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Optimum Plant Design for Relief Safety System

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ABSTRACT

Pressure relief system is a system to prevent overpressure inside protected equipment that exceeds its maximum allowable working pressure (MAWP) to disposal treatment. Relief system is designed on two different plant case studies, which are dimethyl-ether and ethylbenzene plants by using conventional design procedure. Nevertheless, the conventional design steps are not considering cost optimization of plant installed with relief system. Thus, the design pressure of protected equipment, piping diameter, and disposal treatment is set to be manipulated variables to determine the cost minimization. Pressure drops of inlet piping and backpressure are as constraint variables due to standard requirements. The standards state that inlet piping pressure drops should be below 3% of set pressure and outlet piping pressure drop to set pressure percentage based on range to determined types of the relief valve to be used. From that, optimum plant design with consideration of pressure relief system's installation can be achieved by calculating the total cost of plant designed with relief system when set pressures are changed. As a result, dimethyl-ether plant shows a minimum point of the total cost at 170% of set pressure increment, which is lower than its original design. On the other hand, ethylbenzene plant gives optimum point at original design as the total cost is higher at set pressure increment.

Keywords: Dimethyl-ether, ethylbenzene, plant optimization, relief system, set pressure variation

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INTRODUCTION

Background of Study

Pressure relief system is a preventive system to encounter excessive pressures inside pipelines and equipment. The system is designed to automatically relieve liquids or gases to atmosphere or any safe locations depending on properties of discharge materials and close when the pressure is back to the normal condition (Patil & Sondur, 2013). The excessive pressures are mainly due to some reasons such as blocked-outlet, exposed to external fire, thermal expansion, runaway chemical reaction, heat exchanger tube rupture, and cooling system failure (Hellemans, 2009). Pressure relief system consists of pressure relief devices, piping and downstream process equipment to the safe handling process of materials discharged, as shown in Figure 1. Locating relief devices are based on definitive guidelines (Crowl & Louvar, 2011), which pressure relief devices are located at any possible pressure accumulation due to operating failure in the plant. Several scenarios are listed that contribute to overpressure problem and the worst-case is chosen as the governing scenario by comparing the venting area required to reduce the excessive pressure. The sizes of relief devices can be computed by determined relief loads of discharged materials, the physical states of the fluids, and its relieving conditions. In the normal conceptual design of a safety relief system (Crowl & Louvar, 2011) however, consideration of the optimum cost of the plant with pressure relief system installation has been neglected.

Thus, optimum design of plant and pressure relief system is being considered by manipulating design pressure of equipment-also called as relief device's set pressure, sizes of piping diameter and disposal treatment design. As for project case studies, dimethyl-ether (DME) and ethylbenzene (EB) plants are being used.

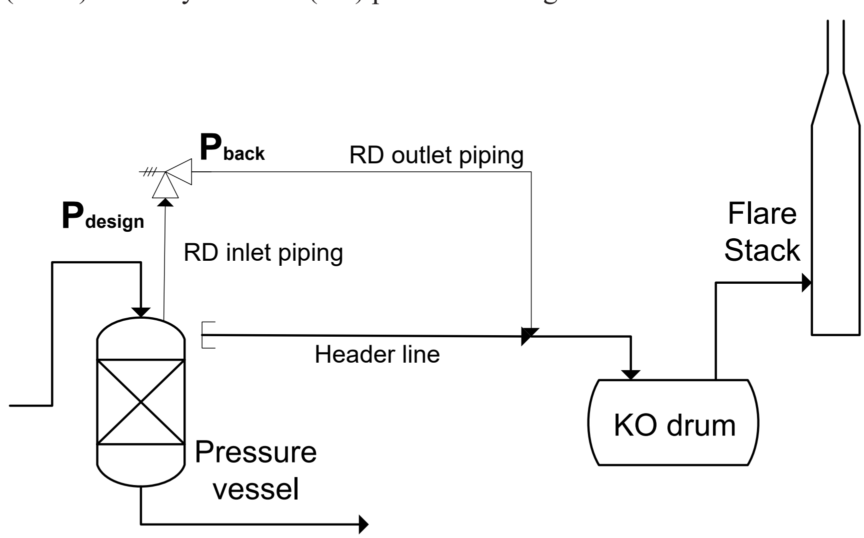

Figure 1. Typical relief system

Problem Statement

This project is proposed to provide an alternative approach to the conventional method of relief and flare system design. In the conventional method, process engineers already decide the design and the set pressure based on the mass balances without considering any cost calculation. This current work aims to include the cost calculation in the design where an optimum set pressure is going to be found that lead to minimum plant cost regarding this safety feature. This study covers the design of the relief system, cost estimations, and varying the design pressure and hence, the set pressure of the relief valves, calculation of backpressure and disposal treatment design to minimize the plant overall cost. The total cost of plant designed, which is the summation of costs including relief device, disposal treatment and protected equipment. Changes of the design pressure of protected equipment will be affecting pressure drops the percentage of inlet piping and backpressure of relief device which are associated with the changes of piping size, equipment strength and Knock-Out (KO) drum. Therefore, the cost of equipment, KO drum and piping is changing. From the manipulative actions, the most optimum plant design with the relief system is chosen based on the most economical total costs calculated.

