Sizing Relief Systems for High Viscosity Two-Phase Flow

Georges A. Melhem
ioMosaic Corp.

Qualify your design methods for high-viscosity two-phase relief systems against these simple benchmarks.

HIGH VISCOSITY TWO-PHASE (HVTP) FLOW occurs in many industrial scale reactors, particularly when runaway reactions (*e.g.*, during polymerizations) are vented through an emergency relief system. The design of a relief system for two-phase discharge can be complicated, as it involves a fluid with a liquid-like density and a gas-like compressibility. Moreover, the fluid may flash as it loses pressure, achieving choked flow at the valve's exit and thus, limiting the flow capacity of the relief system. The literature indicates that 30–40% of the relief devices that are in existence violate industry guidelines for inlet pressure drop and backpressure *(1)*. These studies followed the Occupational Safety and Health Administration. (OSHA; Washington, DC; www.osha.gov) Process Safety Management (PSM) rule, which requires that the relief device design basis be documented and verified. Many of these installed relief systems were designed using best industry practices, such as the American Petroleum Institute's API-520, which details how to design emergency relief devices for low viscosity single-phase (gas, steam, liquid) systems.

However, there are currently no broadly accepted methods for designing a relief system for HVTP flow. Recent editions of API-520 included changes for sizing relief systems for two-phase flow using DIERS technology. API-520 also includes the addition of a viscosity correction factor, K_V, for high viscosity flow systems. In July 2002, the AIChE Design Institute of Emergency Relief Systems (DIERS) Users Group released a consensus-based best practice called SuperChems* for DIERS *(2)* that incorporates two methods for handling high viscosity flow through a relief valve — a volume-based HVTP flow model and a slip-flow model *(3, 4, 5)*.(Slip flow refers to the flow characteristics in a two-phase stream whereby the vapor travels faster than the liquid.) Respectively, these models are referred to as the nozzle method and the pipe method (Box).

A viscosity correction factor, K_V, is used to account for losses in pressure in the nozzle due to entrance effects, and a pipe representation is utilized to account for the impact of wall shear (in the inlet/discharge piping) on pressure drop *(6, 7)*. The calculations are based on the models established by Grolmes *(10, 11)* and Melhem *(3)*. This article addresses the applicability and use of the SuperChems methodology in order to improve and verify the design and performance of existing relief systems made for HVTP flow.

Safety valve representation

Recent DIERS-sponsored research on HVTP flow suggests that a safety relief valve can be represented using a simple pipe representation of a nozzle. This model does not require knowledge of a viscosity correction factor. Rather, it relies on wall shear to produce the effects of viscosity on pressure drop (dP) and flow reduction. SuperChems' pipe flow solutions are calculated by solving differential equations that represent mass, momentum, energy and physical equilibrium. In addition to components that account for the contributions of acceleration, friction and gravity to pressure drop through the valve, SuperChems defines a velocity head loss term, k, to account for entrance, geometry and laminar-flow development effects *(3)*. In Eq. 1, k depends on valve lift and and the velocity head-loss contributions due to laminar and turbulent flow:

$$k = (1/(k_{lift,i})^2)((1/(k_{lift,BP})^2) \times (k_{ent,turb} + 1 + k_{lam}) \qquad (1)$$

where $k_{lift,i}$ represents the reduction in effective flow area of the relief device caused by excessive inlet pressure loss, $k_{lift,BP}$ represents the reduction in effective flow area caused by excessive backpressure (BP), and k_{lam} and

*SuperChems is a trademark of ioMosaic Corp. SuperChems for DIERS is sold by AIChE.

