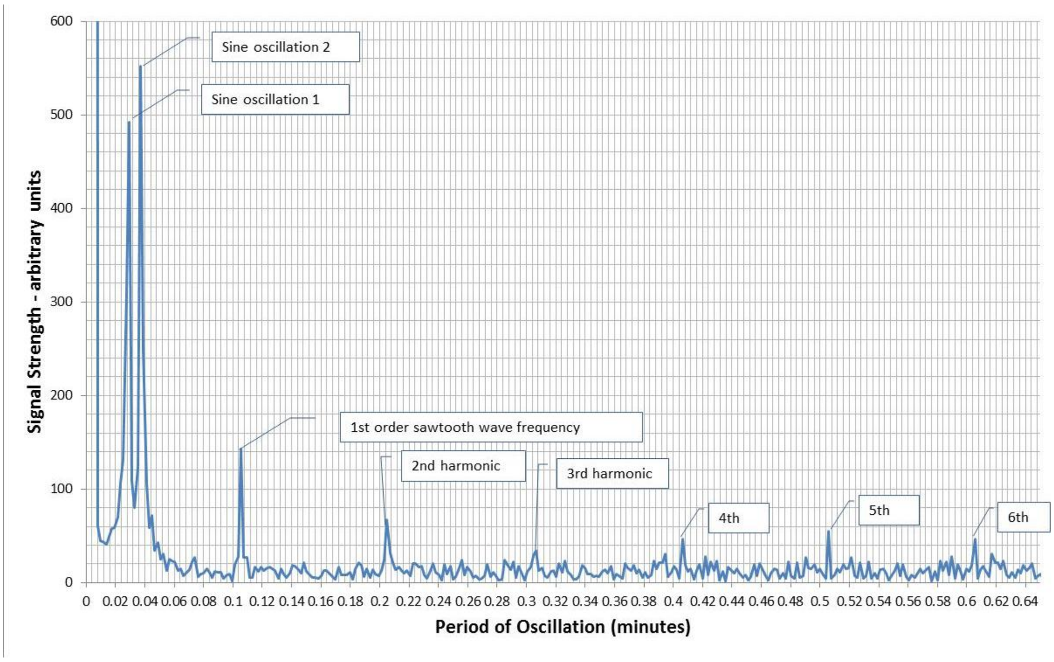


How can we demonstrate that it is safe to operate at a mean throughput of 30 tonnes per hour. Questions to be answered include:-

- if the mean throughput was now to be 30 tonnes per hour, what would the maximum throughput be, given that it is likely still to be variable (and is such a high flow safe?);
- how would we be able to characterise the flow? Would the variability change at a higher flow?
- What values of flow should now be used at the process trip set points?
- What can the variation tell us about its causes?
- Isn't the process shown to be safe at 30 tonnes per hour? Nothing happened when it reached 32.2 tph.

In the above case, the oscillations used to generate the variability can be extracted using Fourier Analysis. The Fourier Analysis showed the contribution from the 3 oscillations that produced the flow variation. On inspection of the above one might guess there were two waves of different frequency that were combining to produce a "beat" effect, but the analysis clearly shows a 3rd saw tooth wave, whose frequency components can be distinguished to quite high harmonic levels. In real process data, such saw tooth behaviour is indicative of "sticky" control systems that can be a driver for other oscillations at different frequencies. At a lower level is the white noise of random oscillation.

Figure 5 – Fourier Analysis of Simulated Data



Such oscillations are common in operating plant. In a separate facility to that reviewed above, changing the valve packing resulted in a sticky valve and degradation of control. Operation changed from an automatic state to essentially a requirement for manual control by the operator. The length of the oscillation cycles so produced was of the order of hours, with the effect that the operators did not even realise that feedback was occurring. As a shift change occurred, the seriousness of the problem was not appreciated from one team to the next. (There was even a technical specialist from the process licensing company permanently allocated to the site.) In the end serious damage was caused to the plant and although some spares were available there was still a high cost from business interruption.

Thus the impact here on the HAZOP is that there needs to be discussion of the causes of the oscillation. Do they come from the control system or other upstream causes? What are the impacts in terms of wear on valves and other equipment and what might the impact on availability be? Are operators having to put the system in manual, ignore alarms or put trips on bypass simply to operate? Much more can be appreciated about the operation of the plant by including the actual process flows in the HAZOP. It should be noted that such instabilities may be more prominent in turndown conditions than in normal operation.

