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Assessment of Safety Instrumented Functions for Delayed Coking Unit Heater by Using LOPA and Risk Graph Approaches

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Abstract

This paper presents a systematic approach for assessment of Safety Instrumented Functions (SIFs) in the heater of a Delayed Coking Unit (DCU). The fired heater of DCU accommodates many safety instrumented functions to prevent any hazard scenarios. Safety performance criteria for SIFs should be defined by Safety Integrity Levels (SILs), To define the SILs of those safety instrumented functions, two techniques are used in this research work; Layer of Protection Analysis (LOPA) and Risk Graph approaches for the total number of 6 SIFs installed on the fired heater. By comparing the results of these two techniques, it is found that SIF 1 and SIF 2 by using the LOPA technique give SIL rating of SIL 2 while in the case of the Risk Graph technique, the SIL rating for the same SIF is reduced to SIL 1 without changing the required safety measures. Such a reduction in SIL rating has a great impact on the total cost reduction of the unit and reduces design, installation, operation, and maintenance complexity.

Keywords: *Delayed coking unit; Safety integrity level; Safety instrumented function; Layer of protection analysis; Probability of failure on demand.*

1. Introduction

In recent years, Delayed Coking Units (DCU) have experienced several series of accidents despite efforts between many refineries to share the best practice information related to DCU safety and reliability. DCU provides a difficult but increasingly important function for the refiner. The DCU, unlike other petroleum refinery process operations, is a semi-batch type operation. That is, one part of the process is a batch-type operation while the remaining portion is a continuous operation. It's the batch portion of the operation (drum switching and coke cutting) that causes unique hazards not only for the batch section but also for the remaining continuous portion of the operation (fractionation and drum charge heating).

The purpose of this unit is to process the vacuum residue (feed) into refinery intermediate products. The liquid products including heavy coker gas oil (HCGO), light coker gas oil (LCGO), stabilized naphtha and LPG are further processed in the downstream units into transportation fuels. The coker fuel gas is treated in the unsaturated off gas amine absorber and used as refinery fuel gas. Fuel grade coke from the DCU is sold primarily as fuel for power generation [1].

The drum switching frequency typically ranges between 10 to 24 hours. While one drum is filling the alternative drum is cooled by steam and water, opened, un-head, and de-coked. Finally, the offline drum is closed (re-headed), purged by steam for air free, leak test, warmed up, and placed on standby, ready to repeat the cycle. The batch portion of DCU drum switching and coke cutting operations creates unique hazards especially for the continuous portion (fractionator and heater) in the plant, resulting in relatively frequent and serious accidents.

Shutdown systems are traditionally recognized as safety systems that contribute to reducing the likelihood and consequences of dangers to personnel, environment, and assets. Therefore, safety instrumented functions need to be managed through a systematic assessment process to determine any requirement for increased reliability and/ or higher integrity, hence reducing risks. Safety Integrity Level (SIL) study is conducted to perform a systematic review of Heater process systems to identify failures in safety related control systems, which have the potential for harm to personnel, environment, and assets.

The SIL study should be scheduled after completion of the Hazard and Operability study (HAZOP) and incorporation of major HAZOP recommendations onto The Piping & Instrumentation Diagrams (P&IDs) and Cause & Effects. The P&IDs, Cause and Effects, HAZOP Report, Qualitative Risk Assessment (QRA) Reports, and Plot plans shall be available before the SIL assessment [2]. SIL Methodology as defined in the International Standard IEC 61511 [3], is a widely used safety performance measure for safety instrumented functions. A Safety Instrumented Function (SIF) is a safety protective function implemented by a Safety Instrumented System (SIS), and composed of any combination of sensors, logic solver, and final elements (e.g., valves) where A SIF must achieve a specific level of integrity.

For each of the safety instrumented functions (SIFs) operating in demand mode, the required SIL shall be specified according to its probability of failure on demand as stated in IEC 61511 and shown in Table 1 [3].

Table 1. Target average probability of failure on demand for SILs.

Safety integrity level (SIL)	Target average probability of failure on demand
SIL 4	$\geq 10^{-5}$ and $< 10^{-4}$
SIL 3	$\geq 10^{-4}$ and $< 10^{-3}$
SIL 2	$\geq 10^{-3}$ and $< 10^{-2}$
SIL 1	$\geq 10^{-2}$ and $< 10^{-1}$

The International Electrotechnical Commission IEC 61511 standard suggests several methods for SIL determination, ranging from fully quantitative to fully qualitative methods. The widely used techniques in the oil and gas industry for Assessment of Safety Instrumented Functions (SIF) are LOPA technique and Risk Graph technique. Both these methods are included in the IEC61508 [2] and IEC61511 [3] international standards and each of these methods has its advantages and disadvantages as presented in reference [2-4].

LOPA and Risk Graph techniques have proven to be valuable tools in assessing and managing process safety risks [2-3,5]. The LOPA technique offers a systematic approach that considers multiple layers of protection and provides a comprehensive analysis of potential hazards. On the other hand, the risk graph technique offers a simplified graphical representation that allows for quick and intuitive decision making. The choice between the LOPA and Risk Graph techniques depends on various factors, including the complexity of the process, the availability of data, and the expertise of the persons. The LOPA may be more suitable for complex systems with abundant data, while the Risk Graph can be a practical option for simpler processes or when time is a constraint. Furthermore, by combining the strengths of both techniques, a more robust and accurate analysis can be achieved. The LOPA technique can be used to conduct a detailed and in-depth analysis, considering various factors and layers of protection. This provides a comprehensive understanding of the risks involved. The risk graph approach, on the other hand, can be used as a complementary tool to quickly assess risks and make initial decisions.

This paper goes on to explain the need to understand the hazards, the ways to prevent or mitigate the risk which can occur from any unidentified events, and to assess the integrity level for all instrumented protection functions that have been provided for the heater section in the delayed cooking unit in petroleum refinery plants by using the LOPA and Risk Graph techniques. Instrument and control systems play a significant role in the controlling of hazards on oil and gas installations.

2. Description of the two used techniques

The two used approaches in this study for determining the safety integrity levels of the safety instrumented functions are fully described in the following subsection.

2.1. LOPA technique

LOPA is one of the techniques developed in response to a requirement within the process industry to be able to assess the adequacy of the layers of protection provided for an activity. Initially, this was driven by industry codes of practice or guidance and latterly by the development of international standards such as IEC61508 and IEC61511 [2-6]. The important steps that shall be addressed during SIL assessment sessions start by identifying and listing all SIFs for the unit(s) from Cause & Effect after HAZOP recommendations are implemented and finalized then for each identified SIF the methodology steps listed below should be followed [7-10].

A. Define the unwanted impact: The unwanted impact may be minor (m), this minor impact is initially limited to the local area of an event with the potential for broader consequences if corrective action is not taken. Serious (s) impact events could cause serious injury or fatality on-site or off-site. Extensive (e) impact event that is five times or more severe than a serious event.

B. Determine the initiating events and their likelihood: The initiating event description (initiating causes) is determined from HAZOP study. Impact events may have many initiating causes, and it is important to list all of them. Likelihood values or frequencies of the initiating causes occurring, in events per year, the experience of the team is very important in determining the initiating cause likelihood (see Table 2).

Table 2. Initiating event likelihood.

