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AP ACROSS UNIFORMLY DISTRIBUTED-LOSS DEVICES

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**PRESSURE LOSS MODULUS FOR CORRELATION OF STEADY-STATE ΔP
ACROSS UNIFORMLY DISTRIBUTED-LOSS DEVICES**

BY

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ABSTRACT

A dimensionless group, called a Pressure Loss Modulus (N_{PL}), is introduced which, in conjunction with an appropriately defined Reynolds number, is of considerable engineering utility in correlating steady-state ΔP vs. flow calibration data and subsequently as a predictor, using the same or a different fluid, in uniformly distributed pressure loss devices. It is particularly useful under operation in the transition regime. Applications of this simple bivariate correlation to three diverse devices of particular interest for small liquid rocket engine fluid systems are discussed: large L/D capillary tube restrictors; packed granular catalyst beds; and stacked vortex-loss disc restrictors.

* Most of the work herein summarized was accomplished in the author's previous position: Group Supervisor, Combustion Devices and Technology Applications, Jet Propulsion Laboratory, Pasadena, California

INTRODUCTION

In utilizing distributed-loss devices (such as tubing, metal and ceramic foams, stacked etched platelets, and granular packings) in fluid systems, one is frequently faced with the necessity of calibrating with one fluid under some convenient set of conditions while the intended use involves a different fluid and a wide range of operating conditions. The question arises of how best to correlate the data to facilitate the conversion among fluids and to different operating conditions. The problem is further compounded when operation in the laminar/turbulent transition region is required, where the functional relationships are not directly amenable to analytic representation and, hence, the form of the requisite density and viscosity corrections is uncertain.

In this paper, a dimensionless pressure loss modulus parameter, N_{PL} , is introduced, which, together with the Reynolds number, forms a pair of engineering correlation parameters adequate to uniquely define the steady-state pressure-drop-versus-flow characteristic of any uniformly distributed-loss device over the laminar and turbulent flow regimes for (effectively) incompressible fluids. Examples of applications to specific devices, with appropriate interpretation of the constituent variables in each case, are subsequently discussed.

NOMENCLATURE

A	- cross-sectional area
D	- Diameter
d	- differential operator
f	- friction factor (dimensionless) in Fanning equation
G	- mass flux per unit cross-sectional area
g_c	- universal force-mass gravitational conversion constant (not needed with SI units)
K, k	- proportionality constants (dimensionless)\
L	- Length
$\overline{M}$	- average molecular weight of gas
m	- exponent (dimensionless)
N_{Re}	- Reynolds number (dimensionless)
N_{PL}	- pressure loss modulus (dimensionless)
n	- exponent (dimensionless)
p	- average static pressure
Δp	- pressure drop
R	- universal gas constant
S	- surface area
S_s	-specific surface of packing (surface per unit volume)
$\overline{T}$	- average absolute temperature of gas
x	- characteristic linear transverse dimension of flow passage
β	- correction term to Fanning equation for a capillary
ϵ	- void fraction of packing (dimensionless)
μ	- viscosity
v	- number of stages in stack (dimensionless)
ρ	- density
σ	- average sphericity of particle (dimensionless)
$\phi()$	- unspecified function
$\psi()$	- unspecified function

Subscripts

e	- effective
f	- exposed to flow
g	- granule
o	- referred to total cross-section (without packing)
p	- passage minimum
S	- one stage of a stack of disks
t	- tube

APPROACH

Given the conditions of a steady flow of an incompressible fluid through a distributed-loss device, wherein

- effects due to surface tension and body forces are negligible, and
- density and viscosity are essentially constant,

then for any geometrically similar flow passage, the classical fluid mechanics treatment by dimensional analysis yields the relationship

$$\frac{\rho x}{G^2} \frac{dp}{dL} = \phi \left[\frac{xG}{\mu}, \frac{L}{x} \right] \quad (1)$$

where the function ϕ must be determined experimentally. For the applications of interest in this analysis, assume that length does not enter the functional relationship as an independent variable; specifically assume that:

- (a) cross-sectional flow area is constant (or repetitive in the "length" dimension);
- (b) entrance effects are negligible, so that dP/dL is constant in a flow passage and may be replaced by the corresponding ratio of finite increments; and
- (c) flow passages are geometrically similar in cross-section.

Under these assumptions, and with the insertion of g_c to permit the use of engineering units, the foregoing relationship becomes more explicitly

$$\frac{g_c \rho x \Delta p}{G^2 L} = \phi \left[\frac{x G}{\mu} \right] \quad (2)$$

Equation (2) is, in fact, just a rearrangement of the well-established Fanning equation for pipe flow if x is taken to be the pipe diameter, D , and ϕ is identified as twice the "friction factor", $2f$. However, the objective in this instance is to isolate the flow parameter from the pressure drop parameter. Recognizing the group xG/μ as the Reynolds number, N_{RE} , and formally multiplying both members of (2) by its square yields,

$$\frac{g_c \rho x^3 \Delta p}{\mu^2 L} = \left(\frac{x G}{\mu} \right)^2 \phi \left[\frac{x G}{m} \right] \quad (3)$$

