



## HIGH INTEGRITY PROTECTION SYSTEMS (HIPS) FOR FLARE LOAD MITIGATION

Angela E. Summers, Ph.D., P.E., President, SIS-TECH Solutions, LP

"Flare Reduction with High Integrity Protection Systems (HIPS)," ISA EXPO 2003, Research Triangle Park, NC: Instrumentation, Systems, and Automation Society, Houston, Texas, October 20-23, 2003.

The American Petroleum Institute (API) and American Society of Mechanical Engineers (ASME) provide criteria for the protection of vessels and pipelines from excess pressure. In conventional design, a Pressure Relief Valve (PRV) is used as the primary means of protection, and a flare is used to safely combust the gases relieved during an overpressure event. Although conventional, the use of a PRV is sometimes an unattractive proposition, particularly where the pressure relief involves a large flare load.

API 521 and Code Case 2211 of ASME Section VIII, Division 1 and 2 allow the use of an SIS in lieu of a PRV as long as the SIS meets or exceeds the protection that would have been provided by the PRV. As an SIS, the design must follow the safety lifecycle provided in the United States standard ANSI/ISA 84.01-1996 or the international standard IEC 61511. The required risk reduction results in the need for high SIS safety availability; therefore, these systems are often called High Integrity Protection Systems (HIPS).

### Code Requirements

Until August 1996, ASME required the installation of pressure relief devices on all pressure vessels for compliance with Section VIII, Division 1, para UG-125(a) Section VIII, Division 2, para, AR-100. With the approval of ASME Code Case 2211, ASME acknowledged that overpressure protection should be provided by the most appropriate engineered option. The Code Case is intended to enhance the overall safety and environmental performance of a facility. It is not intended as a blanket approval of the operation of pressure vessels without pressure relief valves. While no specific performance criteria are included in Code Case 2211, the use of HIPS must result in an installation as safe or safer than conventional design.

HIPS can be used to provide overpressure protection if a quantitative or qualitative risk analysis of the proposed system is made addressing credible overpressure scenarios. It must be demonstrated that the proposed system is independent of the potential causes for overpressure; is as reliable as the pressure relief device it replaces; and is capable of completely mitigating the overpressure event.



12621 Featherwood Drive • Suite 120 • Houston, Texas 77034  
Tel: (281) 922-8324 • Fax: (281) 922-4362  
[www.SIS-Tech.com](http://www.SIS-Tech.com)



## Recommended Practices

API recommends a number of practices for addressing pressure relieving systems in the petroleum production industry. For example, API 521 provides assessment and design methods for determining relief loads based on a credible overpressure scenarios. These relief loads are used for individual vessel relief valve sizing and for main flare header sizing that includes the simultaneous venting of all affected vessels. The fourth edition of API 521 allows credit for a favorable response of the instrumented systems. Despite API 521 permitting design alternatives, API 521 Part 2.2 recommends the use of HIPS only when the use of a pressure relief device is *impractical*.

## HIPS Justification

Successful implementation requires examination of applicable regulations and standards, including local codes and insurer requirements that may mandate the use of pressure relief devices. The justification must be based on a hazard analysis of each potential overpressure scenario, following a structured, systematic approach, using a multidisciplinary team. It should document the event propagation from the initiating cause to the final consequence (also referred to as the "overpressure scenario"). The analysis must examine operating (e.g. start-up, shutdown, and normal operation) and upset conditions that result in overpressure. The "Causes of Overpressure" provided in API Recommended Practice 521 should be reviewed to ensure that the analysis is complete.

In the hazard analysis, the process risk is evaluated in terms of frequency and consequence. The mitigated risk is determined by assessing the initiating cause frequency and risk reduction provided by any independent protection layers, such as HIPS. If the mitigated risk is less than the risk tolerance criteria, the scenario is a candidate for removal from the relief device and flare loading calculations.

## Safety Requirement Specification

A safety requirement specification (SRS) describes how and under what conditions the SIS will mitigate each overpressure scenario, including a functional logic description with trip set points and device fail-safe state. Only those scenarios that can be successfully mitigated by the SIS can be *considered* for removal from the flare loading calculations. Careful consideration must be given when removing a scenario, because if the assumptions used to justify its removal are incorrect, the vessel or pipeline may be under-protected. For example, in hydrocarbon applications, the fire case scenario cannot be removed from the sizing calculations due to the inability of HIPS to mitigate the cause of overpressure.