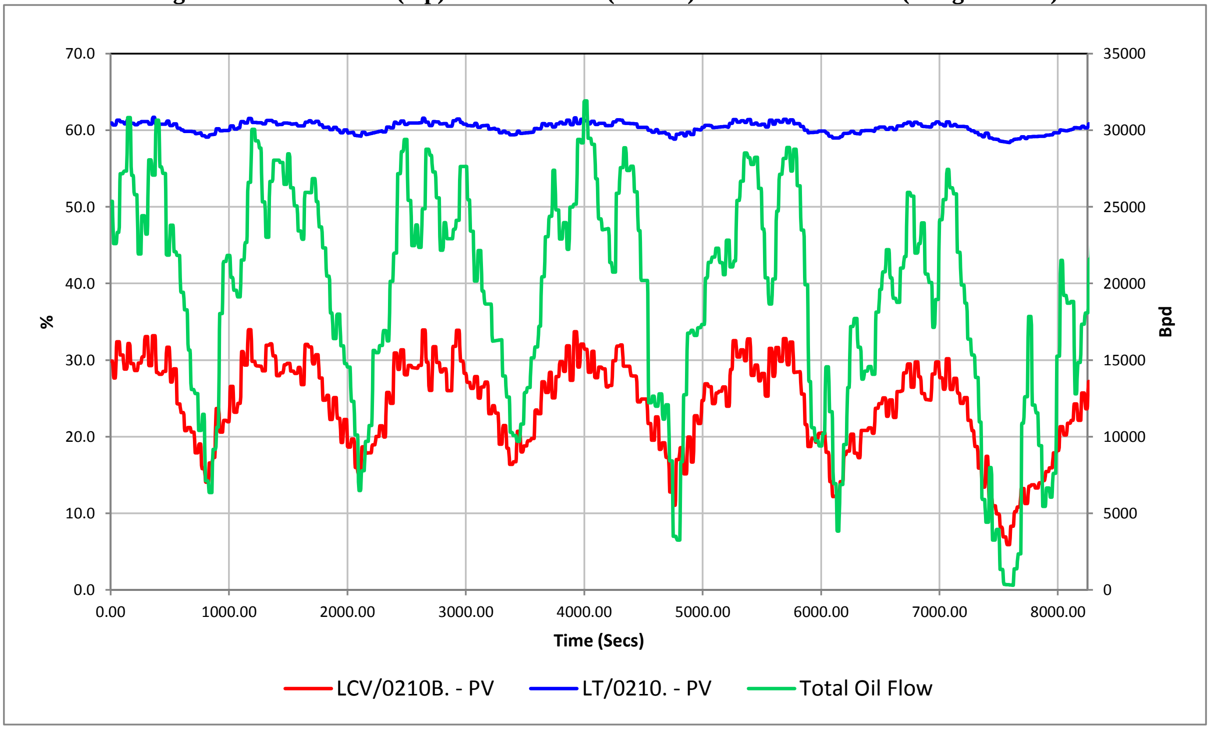
Modelling Process Plant Flow Instability

