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A Homogeneous Equilibrium Model Improved for Pipe Flows

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Abstract—**The homogeneous equilibrium model improved (HEMI) is for determining the mass flux of all fluids in piping systems. Existing homogeneous equilibrium models tend to underpredict overall pressure drops in piping systems as the flow length is increased. HEMI has corrected the fundamental defect and the associated flow equation is both mathematically and physically satisfactory. The HEMI can also be extended to account for non-equilibrium effects. The resulting calculation sets are compared with experimental data.**

Index Terms— **Homogeneous equilibrium model improved, nonequilibrium effects, pipe flows, pressure relief system.**

I. Introduction

This article presents an improved homogeneous equilibrium model for estimating the mass flux of all fluids in piping systems. Piping systems mean fluids flow in pipes, tubes or ducts at a constant diameter or changing diameters, elevations, and directions. Determining accurate flow capacities and pressure drops in piping systems is important in designing pressure relief systems. The goal of pressure relief systems is to prevent excessive pressure accumulation in a pressure vessel for all credible emergency scenarios. An improperly designed relief system may result in catastrophic failures. Therefore the pressure relief system should be sized with as much certainty as possible for proper protection of the pressure vessel or system. Currently, a homogeneous equilibrium model (HEM) is used extensively in designing the pressure relief system because it gives conservative results (smallest flow capacity). However, the existing homogeneous equilibrium model tends to underpredict overall pressure drops in piping systems as the flow length is increased. The homogeneous equilibrium model improved (HEMI) has corrected the fundamental defect which caused this underprediction.

HEMI requires an accurate correlation of pressure–specific volume. This greatly enhances calculation capabilities when handling compressible fluids in piping systems. The correlation of pressure-specific volume can be obtained by flash calculations. The flow path can be either isenthalpic or isentropic. For two-phase flow, the difference between isenthalpic and isentropic is not significant. HEMI provides accurate and conservative results because the associated flow equation is mathematically and physically satisfactory. Additionally, HEMI can be extended to account for nonequilibrium effects for flashing flow. The resulting calculation sets are compared with experimental data.

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II. Uncertainties of Existing Pipe Flow Models

Sozzi and Sutherland [1] performed extensive tests for high pressure saturated and subcooled water. The test data for low stagnation quality ($0.001 < x_0 < 0.005$, mass fraction of vapor in two-phase) saturated water on Nozzle 2 are of particular interest. Nozzle 2 had a well-rounded entrance (no entrance loss). The test data shown in Table I (selected out of 25 tests in Table V) are for a 1.778 m straight horizontal long run of 12.7 mm diameter tubing. Isenthalpic flash calculations were performed to generate correlations of pressure-specific volume. This is to ensure that the calculation results would be conservative since the isenthalpic flow path yields a lower bound estimate of mass flux. The existing HEM results are obtained using a computer program called CCflow, a flow capacity calculation program for use with the CCPS Guidelines book “Pressure Relief and Effluent Handling Systems” [2]. AIChE Design Institute for Emergency Relief Systems (DIERS) recommends the use of the homogeneous equilibrium model in the pressure relief system calculations.

The experimental data in Table I show that the existing homogenous equilibrium model predicts less conservative mass flux results than expected. In addition, the calculated exit pressures deviate significantly from the observed ones. One could easily calculate the critical or choked mass flux G_c using equation (3) if the exit pressure is known. The difference between the observed mass flux and the calculated critical mass flux at observed exit pressure is, as expected, significant. Generally nonequilibrium results in appreciable mass flux increase. The calculated mass flux values at observed exit pressures support the nonequilibrium hypothesis. Consequently, it is obvious that the significant difference in exit pressures is an indication of nonequilibrium.

Table I
Sozzi and Sutherland Nozzle 2 Test Data vs. HEM Predictions

	Test #1	Test #2	Test #3
Tube Diameter (m)	0.0127	0.0127	0.0127
Tube Length (m)	1.778	1.778	1.778
Stagnation Pressure (MPa)	6.722	6.826	6.826
Stagnation Quality (x_0)	0.0035	0.00455	0.0024
Observed Mass Flux (kg/m^2·s)	17528	17577	17870
Observed Exit Pressure (MPa)	2.379	2.337	2.330
Calculated Mass Flux (kg/m^2·s) with HEM	17518	17611	17782
Calculated Exit Pressure (MPa) with HEM	3.882	3.923	3.944
Calculated G_c (kg/m^2·s) at Observed Exit Pressure	12602	12465	12489

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The HEM results in Table I were also in good with the observed values. Therefore, the mass flux of the HEM was definitely overpredicted since equilibrium conditions gave more conservative results.