MATERIALS AND METHODS

The methodology of the project is illustrated in flowchart form in Figure 2.

Relief Sizing

Possible scenarios are listed on located relief device, while calculation of relief loads and relief sizing is referring to American Petroleum Institute Standards, API RP 520 Part I and API Std 521 (API RP 520 Part I & Part II, 2011), based on relieving scenarios and flow states. Data of plants' streams were from iCON process simulation-main reference (Turton, 2012).

Backpressure was calculated as in Equation (1) and pressure drops across piping using Darcy-Weishbach Equation, Equation (2), where the pressure of KO-drum is set at 1.1 barg.

$$\mathrm{P_{back}} = \mathrm{P_{at\ KO\ drum}} + \Delta\mathrm{P_{across\ outlet-piping}} \tag{1}$$

$$\Delta\mathrm{P} = \mathrm{f_D}\frac{\mathrm{L}}{\mathrm{D}}\frac{\rho\mathrm{v}^2}{2} \tag{2}$$

Relief Downstream Design

The worst-case scenario for each relief valve was selected based on the biggest venting area calculated. Thus, the design of the relief downstream system was focused on these worst scenarios. The calculation to design knock-out (KO) drum and flare were referred to API Std 521 (American Petroleum Institute, 2014).

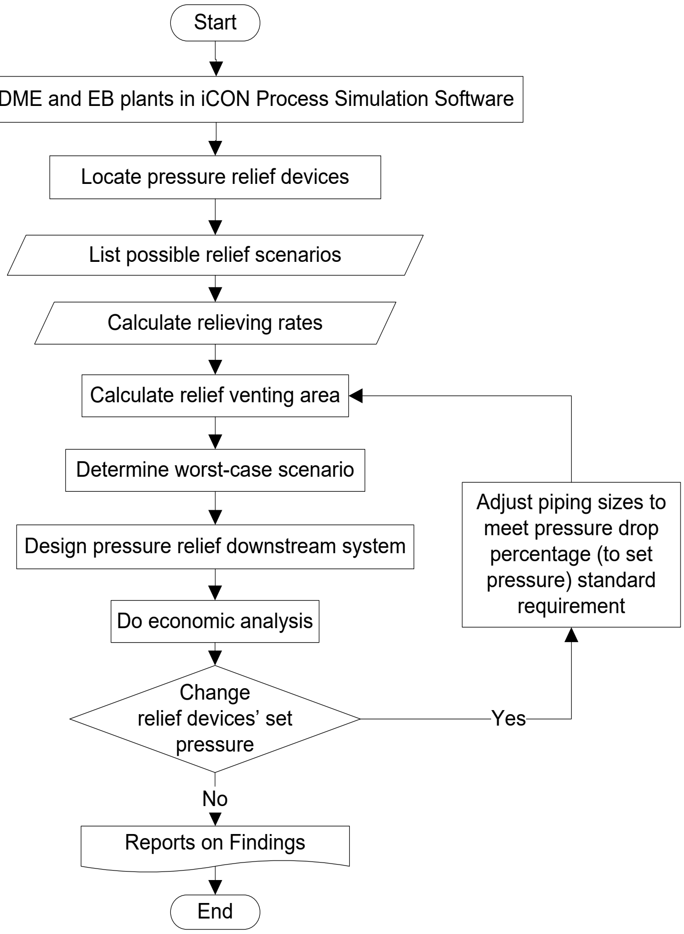

Figure 2. Project methodology

Economic Analysis

Total plant cost was done on the actual plant with relief system design as basis value for analysis. Set pressure of each relief valves was increased by a certain percentage to study the sensitivity of cost elements and total plant cost towards the changes. Equipment cost was estimated using CAPCOST BETA Spreadsheet, available at https://www.eng.famu.fsu.edu/~palanki/design/lectures/CAPCOST/CAPCOST.XLS, with CEPCI value of 591.335 (Jenkins, 2018), flare costing (Stone et.al., 1992) as in Equation (3), while, piping cost and relief valve were referred to products catalogue, US Pipe Fabrication (U.S. PIPE, 2018) and Flomatic Valves (FOLOMATIC, 2020).

$$C_F(\$) = (78.0 + 9.14D + 0.749L)^2 \tag{3}$$

RESULTS AND DISCUSSION

Relief Sizing

Relief valves were located on DME and EB plants and possible scenarios contributed to pressure built-up were determined. The locations are as in Figure 3 and Figure 4.