Nomenclature

A	= valve flow area, in^2
C_d	= discharge coefficient, dimensionless
D	= inside diameter of pipe, in.
FRF	= flow reduction factor or C_d corrected for viscosity effects on flow through a nozzle
I	= inferred viscosity correction based on the pipe representation of the nozzle
k	= velocity head loss, dimensionless
K_V	= viscosity correction factor, dimensionless
L	= length of pipe or nozzle, in.
M	= actual or estimated flowrate, lb/h
M_{ideal}	= flowrate when C_d = 1; equal to 919,221 lb/h in Benchmark 1
N_{Re}	= Reynolds number, dimensionless
P	= pressure, psia or psig
Ps	= system set pressure, psia or psig
SR	= slip ratio; ratio vapor velocity to liquid velocity, dimensionless
T	= temperature, °C
X	= inlet vapor quality, dimensionless

Greek Letters

μ	= fluid viscosity, cP
β	= ratio of nozzle dia. to inlet dia., dimensionless
ρ	= density, lb/gal
ψ	= laminar velocity head loss parameter, dimensionless
ξ	= turbulent velocity head loss parameter, dimensionless

Subscripts

BP	= backpressure
c	= choked conditions
d	= discharge
DM	= Darby-Molavi
ent,turb	= turbulent conditions at entrance of relief device
f	= fluid
Grolmes	= calculated using methods in Refs. 10–11
lift	= relief device lift
i	= inlet
l	= liquid
lam	= laminar
o	= outlet
SuperChems	= calculated using SuperChems software
v	= vapor

Table 1. Values of ξ and ψ as functions of β.

β	ξ	ψ
0.1	1,086	0.192
0.2	858	0.112
0.3	679	0.067
0.4	522	0.037
0.5	380	0.014
0.6	254	0
0.7	146	0
0.8	63	0
0.9	14	0

$k_{ent,turb}$ represent the velocity head-loss contributions due to laminar and turbulent flow, respectively. Lift characteristics are available from valve manufacturers. If manufacturer data is not available for a specific model, one can use data published by API. In this article, the slip flow model is used to extrapolate $k_{ent,turb}$ for HVTP flow after the model is fit from the single-phase data.

The turbulent entrance component, $k_{ent,turb}$, can be estimated from the valve manufacturer's reported discharge coefficient (C_d), or preferably, can be established by requiring the pipe representation of the valve to flow the valve's reported capacity for air or steam.

$$k_{ent,\,turb} \approx (1/C_d^2 - 1)(1 - \beta^4) \qquad (2)$$

Often, using the valve manufacturer's reported C_d works well, even though it only applies to single-phase flows. Recent work by Darby recommends the use of an all-vapor C_d for choked, flashing two-phase flow and an all-liquid C_d for subsonic flow *(6)*. For most systems of practical interest in relief design for flashing two-phase flow, an all-gas C_d will work well.

The laminar velocity head contribution to pressure drop, k_{lam}, is a strong function of the Reynolds number (N_{Re}), but will also depend on the valve geometry to some extent as the flow profile develops *(10, 11)*. k_{lam} is most important for high viscosity liquids and for short pipes (Box). The value of k_{lam} approaches zero at N_{Re} > 3,100 and infinity as N_{Re} approaches zero.

The author will show that k_{lam} is well represented by the Darby-Molavi viscosity correction factor, k_{DM}, for both all-liquid flow and two-phase flow *(8, 9)*:

$$k_{DM} = 0.975(\beta^{0.1}/(0.9 + (950(1 - \beta)^{1.4})/N_{Re}))^{0.5} \qquad (3)$$

However, the Darby-Molavi factor for viscosity correction defined in Eq. 3 is actually a discharge coefficient. It must be converted to the equivalent velocity head-loss correction form. Further, it can be shown that k_{DM} collapses to the k_{lam} velocity head form:

$$k_{lam} = \xi/N_{Re} + \psi \qquad (4)$$

where,

$$\xi = (950(1 - \beta^4)(1 - \beta)^{1.4})/(0.95\beta^{0.1}) \qquad (5)$$

and

$$\psi = 0.9(1 - \beta^4)/(0.95\beta^{0.1}) - (1 - \beta^4) \qquad (6)$$

The data in Table 1, calculated using Eqs. 5 and 6, suggest that k_{DM} is derived from Equation 4, since the contributions of ψ are very small compared to ξ. For β values ranging from 0.1 to 0.9, the value of ψ is negligible.