III. Homogeneous Equilibrium Model

HEM uses the following pipe flow equation for horizontal pipe flows [3], [4]:

$$G^2 = \frac{\int -\frac{dP}{v}}{\frac{(N+1)}{2} + \ln\frac{v_2}{v_0}} = \frac{\int -\rho dP}{\frac{(N+1)}{2} + \ln\frac{\rho_0}{\rho_2}} = -\frac{\rho_{avg}\Delta P}{\frac{(N+1)}{2} + \ln\frac{\rho_0}{\rho_2}}$$

where G is the mass flux in pipe flows, ρ is the density of the fluid, P is the pressure in pipe flow systems, N is the overall loss coefficient, and v is the specific volume of the fluid. $\int \rho dP$, $\rho_{arithmetic-avg}\Delta P$, is satisfactory mathematically, but not physically since actual average density is computed for n data points of density over a constant interval of pressure as the following:

$$\rho_{actual-avg} = \frac{1}{\frac{1}{n}\sum_{i=1}^{n}\frac{1}{\rho_i}}$$

The HEM provides only a mathematical solution. This means that the HEM does not represent the true compressible flow behavior in piping systems. The arithmetic average density should not be used in the pipe flow equation for compressible flow. But $\int \rho dP$ can easily be seen in textbooks for flow equations for compressible flows. Using the arithmetic average density in the pipe flow equation signals overprediction in mass flux. Therefore, the HEM significantly underpredicts overall pressure drop in the piping system for two-phase flow as shown in Table I. The underprediction in overall pressure drop increases with increasing flow lengths (overall loss coefficient). Also if the back pressure of a pressure relief valve is also to be determined for a given rated capacity of the pressure relief valve, then the homogeneous equilibrium method will underestimate the back pressure. Pressure relief valves are the most commonly used relief device. The pressure relief valve performance and stability are affected by the back pressure. Therefore, a mandatory requirement of ASME - Pressure Vessel Code [5] is that the outlet piping of the relief valve be such that the developed back pressure will not reduce the relieving capacity below the flow required to protect the equipment.

IV. Homogeneous Equilibrium Model Improved

On the other hand, HEMI uses the following pipe flow equation for horizontal pipe flows:

$$G^2 = -\frac{\int vdP}{vdv + \frac{Nv_{avg}^2}{2}} \quad (1)$$

$$v_{avg} = \frac{1}{\Delta P}\int vdP \quad (2)$$

$$G_c^{\,2} = -\frac{dP}{dv} \quad (3)$$

$$G^2 - G_c^{\,2} = \sim 0 \text{ at choked conditions} \quad (4)$$

The pipe flow equation (1) is derived from the Bernoulli equation with proper manipulation not to include $\int \rho dP$. HEMI predicts accurate and conservative estimates of homogeneous equilibrium flow conditions because $\int vdP$, $v_{arithmetic-avg}\ \Delta P$, is both mathematically and physically satisfactory. The actual average specific volume is computed for n data points of specific volume over a constant interval of pressure as the following:

$$v_{avg} = \frac{1}{n}\sum_{i=1}^{n} v_i$$

To illustrate the improved homogenous equilibrium model, the calculation procedure shown in Fig. 1 will be explained in detail below. Fig. 2 is a typical configuration of a pressure relief piping system. The pressure relief piping system often includes a rupture disk. A rupture disk is a non-reclosing pressure relief device actuated by the inlet static pressure.