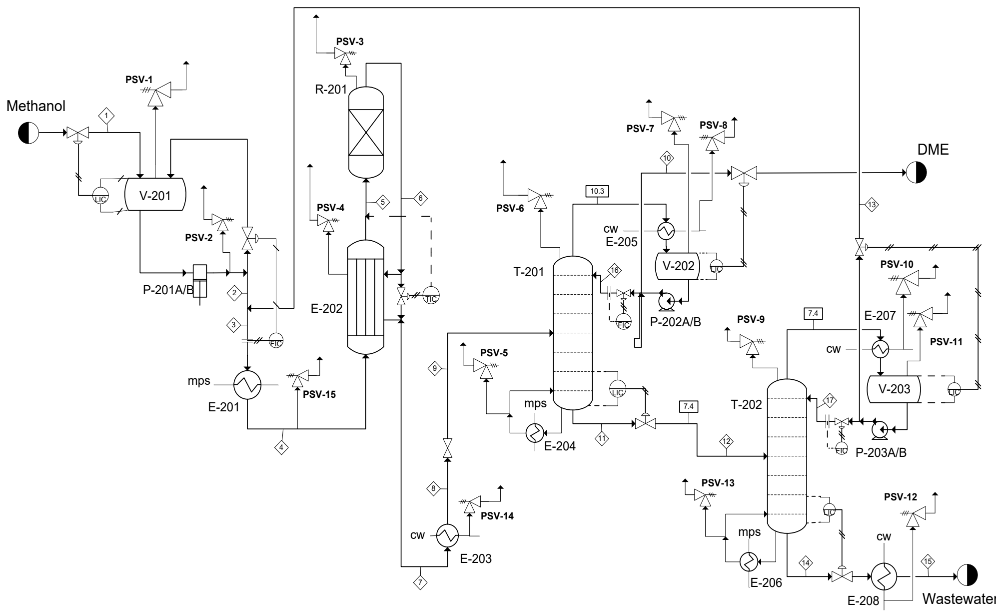

Figure 3. Relief valves location on DME plant

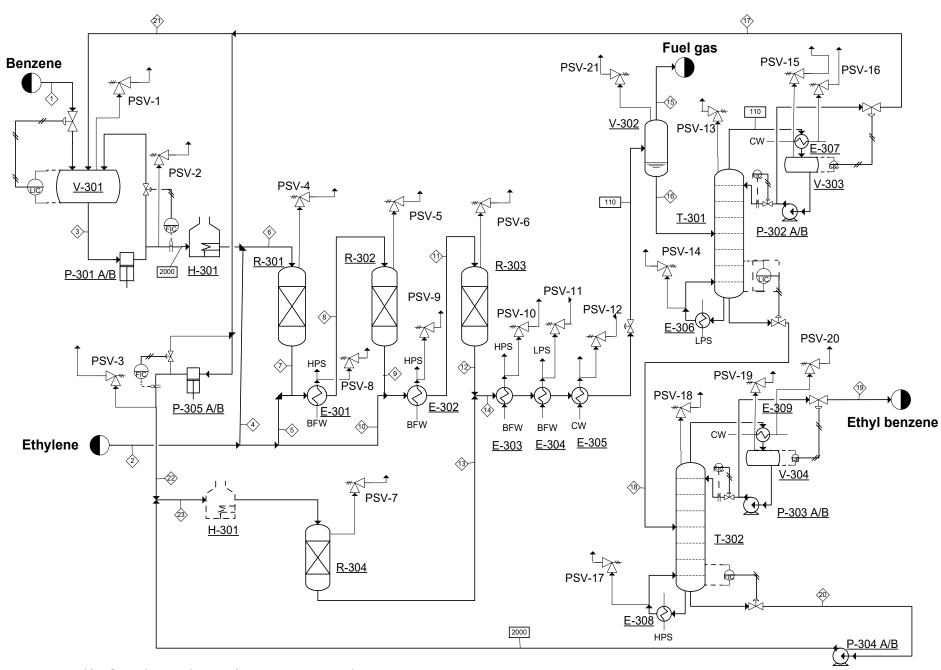

Figure 4. Relief valves location on EB plant

Relief sizing for all possible scenarios of each relief valves is summarized in Table 1 and Table 2 for the DME and EB plants, respectively.

Table 1
The possible and worst-case scenario for each relief valves on DME plant

PRV[1]	PE[2]	Possible scenario	Area (in^2)	Worst-case scenario
1	V-201	Fire Case	0.59	Automatic Control Failure
		Automatic Control Failure	1.18	
		Overfilling storage	1.18	
2	P-201 A/B	Closed-outlet	0.09	Closed-outlet
3	R-201	Chemical reactions	2.71	Chemical reactions
		Abnormal Process Heat Input	2.70	
4	E-202	Abnormal Process Heat Input	2.50	Abnormal Process Heat Input
		Heat-exchanger equipment failure	0.01	
5	E-204- Cold stream	Abnormal Process Heat Input	0.37	Abnormal Process Heat Input
		Hydraulic expansion	0.01	
		Closed-outlet at bottom T-201 outflow	0.05	
6	T-201	Accumulation of non-condensable at condenser	1.24	Accumulation of non-condensable at condenser
		Cooling failure of condenser	1.24	
		Top-tower reflux failure	1.24	
7	V-202	Closed-outlet	0.11	Cooling failure of condenser
		Cooling failure of condenser	0.53	
		Overfilling storage	0.19	
8	E-205 CW[3]stream	Abnormal Process Heat Input	0.41	Abnormal Process Heat Input
		Hydraulic expansion	0.01	
9	T-202	Accumulation of non-condensable at condenser	1.48	Accumulation of non-condensable at condenser
		Cooling failure of condenser	1.48	
		Top-tower reflux failure	1.48	