The initiating event	PFD per year	The initiating event	PFD per year
Control loop	0.1	Hand valve	0.1
Human under stress	0.5: 1	Pump trip	0.33
Vessel pressure above maximum allowable pressure	0.00001	Pressure safety valve/ mech. stop	0.01
Operator response to alarm	0.1	Operator error	0.1
Stuck/Blockage line	0.1	Alarm system	0.1
different fuel	0.1	High skin temp	0.3
Switch ON/OFF Valve	0.03	Snuffing steam	0.1
Seal tandem	0.1	Explosion door (heater natural draft)	0.1
People present	0.1 (0.01)	Lightning strike	0.001

C. Protection layers: Each protection layer consists of a grouping of equipment and/or administrative controls that function in concert with the other layers e.g. (Basic Process Control System BPCS, Alarm). Protection layers that perform their function with a high degree of reliability may qualify as independent protection layers (IPL).

D. Additional mitigation: Mitigation layers are normally mechanical, structural e.g., pressure relief devices, dikes (bunds), and restricted access. Mitigation layers may reduce the severity of the impact event but not prevent it from occurring e.g., deluge systems for fire or fume release, fume alarms, and evacuation procedures. The LOPA team should determine the appropriate PFDavg for all mitigation layers as per experience.

E. Independent protection layers (IPL) and their frequencies: IPL is a device, system, or action that can prevent a scenario from proceeding to its undesired consequence independent of the initiating event or the action of any other layer of protection associated with the scenario. Table 3 illustrates the Probability of Failure on Demand (PFD) for the IPL.

Table 3. IPL probability of failure on demand values.

IPL	PFD Values	IPL	PFD Values
Basic process control system	1×10^{-1}	Underground drainage system	1×10^{-2}
Relief valve	1×10^{-2}	Open vent (no valve)	1×10^{-2}
Rupture disc	1×10^{-2}	Fireproofing	1×10^{-2}
Flame/detonation arrestors	1×10^{-2}	Blast-wall / bunker	1×10^{-2}
Dike	1×10^{-2}		

F. Intermediate event likelihood (IEL).

This likelihood can be described as follows:

- The intermediate event likelihood is calculated by multiplying the initiating likelihood by the PFDavg of the protection layers and mitigating layers. The calculated number is in units of events per year.
- If the intermediate event likelihood is less than the process safety target level for events of this severity level, additional protection layers are not required.
- If the intermediate event likelihood is greater than your corporate criteria for events of this severity level, additional mitigation is required. Inherently safer methods and solutions should be considered before additional protection layers in the form of a Safety Instrumented System (SIS) are applied [2].

G. Target Mitigated Event Likelihood (TMEL): TMEL is considered for the consequence from the HAZOP study and TMEL values are taken from company standards where events are per year. Table 4 shows the standards used in this research work.

Table 4. Target mitigated event likelihood values.

People "Safety"	Financial (assets and production loss)	Environment	TMEL values
Multiple fatalities (6 or more). Numerous	Extensive damage (>US$100m)	Massive offsite effect (widespread permanent or chronic effects / constant high exceedance)	$1.0*10^{-6}$
Up to 2 Fatalities. Serious injuries.	Localized damage (US$1 10m)	Major offsite environmental damage with lasting, national impact e.g., liquid spill into river or sea, lasting damage to plants or and fauna (toxic effects), ground water pollution	$1.0*10^{-5}$
Single or few serious injuries.	Localized damage (US$100k-1m)	Major offsite environmental damage, but which can be completely cleared up within 1month e.g., causes temporary	$1.0*10^{-4}$
Lost time incident	Minor damage (US$10–100k)	Significant on-site environmental damage e.g., toxic vapor cloud extending beyond single unit, large leak or liquid spill, no affect upon ground water	$1.0*10^{-2}$
Minor injury first aid case	Slight damage (< S$10k)	Minor on site environmental damage, large enough to be reported to plant management. e.g., moderate leak from a flange or valve	$1.0*10^{-1}$

2.2. Risk graph technique

This technique aims to determine the SILs of the SIFs. This is a qualitative method that enables the SIL of a SIF to be determined from knowledge of the risk factors associated with the process and basic process control system (BPCS) [3]. The risk graph is based on the principle that risk is proportional to the consequence and frequency of the hazardous event. It starts by assuming that no safety instrumented system (SIS) exists, although typical non-SIS such as a basic process control system BPCS and monitoring systems are in place. Foord *et al.* [4] highlighted the implications of the issues at Safety-Critical Systems Symposium (SSS04) as follows:

- Risk graphs are very useful but imprecise tools for assessing SIL requirements. (It is inevitable that a method with 5 parameters — C, F, P, W, and SIL — each with a range of an order of magnitude, will produce a result with a range of 5 orders of magnitude.)
- They must be calibrated on a conservative basis to avoid the danger of underestimating the unprotected risk and the amount of risk reduction/protection required.
- Their use is most appropriate when several functions protect against different hazards, which are themselves only a small proportion of the overall total hazards. Underestimates

and overestimates of residual risk will likely average out when they are aggregated. Only in these circumstances can the method be realistically described as providing a "suitable" and "sufficient", and therefore legal, risk assessment.

In the process sector, risk is a function of four parameters which are consequence, occupancy, probability of avoiding the hazard, and demand rate as listed in Table 5.

Table 5. Descriptions of process industry risk graph parameters.

Parameter		Description
Consequence	C	Number of fatalities and/or serious injuries likely to result from the occurrence of the hazardous event. Determined by calculating the numbers in the exposed area when the area is occupied considering the vulnerability to the hazardous event.
Occupancy	F	Probability that the exposed area is occupied at the time of the hazardous event. Determined by calculating the fraction of time the area is occupied at the time of the hazardous event. This can consider the possibility of an increased likelihood of persons being in the exposed area to investigate abnormal situations which may exist during the build-up to the hazards
Probability of avoiding the hazard	P	Probability that exposed persons can avoid the hazardous situation which exists if the SIF fails on demand. This depends on there being independent methods of alerting the exposed persons to the hazard prior to the hazard occurring and there being methods of escape.
Demand rate	W	The number of times per year that a hazardous event would occur in the absence of the SIF under consideration. This can be determined by considering all failures which can lead to the hazardous event and estimating the overall rate of occurrence. Other protection layers should be included in the consideration.

Risk graph approaches are based on methods described in the German publication DIN V 19250 (see Figure 1) [11]. It enables SILs to be determined using process risk factors or parameters for hazardous events. Usually, four parameters are employed. The procedures for SIL classification are given with IEC 61511-3, Annex C- F. The risk is to be estimated considering the independent layer of protection.

The risk graph contains four risk parameters which must be determined by the SIL team using the given parameter ranges. Where risk parameters and parameter ranges are described below [3]:

W: Demand rate for SIF or occurrence frequency of deviation considered
- W1: Very low (once per 10 to 100 years), e.g., no case history
- W2: Low (once per 1 to 10 years), e.g., case has happened.
- W3: High (< 1 year), e.g., cases have happened on several occasions.