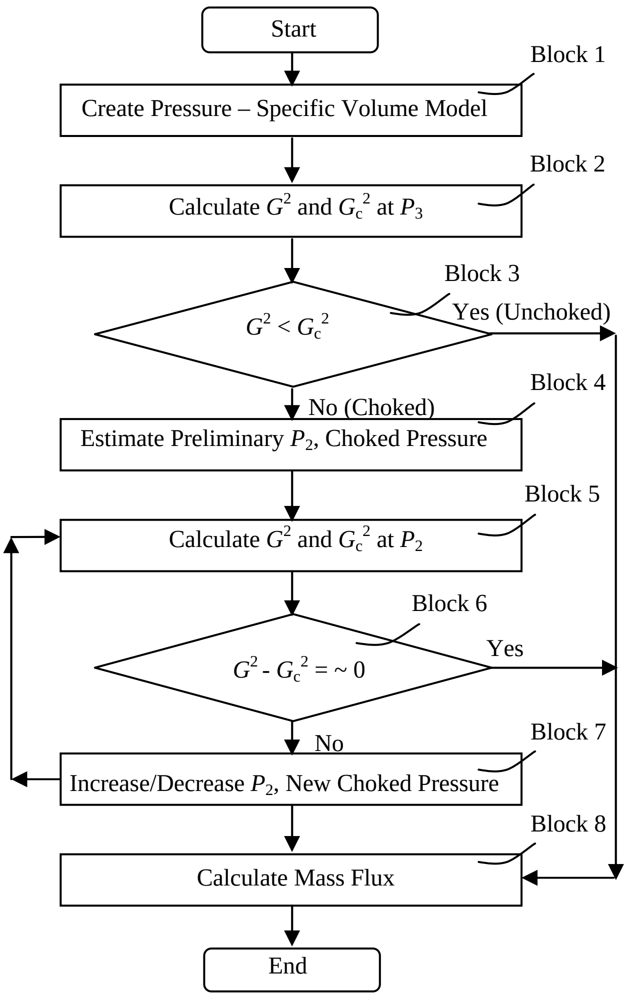

Fig. 1. Procedure Block Diagram for HEMI Calculations

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The flow is from a large upstream reservoir **0** (named station 0), through a constant-area run of pipe entrance **1** (named station 1). The flow from the pipe entrance discharges to the outside **3** (named station 3) of the end of the pipe **2** (named station 2).

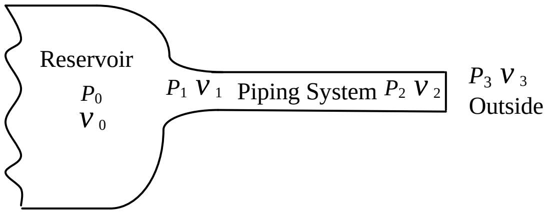

Fig. 2. Typical configuration of a pressure relief piping system.

Fig. 1 is a procedure block diagram for the mass flux and exit pressure calculations. The first step, block **1**, is to create pressure-specific volume models using data obtained by flash calculations on either an isentropic or isenthalpic flow path. A pressure-specific volume model with two constants, α and β, by Simpson [4] as the following fits the data for a broad range of fluids, and is suitable for integration:

$$\frac{v_2}{v_1} - 1 = \alpha\left[\left(\frac{P_1}{P_2}\right)^{\beta} - 1\right] \tag{5}$$

Estimating accurate specific volume is extremely important in calculating the mass flux in piping systems. Single phase or two-phase flow requires at least one *P*-*v* correlation while at least two *P*-*v* correlations in series are recommended for subcooled liquid. For subcooled liquid, the first *P*-*v* model can be based on two data points. The second data point is at the saturated pressure. The pressure-specific volume model "constant" β for the two data points is 1.0. A general guideline for selecting data points is presented as the following:

Table II
Data Point Guidelines for Pressure-Specific Volume Correlations

One P-v model	P_0, $(P_0 + P_3)/2$, P_3
Two P-v models 1st P-v model 2nd P-v model	 P_0, $(P_0 + 0.5P_0)/2$, $0.5P_0$ $0.5P_0$, $(0.5P_0 + P_3)/2$, P_3

The second step, block **2**, is to calculate G^2 and G_c^2 at outside pressure ($P_2 = P_3$). All calculations are often based on two segments: a frictionless section from P_0 to P_1 and a friction section from P_1 to P_2 as shown in Figure 2. More segments can be considered if necessary. Although the pressure range from P_1 to P_2 is more desirable for v_{avg} in flow equation (1), the v_{avg} can be calculated for a pressure range from P_0 to P_2. This does not give significant differences for a large *N* (overall loss coefficient) piping system. The pressure value of P_1 can be obtained by solving the following equation:

$$G^2 = -\frac{\int_{p_0}^{p_1} v dp}{v dv} = -\frac{\int_{p_0}^{p_2} v dp}{v dv + \frac{N v_{avg}^2}{2}} \tag{6}$$