Table 1 (Continued)

PRV[1]	PE[2]	Possible scenario	Area (in^2)	Worst-case scenario
10	E-207 CW stream	Abnormal Process Heat Input	0.98	Abnormal Process Heat Input
		Hydraulic expansion	0.01	
		Cooling failure of condenser	1.49	
		Overfilling storage	0.09	
11	V-203	Closed-outlet on vessel	0.09	Cooling failure of condenser
		Cooling failure of condenser	1.49	
		Overfilling storage/ surge vessel	0.09	
12	E-208 CW Stream	Abnormal Process Heat Input	0.19	Abnormal Process Heat Input
		Hydraulic expansion	0.00	
13	E-206 Cold stream	Abnormal Process Heat Input	0.76	Abnormal Process Heat Input
		Hydraulic expansion	0.01	
		Closed-outlet at bottom T-202 outflow	0.03	
14	E-203 CW stream	Abnormal Process Heat Input	0.28	Abnormal Process Heat Input
		Hydraulic expansion	0.00	
15	E-201 Cold stream	Abnormal Process Heat Input	2.00	Abnormal Process Heat Input
		Hydraulic expansion	0.08	

Notes: [1] Pressure relief valve, [2] Protected equipment, [3] Cooling water

Table 2

Possible and worst-case scenarios for each relief valves on EB plant

PRV	PE	Possible scenario	Area (in^2)	Worst-case scenario
1	V-301	Fire Case	0.47	Automatic Control Failure
		Automatic Control Failure	0.54	
		Overfilling storage	0.54	
2	P-301 A/B	Closed-outlet	0.16	Closed-outlet
3	P-305 A/B	Closed-outlet	0.03	Closed-outlet

Table 2 *(Continued)*

PRV	PE	Possible scenario	Area (in^2)	Worst-case scenario
1	V-301	Fire Case	0.47	Automatic Control Failure
		Automatic Control Failure	0.54	
		Overfilling storage	0.54	
2	P-301 A/B	Closed-outlet	0.16	Closed-outlet
3	P-305 A/B	Closed-outlet	0.03	Closed-outlet
4	R-301	Chemical reactions	2.80	Abnormal Process Heat Input
		Abnormal Process Heat Input	2.81	
5	R-302	Chemical reactions	2.72	Chemical reactions
6	R-303	Chemical reactions	2.77	Chemical reactions
7	R-304	Chemical reactions	0.88	Abnormal Process Heat Input
		Abnormal Process Heat Input	8.56	
8	E-301 BFW[4] stream	Abnormal Process Heat Input	0.16	Abnormal Process Heat Input
		Hydraulic expansion	0.00	
9	E-302 BFW stream	Abnormal Process Heat Input	0.18	Abnormal Process Heat Input
		Hydraulic expansion	0.00	
10	E-303 BFW stream	Abnormal Process Heat Input	0.77	Abnormal Process Heat Input
		Hydraulic expansion	0.01	
11	E-304 BFW stream	Abnormal Process Heat Input	1.04	Abnormal Process Heat Input
		Hydraulic expansion	0.02	
12	E-305 CW stream	Abnormal Process Heat Input	0.40	Abnormal Process Heat Input
		Hydraulic expansion	0.01	
13	T-301	Accumulation of non-condensable	6.45	Accumulation of non-condensable
		Cooling failure	6.45	
14	E-306 Cold stream	Abnormal Process Heat Input	12.31	Abnormal Process Heat Input
		Hydraulic expansion	0.14	
		Blocked-outlet at bottom of T-301	0.41	
15	V-303	Blocked-outlet	0.46	Cooling failure
		Cooling failure	6.37	
		Overfilling storage	0.46	
16	E-307 CW Stream	Abnormal Process Heat Input	3.03	Abnormal Process Heat Input
		Hydraulic expansion	0.04	

Table 2 *(Continued)*

PRV	PE	Possible scenario	Area (in^2)	Worst-case scenario
17	E-308 Cold Stream	Abnormal Process Heat Input	8.41	Abnormal Process Heat Input
		Hydraulic expansion	0.08	
		Blocked-outlet at bottom of T-302	0.06	
18	T-302	Accumulation of non-condensable	4.76	Accumulation of non-condensable
		Cooling failure	4.76	
19	V-304	Blocked-outlet	0.42	Cooling failure
		Cooling failure	4.76	
		Overfilling storage	0.42	
20	E-309 CW Stream	Abnormal Process Heat Input	2.13	Abnormal Process Heat Input
		Hydraulic expansion	0.03	
21	V-302	Overfilling storage	0.71	Overfilling storage
		Fire Case	0.33	

Notes: [4] Boiler feed water

Worst-case scenarios relief system details for the DME and EB plants are tabulated in Table 3 and Table 4, respectively.

