TWO-PHASE BLOWDOWN FROM HIGH-PRESSURE LIQUID PIPELINES

Jayant Sabnis,* Wayne Simmons,† Lynn Teuscher*

A computational procedure is presented for predict‐ ing the two-phase blowdown rate resulting from the Guillotine break of a high-pressure pipeline. The procedure presented is suitable only for a pure substance, such as propane, because no analyt‐ ical model is available at present to predict the mass-limited, two-phase flow rate of multicomponent mixtures, which governs the boundary condition at the exit. If such a model becomes available, however, the procedure can be modified easily to accommodate the multicomponent mixtures, such as NGL.

INTRODUCTION

The failure of a high-pressure liquid pipeline would result in rapid depressurization of the liquid in the pipeline, causing part of the liquid in the pipeline to flash into vapor phase. The resultant liquid-vapor mixture would initially consist of vapor bubbles entrained in the flow of liquid and the size of these bubbles will increase as the pressure reduces. Eventually, the bubbles will coalesce forming a core of vapor, and the flow will become essentially annular.

The flow at the exit will be mass limited or critical. Mass limiting flow is characteristic of compressible fluid systems. In single phase flow, the critical velocity is such that it is no longer possible to transmit a pressure disturbance upstream of the point where the flow becomes critical (1). In two-phase flow, a maximum flow rate is also observed. The critical phenomena (e.g., pressure propagation) are less well-defined, however. The impor‐ tance of critical two-phase flow in process engineering equipment has led to a fairly extensive literature on the subject and to the development of a variety of analytical models for the calculation of the appropriate critical flow rates. Hewitt and Hall-Taylor (2) have provided a breif survey of various models. The critical flow rate will determine the boundary condition at the exit during the two-phase blowdown from high-pressure liquid pipelines, until the pressure in the pipe at the exit equals the ambient pressure.

* Energy Resources Co. Inc., La Jolla, CA, U.S.A.
† Department of Mechanical Engineering, University of Wisconsin, Madison, WI, U.S.A.

GOVERNING EQUATIONS

The flow pattern most likely to prevail inside the pipelines of interest is annular because of the high vapor velocities that are expected to exist. The flow will be assumed to be one dimensional; i.e., the velocity of either phase will be assumed to change along the pipe axis only. At any section, the velocity of the vapor phase will, in general, be different from that of the liquid phase at that section. The two phases will also be assumed to be in the thermodynamic equilibrium.

The mass conservation equation for the liquid phase can now be written as:

$$\frac{\partial}{\partial t}(\rho_\ell A_\ell) = -\frac{\partial}{\partial z}(\rho_\ell A_\ell u_\ell) - \dot{m}_{evp} \tag{1}$$

where ρ_ℓ = density of the liquid phase

A_ℓ = area occupied by the liquid phase

u_ℓ = velocity of the liquid phase

$\dot{m}_{evp}$ = evaporation rate per unit length

Dividing (1) by the cross sectional area of the pipe, A, results in:

$$\frac{\partial}{\partial t}(\rho_\ell \alpha_\ell) = -\frac{\partial}{\partial z}(\rho_\ell \alpha_\ell u_\ell) - \frac{\dot{m}_{evp}}{A} \tag{2}$$

where $\alpha_\ell = \dfrac{A_\ell}{A}$ = liquid volume fraction

Similarly, the mass conservation equation for the vapor phase may be written as:

$$\frac{\partial}{\partial t}(\rho_g \alpha_g) = -\frac{\partial}{\partial z}(\rho_g \alpha_g u_g) + \frac{\dot{m}_{evp}}{A} \tag{3}$$

where the subscript g now denotes the corresponding quantities for the vapor phase, so that

$$\alpha_g + \alpha_\ell = 1 \tag{4}$$

Adding (2) and (3) gives the global mass conservation equation as:

$$\frac{\partial}{\partial t}(\rho_\ell \alpha_\ell + \rho_g \alpha_g) = -\frac{\partial}{\partial z}(\rho_\ell \alpha_\ell u_\ell + \rho_g \alpha_g u_g) \tag{5}$$

The momentum equations for the liquid and vapor phases, respectively, can be written as:

$$\frac{\partial}{\partial t}(\rho_\ell \alpha_\ell u_\ell) = -\frac{\partial}{\partial z}(p\alpha_\ell) + p\frac{\partial \alpha_\ell}{\partial z} - \frac{\partial}{\partial z}(\rho_\ell \alpha_\ell u_\ell^2)$$
$$- u_\ell \frac{\dot{m}_{evp}}{A} - \frac{F_\ell}{A} \tag{6}$$

$$\frac{\partial}{\partial t}(\rho_g \alpha_g u_g) = -\frac{\partial}{\partial z}(p\alpha_g) + p\frac{\partial \alpha_g}{\partial z} - \frac{\partial}{\partial z}(\rho_g \alpha_g u_g^2)$$
$$+ u_\ell \frac{\dot{m}_{evp}}{A} - \frac{F_g}{A} \tag{7}$$

where F_ℓ and F_g are the net frictional forces per unit length on liquid and vapor phases, respectively.