C: Consequences parameter (health and safety)
- C1: Minor injury
- C2: Serious permanent injury to 1 or 2 persons; death to 1 person
- C3: Death to several persons
- C4: Many deaths/catastrophe

F: Frequency that the exposed area is occupied
- F1: Rare to frequently
- F2: Frequently to continuously

P: Probability of avoiding the hazard
- P1: Under certain circumstances, e.g., operator invention (independent alarm and possibility for operator intervention in ~ five minutes), escape, and emergency stop
- P2: Almost impossible

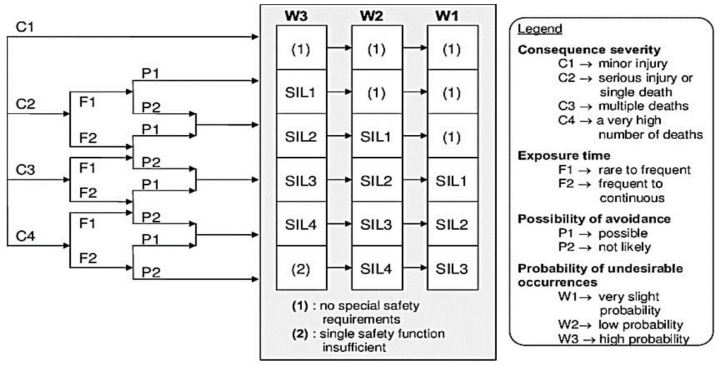


Figure 1. Safety integrity level (SIL) risk graph [3].

3. Case study

The case study of this work is used to carry out the SIL by LOPA & Risk Graph techniques for 6 Safety Instrument Functions (SIFs) of the considered DCU heater as it is considered the important and critical node of the DCU plant [12-13]. Then comparing the results between both methods for further study can affect the SIL rating and so on the total cost reduction of the unit and reduce design and installation, operation, and maintenance. SIL rating is only for the Safety Instrument System "SIS" which contains SIFs, any Safety trip signals in the design connected to the emergency shutdown system shall be subjected to SIL study.

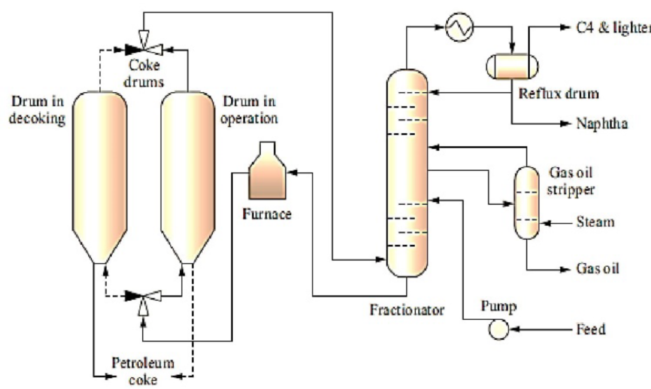


Figure 2. Schematic of the considered delayed coking unit [12].

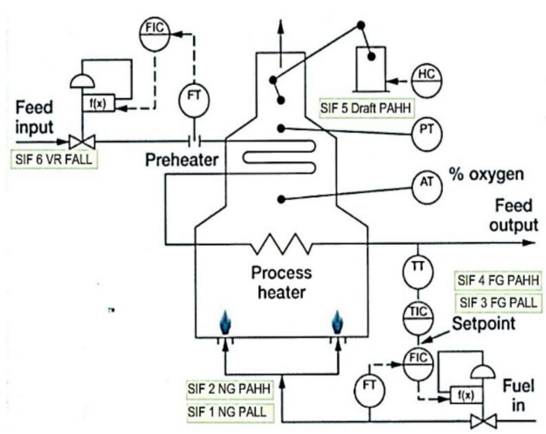


Figure 3. The Considered Coker Heater control scheme [13].

The vacuum residue (VR) feed comes from VDU to the DCU fractionator bottom mixing with DCU distillate and natural recycle then through the Feed pump directly to the heater passes controlled by flow control valves and flow ESD transmitter (SIF 6). In case of No/Less flow from the feed pump to the heater, passes must trip the heater to avoid the coking in the tubes. Then mixing the reside flow with velocity steam helps in delaying the cracking reaction in the tubes to be in the coke drum by reducing the residence time in the heater tubes.

The investigated coker heater accommodates 6 SIFs as per cause & effect and HAZOP final report as indicated in Figure 3 which included pilot natural gas alarm Low-Low and High-High pressure (PALL SIF1&PAHH SIF2), burners fuel gas alarm Low-Low and High-High pressure (PALL SIF3&PAHH SIF4), Firebox pressure alarm High-High (PAHH SIF5) and the feed flow alarm Low-Low (FALL SIF6) to the heater passes flow which is shown below in detail in Table 6.

Table 6. Safety instrumented functions (SIFs) list.

SIF No.	Interlock number (SIF)	Causes of failure	Consequence
1	PALL (pressure alarm Low-Low for pilot natural gas)	1. Strainer in natural gas line blocked 2. PCV (NG pressure control valve) fails to close 3. The suction 1st shut-off valve fails to close 4. The suction 2nd shut-off valve fails to close 4. shut off valve to flare fails to open 5. Inadvertent closure of any of the manual valves	Pilot burners flame out resulting in loss of pilots and escalating to heater total trip, Potential issue during process parameters deviation in main burners leading to fire and explosion
2	PAHH (pressure alarm High-High for pilot natural gas)	1. PCV (NG pressure control valve) fails to open 2. Nozzles on one or more pilots blocked 3. High pressure natural gas supply suddenly 4. Inadvertent closure of the last manual valves of the pilot	Pilot burners flame out resulting in loss of pilots and escalating to heater total trip, Potential issue during process parameters deviation in main burners leading to fire and explosion
3	PALL (FG pressure switch Low-Low to burners)	1. Strainer in fuel gas line blocked 2. Fail close of the 1st shut-off valve 3. Fail close of the 2nd shut-off valve 4. Fail open of the flare shut-off valve 5. Fail close of the regulating control valve 6. Inadvertent closure of any of the manual valve	the potential of the tarry drum due to introducing cold reside flow to the coke drum and potential fire during decoking the drum
4	PAHH (pressure alarm High-High for burner fuel gas)	1. FG control valve fails to open 2. Inadvertent opening of the manual bypass of the control 3. High pressure natural gas supply suddenly 4. Inadvertent closure of the last manual valve of the burner	1. potential coking in furnace tubes due to uneven heating 2. excess fuel gas inside the firebox can cause an explosion if not controlled well 3. overheating the tubes and tube rupture
5	PAHH (pressure alarm High-High for heater firebox)	1. Arch damper pressure control valve malfunction close 2. BL blower fan trip 3. Inadvertent closure of the manual arch damper 4. Combustion air damper flow control valve fails to open	High pressure in firebox leads to potential backfire and explosion

SIF No.	Interlock number (SIF)	Causes of failure	Consequence
6	FALL (Flow alarm Low-Low of Vacuum residue)	1. The battery limit feed control valve fail close 2. The suction shut-off valve for the feed pump fail close 3. The heater feed pump is a trip 4. The control valve at the heater pass is fail close 5. Inadvertent closure of any of the manual valves to the pass flow	Interruption of Vacuum residue flow to the heater passes leading to potential over temperature causing coke formation inside the heater coils and potential hot spots leading to coil rupture

4. Results and discussion

As mentioned before, the objective of this current work begins with the determination of the DCU heater SIF integrity levels SIL by using two methods and then comparing the results obtained by applying the LOPA and Risk Graph techniques. The following subsections present the results and discussion of these two applied methods.