Relief Downstream System

Relief downstream system consisted of RV outlet piping, KO drum and flare which required to safe handling relieving materials. Thus, the design of the KO drum and flare was based on worst vapour and the liquid case of scenarios. The design is tabulated in Table 5.

Economic Analysis

An economic analysis of actual designs was performed to be used as baseline values. As a side note, 100% increment was the actual set pressure of the plants. Results and discussion of economic analysis are summarized and illustrated in Table 6 and Table 7.

Based on the obtained data, graphs of elements' cost and total cost were plotted for both plants. As illustrated, along with setting pressure increment, costs of piping and flare were deflating while costs of the protected equipment and KO drum were inflating, refer to Figure 5 and Figure 6.

Table 3

Relief sizing for DME plant, header diameter is set at 350 mm

PSV	Protected equipment	Worst-Case Scenario	W kg/h	T_{relief} °C	P_{relief} bar	P_{set} bar	Piping Diameter D_{inlet} mm	Piping Diameter D_{outlet} mm	%PD $_{inlet}$	%PD $_{outlet}$	Relief Device Type	A in^2
1	V-201	Automatic Control Failure	8370.00	69.06	1.21	1.10	150	250	1.59	3.32	Conventional	1.18
2	P-201A/B	Closed outlet at piping	8370.00	159.15	17.05	15.50	100	150	2.14	5.17	Conventional	0.09
3	R-201	Chemical reactions	10490.00	400.00	16.17	14.70	250	400	2.26	7.70	Conventional	2.71
4	E-202	Abnormal Process Heat Input	10490.00	400.00	16.50	15.00	250	350	2.84	10.10	Balanced-Bellow	2.50
5	E-204 - Cold stream	Abnormal Process Heat Input	1331.68	173.70	12.10	11.00	125	200	1.97	2.57	Conventional	0.37
6	T-201	Accumulation of non-condensable at condenser	8140.00	51.95	11.66	10.60	200	300	2.33	7.61	Conventional	1.24
7	V-202	Cooling failure of condenser	8140.00	50.81	11.33	10.30	200	350	2.40	4.14	Conventional	0.53

Table 3 *(Continued)*

PSV	Protected equipment	Worst-Case Scenario	W kg/h	T_{relief} °C	P_{relief} bar	P_{set} bar	Piping Diameter		$\%PD_{inlet}$	$\%PD_{outlet}$	Relief Device Type	A in^2
							D_{inlet} mm	D_{outlet} mm				
8	E-205 CW stream	Abnormal Process Heat Input	1391.36	183.68	11.00	10.00	150	200	1.28	8.06	Conventional	0.41
9	T-202	Accumulation of non-condensable at condenser	5750.00	129.87	8.36	7.60	250	350	1.50	7.83	Conventional	1.48
10	E-207 CW stream	Abnormal Process Heat Input	2640.92	168.51	7.70	7.00	200	300	2.19	8.30	Conventional	0.98
11	V-203	Cooling failure of condenser	5750.00	128.36	8.03	7.30	250	350	1.51	9.13	Conventional	1.49
12	E-208 CW Stream	Abnormal Process Heat Input	531.73	174.06	8.80	8.00	125	150	1.07	4.78	Conventional	0.19
13	E-206 Cold stream	Abnormal Process Heat Input	2572.57	187.73	12.10	11.00	200	250	0.86	4.82	Conventional	0.76
14	E-203 CW stream	Abnormal Process Heat Input	1103.34	199.10	15.40	14.00	125	150	1.48	4.38	Conventional	0.28
15	E-201 Cold stream	Abnormal Process Heat Input	12959.11	157.77	16.50	15.00	250	350	1.88	8.68	Conventional	2.00

Table 4
Relief sizing for EB plant, header diameter is set at 900 mm

PSV	Protected equipment	Worst-Case Scenario	W kg/h	T_{relief} °C	P_{relief} bar	P_{set} bar	Piping Diameter D_{inlet} mm	Piping Diameter D_{outlet} mm	%PD inlet	%PD outlet	Relief Device Type	A in^2
1	V-301	Automatic Control Failure	17952.20	117.04	2.75	2.50	200	300	0.73	9.66	Conventional	0.54
2	P-301 A/B	Closed outlet	17952.20	227.90	21.84	19.85	125	250	1.27	3.38	Conventional	0.16
3	P-305 A/B	Closed outlet	3130.60	228.25	22.00	20.00	65	100	2.38	5.69	Conventional	0.03
4	R-301	Abnormal Process Heat Input	18797.90	480.00	24.20	22.00	200	350	2.34	7.75	Conventional	2.81
5	R-302	Chemical reactions	19784.70	500.00	24.20	22.00	200	350	2.31	8.57	Conventional	2.72
6	R-303	Chemical reactions	20771.50	500.00	24.20	22.00	200	350	2.39	8.76	Conventional	2.77
7	R-304	Abnormal Process Heat Input	4616.50	525.00	24.20	22.00	300	450	0.02	0.04	Conventional	1.03
8	E-301 BFW stream	Abnormal Process Heat Input	871.59	221.71	24.20	22.00	80	125	1.24	5.22	Conventional	0.16