Finally, the combined energy equation for the two phases can be written as:

$$\frac{\partial}{\partial t}\left[\rho_g \alpha_g \left(h_g + \frac{u_g^2}{2}\right) + \rho_\ell \alpha_\ell \left(h_\ell + \frac{u_\ell^2}{2}\right)\right]$$

$$= -\frac{\partial}{\partial z}\left[\rho_g \alpha_g u_g \left(h_g + \frac{u_g^2}{2}\right) + \rho_\ell \alpha_\ell u_\ell \left(h_\ell + \frac{u_\ell^2}{2}\right)\right] + \frac{\dot{Q}w}{A} + \frac{\partial p}{\partial t} \tag{8}$$

where $\dot{Q}w$ is the rate of heat transfer from the ambient to the fluid inside the pipeline.

For the portion of the pipeline which is in compressed liquid regime, the governing equations simplify into the following form:

Continuity:

$$\frac{\partial \rho_\ell}{\partial t} = -\frac{\partial}{\partial z}(\rho_\ell u_\ell) \tag{9}$$

Momentum:

$$\frac{\partial}{\partial t}(\rho_\ell u_\ell) = -\frac{\partial p}{\partial z} - \frac{\partial}{\partial z}(\rho_\ell u_\ell^2) - \frac{F_\ell}{A} \tag{10}$$

Energy:

$$\frac{\partial}{\partial t}\left[\rho_\ell \left(h_\ell + \frac{u_\ell^2}{2}\right)\right] = -\frac{\partial}{\partial z}\left[\rho_\ell u_\ell \left(h_\ell + \frac{u_\ell^2}{2}\right)\right] + \frac{\dot{Q}w}{A} + \frac{\partial p}{\partial t} \tag{11}$$

BOUNDARY CONDITIONS

As mentioned earlier, the flow at the exit will be mass limited. Several models have been developed to predict the mass limited flow rate for annular two-phase flow. Different models predict

different values for the mass-limited flow rate; however, the experimental measurements themselves show such scatter that it is not possible to distinguish unequivocally between the models. The model used in the present work is after Moody (3). Following this model, the slip ratio, $k \left(= \dfrac{u_g}{u_\ell} \right)$, at exit under conditions of mass-limited flow is not an independent variable and is given by:

$$K_M = \left(\frac{\rho_\ell}{\rho_g} \right)^{1/3} \tag{12}$$

and the mass-limited flow rate per unit cross-sectional area is given by:

$$G_M = \left[\frac{2c}{a\,(ad\,+\,2be)} \right]^{1/2} \tag{13}$$

where,

$$a = \frac{K_M}{\rho_g} + x_M \left(\frac{1}{\rho_g} - \frac{K_M}{\rho_\ell} \right) \tag{14}$$

$$b = \frac{1}{K_M{}^2} + x_M \left(1 - \frac{1}{K_M{}^2} \right) \tag{15}$$

$$c = - \left(\frac{1-x}{\rho_\ell} + \frac{x}{\rho_g} \right) \tag{16}$$

$$d = \frac{s_\ell{}'}{s_{fg}} \left(\frac{1}{K_M{}^2} - 1 \right) \, (1-x_M) - \frac{2K_M{}'}{K_M{}^3} \, (1-x_M) + \frac{x_M s_g{}'}{s_{fg}{}^2} \left(\frac{1}{K_M{}^2} - 1 \right) \tag{17}$$

$$e = \left(\frac{K_M}{\rho_\ell} \right)' (1-x_M) + \frac{s_\ell{}'}{s_{fg}} (1-x_M) \left(\frac{K_M}{\rho_\ell} - \frac{1}{\rho_g} \right)$$

$$+ x_M \left[\frac{s_g{}'}{s_{fg}} \left(\frac{K_M}{\rho_\ell} - \frac{1}{\rho_g} \right) + \left(\frac{1}{\rho_\ell} \right)' \right] \tag{18}$$

$$x_M = \frac{1}{1 + K_M \dfrac{\rho_\ell}{\rho_g} \dfrac{\alpha_\ell}{\alpha_g}} \tag{19}$$

Primes indicate the derivative with respect to the pressure. Thus, given local static conditions at the exit, K_M can be evaluated from equation (12) and then x_M can be evaluated from equation (19). Then, a, b, c, d, and e can be evaluated from equations (14) through (18), thus making it possible to evaluate G_M from equation (13). It should be noted, however, that Moody's model is valid for a pure substance only.

The boundary condition at the other end of the pipeline would be dependent upon the assumption made about the operation of the block value. A closed block value would imply zero pressure gradient and velocities at that end.

SOLUTION PROCEDURE

The nature of the governing equations makes it necessary to use an iterative, explicit procedure for the numerical solution. The steps of the solution procedure are outlined below.

a. Guess the updated temperature in the cell

b. Compute the corresponding pressure from the vapor pressure equation and the densities and enthalpies for the liquid and vapor phases using equation of state

c. Compute the corresponding updated volume fractions, α_ℓ and α_g, using equations (5) and (4)

d. Compute the evaporation rate, $\dot{m}_{evp}$, using equation (3)

e. Compute the update velocities for liquid and vapor phase using equations (6) and (7), respectively

f. Check if the update values of the flow and state variables satisfy equation (8) within a specified accuracy, failing which, compute a better guess for updated temperature and return to step b.