4.1. LOPA scenario

According to the LOPA method, For SIL calculation, the PFD for the SIF should be calculated by using Equations 1 and 2.

$$\mathrm{PFD\ (SIF)} = \frac{\mathrm{TMEL}}{\mathrm{IEL_{total}}} \tag{1}$$

where PFD is the probability of failure on demand for each SIF; TMEL is the highest likelihood value of TMEL for the asset, environmental, and safety aspects as indicated in Table 4 for CCPS; TMEL values taken from company standard and determined during HAZOP study as per Scenario Consequences; IEL sums the intermediate event likelihood of all causes e.g. IEL of cause1+ IEL of cause2+…, etc.

$$\mathrm{IEL} = \mathrm{ICL} * \mathrm{PFD} \tag{2}$$

where ICL is the initiating cause likelihood as per IEC-61508 as addressed in Table 2; PFD is for the protection layers “PLs” as shown in Table 3.

SIL calculations for the 6 SIFs listed in Table 6 and shown in Figure 3 are described in the following paragraphs.

SIF 1; PALL (pressure alarm Low-Low for pilot natural gas): Interruption of natural gas to the pilot of the burners leading to pilot burners flame out resulting in loss of pilots and escalating to heater total trip, Potential issue during process parameters deviation in main burners leading to fire and explosion, loss of containment, injuries, fatalities, asset damage and environmental impact.

From Table 7 we can get that the calculation of IEL for each cause presented in Table 6 and using the data in Table 2 that calculated using Equation 2.

IEL for cause 1 = 0.1*0.1*0.1= $1*10^{-3}$ event/year
IEL for cause 2 = 0.1*0.1*0.1= $1*10^{-3}$ event/year
IEL for cause 3 = 0.03*0.1*0.1=$3*10^{-4}$ event/year
IEL for cause 4 = 0.03*0.1*0.1=$3*10^{-4}$ event/year
IEL for cause 5 = 0.03*0.1*0.1=$3*10^{-4}$ event/year
IEL for cause 6 = 0.1*0.1*0.1= $1*10^{-4}$ event/year
Thus, $\mathrm{IEL}_{\mathrm{Total}} = (1*10^{-3}) + (1*10^{-3}) + (3*10^{-4}) + (3*10^{-4}) + (3*10^{-4}) + (1*10^{-3}) = 3.9*10^{-03}$ event/year

PFD required for the studied SIF is calculated by using Equation 1. Since TMEL is getting from Table 3, considering the TMEL Value for this scenario is $1*10^{-5}$ (personnel safety is the target). PFD of SIF1 = $(1*10^{-5}) / (3.9*10^{-3}) = 2.56*10^{-03}$. According to the data of Table 1, the SIL required for the studied SIF1 is SIL 2

Table 7. LOPA SIF 1 SIL calculation.

	1	2	3	4	5			6	7	8	9	10
					Protection layers (PLs) and							
Impact event assessment	Impact Event Description	Severity Level	Initiating Cause	Initiation Likelihood (events per year)	General Process Design (probability)	Basic process control system (probability)	Alarms, Etc. (probability)	Additional Mitigation, Restricted Access (probability)	IPL Additional Mitigation Dikes, Pressure Relief (probability)	Intermediate Event Likelihood (per year)	SIF Integrity Level & PFD	Target Mitigated Event Likelihood (events per year)
			Sensors	PT pressure transmitter								
			Final Elements	Shut off valves UV for the Fuel gas close, Partial trip will be activated								
Safety Risk	PAHH (FG to main burner pressure high high) resulting in increase duty and process upset, Potential coking due to overheating and thermal stress of furnace tube leading to tube rupture, followed by potential fire and explosion, loss of containment, injuries, fatalities, asset damage and environmental impact.	Serious	Cause 1	FG control valve fails open		BPCS	Independent alarm		Explosion door to open		SIL 1	People "Safety": Up to 2 Fatalities. Serious Injuries. Financial (Assets and Production Loss): Localized Damage (US$1 -10m) Environment: Major offsite environmental damage, national impact
				0.10	1.00	0.10	0.10	1.00	0.10	1.00E-04		
			Cause 2	Inadvertent opening of the manual bypass of the control		BPCS	Independent alarm		Explosion door to open			
				0.10	1.00	0.10	0.10	1.00	0.10	1.00E-04		
			Cause 3	High pressure fuel gas supply suddenly		BPCS	Independent alarm		Explosion door to open			
				0.10	1.00	0.10	0.10	1.00	0.10	1.00E-04		
			Cause 4	Inadvertent closure of the last manual valve of the Burner		BPCS	Independent alarm		Explosion door to open			
				0.10	1.00	0.10	0.10	1.00	0.10	1.00E-04		
										4.00E-04	2.50E-02	1.00E-05
								Overall Target Average PFD			1.00E-02	SIL 1

SIF 2; PAHH (pressure alarm High-High in natural gas): Pressure alarm High-High in natural gas pilot burners leading to flame out resulting in loss of pilots and escalating to heater total trip, Potential issue during process parameters deviation in main burners leading to fire and explosion, loss of containment, injuries, fatalities, asset damage, and environmental impact. Then by using the same manner of SIF1, the SIL required for the studied SIF 2 is SIL 2 as illustrated in Table 8.

Table 8. LOPA SIF2 SIL calculation.

	1	2	3	4	5			6	7	8	9	10
					Protection layers (PLs) and							
Impact event assessment	Impact Event Description	Severity Level	Initiating Cause	Initiation Likelihood (events per year)	General Process Design (probability)	Basic process control system (probability)	Alarms, Etc. (probability)	Additional Mitigation, Restricted Access (probability)	IPL Additional Mitigation Dikes, Pressure Relief (probability)	Intermediate Event Likelihood (per year)	SIF Integrity Level & PFD	Target Mitigated Event Likelihood (events per year)
			Sensors	PT pressure transmitter								
			Final Elements	Shut off valves UV for the Fuel gas close, Partial trip will be activated								
Safety Risk	PAHH (FG to main burner pressure high high) resulting in increase duty and process upset, Potential coking due to overheating and thermal stress of furnace tube leading to tube rupture, followed by potential fire and explosion, loss of containment, injuries, fatalities, asset damage and environmental impact.	Serious	Cause 1	FG control valve fails open		BPCS	Independent alarm		Explosion door to open		SIL 1	People "Safety": Up to 2 Fatalities. Serious Injuries. Financial (Assets and Production Loss): Localized Damage (US$1 -10m) Environment: Major offsite environmental damage, national impact
				0.10	1.00	0.10	0.10	1.00	0.10	1.00E-04		
			Cause 2	Inadvertent opening of the manual bypass of the control		BPCS	Independent alarm		Explosion door to open			
				0.10	1.00	0.10	0.10	1.00	0.10	1.00E-04		
			Cause 3	High pressure fuel gas supply suddenly		BPCS	Independent alarm		Explosion door to open			
				0.10	1.00	0.10	0.10	1.00	0.10	1.00E-04		
			Cause 4	Inadvertent closure of the last manual valve of the Burner		BPCS	Independent alarm		Explosion door to open			
				0.10	1.00	0.10	0.10	1.00	0.10	1.00E-04		
										4.00E-04	2.50E-02	1.00E-05
								Overall Target Average PFD			1.00E-02	SIL 1

SIF 3; PALL (FG pressure alarm Low-Low to burners): PALL of fuel gas leading to Interruption of fuel gas to the heater resulting in the potential of the tarry drum due to introducing cold reside flow to the coke drum and potential fire during decoking the drum leading to injuries, fatalities, asset damage, and environmental impact. Then by using the same manner of SIF1, the SIL required for the studied SIF 3 is SIL 1 as illustrated in Table 9.