Grolmes arrived independently at a velocity head correction factor for a fully open valve that is very similar to the form used by SuperChems *(10, 11)*. Eqs. 1–4 illustrate that the SuperChems form of the velocity correction factor is based on k_{DM}.

$$k_{Grolmes} = k_{ent,turb} + 1 + 0.576/N_{Re} \qquad (7)$$

$$k_{SuperChems} = k_{ent,turb} + 1 + \xi/N_{Re} \qquad (8)$$

Key Modeling Questions

These questions form the basis of understanding the requirements for accurate modeling of high viscosity two-phase (HVTP) flow.

1. How does one calculate a two-phase viscosity (μ) to use for the estimation of two-phase pressure drop (*dP*) in the inlet line and the outlet line? A common technique is the use of volume averaging — *i.e.*, take the volume fraction occupied by the vapor and multiply that by the vapor viscosity (μ_v) and then take the volume fraction occupied by the liquid and multiply that by the liquid viscosity (μ_l). This is shown to work well for low viscosity flow. However, high viscosity flow will exhibit flow separation. The vapor will travel faster than the liquid, and as a result there is no easy way of estimating a two-phase Reynolds number (N_{Re}).

2. How do I compute N_{Re} *for* a two-phase mixture? This depends on how one computes the two-phase viscosity, density and velocity. One has to invoke a mixing/combining rule or treat the phases separately and address mass, momentum and energy transfer across phases.

3. For a two-phase mixture, the choke point is influenced by the vapor mass fraction, *X* (also known as vapor quality), the volume (or void) fraction occupied by vapor, and μ_v. How do I estimate the *X* and associated *dP* at the right location? For single-phase flow, a common practice is to take all the fittings and piping segments and reduce them into a combined flow-resistance factor and then estimate *dP*. In two-phase flow, this will lead to erroneous designs. Pressure drop due to piping fittings and flow resistance must be taken as a function N_{Re} and at the right location in the flow path.

For a fixed upstream pressure, as the downstream pressure is decreased, the flowrate of a fluid across a conduit will increase until it reaches sonic flow at the duct exit. At this point, the flow is said to be choked, and further reduction in downstream pressure will have no effect on the mass flowrate (*M*). Pressure drop leads to the formation of more vapor, which alters the choke point, and thus, the flow capacity estimates. For choked flow, the upstream conditions and the size of the orifice or flow duct determine *M*. An interesting example showing the choking effect has been published by Melhem and Fisher *(4, 5)*.

4. Does a HVTP mixture separate in the relief valve or discharge pipe? Yes, high viscosity flow will separate in the discharge line due to vapor-liquid slip (slip is when a vapor travels at a faster velocity than the liquid). Slip increases as μ_l increases, and thus, can lead to higher pressure drops in the discharge line. For the discharge pipe, a length exceeding 35 *L*/*D* (at which point flow profiles are fully established) may result in separated flow, depending on the fluid's overall viscosity and the velocity of the flowing liquid-vapor mixture.

However, establishing a viscosity cutoff number is difficult. It is best to use a correction factor that depends on the inverse of the fluid's N_{Re}, so that as the N_{Re} decreases the correction becomes more significant. This does not apply to non-Newtonian fluids.

Recent research sponsored by the Design Institute for Emergency Relief Systems (DIERS) Users Group has shown that HVTP flow through relief valves is best represented by the homogeneous equilibrium (or no slip) flow and viscosity model, also called a volume-averaged, two-phase viscosity system. In fact, several publications over the past 30 years suggest that a volume-averaged, two-phase viscosity should be used in design problems involving HVTP flow.

Variations on the volume-averaged theme have also been published, assigning different weighting factors to the vapor or liquid portion of the flow. But, these various techniques can lead to pressure losses for HVTP flows in short pipes that differ by 25–50%. Darby recommends the use of two times the slip ratio (*SR*) calculated by the Hughmark slip model for nozzle flow *(6)*.

