

BLEVE prevention using vent devices

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Abstract

The main features of accidents involving BLEVEs (boiling liquid expanding vapour explosions) occurring in fires are analysed. Simple mathematical models describing liquid or vapour discharge through the perforation of vessels by superheated liquids and liquid behaviour in a vessel heated by fire are formulated. These models are verified on the basis of experimental data available in the literature. An analysis of the well known accident at Alma-Ata in 1989 is presented. This accident involved the BLEVE of a railway tank engulfed by fire, with the resulting formation of a fireball. It is shown that this accident could have been prevented by means of a vent device (safety valve or breaking diaphragm) with cross-sectional area greater than 77 cm² and operational pressure below 1.6 MPa. The use of such vent devices could prevent the occurrence of BLEVEs when tanks containing superheated liquids or liquefied gases are subjected to thermal loading from fires.

Keywords: BLEVE; Mathematical model; Vent device

1. Introduction

If a tank containing liquid or a liquefied gas is subjected to thermal loading from a fire, an explosion of the tank is possible. Such an event is called a BLEVE (boiling liquid expanding vapour explosion) [1–3]. If the liquid or liquefied gas is combustible, a fireball (a large-scale diffusion flame with strong thermal radiation) is formed. During the destruction of the tank, shock waves of high amplitude are produced. Accidents involving BLEVE are characterised by the severe destruction of plant with people being killed. Such accidents have taken place in Fazen, France (1966), Mexico (1984), and

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Alma-Ata, Kazakhstan (1989) [4]. The serious consequences of BLEVE and the damage to vessels containing LPG subjected to fire have stimulated the attention of many investigators (see for example [5–23]). Some phenomenological peculiarities of accidents involving BLEVE have been revealed during these investigations.

Some methods for the prevention of BLEVE are known [13–15]:

- prevention of fire;
- cooling of tank walls in a fire by dispersed water;
- thermal isolation of tank walls;
- fitting of metal net inside the tank to produce improved heat transfer;
- addition of substances preventing homogeneous nucleation of liquid during the rapid pressure drop in the tank in the initial step of its destruction;
- use of appropriate vent devices.

The first three methods are widely used in practice, but the statistics for occurrence of BLEVE suggest that they are not reliable enough. The application of metal net inside the tanks is rather difficult, and in any case the effectiveness of such nets has not been proved, either theoretically or experimentally. The same can be said of chemical additives preventing homogeneous nucleation. This work addresses the application of vent devices.

2. Preliminary considerations

Let us first analyse the mechanism of occurrence of BLEVE. Heating of a closed vessel in a fire leads to elevation of the temperature of the liquid to values exceeding the normal boiling temperature. The vapour pressure is increased to a value much greater than atmospheric pressure. Heating of the walls of dry tanks causes reduction in the tensile strength of the metal, leading to destruction of the tank. A rapid pressure decrease causes propagation into the liquid of a rarefaction wave, which is followed by a liquid boiling wave with the associated pressure elevation. The accident scenario depends strongly on the liquid temperature T_0 at the instant of destruction of the tank. If T_0 value exceeds the so-called superheating limit [13,14], homogeneous nucleation occurs in the liquid, and the liquid boils in an explosive regime with an associated rapid rise in pressure, so that an explosion occurs. It should be noted that the shock wave that is produced by relief of the pressure in the vessel can be damaging in itself but its maximum overpressure is usually rather small. If T_0 is lower than the superheating limit mentioned above, the liquid evaporates much more slowly. Because the boiling liquid wave is followed by a shock wave and the liquid ceases to boil [10], a violent explosion does not occur.

The superheating limit at various pressures can be drawn on the p - T diagram of a liquid by a spinodal curve (see Fig. 1 for propane). This curve defines the geometric locus described by the formula

$$(\partial p / \partial v)_T = 0 \quad (1)$$

where p is pressure, T is temperature and v is a specific volume.