For most pressure relief piping systems, the pipe length *L*, the inside pipe diameter *D*, and the total frictional loss coefficient *K* are known. An overall loss coefficient *N* is calculated as the following:

$$N = 4f\frac{L}{D} + K \tag{7}$$

A Fanning friction factor of f can be taken for fully turbulent flow. More rigorous calculation methods for the friction factor can be considered if the overall loss coefficient is sensitive to the friction factor. For such a case, a simple polynomial equation for fluid viscosities varying with pressure can be developed in the same manner as the P-v model. G^2 is estimated by solving flow equation (1) by either direct integration, numerical integration, analytical integration, or a simple way using direct data points. G_c^2 is estimated using equation (3) by taking the pressure derivative with respect to the specific volume at P_3. G_c^2 can also be estimated at a small pressure increment *dP* as well as the corresponding specific volume change at the small pressure increment.

The third step, block **3**, is to determine if the piping system is choked at station 3 pressure. If the estimated G^2 is greater than the estimated G_c^2, it means the system is choked. Thus, it is required to proceed with the next step. If not, the mass flux is calculated at block **8**. Incompressible fluid is generally not choked. Thereafter, a preliminary choked pressure is estimated, block **4.** A good initial choked pressure ensures fast convergence. The choked pressure can be determined by the point of intersection between curves G^2 and G_c^2. The point of intersection indicates choked conditions. Or simply take around 50% of P_0 as the initial choked pressure estimate.

For the accurate choked pressure, equation (4) is solved by trial and error by changing P_2 until G^2-G_c^2 = ~0 (within error tolerance), blocks **5 - 7**. New P_2 (choked pressure) is (G^2-G_c^2) (Previous P_2) / [(2.0)(G^2)] + Previous P_2. Factor 2, which is adjustable, is to achieve stable convergence. The Newton-Raphson Method can also be used as an alternative for the partial substitution method.

The final step, block **8,** is to determine the homogeneous equilibrium mass flux along with unit conversion as the following:

$G = [G^2(10^6)]^{0.5}$

However, experimental data show that significant nonequilibrium exists for two-phase flow. The nonequilibrium is generally believed to be mostly from thermal nonequilibrium. HEMI can be extended to account for the nonequilibrium effects. The authors are proposing the following preliminary nonequilibrium factor until complete comprehensive correlations are available.

The overall pressure drops in piping systems consist of the following three pressure drop terms:

$$\Delta P_{total} = P_0 - P_2 \tag{8}$$

$$\Delta P_{kinetic} = \frac{1}{2} G^2 v \tag{9}$$

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$$\Delta P_{friction} = \frac{N}{2} G^2 v_{avg} \quad (10)$$

$$\Delta P_{\exp ansion} = \Delta P_{total} - \Delta P_{kinetic} - \Delta P_{friction} \quad (11)$$

Out of the three pressure loss terms, the pressure drop for expansion contributes the most to the nonequilibrium effects in the piping systems. The nonequilibrium factor (NF) can be defined as the following:

$$\text{NF} = (1+ \Delta P_{\exp ansion} \ / \ \Delta P_{total})^{0.5} \quad (12)$$

Thus, the nonequilibrium mass flux (NEMF) and the exit pressure (NEEP) are calculated as the following:

$$\text{NEMF} = (\text{Equilibrium Mass Flux})\ (\text{NF}) \quad (13)$$

$$\text{NEEP} = P_0 - (\Delta P_{total})(\text{NF})^{2.3} \quad (14)$$

The actual pressure drop for nonequlibrium two-phase flow is likely to be greater than the pressure drop which is generally proportional to the square of NF. However, equations (12) and (14) are temporary and valid for the specific test conditions ($0.001 < x_0 < 0.005$ and N=2.61).

V. Example Using HEMI

The following example uses a simplified calculation approach for Test #1 in order to make the calculation procedure easier to follow. The fluid is saturated water with stagnation quality of 0.0035 at 6.722 MPa in a pressure vessel. The following three data points are prepared based on an isenthalpic flow path.