Two-phase mixture separation has a profound influence on how pressure drops are estimated in the discharge line for HVTP flow. Preliminary DIERS findings suggest that short discharge lines (< 500 *L*/*D*) can be undersized by one to two pipe sizes if *dP* is estimated with no slip. This design flaw can lead to valve chatter and inadequate venting capacities.

Valve geometry also comes into play when dealing with *dP* for two-phase flow. For example, when using the homogeneous no-slip flow and viscosity model, a valve with a constant diameter bore that is 4 in. or longer should be the valve of choice for HVTP flow because this is the length required to establish equilibrium *(7)*. DIERS has indicated that homogeneous equilibrium for low viscosity two-phase flow is likely to be established in less than 4 in.

5. How sensitive is the final design to small changes in inlet vapor quality? Very sensitive. At *X* values ranging from 0 to 1%, the void fraction ranges approximately from 0 to 90%. Even for non-viscous two-phase systems, the presence of slip in the inlet line can result in higher pressure drops and larger inlet-line size requirements.Therefore, inlet pressure-loss estimates should consider slip for non-viscous flow. One must always account for slip when estimating the pressure loss in the inlet line for HVTP flow. The allowable inlet *dP* is restricted by industry guidelines to 3% of the set differential pressure of the valve.

The examples and benchmarks in the next section will illustrate that when modeling relief systems for HVTP flow, a nozzle representation (corrected for high viscosity flow) of the flow system and a pipe representation of a flow system will yield essentially the same flowrates and reproduce the same viscosity correction factors.

DIERS benchmarks

The following benchmarks are designed to help operating companies determine if their current design methods will work for systems with HVTP flow. The benchmarks also illustrate the impact of *X* and slip on flow estimates, relief capacity and discharge pipe backpressure. These benchmarks arbitrarily use water as the flow medium and adjust its viscosity for illustrative purposes.

Benchmark 1 — All-liquid viscous flow. The benchmark system uses a 4P6 bellows safety relief valve with: $A = 6.38$ in^2; $C_d = 0.71$; and $P = 52.5$ psig. The operating conditions of the fluid are: $T = 40°C$; $P_f = 72.6$ psia; and $P_{BP} = 14.7$ psia. Fluid ρ = 8.274 lb/gal and its μ will be varied arbitrarily from 1 cP to 100,000 cP. The value of $k_{ent,turb}$ is estimated from Eq. 2 and is adjusted by adding 0.191 to the result to account for geometry effects. The characteristics of the pipe representation of the valve are $L = 6$ in., $D = 2.85$ in. and a pipe surface roughness of 0.0018 in. The results were generated using SuperChems for DIERS.

An ideal nozzle ($C_d = 1$) flowing low viscosity water will produce an ideal flowrate (M_{ideal}) of 919,221 lb/h. Table 2 data are reported for both the pipe solution and a simple nozzle representation using $K_{V,I}$ (a "corrected" form of C_d that accounts for viscosity effects when a pipe representation of the nozzle is used), and k_{DM}. This benchmark shows that both pipe and nozzle solutions predict the same viscosity correction factors ($K_{V,I}$ and k_{DM}) over a wide range of viscosities. Note that the laminar velocity head contributions become negligible at $N_{Re} > 1{,}500$. Also note the significant flow reduction at high liquid viscosity.

Table 2. Comparing a 4P6 pipe representation solution with nozzle solution ($C_d = 0.71$).

	Pipe Solution			Nozzle Solution		
μ, cP	M, lb/h	$K_{V,I}$	N_{Re}	M, lb/h	k_{4DM}	N_{Re}
1	655,692	0.994	2,148,030	652,580	1.000	2,531,943
10	655,297	0.992	145,209	652,580	1.000	171,617
100	651,863	0.988	14,444	652,580	1.000	17,161
1,000	633,583	0.960	1,404	624,295	0.956	1,641
5,000	522,570	0.792	232	503,818	0.772	265
10,000	416,196	0.630	93	391,785	0.600	103
100,000	67,819	0.102	1.5	60,073	0.092	1.57

Benchmark 2 — Two-phase flashing viscous flow and all gas flow. Benchmark 2 uses the same safety relief valve as Benchmark 1. The fluid is saturated at $T = 151.85°C$ and $P = 72.6$ psia. C_d is arbitrarily specified as 0.91 and the water-like fluid viscosity is held constant at μ = 5,000 cP. All other process parameters remain the same as in Benchmark 1.