Table 4 *(Continued)*

PSV	Protected equipment	Worst-Case Scenario	W kg/h	T_{relief} °C	P_{relief} bar	P_{set} bar	Piping Diameter		%PD $_{inlet}$	%PD $_{outlet}$	Relief Device Type	A in^2
							D_{inlet} mm	D_{outlet} mm				
9	E-302 BFW stream	Abnormal Process Heat Input	1148.54	221.71	24.20	22.00	80	125	2.15	9.80	Conventional	0.18
10	E-303 BFW stream	Abnormal Process Heat Input	4466.53	221.71	24.20	22.00	150	250	1.40	7.18	Conventional	0.77
11	E-304 BFW stream	Abnormal Process Heat Input	5479.92	221.71	24.20	22.00	150	300	2.11	4.36	Conventional	1.04
12	E-305 CW stream	Abnormal Process Heat Input	2190.28	221.71	24.20	22.00	100	200	2.56	5.30	Conventional	0.40
13	T-301	Accumulation of non-condensable	18482.25	125.28	3.30	3.00	400	600	1.84	33.82	Balanced-Bellow	6.45
14	E-306 Cold stream	Abnormal Process Heat Input	26850.37	176.59	2.20	2.00	500	750	2.33	21.31	Balanced-Bellow	12.31
15	V-303	Cooling failure	18482.25	125.28	3.30	3.00	400	600	1.95	32.20	Balanced-Bellow	6.37
16	E-307 CW Stream	Abnormal Process Heat Input	3224.05	123.15	2.20	2.00	300	600	2.43	9.14	Conventional	3.03

Table 4 *(Continued)*

PSV	Protected equipment	Worst-Case Scenario	W kg/h	T_{relief} °C	P_{relief} bar	P_{set} bar	Piping Diameter D_{inlet} mm	Piping Diameter D_{outlet} mm	$\%PD_{inlet}$	$\%PD_{outlet}$	Relief Device Type	A in^2
17	E-308 Cold Stream	Abnormal Process Heat Input	15485.01	216.90	2.20	2.00	400	500	2.09	37.75	Balanced-Bellow	8.41
18	T-302	Accumulation of non-condensable	15841.71	186.63	3.30	3.00	350	750	2.20	6.84	Conventional	4.76
19	V-304	Cooling failure	15841.71	186.63	3.30	3.00	350	750	2.20	6.96	Conventional	4.76
20	E-309 CW Stream	Abnormal Process Heat Input	2331.63	123.15	2.20	2.00	300	600	1.27	4.47	Conventional	2.13
21	V-302	Overfilling storage/ surge vessel	25387.90	60.68	2.75	2.50	200	350	1.37	5.02	Conventional	0.71

Table 5

Relief downstream design

Plant		DME	EB
KO Drum	Length (m)	3.0	3.5
	Diameter (m)	1.7	2.5
	Materials of Construction	SS	SS
	Design Pressure (Barg)	16.5	24.2
Flare	Stack Diameter (m)	0.26	0.29
	Stack Height (m)	50.00	100.00
	Minimum distance from the epicentre of the flame to the object considered (m)	568.83	1104.46
	Grade distance from flare (m)	575.90	1117.40

Table 6

DME's cost fluctuation due to set pressure increment, in mil RM

P_{set} Increment	Piping Cost	PRV cost	PE Cost	KO Drum Cost	Flare Cost	Total Cost
100%	1.969	0.122	1.388	0.207	0.373	4.061
110%	1.969	0.122	1.388	0.219	0.373	4.073
130%	1.836	0.123	1.422	0.269	0.379	4.028
170%	1.675	0.123	1.465	0.343	0.361	3.966
200%	1.593	0.123	1.496	0.395	0.359	3.967
230%	1.520	0.123	1.527	0.566	0.361	4.096
300%	1.455	0.123	1.598	0.566	0.352	4.093

Table 7

EB's cost fluctuation due to set pressure increment, in mil RM

P_{set} Increment	Piping Cost	PRV Cost	PE Cost	KO Drum Cost	Flare Cost	Total Cost
100%	12.277	0.291	7.260	0.694	0.789	21.311
110%	13.335	0.245	7.583	0.754	0.780	22.697
120%	13.244	0.233	7.913	0.814	0.807	23.011
130%	13.105	0.225	8.250	0.874	0.807	23.263
140%	12.882	0.218	8.595	0.934	0.807	23.436
150%	11.670	0.212	8.947	0.994	0.807	22.632
155%	11.445	0.210	9.126	1.025	0.808	22.615
160%	11.445	0.209	9.307	1.055	0.818	22.833