Table III
Physical Properties for Example

Data	P, MPa	v, m^3/kg
#1	6.722	0.0014378
#2	4.568	0.0045886
#3	2.413	0.0146616

Using the three data points and equation (5), the following two simultaneous equations are obtained:

$$\frac{v_2}{v_1} - 1 = \alpha\left[\left(\frac{P_1}{P_2}\right)^{\beta} - 1\right] \qquad \frac{v_3}{v_1} - 1 = \alpha\left[\left(\frac{P_1}{P_3}\right)^{\beta} - 1\right]$$

A simple bisection routine could solve for α and β. The results are:

$\alpha = 3.509350 \quad \beta = 1.255901$

Therefore, the specific volume at any pressure point can be obtained easily with the following pressure-specific volume model:

$$v = 0.0014378\ [3.509350((6.722/P)^{1.255901} - 1) + 1]$$

The following specific volume data are prepared using the pressure-specific volume model above. Average specific volume is computed for n data points of specific volume over a constant interval of 6.8947 Pa (1 psi) as the following:

$$v_{avg} = \frac{1}{n}\sum_{i=1}^{n} v_i$$

Table IV
Specific Volume Data for Example

Pressure, MPa	Specific Volume, m^3/kg	Average Specific Volume, m^3/kg
3.544	0.0076659	0.0038165
3.537	0.0076939	0.0038248
3.530	0.0077221	0.0038333
3.523	0.0077504	0.0038416
-	-	-
2.420	0.0145952	0.0056008
2.413	0.0146616	0.0056153

It is assumed that the overall loss coefficient in the piping system is independent of the Reynolds number and the tube absolute roughness is 0.01016 mm. The value of the absolute roughness of the test tube was not available. Darby *et al.* [3] also used absolute roughness of 0.01016 mm for the test tube. Based on the tube length of 1.778 m and the typical Fanning friction factor of 0.00466, the overall loss coefficient of 2.61 is calculated as the following:

$$N = 4f\frac{L}{D} + K = 4(0.00466)\frac{1.778}{0.0127} + 0 = 2.61$$

First, it is required to check if the piping system is choked at 2.413 MPa assumed as P_3 pressure (tube outside).

$$G^2 = -\frac{v_{avg}\Delta P}{vdv + \frac{Nv_{avg}^2}{2}} = -\frac{0.0056153(2.413 - 6.722)}{\frac{0.0146616^2}{2} + \frac{2.61(0.0056153)^2}{2}} = 162.8$$

$$G_c^{\,2} = -\frac{dP}{dv} = -\frac{(2.420 - 2.413)}{(0.0145952 - 0.0146616)} = 105.4$$

Since G^2 is greater than $G_c^{\,2}$, the flow is choked and it is required to calculate the accurate choked pressure.
For the first trial, the calculated results at P_2 = 3.523 MPa (assuming P_2 = ~ 52% of 6.722 MPa) are:

$$G^2 = -\frac{v_{avg}\Delta P}{vdv + \frac{Nv_{avg}^2}{2}} = -\frac{0.0038416(3.523 - 6.722)}{\frac{0.0077504^2}{2} + \frac{2.61(0.0038416)^2}{2}} = 249.3$$

$$G_c^{\,2} = -\frac{dP}{dv} = -\frac{(3.530 - 3.523)}{(0.0077221 - 0.0077504)} = 247.3$$

$G^2 - G_c^{\,2}$ = 249.3 – 247.3 = 2.0

Further trials are required to refine the results. The new choked pressure can be estimated as the following:

New P_2 = (G^2-$G_c^{\,2}$) (Previous P_2) / [(2.0)(G^2)] + Previous P_2
= (249.3 – 247.3)(3.523) / [(2)(249.3)] + 3.523
= 3.537

For the second trial, the calculated results at 3.537 MPa are:

$$G^2 = -\frac{v_{avg}\Delta P}{vdv + \frac{Nv_{avg}^2}{2}} = -\frac{0.0038248(3.537 - 6.722)}{\frac{0.0076939^2}{2} + \frac{2.61(0.0038248)^2}{2}} = 250.2$$

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$$G_c^{\,2} = -\frac{dP}{dv} = -\frac{(3.544 - 3.537)}{(0.0076659 - 0.0076939)} = 250.0$$

$G^2 - G_c^{\,2} = 250.2 - 250.0 = 0.2$

The difference is so small that further trials are unlikely to improve the results appreciably. Finally, the homogeneous equilibrium mass flux can be determined from the second trial results as the following:

$G = [G^2(10^6)]^{0.5} = [(250.2)(10^6)]^{0.5} = 15818$ kg/m^2·s

The mass flux of 15818 kg/m^2·s at 3.537 MPa exit pressure is essentially a converged solution. It is proven that HEMI predicts conservative results (smallest flow capacity) because the observed mass flux is 17528 kg/m^2·s. In general, using direct integration of the flow equation is preferred over simpler calculation methods such as the one shown in this example. However, the results are reasonably similar.