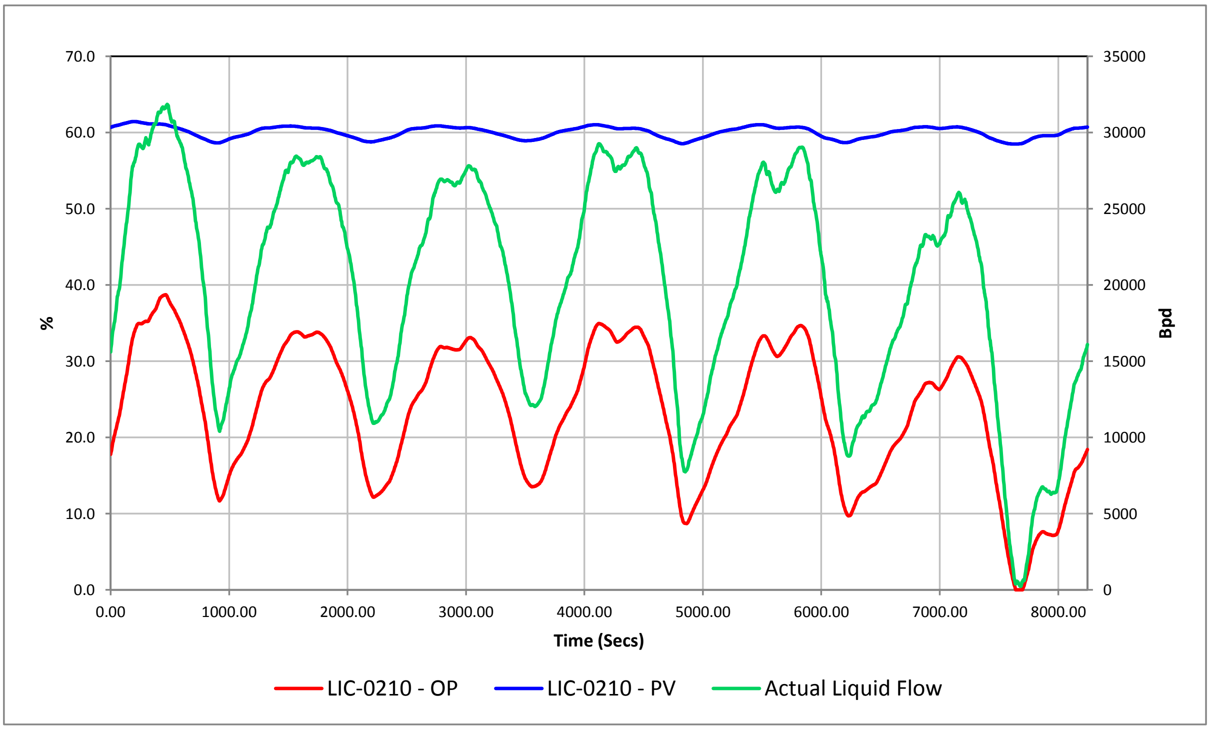
Process control feedback is another source of unstable flow that needs to be picked up for the HAZOP to have a realistic view of the characteristic of the process control system. In this case we seek information about how the process behaves (with and without disturbance). What information the operator sees (to instigate manual intervention) is another important topic.

The software HYSYS Dynamic (Ref. 4) can model the variation in process plant behaviour, giving results that are reasonably representative of the real behaviour of an operational plant, as shown below in Figures 6 a and b.

In a more complete analysis of an operational plant, additional information, such as valve positions and other process variables, would also be available via data historians. These would give a more complete picture of the behaviour of the plant and can be matched with models built in HYSYS dynamic for example.

Figure 6 a and b – Real (top) and Simulated (bottom) Plant Performance (using HYSYS)





For a plant in design, the challenge is obviously to ensure that the modelled behaviour matches what will result when the plant is built. Even so, the use of HYSYS dynamic gives reasonable confidence that response times and feedback behaviour are being captured.

Once again, the ability to interrogate the behaviour of the real plant has benefits, and this time the feedback is from the HAZOP into the model for assessing the plant. HYSYS dynamic can be used to identify the cause of the flows brought to the HAZOP meeting.

The Role of Unstable Flow in the Chernobyl Accident

Whilst discussing a nuclear accident may seem to be a departure from the examples above in the hydrocarbon industry, it could not, of course, be more relevant to the topic of this paper.

It is often thought that the Chernobyl Accident was the result of a runaway nuclear reaction – and in terms of the seconds immediately preceding the power excursion that is undoubtedly true. The prompt criticality led to the number 4 reactor generating power at a rate roughly half the UK electricity demand, about 33,000 MW. The entire upper structure of the reactor, weighing about 2,000 tonnes, was ejected. Reviewing the hour and a half previous to the explosion, however, one finds the interplay of the at least 4 reaction rate controlling influences, all of which have complex effects:-

- The control rods – these reduce the nuclear reaction rate by absorbing neutrons, however, the tips consist of graphite, whose neutron moderating (slowing) effect actually enhances the reaction. The insertion of the control rods to a critical reactor led to the prompt criticality;
- When the control rods were first inserted an hour and a half previously, they led to "Xenon poisoning" a nuclear decay product whose effect was to reduce the reaction rate significantly. Usually on a trip one has to wait several hours before being able to restart safely as this effect seriously upsets a possible operating regime;
- Steam take off – the experiment was all about investigating the potential to use a running down turbine to power some pumps, however, in shutting off the steam to the turbine the water in the reactor began to boil and the voids created increased the nuclear reaction rate (due to the "positive void coefficient of the RBMK design);
- Feed water flow. Increased feed water flow reduces reactor power by absorbing neutrons and also cools the reactor.