Step b requires computation of the state properties of the vapor and liquid phases as functions of temperature and pressure. For light petroleum substances, the relations provided by Starling (4) are suitable for this purpose.

For cells which are in compressed liquid regime, the solution procedure has to be modified as follows:

a. Compute the updated density from equation (9)

b. Compute the updated velocity from equation (10)

c. Guess the updated temperature

d. Compute the updated pressure and enthalpy using the equation of state and the updated density and temperature.

3. Check if the updated flow and state variables satisfy equation (11), failing which, compute a better guess for updated temperature and return to step d.

DISCUSSION

A computer program developed to follow the solution procedures outlined above is currently being used to simulate the two-phase blowdown from a liquid propane pipeline. Various differencing schemes are being tested with the objective of obtaining one that would optimize the solution procedure in terms of computation speed as well as stability. With the schemes that have been used, the time step has been limited to only a fraction of a millisecond and the simulation of the blowdown process has been accomplished for about 400 milliseconds. The void fraction and pressure at the exit are found to oscillate with time from the results of this simulation. While there is some experimental evidence that such oscillations are indeed present (5), it is considered desirable to test the program further to ensure that the predicted oscillations are not caused by numerical instabilities.

Along the pipe axis, the liquid-vapor interface is observed to have waves. Further, pressure gradient is rather steep at the point where two-phase region ends and one-phase region begins. Due to these reasons, it appears to be necessary to use a rather fine spatial grid. The stability criterion, then, restricts the time step to very small values and the computation speed has, therefore, been very slow. To simulate the two-phase blowdown of multicomponent mixtures, such as NGL, another computer program has been developed assuming that Moody's model can be used to predict the boundary condition at exit. The computational speed for this program is even slower because phase equilibrium calculations (which are iterative in nature) have to be performed at each stage of the iteration. Further improvement of the existing program is, thus, necessary in order to make it practical to be used for the analysis of hazard resulting from breaks in liquified hydrocarbon pipelines.

REFERENCES

1. Shapiro, A.H., 1953, "The Dynamics and Thermodynamics of Compressible Fluid Flow", Vol. 1, The Ronald Press Co., New York, U.S.

2. Hewitt, G.F., and Hall-Taylor, N.S., 1970, "Annual Two-Phase Flow", Pergamon Press, U.S.

3. Moody, F.J., 1965, "Maximum Flow Rate of a Single Component, Two-Phase Mixture", Journal of Heat Transfer 87, 134.

4. Starling, K.E., 1973, "Fluid Thermodynamic Properties of Light Petroleum Substances", Gulf Publishing Co., U.S.

5. Edwards, A.R., and O'Brien, T.P., 1970, "Studies of Phenomena Connected with Depressurization of Water Reactors", Jour. BNES 9, 125.

THE EVAPORATION OF HAZARDOUS MATERIALS FROM SOLUTION.

M.H. Hilder[*]

Conventional procedures for estimating the rate of evaporation of hazardous materials are based on the evaporation of pure liquids. With materials in solution one must not only take account of the reduced vapour pressure but also of the liquid phase resistance to mass transfer; this can have a marked influence on the rate of release and on the surface concentration of the material and hence on its dispersion in the atmosphere. Experimental and theoretical evidence is presented which should help in estimating the magnitude of these effects.

INTRODUCTION

Conventional procedures for estimating the rate of evaporation of potentially hazardous material (3, 4) are based on the evaporation of pure liquids and obviously do not need to take account of any liquid phase resistance to mass transfer.
The evaporation of (soluble) flammable or toxic materials from solution, although normally less of a potential hazard, sometimes require evaluation as well. If the same conventional procedures are applied, without making any allowance for liquid phase resistance, the results could be wildly inaccurate and could therefore lead to the wrong conclusions being drawn.

The present paper stems from a practical example where the liquid phase resistance was neglected, although the consequences in this particular case were negligible.
The potential emission of acetone from an open tank of 2×8.8 m surface area containing aqueous effluent at $20^{\circ}C$ assumed to be contaminated with 0.1 wt.% acetone was estimated by a conventional procedure (3) to be 5.4 mg/m^{2}s at a wind speed of 2 m/s. The corresponding acetone-in-air concentration at the water-air surface was taken to be equal to the equilibrium concentration, which is 1270 mg/m^3.
This estimate was considered dubious because it did not take account of any liquid phase resistance. Measurements and calculations of the plate efficiency for acetone-water distillations (8) have shown that the liquid phase resistance can be considerable at low acetone concentrations.
Another indication was that a 0.1 wt.% solution of acetone in an open sample bottle did not produce a tell-tale smell of acetone, except shortly after being shaken, although the equilibrium concentration is above the smell threshold.

[*] CROKLAAN B.V., P.O. Box 4, 1520 AA WORMERVEER, The Netherlands