Using the calculation results from HEMI, the nonequilibrium factor is:

$$\Delta P_{total} = P_0 - P_2 = 6.722 - 3.537 = 3.185$$

$$\Delta P_{kinetic} = \frac{1}{2} G^2 v = \frac{1}{2}(250.2)(0.0076939) = 0.963$$

$$\Delta P_{friction} = \frac{N}{2} G^2 v_{avg} = \frac{2.61}{2}(250.2)(0.0038248) = 1.249$$

$$\Delta P_{\exp ansion} = \Delta P_{total} - \Delta P_{kinetic} - P_{friction}$$
$$= 3.185 - 0.963 - 1.249 = 0.973$$

$$\text{NF} = (1 + \Delta P_{\exp ansion} / \Delta P_{total})^{0.5}$$
$$= (1 + 0.973/3.185)^{0.5} = 1.143$$

The nonequilibrium mass flux and the exit pressure are:

NEMF = (Equilibrium Mass Flux) (NF)
= (15818)(1.143) =18080 kg/m^2·s

NEEP = $P_0 - (\Delta P_{total})(\text{NF})^{2.3}$
= $6.722 - (3.185)(1.143)^{2.3} = 2.391$ MPa

The values for the homogeneous nonequilibrium calculations are in good agreement with the observed ones.

VI. Comparison of Model Predictions

Fig. 3 shows changes in dimensionless mass flux (G/G_{max}) with overall loss coefficient for the Test #1. G_{max} in the dimensionless mass flux is the calculated mass flux for a perfect nozzle. The mass flux of the HEM is clearly shown to overpredict with larger overall loss coefficients (longer runs of pipe). Predictions from HEM and HEMI indicate there are significant differences. Employing an incompatible mathematical model with fluid physics results in less conservative results as shown in Fig. 3. Thus, it is very important that every homogeneous equilibrium model should be confirmed to satisfactorily represent the flow behavior in pipes both mathematically and physically.

Table V is a summary of prediction results for Sozzi and Sutherland test data on Nozzle 2 with the 1.778 m long straight tube (longest flow length) for low stagnation quality ($0.001 < x_0 < 0.005$) saturated water. The experimental data indicate thermal nonequilibrium because the observed exit pressures are significantly lower than the calculated exit pressures of both of the homogeneous equilibrium models.

Table V
Sozzi and Sutherland Nozzle 2 (L= 1.778 m) Data vs. HEM and HEMI Predictions