Table 3 summarizes estimates of M for an ideal theoretical nozzle without frictional losses, where M is the maximum possible flow through the valve. X is specified as an input value at saturation conditions. The choke quality, X_c, is estimated using SuperChems for DIERS via an isentropic flow path that employs the Melhem modification of the Peng-Robinson equation of state *(3)*. Note that the huge impact of X on M is most significant from $X = 0.0001$ to $X = 0.01$. But at $X > 0.5$, the impact of X is much smaller, indicating that the fluid is behaving like a vapor.

Table 3. Ideal nozzle estimate of low-viscosity, flashing two-phase flow ($C_d = 1$)

X	M, lb/h	P_c, psia	T_c, °C	X_c	$N_{Re,c}$
0.0001	123,957	65.75	148.26	0.007	197
0.001	121,380	65.62	148.19	0.008	213
0.01	104,673	58.49	143.99	0.025	525
0.1	62,454	48.74	137.52	0.122	1,883
0.5	32,962	43.44	133.55	0.502	7,947
0.8	26,773	42.74	132.99	0.782	23,243
0.95	24,771	42.61	132.88	0.922	70,297
0.98	24,426	42.59	132.87	0.950	110,413
0.9999	24,223	42.57	132.85	0.967	266,928

Table 4 summarizes estimates of M for a nozzle representation of a valve, with velocity losses represented by C_d and a K_V. These flow estimates would be used for design purposes if the corresponding correction factors are utilized. The API-520 K_V correction is often used in lieu of the Darby-Molavi form (k_{DM}). However, the Darby-Molavi correlation is in agreement with the API curve, which has large error bounds, but is more general because it accounts for the effect of nozzle size on K_V. For this reason, the author prefers the Darby-Molavi correlation.

Table 4. Ideal nozzle estimates of high-viscosity, flashing two-phase flow with viscosity correction ($C_d = 0.91$).

X	M, lb/h	P_c, psia	T_c, °C	X_c	$N_{Re,c}$	μ_c, cP	$K_{V,c}$
0.0001	82,530	55.99	142.41	0.018	333	549.538	0.807
0.001	82,713	56.09	142.48	0.019	345	530.572	0.812
0.01	80,622	55.17	141.88	0.029	503	355.131	0.861
0.1	54,706	46.94	136.20	0.124	1,801	67.283	0.961
0.5	29,978	43.27	133.41	0.502	7,532	8.818	0.997
0.8	24,230	42.84	133.06	0.782	21,812	2.461	1.000
0.95	22,358	42.65	132.91	0.922	65,918	0.751	1.000
0.98	22,033	42.61	132.88	0.950	103,554	0.471	1.000
0.9999	21,845	42.59	132.87	0.967	323,231	0.311	1.000

Table 5 summarizes the estimates of M obtained by using a piping representation of the relief valve, as opposed to a nozzle estimate, as shown in Table 4. The flow reduction factor (FRF) is calculated by dividing M from Table 5 (second column) by the ideal nozzle flowrate estimate, represented by M in Table 3, with no loss corrections. This FRF would be equal to the product of K_V and C_d. $K_{V,I}$ is obtained by dividing FRF by C_d:

$$K_{V,I} = FRF / 0.91 \tag{9}$$

The last column of Table 5 shows that the viscosity correction inferred for the pipe solution is essentially the same as the Darby-Molavi viscosity correction and produces the same M values as the nozzle solution with viscosity correction, as shown in Table 4.

These benchmarks show that a one-dimensional piping representation of a relief device produces the same solution as a nozzle representation for HVTP and low viscosity two-phase (LVTP) flow. A piping representation of a relief device has significant advantages over a nozzle representation because a two-phase discharge coefficient does not need to be specified for two-phase flow covering a wide range of flow types (*e.g.*, flashing and frozen flows, where frozen refers to a two-phase stream in which the liquid does not flash, such as air and water at ambient conditions) and because a pipe solution more accurately describes the relief-device flow path.