The experiment had been tried 3 times before and failed each time. There were also significant management failures, the most critical of which was possibly not informing the onsite Nuclear Oversight Regulator representative that the test was taking place, although the Emergency Core Cooling System, neutron alarms, SCRAM button and other alarms were also disabled. Fundamentally, however, this accident was one of failing to understand the interaction of controls available to the operator and their effect on the reactor state. Essentially the reactor had moved from a regime where the control rods controlled the reaction rate, which was normal at the power rates the test was intended to be conducted at (>700 MW output), to a state where control rods were fully withdrawn and the operators were using feed pumps to balance the output, with Xenon poisoning "functioning" as the control rods at the much lower power of 160 MW. The sequence of events is given in Table 1 and is taken from ref. 6.

Table 1 – Sequence of Events at Chernobyl RBMK Reactor 4, 25-26th Aprill 1986

Date	Time	Shift	Power	Action	Procedure	Nuclear Oversight Regulator Representative	ECCS	Neutron Alarms	Water levels in separator	SCRAM
24-Apr-86		Day			Yes. Lower safe limit of power is 700 MW	Not informed				
25-Apr-86		Day	1600MW		Yes	Not informed	Disabled			
25-Apr-86		Day / Evening	3200MW	Test Halted		Not informed	Disabled			
25-Apr-86	23:04	Night	<1600MW	during shift changeover	Not trained	Not informed	Disabled			
26-Apr-86	00:05	Night	700 MW		Not trained	Not informed	Disabled			
		Night	500 MW	Xenon poisoning - control rods inserted too far	Not trained	Not informed	Disabled			
		Night	30 MW	Control rods withdrawn	Not trained	Not informed	Disabled			
		Night	160MW	More Xe poisoning - more control rods withdrawn	Not trained	Not informed	Disabled			
26-Apr-86	00:35 - 00:45	Night	160MW	Unstable flows in core and of feed water, relief valves	Not trained	Not informed	Disabled	Alert - disabled	Alert - disabled	
	01:05	Night	160MW	Extra water pumps activated. More boiling in core	Not trained	Not informed	Disabled	Alert - disabled	Alert - disabled	
	01:19	Night	<160 MW	Excess water reduces reactor power. More control rods withdrawn. Loss of steam pressure to drive turbines	Not trained	Not informed	Disabled	Alert - disabled	Alert - disabled	Disabled
	01:23	Night	<160 MW	Experiment begins. Steam to turbines shut off - power to pumps decreased, steam formed stays in reactor, positive void coefficient.	Not trained	Not informed	Disabled	Alert - disabled	Alert - disabled	Disabled
	01:23	Night	33,000 MW	40 seconds later shutdown triggered but tips of control rods are graphite, which causes prompt criticality	Not trained	Not informed	Disabled	Alert - disabled	Alert - disabled	Manual activation

It is interesting to ask how a HAZOP would have identified the dangerous elements of this situation. We would need to know that certain power regimes are more stable and allow control by one device (the control rods). We would also need to know that operating below a certain "turndown" would require a complete shutdown. As noted above, the engineers who proposed the test had required a minimum power load of 700 MW for stability essential to success. Would a HAZOP simply have stated that below 700 MW the reactor would be shutdown and operation would not have been countenanced?

Other Process Flow Lessons

Two further examples show the need for the HAZOP to learn from how designs are operated in response to flow changes:-

- A client had installed a surge drum to cope with excess propane produced. As the surges were more common than anticipated, the high level alarm in the drum was isolated by the operator, eroding a vital safety margin.