P_0, MPa	Quality, x_0 at P_0	Observed G, kg/m2·s	Observed P2, MPa	Calculated HEM G, kg/m^2·s	Calculated HEM P_2, MPa	Calculated HEMI G, kg/m^2·s	Calculated HEMI P_2, MPa	Calculated HEMI Nonequilibrium G, kg/m^2·s	Calculated HEMI Nonequilibrium P_2, MPa
6.867	0.00200	19286	2.455	17884	3.971	15970	3.592	18251	2.413
6.833	0.00280	18358	2.441	17762	3.944	15873	3.565	18128	2.399
6.791	0.00310	17919	2.413	17670	3.916	15790	3.544	18036	2.379
6.757	0.00250	17919	2.379	17660	3.909	15770	3.537	18021	2.379
6.722, Test #1	0.00350	17528	2.379	17518	3.882	15658	3.509	17884	2.358
6.681	0.00330	17528	2.365	17464	3.861	15604	3.496	17826	2.358
6.647	0.00340	17528	2.344	17401	3.840	15546	3.475	17757	2.337
6.605	0.00330	17528	2.344	17338	3.820	15487	3.454	17694	2.324
6.936	0.00188	19334	2.482	18011	4.006	16092	3.627	18382	2.441
6.915	0.00301	19334	2.413	17884	3.985	15990	3.606	18260	2.427
6.902	0.00405	19334	2.351	17777	3.964	15907	3.592	18163	2.413
6.881	0.00428	18309	2.358	17728	3.951	15863	3.578	18109	2.399
6.853	0.00430	17772	2.344	17679	3.937	15819	3.565	18055	2.392
6.826, Test #2	0.00455	17577	2.337	17611	3.923	15760	3.551	17992	2.386
6.860	0.00131	19481	2.448	17928	3.971	16005	3.592	18290	2.420
6.833	0.00162	18944	2.448	17826	3.951	15902	3.565	18177	2.386
6.812	0.00172	18944	2.420	17816	3.944	15902	3.565	18172	2.399
6.784	0.00178	18553	2.427	17762	3.930	15853	3.551	18119	2.386
6.757	0.00177	18553	2.413	17718	3.916	15814	3.537	18075	2.379
6.709	0.00197	18553	2.344	17621	3.889	15726	3.516	17972	2.372
6.709	0.00209	18553	2.330	17611	3.889	15717	3.509	17962	2.358
6.860	0.00240	19334	2.386	17840	3.958	15936	3.585	18207	2.413
6.826, Test #3	0.00240	17870	2.330	17782	3.944	15883	3.565	18148	2.392
6.791	0.00280	17870	2.310	17689	3.923	15809	3.551	18060	2.392
6.757	0.00290	17577	2.310	17640	3.902	15760	3.530	18006	2.372

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Although homogenous equilibrium models predict smaller mass flux with lower exit pressure, thermal nonequilibrium ultimately leads to higher flow values. Therefore, it is evident that there is something wrong if the homogeneous equilibrium predictions are quite close to the experimental data. However, preliminary HEMI's nonequilibrium predictions for both the mass flux and the exit pressure are in good agreement with the experimental data. This also verifies that HEMI is a real homogenous equilibrium model that can be extended to account for the non-equilibrium effects.

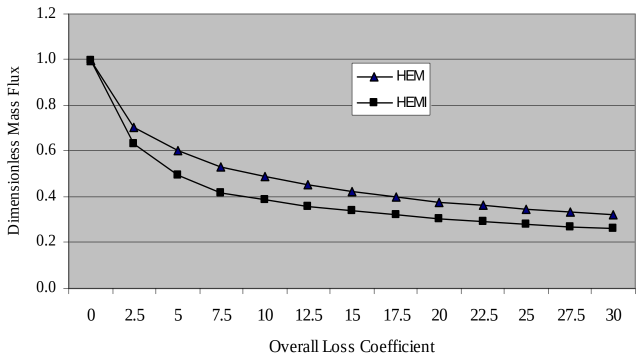

Fig. 3. Changes in dimensionless mass flux with overall loss coefficient for Test #1.

VII. Conclusion

HEMI has several practical advantages over existing homogeneous equilibrium models: it predicts accurate mass flow capacity and pressure drop for an equilibrium flow using a theoretically developed flow equation which is mathematically and physically satisfactory, it yields conservative results as supposed, and it is simple and easy to apply for all fluids at any conditions. HEMI also offers opportunities to revisit previous experimental data to draw out a better solution for nonequilibrium effects.

References

[1] G. L. Sozzi and W. A. Sutherland, "Critical flow of saturated and subcooled water at high pressure," Report NEDO-13418, General Electric Company, San Jose, CA (July 1975).

[2] "Guidelines for pressure relief and effluent handling systems," Center for Chemical Process Safety (CCPS), AIChE, New York (1998).

[3] R. Darby, P. R. Meiller, and J. R. Stockton, "Select the best model for two-phase relief sizing," Chem. Eng. Progress, pp. 56 – 64 (May 2001).

[4] L. L. Simpson, "Navigating the two-phase maze" in International Symposium on Runaway Reactions and Pressure Relief Design, G. A. Melhem and H. G. Fisher, eds., Design Institute for Emergency Relief Systems (DIERS), AIChE, New York, meeting held in Boston, MA, pp. 394 – 417 (August 2-4 1995).

[5] "ASME Boiler and Pressure Vessel Code, Section VIII Div. 1, Pressure Vessels", American Society of Mechanical Engineers, NY (2007 Edition).

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