Both sides of equations (3) are dimensionless. Note that nothing has as yet been assumed about the nature of the flow regime and hence (3) is valid for laminar, transition, or turbulent flow. The nature of the function, ϕ , depends upon the character of the flow, but the right hand member is a unique function of Reynolds number only for a given device, that is

$$\frac{g_c \rho x^3 \Delta p}{\mu^2 L} = N_{Re}^2 \phi(N_{Re}) = \psi(N_{Re}) \quad (4)$$

If the group on the left of equation (4) is defined as dimensionless pressure loss modulus, N_{PL}

$$N_{PL} = \frac{g_c \rho x^3 \Delta p}{\mu^2 L}$$

equation (4) becomes simply

$$N_{PL} = \psi(N_{Re}) \quad (5)$$

Equation (5) indicates that, providing the few mildly restrictive assumptions are valid, experimental steady-state pressure drop data—taken with varying fluids under different conditions for a uniformly distributed loss device—should all be groupable on a single curve using the parameters N_{PL} and N_{Re} . This curve can be used as a predictor, throughout the flow regime spanned by the data. It obviates questions as to what exponents should be used in density and viscosity corrections, if a different fluid or temperature is of interest.

Frequently, the functional dependence of ϕ upon Reynolds number can be adequately approximated by

$$\phi(N_{Re}) \approx K N_{Re}^{-m}$$

and equation (5) becomes

$$N_{PL} \approx K N_{Re}^n \quad (5a)$$

where $n=2-m$. Form (5a) is particularly convenient because it plots linearly in bilogarithmic coordinates, thus facilitating limited extrapolations to predict behavior beyond the range studied experimentally. The applicability of form (5a) over the more general form (5) must, of course, be experimentally determined in any given instance.

APPLICATIONS

In the sections that follow, the application of the foregoing generalized correlation technique to three specific distributed-loss devices, frequently used in small liquid rocket fluid systems, is discussed: a capillary tube operating in the transition regime; a granular-type catalytic reactor; and a vortex-loss stacked-disk metering device.

1. Capillary Tube

A length of capillary tubing, with an inside diameter of 8.5 mils, was used as both an "orifice" and a flow metering device in a miniature rocket engine. The normal operating range was in the laminar/turbulent transition regime. The working fluid was hydrazine, but calibration was effected with water, and the N_{PL} -versus- N_{Re} correlation technique was used with excellent success. The characteristic x dimension used in both N_{Re} and N_{PL} is the internal tube diameter, D_t , and L_t is the physical length of the tube. Thus,

$$N_{Re} \equiv \frac{D_t G}{\mu}$$

and

$$N_{PL} \equiv \frac{g_c \rho D_t^3 \Delta p}{\mu^2 L_t}$$

Riebling and Powell⁽¹⁾ have shown, through a detailed analysis of a long capillary passage, that equation (2) should contain an L/D correction through an additional term, β , in the form

$$\frac{g_c \rho x \Delta p}{G^2} = (1 + \beta) \phi(N_{Re}) \quad (6)$$

where β is a function of the passage L/D , entrance geometry, and N_{Re} , for the flow regime in question. In this application, however, the tube L/D was large ($\sim 10^3$) resulting in a range of β from 0.02 to 0.03 ($\ll 1$). Hence, the additional complexity of (6) is not warranted and the correlation parameters derived from the simpler form (2) were employed.

The experimental data, plotted in Figure 1, fall on a single curve with minimal scatter, confirming the validity of the assumptions. The curve is nonlinear over most of the range of interest, as anticipated from the range of Reynolds numbers, indicating that equation (5a) does not apply. However, the curve itself is perfectly adequate for correlating the data and calculating desired flow rates, given Δp and the fluid properties.

2. Catalyst Bed

Catalyst beds packed with granular material have wide industrial application. One such is as a combustion chamber for monopropellant rocket engines. The flow characteristics of the beds are of major importance. Data of this type have been correlated by other (2) investigators using semiempirical dimensional equations and it was therefore of interest to compare the use of the correlating parameters described herein. For a packed bed, some interpretation is required in defining N_{Re} and N_{PL} appropriately to account for the packing characteristics. To aid in this interpretation, we temporarily write equation (2) in Fanning form. For pipes with $x = D$

$$\Delta p = \phi(N_{Re}) \frac{G^2}{g_c \rho} \frac{L}{D}$$

and multiplying numerator and denominator of the right member by $\pi D/4$ gives the alternative form

$$\Delta p = \frac{\phi(N_{Re}) G^2}{4 g_c \rho} \left[\frac{\pi D L}{\pi D^2} \right] = \frac{\phi(N_{Re}) G^2}{4 g_c \rho} \frac{S_f}{A_f} \quad (7)$$