Table 7 *(Continued)*

P_{set} Increment	Piping Cost	PRV Cost	PE Cost	KO Drum Cost	Flare Cost	Total Cost
170%	11.496	0.205	9.674	1.115	0.818	23.308
180%	11.431	0.203	10.049	1.175	0.818	23.675
200%	10.230	0.197	10.820	1.296	0.818	23.360
210%	10.113	0.195	11.217	1.357	0.818	23.700
220%	10.054	0.194	11.621	1.417	0.818	24.105

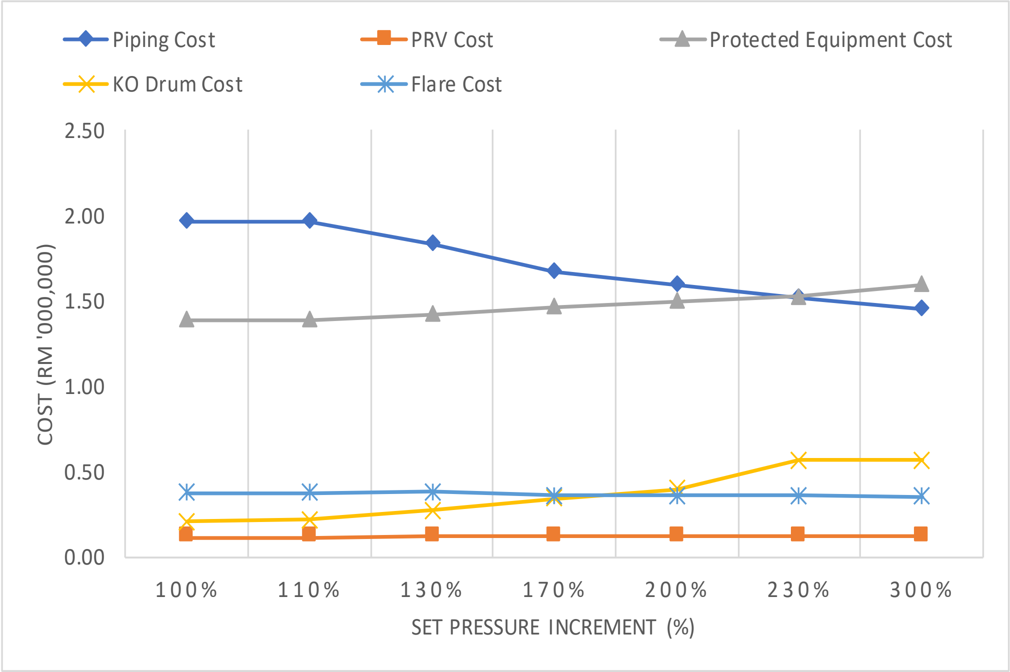

Figure 5. Elements' cost inflation and deflation with an increment of set pressure for DME plant.

This can be explained, when the set pressure is increased, more tolerable pressure drops across piping is acceptable to have within the allowable percentage of pressure drop to set pressure, Equation (4). Therefore, the piping diameter was altered to be smaller than actual design and it reduced the cost of piping. On the other hand, when the set pressure was increased, the design pressure of protected equipment was also increased to withstand higher pressure than the usual design. Hence, its price was raised. Same goes to KO drum, which was designed to withstand the maximum pressure of vapour case. Thus, design pressure of the drum was changed and so did its cost.