Benchmark 3 — Pressure drop in inlet and discharge piping for viscous two-phase flow. A key finding of the DIERS research program is that HVTP flow will separate in the discharge line, resulting in slip flow and thus a higher-pressure drop in the discharge line. But, this is not true for all flow types. Inlet pressure-loss estimates should consider slip even for non-viscous flow. Preliminary findings suggest that a short discharge (< 500 L/D) line can be undersized by one or two pipe sizes if dP were estimated with no slip.

With this in mind, a discharge line is added to Benchmark 2. The required discharge line diameter is estimated in order to reach 30% backpressure (maximum recommended backpressure for a bellows device). The discharge line is composed of a horizontal segment (1-ft long), one 90-deg. elbow ($k = 800/N_{Re} + 0.3$) and a vertical segment (L = 7 ft long). The fluid viscosity is μ = 5,000 cP at T = 151.8°C and μ = 14,305 cP at 100°C.

Table 6 shows the impact of slip on dP in the discharge line for X = 0.0001. The slip estimates were generated in SuperChems for DIERS, using the Duns and Ross pressure drop correlation *(3)*. This correlation is described in detail by Aziz and Govier as best in class, but complex to implement *(12)*. The Hughmark slip model *(6)* is also commonly used. The Moody slip model is found to predict nearly the same flowrates for pipe flow as those calculated using low viscosity experimental data *(4)*. Darby recommends the use of two times the slip ratio (SR) predicted using the Hughmark correlation for nozzle flow *(6)*. Grolmes *(10, 11)* suggests that $SR \approx 10$ is reasonable for high viscosity flow.

The results in Table 6 show that the discharge line nominal diameter should be 6 in. using the homogeneous (or no-slip model, 8 in. using the Moody slip model, and 10 in. using the Fauske slip model. These are the line sizes at which the 30% BP is reached. Table 6 is not meant to suggest a heuristic of adding two line sizes to the discharge line when using a no-slip model; rather it illustrates the importance of slip on the required discharge diameter. of the pipe.

Table 5. Flow estimates of high-viscosity flashing two-phase flow using pipe representation with viscosity correction (K_V).

X	M, lb/h	P_c, psia	T_c, °C	X_c	FRF	$K_{V,I}$	$K_{V,I}/K_{DM}$
0.0001	83,259	47.20	136.40	0.029	0.67	0.74	0.915
0.001	82,521	46.86	136.14	0.031	0.68	0.75	0.921
0.01	75,674	44.68	134.50	0.042	0.72	0.80	0.920
0.1	48,019	37.42	128.52	0.137	0.77	0.84	0.876
0.5	27,477	35.95	127.20	0.515	0.83	0.92	0.917
0.8	23,050	36.07	127.31	0.796	0.86	0.95	0.953
0.95	21,456	36.80	127.97	0.937	0.87	0.96	0.962
0.98	21,161	36.80	127.97	0.966	0.87	0.96	0.963
0.9999	20,984	36.80	127.97	0.981	0.87	0.96	0.963

Table 6. Impact of slip ratio (SR) on percent backpressure (%BP) for high-viscosity flow in a discharge line.

Slip Model	SR_i	SR_o	D, in.	%BP	M, lb/h
Homogeneous	1	1	6	33.52	83,909
Moody	8.2	11.4	6	50.50	83,672
Moody	9.1	11.4	8	26.67	83,909
Fauske	24.9	40.07	8	42.54	83,909
Fauske	27.97	40.08	10	28.32	83,909

The viscosity values and physical properties used here may be very different for other high viscosity flow systems such as those involving polymers. (Note: polymeric fluids with high viscosities are inevitably non-Newtonian). These results should also apply to frozen flows. The total discharge line length used here is 8 ft. Actual installations typically have discharge lines of 50 ft or 100 ft connecting into flare headers, vent-containment systems or other equipment. For situations with high viscosity flows, the use of rupture disks should be considered and may even be preferred.