Figure 1. Capillary Tube Correlation

Thus, the pressure drop is proportional to the ratio of wetted surface area of the pipe to its transverse flow area. It is reasonable to assume that a similar proportionality holds for the packed bed. In terms of the bed void fraction, ϵ , the transverse flow area is related to the total bed cross section by

$$A_f = \epsilon A$$

and the actual mass flux is consequently related to the mass flux defined for the empty chamber by

$$G = G_0/\epsilon$$

The wetted surface of packing is given by the product of its specific surface and bulk volume

$$S_f = A L S_s.$$

For quasispherical (sphericity, $\sigma \geq 0.7$) granules, the specific surface is related to the mean effective granule diameter, $\bar{D}_g$, by

$$S_s = \frac{6 (1-\epsilon)}{\sigma \bar{D}_g}$$

The characteristic passage dimension, x , should be the mean hydraulic diameter of the interstices between granules. For monosize spherical particles, this dimension is directly proportional to the particle diameter, and for the quasispherical granules employed here this is still a reasonably good assumption. Hence assume

$$x = k \bar{D}_g.$$

Substituting all these relationships back into (7), rearranging, and multiplying through by $(\bar{D}_g/\mu)^2$ yields

$$\left(\frac{\epsilon}{1-\epsilon}\right) \left(\frac{g_c \rho \bar{D}_g^3 \Delta p}{\mu^2 L}\right) = \frac{3}{2\sigma} \left(\frac{\bar{D}_g G_o}{\epsilon \mu}\right)^2 \phi \left[\frac{k \bar{D}_g G_o}{\epsilon \mu}\right] \quad (8)$$

For the packed bed, it is therefore logical on the basis of the groupings in (8) to define a modified Reynolds number and Pressure Loss Modulus by

$$N_{PL} = \left(\frac{\epsilon}{1-\epsilon}\right) \left[\frac{g_c \bar{p} \bar{D}_g^3 \Delta p}{\bar{\mu}^2 L}\right]$$

and

$$N_{Re} = \frac{\bar{D}_g G_o}{\epsilon \bar{\mu}}$$

Wherein the average gas properties

$$\bar{p} = \frac{\bar{P} \bar{M}}{R \bar{T}}, \quad \bar{\mu} = \mu(\bar{T})$$

are used to allow for their variation through the bed.

Data from several reactors using catalysts of various mesh sizes are shown plotted in this form in Figure 2, with Reynolds numbers ranging from about 200 to 2000. The scatter is greater than that of Figure 1, and is probably due primarily to: (a) difficulty in accurately measuring the small differential pressures across the beds in an environment of high absolute pressure, and (b) small L/D for some of the beds, weakening the assumption of negligible entrance effects. Instrument error, compressibility and varying properties of the gas, and uncertainties in packing parameters also contribute to the scatter. The correlation is nevertheless good and it appears that the data can be adequately represented by an equation of form (5a) with $K \approx 20.5$ and $n \approx 1.75$, which result may be compared directly with the pressure-drop equations of reference (2). Additional empirical data are required to firmly establish the validity and Reynolds number limits of applicability of this correlation.

3. Vortex-Loss Metering Device

A convenient device for inserting a known and relatively high pressure drop in a liquid flow system, without the inherent susceptibility to contaminant plugging of a capillary tube or other small flow restriction, is a stacked-disk vortex-loss fluid resistor, such as the "Visco Jet".⁽³⁾ Each disk or "stage" in the unit contains a multiplicity of tiny photo-etched accelerate/decelerate spin chambers series-connected by passages of carefully controlled dimensions, as shown in Figure 3. One such unit, as received from the manufacturer, was calibrated with the intended working fluid (hydrazine) over a range of temperature conditions. It was decided to employ the correlation technique discussed herein to facilitate conversion to operating conditions since the analytic form of the corrections required was not known. In defining the correlating parameters, the characteristic transverse linear dimension, x , was taken to be the minimum passage dimension, x_p , equivalent to its hydraulic diameter. The effective length, L_e , is not immediately apparent. The physical length of the unit could have been used, but a length related to the number of accelerate/decelerate chambers was considered more meaningful. Since the spin chambers in each stage are laid out essentially on a spiral, the flow path length through one stage should be proportional to the overall diameter of a stage, D_s , and for v stages should v times as long.

Thus, taking the definitions

$$x_e = x_p$$

$$L_e = vD_s$$

the working form of the correlating parameters is

$$N_{Re} \equiv \frac{x_p G}{\mu}$$

and

$$N_{PL} \equiv \frac{g_c \rho x_p^3 \Delta p}{v \mu^2 D_s}$$

Figure 2. Catalyst Bed Correlation

Figure 3. LEE VISCO JET[®]

The calibration data are plotted in this form in Figure 4 with negligible scatter. It is evident that these data are well represented by an equation of the form (5a)

$$N_{PL} = 0.25 N_{Re}^{2.37}$$

which is extremely useful as either a pressure-drop predictor, or as a flow-monitor calibration, and considerably more convenient than the treatments recommended elsewhere.

CONCLUSION

The success attained with the N_{PL} - N_{Re} correlation parameters in the three diverse applications cited tends to confirm the general utility of this approach for any uniformly-distributed-loss device. Additional data in these and other applications are required to strengthen this tentative conclusion.

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Figure 4. "Visco Jet" Correlation

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