Table 9. LOPA SIF 3 SIL calculation.

	1	2	3	4	5			6	7	8	9	10
					Protection layers (PLs) and							
Impact event assessment	Impact Event Description	Severity Level	Initiating Cause	Initiation Likelihood (events per year)	General Process Design (probability)	Basic process control system (probability)	Alarms, Etc. (probability)	Additional Mitigation, Restricted Access (probability)	IPL Additional Mitigation Dikes, Pressure Relief (probability)	Intermediate Event Likelihood (per year)	SIF Integrity Level & PFD	Target Mitigated Event Likelihood (events per year)
			Sensors	PT pressure transmitter								
			Final Elements	Shut off valves UV for the Fuel gas close, Partial trip will be activated								
Safety Risk	PAHH (FG to main burner pressure high high) resulting in increase duty and process upset, Potential coking due to overheating and thermal stress of furnace tube leading to tube rupture, followed by potential fire and explosion, loss of containment, injuries, fatalities, asset damage and environmental impact.	Serious	Cause 1	FG control valve fails open		BPCS	Independe nt alarm		Explosion door to open		SIL 1	People “Safety”: Up to 2 Fatalities. Serious Injuries. Financial (Assets and Production Loss): Localized Damage (US$1 -10m) Environment: Major offsite environmental damage, national impact
				0.10	1.00	0.10	0.10	1.00	0.10	1.00E-04		
			Cause 2	Inadvertent opening of the manual bypass of the control		BPCS	Independe nt alarm		Explosion door to open			
				0.10	1.00	0.10	0.10	1.00	0.10	1.00E-04		
			Cause 3	High pressure fuel gas supply suddenly		BPCS	Independe nt alarm		Explosion door to open			
				0.10	1.00	0.10	0.10	1.00	0.10	1.00E-04		
			Cause 4	Inadvertent closure of the last manual valve of the Burner		BPCS	Independe nt alarm		Explosion door to open			
				0.10	1.00	0.10	0.10	1.00	0.10	1.00E-04		
										4.00E-04	2.50E-02	1.00E-05
								Overall Target Average PFD			1.00E-02	SIL 1

SIF 4; PAHH (pressure alarm High-High for burner fuel gas): PAHH of FG to the main burner resulting in increasing duty and process upset, Potential coking due to overheating and thermal stress of furnace tube leading to tube rupture, followed by potential fire and explosion, loss of containment, injuries, fatalities, asset damage, and environmental impact. Then by using the same manner of SIF1, the SIL required for the studied SIF 4 is SIL 2 as illustrated in Table 10.

SIF 5; PAHH (pressure alarm High-High for heater firebox): PAHH in firebox resulting in potential flame out followed by potential backfire and explosion leading to loss of containment, injuries, fatalities, asset damage, and environmental impact. Then by using the same manner of SIF1, the SIL required for the studied SIF 3 is SIL 1 as illustrated in Table 11.

SIF 6; FALL (Flow alarm Low-Low of Vacuum residue): Interruption of Vacuum residue flow to the heater passes leading to potential over temperature causing coke formation inside the heater coils and potential hot spots leading to coil rupture, loss of containment, fires, explosions, injuries, fatalities, asset damage, and environmental impact. Then by using the same manner of SIF1, the SIL required for the studied SIF 3 is SIL 2 as illustrated in Table 12.

Table 10. LOPA SIF 4 SIL calculation.

	1	2	3	4	5			6	7	8	9	10
					Protection layers (PLs) and							
Impact event assessment	Impact Event Description	Severity Level	Initiating Cause	Initiation Likelihood (events per year)	General Process Design (probability)	Basic process control system (probability)	Alarms, Etc. (probability)	Additional Mitigation, Restricted Access (probability)	IPL Additional Mitigation Dikes, Pressure Relief (probability)	Intermediate Event Likelihood (per year)	SIF Integrity Level & PFD	Target Mitigated Event Likelihood (events per year)
			Sensors	PT pressure transmitter								
			Final Elements	Shut off valves UV for the Fuel gas close, Partial trip will be activated								
Safety Risk	PALL of fuel gas leading to Interruption of fuel gas to the heater resulting in potential of tarry drum due to introducing reside flow under the cracking temperature to the coke drum and potential fire during decoking the drum leading to injuries, fatalities, asset damage and environmental impact.	Serious	Cause 1	Strainer in fuel gas line blocked	stand by straner in place	BPCS	Independent alarm		Explosion door to open		SIL 2	People "Safety": Up to 2 Fatalities. Serious Injuries. Financial (Assets and Production Loss): Localized Damage (US$1 -10m) Environment: Major offsite environmental damage, national impact
				0.10	0.10	0.10	0.10	1.00	0.10	1.00E-05		
			Cause 2	Fail close of the 1st shut off valve UV			Independent alarm		Explosion door to open			
				0.03	1.00	1.00	0.10	1.00	0.10	3.00E-04		
			Cause 3	Fail close of the 2nd shut off valve UV			Independent alarm		Explosion door to open			
				0.03	1.00	1.00	0.10	1.00	0.10	3.00E-04		
			Cause 4	Fail open of the flare shut off valve UV			Independent alarm		Explosion door to open			
				0.03	1.00	1.00	0.10	1.00	0.10	3.00E-04		
			Cause 5	Fail close of the regulating control valve			Independent alarm		Explosion door to open			
				0.10	1.00	1.00	0.10	1.00	0.10	1.00E-03		
			Cause 6	Inadvertent closure of any of the manual valve			Independent alarm		Explosion door to open			
				0.10	1.00	1.00	0.10	1.00	0.10	1.00E-03		
										2.91E-03	3.44E-03	1.00E-05
								Overall Target Average PFD			3.44E-03	SIL 2

Table 11. LOPA SIF 5 SIL calculation.

	1	2	3	4	5			6	7	8	9	10
					Protection layers (PLs) and							
Impact event assessment	Impact Event Description	Severity Level	Initiating Cause	Initiation Likelihood (events per year)	General Process Design (probability)	Basic process control system (probability)	Alarms, Etc. (probability)	Additional Mitigation, Restricted Access (probability)	IPL Additional Mitigation Dikes, Pressure Relief (probability)	Intermediate Event Likelihood (per year)	SIF Integrity Level & PFD	Target Mitigated Event Likelihood (events per year)
			Sensors	PT pressure transmitter								
			Final Elements	Forced and Induced Blowers stop, Arch damper control valves fully open , All air doors Fully open, Total trip activated								
Safety Risk	PAHH in fire box, potential flame out followed by potential back fire and explosion leading to loss of containment, injuries, fatalities, asset damage and environmental impact.	Serious	Cause 1	Arch damper pressure control valve malfunction close		BPCS	Independent alarm		Explosion door to open		SIL 1	People "Safety": Up to 2 Fatalities. Serious Injuries. Financial (Assets and Production Loss): Localized Damage (US$1 -10m) Environment : Major offsite environmental damage, national impact
				0.10	1.00	0.10	0.10		0.10	1.00E-04		
			Cause 2	BL blower fan trip		BPCS	Independent alarm		Explosion door to open			
				0.33	1.00	0.10	0.10		0.10	3.30E-04		
			Cause 3	Inadvertent closure of the manual arch damper		BPCS	Independent alarm		Explosion door to open			
				0.10	1.00	0.10	0.10		0.10	1.00E-04		
			Cause 4	Combustion air damper flow control valve fails open		BPCS	Independent alarm		Explosion door to open			
				0.10	1.00	0.10	0.10		0.10	1.00E-04		
										6.30E-04	1.59E-02	1.00E-05
								Overall Target Average PFD			1.59E-02	SIL 1

Table 12. LOPA SIF 6 SIL calculation.