The process dynamics must be evaluated to ensure that the HIPS response time is fast enough to prevent overpressure of the vessel or pipeline. The response time must be evaluated by considering the time it takes to



sense that there is an unacceptable process condition; the scan rate and data processing time of the logic solver; and the time required for the final element action to influence the system pressure. The required closure speed must be determined for each installation. The valve specification must include the acceptable leakage rate, since this affects downstream pressure and relief loading. The valve actuator must provide sufficient driving force to close the final element under the worse case, upset pressure condition.

The SRS also includes documentation of the safety integrity requirements, including the Safety Integrity Level (SIL) and anticipated testing interval. At a minimum, the target SIL for the HIPS should be equivalent to the performance of a pressure relief device. Reliability information for a single-valve relief system is provided in "Guidelines for Process Equipment Reliability Data" by the Center for Chemical Process Safety. Table 1 provides a summary of this data, which shows substantial uncertainty in the failure to open on demand. Additionally, the failure modes of the PRV and the SIS are different. A PRV that fails to operate at the set pressure may nevertheless operate at a higher pressure, whereas an SIS is more likely to fail completely. The failure to open on demand uncertainty coupled with the difference in the failure modes results in most users setting an SIL 3 target for the SIS.

Table 1. Pressure Relief Device Failure to Open on Demand

Pressure Relief Device Type	Failure to Open on Demand		
	Lower	Mean	Upper
Spring Operated	7.9E-06	2.12E-04	7.98E-04
Pilot Operated	9.32E-06	4.15E-03	1.82E-02

#### **"As Safe or Safer" Verification**

The HIPPS must provide an installation that is as safe or safer than the pressure relief device that it replaces. In order to document that this has been achieved, the HIPS design should be quantitatively verified to ensure it meets the required PFDavg. HIPS are SIL 3 SIS and are often the only layer of protection against the overpressure event. Consequently, many users require that an independent, third party perform the evaluation of the appropriateness of the design and the determination of the PFDavg.

#### **HIPS Example**

Plant personnel were working on a major unit expansion that involved debottlenecking one of the process units. The initial design showed favorable capital cost relative to the increased plant production. A Layer of Protection Analysis (LOPA) was performed to identify various overpressure scenarios associated with the proposed expansion. One scenario involved a pipeline supplying gas to the process unit. The pipeline had a pipe specification change from



750-psig to 150-psig. During normal operation, the process pressure was reduced using a pressure control valve. If this pressure control valve failed, the downstream process pressure exceeded the 150-psig pipe specification, resulting in a potential release of a large volume of gas within the process unit. To prevent pipeline rupture, a Pressure Relief Valve (PRV) was located downstream of the pressure control valve. The PRV was designed to safely relieve the pressure to a flare system.

Flare loading was then calculated and a serious problem became immediately apparent. The flare system could not handle the load of the expanded process unit. The largest load was determined to be the pipeline PRV. The cost of expanding the flare system was very high, since the flare header and flare would need to be replaced.

The team began evaluating various options. One option was to relieve the pipeline pressure to the atmosphere rather than to the flare system. Unfortunately, the PRV was located too close to the process unit for safe relief of the large gas volume. The team considered relocating the relief point to reduce the safety risk, but the environmental department vetoed this approach. The release would result in excessive emissions and potential off-site impact. The team then examined the use of a HIPS and determined that the cost was favorable. Consequently, the decision was made to install a HIPS in order to remove the pipeline relief scenario from the flare load case.

LOPA had determined that there were no SISs assigned an SIL higher than SIL 1. The HIPS required SIL 3. Based on these results, the design team decided to utilize relays for the HIPS and a less costly PES for the other SISs. Further, the team reasoned that the use of the separate relay system would eliminate potential common cause faults between the main unit SIS and the HIPS. By placing each HIPS in a separate relay system, the design was mimicking the separation provided by the independent pressure relief device.