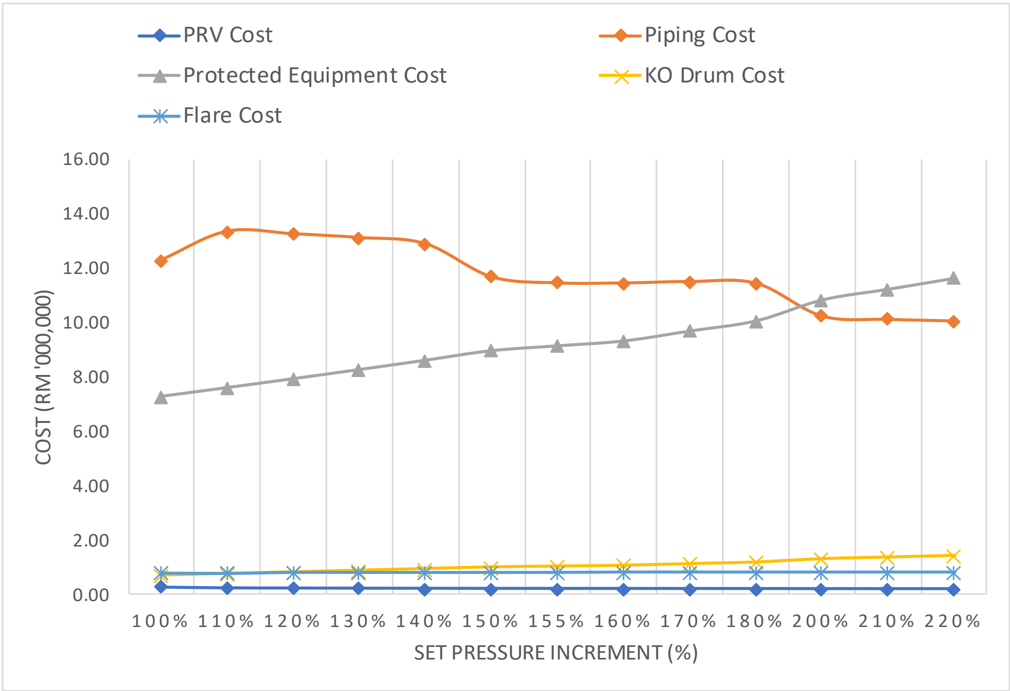

Figure 6. Elements' cost inflation and deflation with an increment of set pressure for EB plant

$$P_{drop}\% = \frac{P_{drop}}{P_{set}} \times 100\% \tag{4}$$

However, the trend of PRV costs behaved differently between the DME plant and EB plant. PRV cost of DME plant was slightly increasing while EB plant was slightly reducing when the set pressure was increased. The reason is PRV sizing (refer to relief sizing Equation (American Petroleum Institute, 2000)) is depending on two variables, upstream pressure, P_1 and backpressure, P_2 of PRV. Thus, the sizing is affected by domineering of these two variables. For case 1, when upstream pressure was dominant, increases of upstream pressure would reduce the size of the relieving area, thus reducing the PRV cost. In contrast, case 2 was when backpressure was more dominant than upstream pressure, increases of backpressure resulting increased of relieving area. For a side note, the backpressure was increased when pressure drops throughout the piping were raised throughout reducing of piping diameter process. So, by looking at the trends given by both plants, PRV cost of DME is falling under case 2 while PRV cost of EB is case 1.

Following, the total cost of DME and EB plants are plotted as in Figure 7 and Figure 8. As for the DME plant, it shows good expecting results when setting pressure was changed. In Figure 7, it shows that at 170% increment of set pressure, the total cost of the plant was at the minimum point compared to other increments including the original plant's cost. Despite DME plant's result, the minimum point of EB plant remained at the actual design of the plant even though there was a peak at range of 150% to 155%, but it was not the lowest point compared to the original design. In conclusion, EB actual plant design is the optimum design and does not require to increase set pressure of all PRVs to have lower total cost while DME shows its optimum plant design at set pressure increment of 170%.

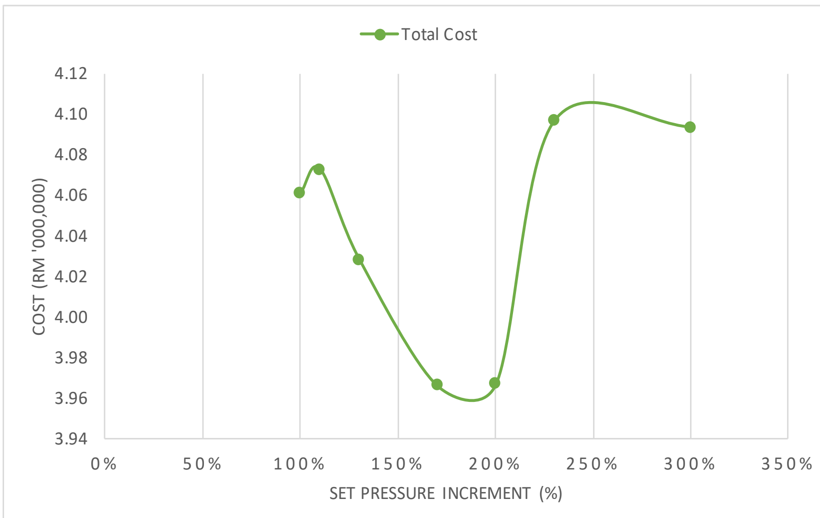

Figure 7. Changes of DME plant's total cost vs increment of set pressure

Figure 8. Changes of EB plant's total cost vs increment of set pressure

CONCLUSION

Pressure relief system is a precautionary system towards unwanted incidence inside the plant which dealing with excessive pressure built inside the plant system. The system consists of a relief device, piping and disposal system. In this project, the pressure relief system is designed by using a different approach than the conventional method. The different approach is by increasing pressure for protected equipment, changes of backpressure of PRV and disposal system design. The manipulation process affects the total cost of plant design due to higher allowable pressure drops across piping hence, the smaller pipe diameter is required and increase of protected equipment cost due to increment of design pressure. From that, a comparison of DME's total cost gives a minimum point at 170% of set pressure increment. On the other hand, EB plant's minimum point is at the original plant design despite a low peak at range of 150% to 155%, which still has a higher cost than the original design.

Recommendation

Based on the study done, there is an issue to determine a reasonable interval of set pressure increment which is to be implemented in industrial equipment design. This is because increasing the design pressure of equipment for optimization of the relief system may not be feasible anymore when the set pressure is increased further. Secondly, we recommend postulating a maximum point of pressure increment for feasible relief system.

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