	1	2	3	4	5			6	7	8	9	10
					Protection layers (PLs) and							
Impact event assessment	Impact Event Description	Severity Level	Initiating Cause	Initiation Likelihood (events per year)	General Process Design (probability)	Basic process control system (probability)	Alarms, Etc. (probability)	Additional Mitigation, Restricted Access (probability)	IPL Additional Mitigation Dikes, Pressure Relief (probability)	Intermediate Event Likelihood (per year)	SIF Integrity Level & PFD	Target Mitigated Event Likelihood (events per year)
			Sensors	FT Flow transmitter								
			Final Elements	Velocity steam control valve will be fully open , All suction on/off shut off valves UV for fuel gas will close								
Safety Risk	FALL Interruption of Vacuum residue flow leading to potential over temperature causing coke formation inside the heater coils and potential hot spots leading to coil rupture, loss of containment , fires, explosions, injuries, fatalities, asset damage and environmental impact.	Serious	Cause 1	battery limit feed control valve fail close			High skin temperature alarms		Snuffing steam		SIL 2	People “Safety”: Up to 2 Fatalities. Serious Injuries. Financial (Assets and Production Loss): Localized Damage (US$1 -10m) Environment: Major offsite environmental damage, national impact
				0.10	1.00	1.00	0.30	1.00	0.01	3.00E-04		
			Cause 2	suction shut off UV valve for the feed pump fail close			High skin temperature alarms		Snuffing steam			
				0.03	1.00	1.00	0.30	1.00	0.01	9.00E-05		
			Cause 3	Heater feed pump is trip	Stand pump auto start		High skin temperature alarms		Snuffing steam			
				0.33	0.10	1.00	0.30	1.00	0.01	9.90E-05		
			Cause 4	control valve at heater pass is fail close	Min. Mechanical Stop		High skin temperature alarms		Snuffing steam			
				0.10	0.01	1.00	0.30	1.00	0.01	3.00E-04		
			Cause 5	Inadvertent closure of any of the manual valve to the passes flow			High skin temperature alarms		Snuffing steam			
				0.10	1.00	1.00	0.30	1.00	0.01	3.00E-04		
										1.09E-03	9.18E-03	1.00E-05
								Overall Target Average PFD			9.18E-03	SIL 2

The LOPA SIL determination study output SIL targets summary for the heater 6 SIFs is presented in Table 13.

Table 13. LOPA SIL target for the heater 6 SIFs.

Classification	PFD	No. of SIF(s)
Non-classified SIL	-	0
SIL 1	$\geq 10^{-2}$ to $< 10^{-1}$	4
SIL 2	$\geq 10^{-3}$ to $< 10^{-2}$	2
SIL 3	$\geq 10^{-4}$ to $< 10^{-3}$	
Total SIFs		6

4.2. Risk graph scenario

To calculate SIL by the risk graph approach, the risk parameters C, F, P, and W described above in Figure 1 should be calculated. SIL calculations for 6 SIFs as listed in Table 6 and shown in Figure 3 are as illustrated in Tables 14-19.

Table 14. Risk Graph SIF 1 SIL calculation.

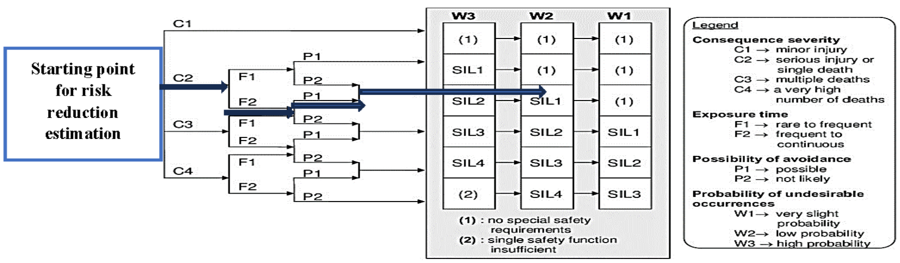
Interlock No.	PALL (pressure alarm Low-Low for pilot natural gas)
Initiation Device Tag	PT (pressure transmitter) 2 out of 3 with low pressure alarm and trip to NG supply on Low-Low pressure
Instrument Loop Description	PALL of NG will trip natural gas supply to pilot burners resulting in Flame out of pilots and accumulation of uncombusted gases in the furnace leading to fire and explosion, loss of containment, injuries, fatalities, asset damage, and environmental impact.

1. Classification: SIL 1

2. Risk assessment for safety instrumentation [3].

2.1 Covered risk:
PT (pressure transmitter) 2 out of 3 to trip natural gas supply to pilot burners at low-low pressure due to: -
1.PCV (NG pressure control valve) fails close,
2. Strainer in natural gas line blocked,
3. suction 1st shut off valve fails close,
4. suction 2nd shut off valve fails close,
5. shut off valve to flare fails open,
6. Inadvertent closure of any of the manual valves on the NG line.

2.2 Risk graph and classes demanded. The results shall be marked in the risk graph (Figure 1).



3. Risk parameters acc. to IEC 61511

3.1 Consequence parameter

C_1	slight injury	
C_2	severe irreversible injury of one or more persons or death of a person	■
C_3	death of several persons	
C_4	catastrophic effect, a great number of dead	

3.2 Frequency and exposure time parameter

F_1	seldom up to often	
F_2	frequently up to permanently	■

3.3 Possibility of avoiding the hazard event

P_1	possible under certain circumstances	■
P_2	not very possible	

3.4 Probability of the unwanted occurrence

W_1	very improbable	
W_2	probable	■
W_3	rather probable	

4. Reasons in brief for the selection of the risk parameters:

4.1 Kind of damage (C)

trip of natural gas supply to pilot burners leading to flame out of pilots and accumulation of uncombusted gases in furnace leading to explosion then severe irreversible injury of one or more persons or death of a person

4.2 Duration of stay (F)

frequently up to permanently where two operators are following the heater frequently due to drum cycle activity which affecting all the plant

4.3 Avoidance of risk (P)

Possible under certain circumstances where the operator is competence and trained well to handle any type of risk.

4.4 Probability of the undesired event (W)

Probable due to semi batch process which affecting the whole process.

Table 15. Risk Graph SIF 2 SIL calculation.

Interlock No.	PAHH (pressure alarm High-High in natural gas)
Initiation Device Tag	PT (pressure transmitter) 2 out of 3 with high pressure alarm and trip to NG supply on High-High pressure
Instrument Loop Description	PAHH of NG will trip natural gas supply to pilot burners leading to potential loss of flame detector 10 out of 12 resulting in Flame out of pilots and accumulation of uncombusted gases in furnace leading to fire and explosion, loss of containment, injuries, fatalities, asset damage and environmental impact.