The HIPS design initially incorporated dual (1oo2) pressure transmitters, triplicated relays (packaged SIS-HIPS), and dual block valves. Transmitter diagnostics were provided by sharing the analog signal between the relay system and the Basic Process Control System (BPCS). The BPCS was then used to generate a deviation alarm. Each block valve was actuated using a simplex solenoid. Figure 1 shows a simplified drawing of the initial design.

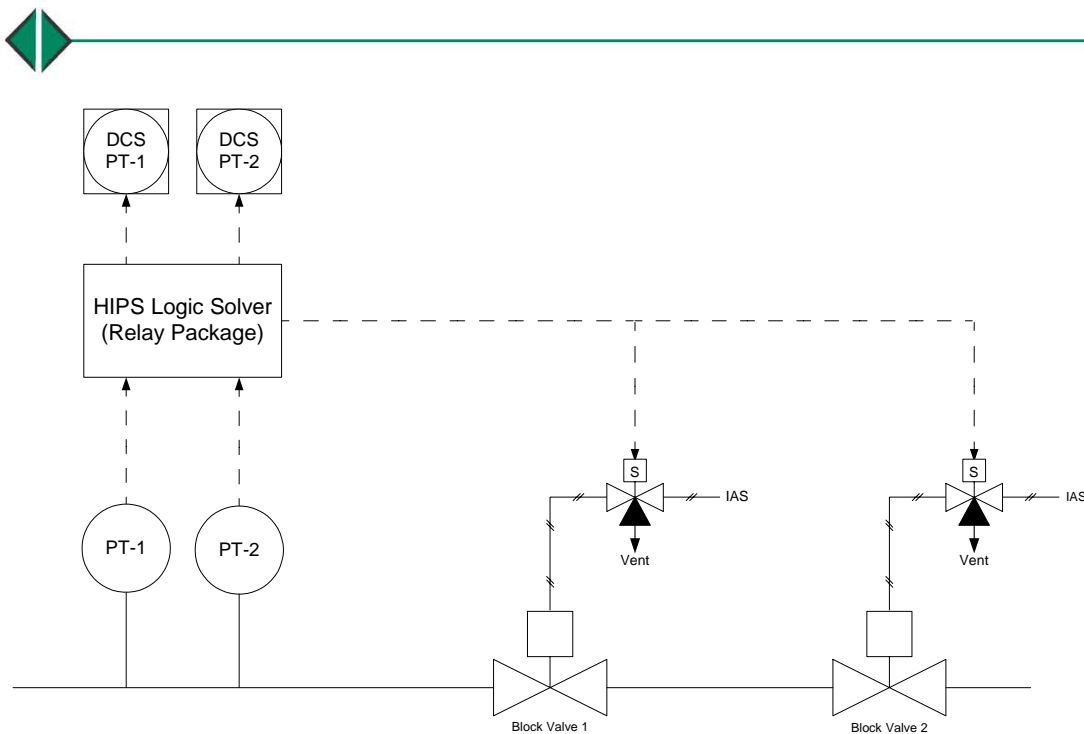


Figure 1. The initial design for the HIPS

This architecture required a one-year testing interval to achieve a  $PFD_{avg}$  in the SIL 3 range. Since the refinery unit was on a five-year turnaround schedule, provisions for on-line testing had to be provided. This included the installation of a bypass line and bypass valve (car sealed closed) and a bypass for each transmitter input signal. Further, this architecture had a predicted mean time to failure spurious ( $MTTF_{spurious}$ ) of 7 years. This  $MTTF_{spurious}$  was considered unacceptably high by operations management due to the potential production losses.

Consequently, the architecture was changed as follows:

- 2003 pressure transmitters
- Triplicated relays (packaged SIS-HIPS)
- ASCO RCS, normally closed version, used for valve actuation and partial stroke testing
- Double block valves



The revised architecture is shown in Figure 2.