Recommendations

The following design recommendations apply to both HVTP and LVTP flow systems:

1. Use a homogeneous equilibrium model (*i.e.*, no slip) to represent a safety relief valve that has a constant diameter bore of four inches or greater; otherwise, use a homogeneous non-equilibrium model or a slip model such as the one recommended by Darby *(6)*.

2. Use a slip flow model to estimate dP and BP for the inlet and discharge lines.

3. A piping representation of a relief valve is preferred over an ideal nozzle representation. Estimate $k_{ent,turb}$ using published manufacturer air or steam flow data with Eq. 2. If flow data are not available, estimate $k_{ent,turb}$ using the

manufacturer's published C_d for your safety relief system.

The appropriate C_d depends on whether the flow in the nozzle is choked. Since high viscosity flows are "slower" than low viscosity flows, choking may be less likely to occur, and the use of a liquid C_d may be more appropriate. Using a piping representation eliminates the guess work in establishing C_d values for different types of flows, such as hybrid, flashing, etc. CEP

GEORGES MELHEM is president of ioMosaic Corp. (93 Stiles Road; Salem, NH 03079; Tel: (603) 893-7009; Fax: (603) 893-7885; E-mail: melhem@iomosaic.com). Prior to working with ioMosaic, Melhem was president of Pyxsys Corp., a technology subsidiary of Arthur D. Little (ADL), and vice president of ADL's Global Safety and Risk Management Practice. He is an internationally known pressure relief design, chemical reaction systems, and fire and explosion dynamics expert. In this capacity, he provides consulting and design services, expert testimony and incident investigation support and reconstruction for a large number of clients. Melhem holds BS, MS and PhD degrees in chemical engineering, as well as a minor in industrial engineering, from Northeastern Univ., and has completed executive training in finance and sales management at the Harvard Business School.

Literature Cited

1. **Shelly, S.,** "Beware: Your Overpressure Protection May be Inadequate," *Chem. Eng.*, **106** (4), p. 58 (April 1999).
2. **Fisher H. G., *et al.*,** "Emergency Relief Systems Design Using DIERS Technology," AIChE, New York, p. 84 (1992).
3. **Melhem, G. A.,** "SuperChems Expert Version 5.4 Reference Manual," ioMosaic Corp. (2003).
4. **Melhem, G. A. and H. G. Fisher,** "An Overview of SuperChem for DIERS for Emergency Relief Systems and Vent Containment Design," *Process Safety Progress*, **16** (3), pp. 185–197 (July 1997).
5. **Melhem, G. A., H. G. Fisher and D. A. Shaw,** "An Advanced Method for the Estimation of Reaction Kinetics, Scale-up and Pressure Relief Design," *Process Safety Progress*, 14 (1) (1995)
6 **Darby, R., F. Self and V. Edwards,** "A Direct Integration Method for Sizing Relief Devices For Two-Phase (Gas/Liquid) Flow," Paper Presented at the DIERS Users Group Meeting, Las Vegas, NV (2003).
7. **Fletcher, B.,** "Flashing Through Orifices and Pipes," AIChE Loss Prevention Symposium, Denver, CO (1983).
8. **Darby, R.,** "Viscous Two-Phase Flow in Safety Relief Valves. Phase I. Present Status and Recommended Design Procedures," DIERS Users Group Final Report, AIChE/DIERS, New York (Nov. 20, 1997).
9. **Darby, R., and K. Molavi,** "Viscosity Correction Factor for Emergency Relief Valves," *Process Safety Progress*, **16** (2), pp. 80–82 (1997).
10. **Grolmes, M.,** "Progress Report on High-Viscosity Two-Phase Flow Modeling," DIERS Users Group (April 1997).
11. **Grolmes, M.,** "High-Viscosity Two-Phase Flow Modeling," DIERS Users Group Draft Final Report (June 2002).
12. **Govier, G. W., and K. Aziz,** "The Flow of Complex Mixtures in Pipes," Robert E. Krieger Publishing Co. (1987).