- An analysis of problems in an existing plant was able to simulate the information gained from a virtual temperature indicator that did not exist on the real plant. It was found that the information this transmitter gave solved a key interpretation issue for the operator. A temperature transmitter was added in due course.

In the first case, the designation of a drum as a "surge" drum should have led to the appreciation that operators might find it somewhat bothersome. In the second, it illustrates the value of a link between process control and situation diagnosis; a wide topic to which this paper can only allude. Ref. 5 gives more background.

Recommendations

Presenting flow assurance results in a HAZOP to inform the team, even at a high level, could require more time and cost, but the information is valuable and highly pertinent and fulfil the HAZOPs requirement to audit the design. It would help prevent future loss of production, equipment damage and personnel hazard.

Equally it is useful a HAZOP of live plant references and documents lessons learnt from operations. It needs to recognise what is happening, how operators are coping and where extra information may be needed. For such reasons, HAZOPs, in particular revalidation HAZOPs, are much more complex than just checking that the plant is physically the same as the one that the previous HAZOP is based on. Existing results may not be applicable.

In addition, it is worth noting that process instability has a similar impact on Layer of Protection Analysis (LOPA) studies. As flow variation is a driver of the spurious trip rate it may also change the results of these studies. The impact of such trips on the availability assessments should also be considered.

Identifying and Recording Flow Instability in HAZOPs

From the above we conclude a HAZOP should document the degree to which flow is inherently prone to instability and / or fluctuating.

The process engineers therefore should come to the meeting with a more extensive design dossier, in fact much more work than would currently be expected. It is not easy to identify from a P&ID or even a process control scheme issues such as:-

- Change in flow control regimes;
- Response times of process controllers;
- Non-linearities and instabilities.

Issues to be included are:-

- The subtleties of operating, is the response time of instruments an issue, what type of feedback is expected;
- How constant will the flow be even if there are no upsets;
- How frequent would the demand on the safety system be (which is a necessary input to Safety Integrity Level (SIL) / LOPA rating);
- The consequences in terms of loss of availability (and some plants like LNG liquefaction systems incur a significant down time penalty for long shutdowns).

Without an assessment of operability the HAZOP is incomplete. It also leaves other project deliverables, such as the availability assessment, with a questionable basis. Such assessments are usually model failure of components, rather than the occurrence of spurious trips (which would arise from flow variation).

The New HAZOP Guidewords

For the cases where the process is in design the HAZOP should check:-

- What flow regimes create particular concerns? Which are unstable?
- What are design flow margins during operational life?
- Are there any through life issues with flow, due to fluid composition, flow at particular temperatures, scaling, coking, sludges, gas oil ratio, gas lift?
- Are there anomalous cases in modelling of flow assurance?
- Anticipated trip rates?

For the case where the process is in operation the following words would promote inquiry:-

- Are there any response time tuning issues within the process that are the source of control upsets?
- Are there any end of life issues for catalyst / coking / scaling / sludges / heat transfer that produce a throughput response?

- What operational cases are there?
- Current feedstock, in comparison with design, effects of changes;
- Operations in comparison with design intent;
- Utilisation of manual control for any reason;
- Interrogation of records;
- Instrument deficiencies, alarms / trips disabled;
- Effects of alarm disablement on plant bow ties;
- Flow stability – what is the maximum and minimum steady and transient flow;
- Flow stability – what compositional changes have a major effect on throughput;
- Positive flow feedback within the process;
- Adherence to design procedures for start up; and
- Records of trips.

One may well ask “what of the other guidewords?”, such as pressure temperature and so on. Does this paper’s focus on flow mean that there is, in the mind of the authors, no need to augment “pressure, temperature, etc.”? Of course not, by focussing on flow in depth we wish to stimulate equal inquiry into other areas. Further, the aim of the whole endeavour is to broaden the HAZOP remit so that it functions to retain more of the design rationale for the benefit of users in the future.

This paper argues that the modelling and documenting of the time dependent throughput of process systems should be available for all characterising conditions that the system is expected or required to operate under. Flow assurance and HAZOP together then give the designer and operator a sound assessment of what may occur and whether the design is, or remains, fit for purpose.

Further the performance of control equipment against upsets should also be documented in these regimes, in order to demonstrate that the control systems will work as intended and that there are no uncontrolled positive feedback loops.

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