1. Classification SIL 1
2. Risk assessment for safety instrumentation [3]
2.1 Covered risk:
PT (pressure transmitter) 2 out of 3 to trip natural gas supply to pilot burners with High-High pressure due to: - 1. PCV (NG pressure control valve) fails to open, 2. Nozzles on one or more pilots blocked, 3. High pressure natural gas supply suddenly, 4. Inadvertent closure of the last manual valves of the pilot.

2.2 Risk graph and classes demanded. The results shall be marked in the risk graph (Figure 1).

3. Risk parameters acc. to IEC 61511

3.1 Consequence parameter

C_1	slight injury	□
C_2	severe irreversible injury of one or more persons or death of a person	■
C_3	death of several persons	□
C_4	catastrophic effect, a great number of dead	□

3.2 Frequency and exposure time parameter

F_1	seldom up to often	□
F_2	frequently up to permanently	■

3.3 Possibility of avoiding the hazard event

P_1	possible under certain circumstances	■
P_2	not very possible	□

3.4 Probability of the unwanted occurrence

W1	very improbable	□
W2	probable	■
W3	rather probable	□

4. Reasons in brief for the selection of the risk parameters:

4.1 Kind of damage (C)

trip of natural gas supply to pilot burners leading to flame out of pilots and accumulation of uncombusted gases in furnace leading to explosion then severe irreversible injury of one or more persons or death of a person

4.2 Duration of stay (F)

frequently up to permanently where two operators are following the heater frequently due to drum cycle activity which affecting all the plant

4.3 Avoidance of risk (P)

Possible under certain circumstances where the operator is competence and trained well to handle any type of risk.

4.4 Probability of the undesired event (W)

Probable due to semi batch process which affecting the whole process.

Table 16. Risk Graph SIF 3 SIL calculation.

Interlock No.	PALL (FG pressure switch Low-Low to burners)
Initiation Device Tag	PT (pressure transmitter) for side and centre burners 2 out of 3 with low pressure alarm and trip to fuel gas supply on Low-Low pressure
Instrument Loop Description	PALL of fuel gas leading to Interruption of fuel gas to the heater resulting in potential of tarry drum due to introducing cold reside flow to the coke drum and potential fire during decoking the drum leading to injuries, fatalities, asset damage and environmental impact.

1. Classification: SIL 2
2. Risk assessment for safety instrumentation
2.1 Covered risk:
PT (pressure transmitter) for side and centre burners 2 out of 3 with low pressure alarm and trip to fuel gas supply on Low-Low pressure due to: - 1. Strainer in fuel gas line blocked, 2. Fail close of the 1st shut off valve, 3. Fail close of the 2nd shut off valve, 4. Fail open of the flare shut off valve, 5. Fail close of the regulating control valve, 6. Inadvertent closure of any of the manual valve

2.2 Risk graph and classes demanded. The results shall be marked in the risk graph.

3. Risk parameters acc. to IEC 61511

3.1 Consequence parameter

C_1	slight injury	☐
C_2	severe irreversible injury of one or more persons or death of a person	■
C_3	death of several persons	☐
C_4	catastrophic effect, a great number of dead	☐

3.2 Frequency and exposure time parameter

F_1	seldom up to often	☐
F_2	frequently up to permanently	■

3.3 Possibility of avoiding the hazard event

P_1	possible under certain circumstances	■
P_2	not very possible	☐

3.4 Probability of the unwanted occurrence

W1	very improbable	☐
W2	probable	☐
W3	rather probable	■

4. Reasons in brief for the selection of the risk parameters:

4.1 Kind of damage (C)

potential of tarry drum due to introducing cold reside flow to the coke drum and potential fire during decoking the drum leading to severe irreversible injury of one or more persons or death of a person

4.2 Duration of stay (F)

Frequently to Permanently where operators are following the activities of the drum cycle and other maintenance

4.3 Avoidance of risk (P)

Possible under certain circumstances where the operator is competence and trained well to handle any type of risk.

4.4 Probability of the undesired event (W)

rather probable due to the heater coil outlet temperature drop and introducing more fuel gas to compensate the temperature dropping

Table 17. Risk Graph SIF 4 SIL calculation.

Interlock No.	PAHH (pressure alarm High-High for burner fuel gas)
Initiation Device Tag	PT (pressure transmitter) for side and centre burners 2 out of 3 with High pressure alarm and trip to fuel gas supply on High-High pressure
Instrument Loop Description	PAHH of fuel gas resulting in increasing duty and process upset, potential tube coking due to overheating the tubes, and thermal stress leading to tube rupture, followed by potential fire and explosion, loss of containment, injuries, fatalities, asset damage, and environmental impact.

1. Classification SIL 1
2. Risk assessment for safety instrumentation
2.1 Covered risk:
PT (pressure transmitter) for side and centre burners 2 out of 3 with High pressure alarm and trip to fuel gas supply on High-High pressure due to: - 1. FG control valve fails open, 2. Inadvertent opening of the manual bypass of the control, 3. High pressure natural gas supply suddenly, 4. Inadvertent closure of the last manual valve of the Burner

2.2 Risk graph and classes demanded. The results shall be marked in the risk graph.

3. Risk parameters acc. to IEC 61511

3.1 Consequence parameter

C_1	slight injury	
C_2	severe irreversible injury of one or more persons or death of a person	■
C_3	death of several persons	
C_4	catastrophic effect, a great number of dead	

3.2 Frequency and exposure time parameter

F_1	seldom up to often	
F_2	frequently up to permanently	■

3.3 Possibility of avoiding the hazard event

P_1	possible under certain circumstances	■
P_2	not very possible	

3.4 Probability of the unwanted occurrence

W1	very improbable	
W2	probable	■
W3	rather probable	

4. Reasons in brief for the selection of the risk parameters:

4.1 Kind of damage (C)

potential coking in furnace tubes due to overheating, excess of fuel gas inside the fire box can cause fire and explosion leading to severe irreversible injury of one or more persons or death of a person

4.2 Duration of stay (F)

Frequently to Permanently operators are following the activities of the drum cycle and other maintenance

4.3 Avoidance of risk (P)

Possible under certain circumstances where the operator is competence and trained well to handle any type of risk.

4.4 Probability of the undesired event (W)

Probable due to semi batch process and disturbance of the plant due to drum cycle activities

Table 18. Risk Graph SIF 5 SIL calculation.

Interlock No.	PAHH (pressure alarm High-High for heater firebox)
Initiation Device Tag	PT (pressure transmitter) 2 out of 3 with High-High pressure alarm to shut off fuel gas supply to the heater and introduce natural draft mode
Instrument Loop Description	PAHH in a firebox leads to potential flame out followed by potential backfire and explosion leading to loss of containment, injuries, fatalities, asset damage, and environmental impact.

1. Classification: SIL 1

2. Risk assessment for safety instrumentation

2.1 Covered risk:

PT (pressure transmitter) 2 out of 3 with High-High pressure alarm to shut off fuel gas supply to the heater and introduce natural draft mode due to: -
1. Arch damper pressure control valve malfunction close,
2. BL blower fan trip,
3. Inadvertent closure of the manual arch damper
4. Combustion air damper flow control valve fails to open

2.2 Risk graph and classes demanded*

*) Explanation of the graph see below. The results shall be marked in the risk graph.