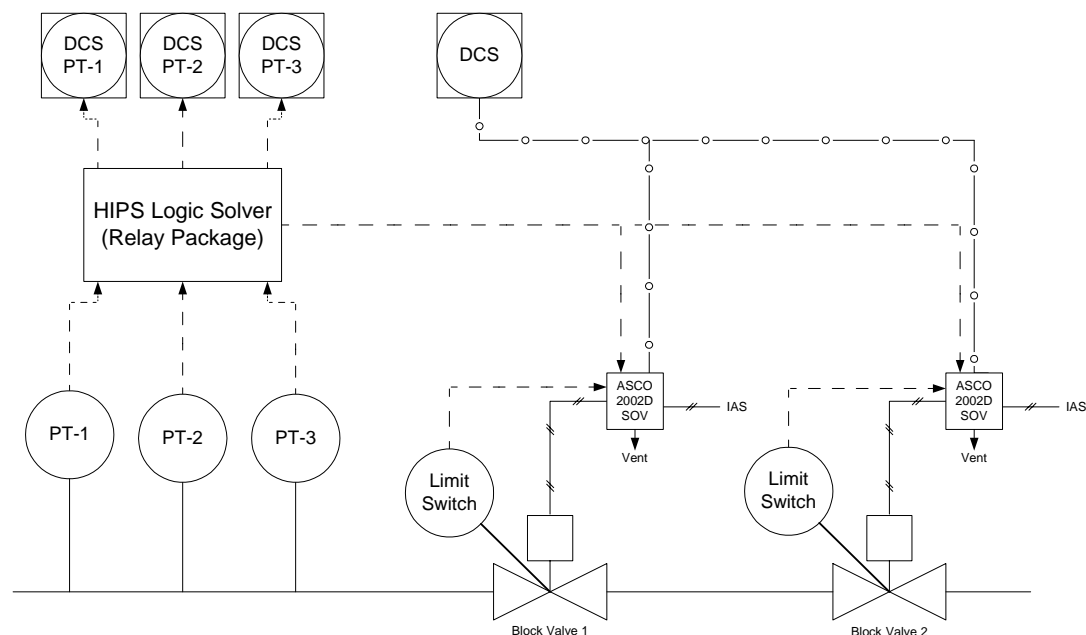


Figure 2. Final design for the pipeline HIPS

Transmitter diagnostics were still provided by generating a deviation alarm in the BPCS. The additional transmitter provided an architecture that could be tested without bypass. The use of the ASCO RCS provided for monthly, automatic testing of the solenoids and partial stroke testing of each block valve without bypass.

The revised architecture achieves a  $PFD_{AVG}$  in the SIL 3 range with annual testing of the pressure transmitters, monthly solenoid testing, monthly partial stroke testing of the block valves, and five-year full stroke testing of the block valves. Since full-stroke testing was extended to unit turnaround, full flow bypass lines and valves were eliminated, yielding substantial reduction in capital cost. The  $MTTF^{spurious}$  was also improved to more than 200 years, significantly reducing the potential of a spurious trip.

### Advantages and Disadvantages of HIPS

HIPS are becoming the option of choice to alleviate the need to replace major portions of the flare system in existing facilities when adding new equipment or units. If the header and flare system must be enlarged, significant downtime is incurred for all of the units that discharge to that header. The capital and installation cost associated with HIPS is attractive when compared to the downtime or equipment cost of flare modification. Another benefit is that the



process unit will not flare as much as a process unit designed for full flare loading. In some areas of the world, this is becoming important as regulatory agencies place greater restrictions on flaring of process gases.

The main disadvantage of HIPS is the careful documentation, design, operation, maintenance, and testing to ensure standard's compliance. Specific regulatory and enforcement jurisdiction requirements must be determined. In some instances, approval of local authorities is required. Regulatory and standards requirements must be understood by all parties, including facility management and instrumentation and electrical, operations, and maintenance personnel.

Any justification for HIPS must be thoroughly documented through a hazard analysis, which identifies all potential overpressure scenarios and demonstrates that the HIPS can adequately address each scenario. The user must verify that HIPS will work from a process standpoint and that the HIPS design results in an installation as safe or safer than a conventional design. The effectiveness of the system is highly dependent on the field design, device testing, and maintenance program. When a pressure relief device is not installed or is undersized based on conventional design, the HIPS becomes the "last line of defense," whose failure potentially results in vessel or pipeline rupture.



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