3. Risk parameters acc. to IEC 61511

3.1 Consequence parameter

C_1	slight injury	
C_2	severe irreversible injury of one or more persons or death of a person	■
C_3	death of several persons	
C_4	catastrophic effect, a great number of dead	

3.2 Frequency and exposure time parameter

F_1	seldom up to often	
F_2	frequently up to permanently	■

3.3 Possibility of avoiding the hazard event

P_1	possible under certain circumstances	■
P_2	not very possible	

3.4 Probability of the unwanted occurrence

W1	very improbable	
W2	probable	■
W3	rather probable	

4. Reasons in brief for the selection of the risk parameters:

4.1 Kind of damage (C)

High pressure in fire box resulting in potential backfire and explosion leading to severe irreversible injury of one or more persons or death of a person

4.2 Duration of stay (F)

Frequently to Permanently operators are following the activities of the drum cycle and other maintenance

4.3 Avoidance of risk (P)

possible under certain circumstances where the operator is competence and trained well to handle any type of risk.

4.4 Probability of the undesired event (W)

Probable due to semi batch process and disturbance of the plant due to drum cycle activities

Table 19. Risk Graph SIF 6 SIL calculation.

Interlock No.	FALL (Flow alarm Low-Low of Vacuum residue)
Initiation Device Tag	FT (Flow transmitter) at Low-Low flow will trigger partial trip for the furnaces and inject emergency steam full flow
Instrument Loop Description	FALL of Vacuum residue leading to potential over temperature causing coke formation inside the heater coils and potential hot spots leading to coil rupture, loss of containment, fires, explosions, injuries, fatalities, asset damage, and environmental impact.
1. Classification : SIL 2	
2. Risk assessment for safety instrumentation	
2.1 Covered risk:	
FT (Flow transmitter) at Low-Low flow will trigger partial trip for the furnaces and inject emergency steam full flow due to: - 1. battery limit feed control valve fails close, 2. Suction shut off valve for the feed pump fails to close, 3. The heater feed pump is trip, 4. The control valve at the heater pass is fail close, 5. Inadvertent closure of any of the manual valves to the pass flow	

2.2 Risk graph and classes demanded. The results shall be marked in the risk graph.

3. Risk parameters acc. to IEC 61511

3.1 Consequence parameter

slight injury	C_1	□
severe irreversible injury of one or more persons or death of a person	C_2	■
death of several persons	C_3	□
catastrophic effect, a great number of dead	C_4	□

3.2 Frequency and exposure time parameter

seldom up to often	F_1	□
frequently up to permanently	F_2	■

3.3 Possibility of avoiding the hazard event

possible under certain circumstances	P_1	■
not very possible	P_2	□

3.4 Probability of the unwanted occurrence

very improbable	W1	□
probable	W2	□
rather probable	W3	■

4. Reasons in brief for the selection of the risk parameters:

4.1 Kind of damage (C)

Potential of tube rupture due to thermal stress leading to potential fire and explosion inside the fire box leading severe irreversible injury of one or more persons or death of a person.

4.2 Duration of stay (F)

Frequently to Permanently operators are following the activities of the drum cycle and other maintenance

4.3 Avoidance of risk (P)

Possible under certain circumstances where the operator is competence and trained well to handle any type of risk.

4.4 Probability of the undesired event (W)

Rather probable due to semi batch process and the flow transmitter is very sensitive to any disturbance in the flow of the feed pump.

The Risk Graph SIL determination study output SIL targets summary for the heater 6 SIFs is presented in Table 20.

Table 20. Risk Graph SIL target for the heater 6 SIFs by Risk Graph Scenario.

Classification	PFD	No. of SIF(s)
Non-classified SIL	-	0
SIL 1	≥ 10^{-2} to < 10^{-1}	4
SIL 2	≥ 10^{-3} to < 10^{-2}	2
SIL 3	≥ 10^{-4} to < 10^{-3}	
Total SIFs		6

The comparison of the results obtained by applying LOPA and risk graph techniques for the same 6 SIFs of the delayed Coker fired heater is illustrated in Figure 4. These comparison shows that SIF1 and SIF2 have different rates by applying both methods without changing the required safety measures and keeping all safety requirements.

The results showed that the SIL rating is reduced to SIL 1 for SIF 1 and SIF 2 in the case of Risk Graph assessment. Such reduction has a good impact on the total cost of the unit and reduces design, installation, operation, and maintenance complexity. Improper SIL determination will affect the safety integrity of the asset protection and add more cost, in contrast, properly rating the SIL levels will lead to safety and cost improvements.

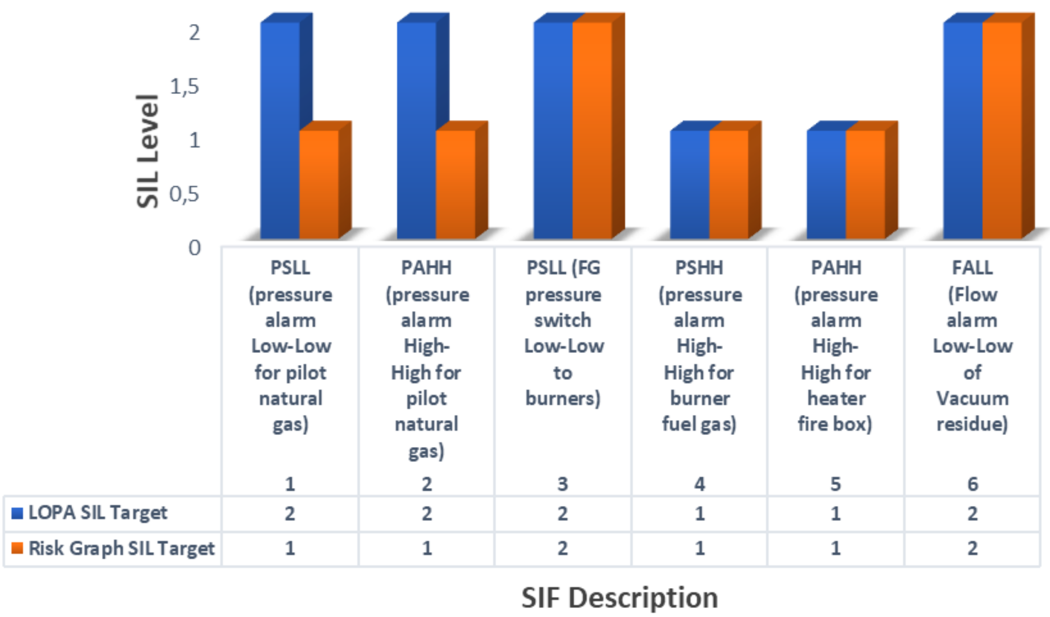


Figure 4. Final SIL target comparison of LOPA and Risk Graph techniques.

5. Conclusion

In this paper, the LOPA and Risk Graph techniques are used for studying the same six instrumented safety devices installed for the delayed coking fired heater. Using both techniques allows for a more holistic approach to risk assessment and a potential biases and oversights can be minimized, leading to a more accurate determination of the SIL target. It is recommended to utilize both techniques in conjunction to enhance the accuracy and reliability of the assessment. Practitioners and decision-makers should carefully evaluate the strengths and limitations of each technique before making a final determination.

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