From the diagram of Fig. 1 it is easy to determine the minimum value of the

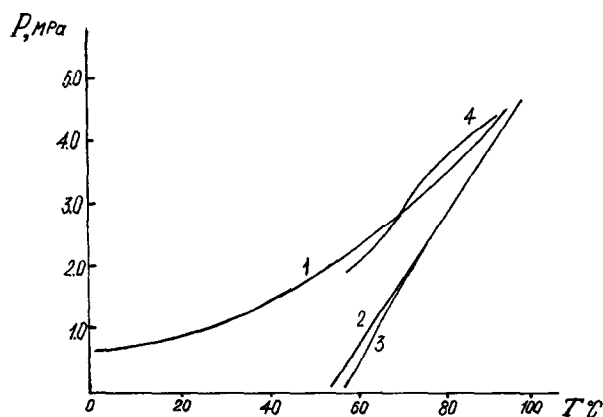


Fig. 1. State diagram of pressure P vs. temperature T for propane. 1, dependence of vapour pressure on temperature; 2, spinodal curve (data from [14]); 3, spinodal curve (calculations by means of Redlich–Quang equation); 4, spinodal curve (calculations by means of Van der Waals equation).

superheating limit, at temperatures below which BLEVE does not take place. This value corresponds to the intersection of the spinodal curve with the abscissa. For propane this minimum superheating limit is equal to 53°C (the corresponding pressure is 1.6 MPa). At propane temperatures below 53°C (pressures lower than 1.6 MPa) BLEVE is impossible.

From this viewpoint, the principal possibility of the prevention of BLEVE can be seen to be the application of vent devices (safety valves or breaking diaphragms) with operational pressures no greater than the pressure P_1 , which corresponds to the minimum superheating limit and having cross-sectional areas large enough to prevent the elevation of pressure in a fire to values higher than P_1 .

To achieve such a method of BLEVE prevention, as a minimum, it is necessary to formulate a reliable and simple mathematical model for a tank holding a liquid or a liquefied gas in the case of a fire. In our model we will not consider in detail heating of the tank walls and their loss of strength, but we shall consider accurate models for processes in the liquid and vapour phases.

Let us first consider calculations of spinodal curves. We have evaluated Eq. (1) for various temperatures using the liquid state Redlich–Quang equation [24]. Note that the application of a simple van der Waals equation gives incorrect results, which differ appreciably from literature data. The results of our calculation for propane are presented in Fig. 1. A good coincidence between calculated and literature data is observed for the case of the Redlich–Quang equation (especially at high temperatures); that is, Eq. (1) with the state Redlich–Quang equation can be used for the evaluation of limiting superheating temperatures.

Because of the possibility of a tank cracking in its lower part, where the liquid is stored, it is necessary to evaluate quantitatively the operation of vent devices and the discharge processes for liquid and vapour phases through them. Therefore let us consider models of the processes mentioned above.

3. Mathematical model

3.1. Modelling of liquid phase discharge

During the discharge of a superheated liquid a two-phase medium is formed at the exit of a tube or perforation. The mass flux G is described in this case by the expression [28]

$$G = \frac{\mu F}{v_a} \sqrt{2(W_0 - W_a)} \quad (2)$$

where μ is a discharge coefficient, F is the perforation area, $v_a = 1/\rho_a$ is the specific volume of the two-phase medium at the perforation exit, W_0 and W_a are the specific enthalpies of the liquid phase and the two-phase medium respectively, and ρ_a is the density of the two-phase medium. It must be noted that in a real fluid phase disequilibrium can occur, or the phases can move at different velocities depending on the flow regime, but these effects are not taken into account in our simple model.

The values of ρ_a , W_0 and W_a can be determined from the reference literature (see for example [25]) taking into account that the value of W_0 value is taken at the initial liquid temperature T_0 and those of ρ_a and W_a are taken at the normal boiling temperature T_b . It is proposed that a discharge occurs under adiabatic conditions, and the fraction η of the liquid phase that evaporates instantaneously is described by the formula [1]

$$\eta = C_{pl}(T_0 - T_b)/L_{ev} \quad (3)$$

where C_{pl} is the specific liquid thermocapacity coefficient and L_{ev} is the specific heat of evaporation.

Let us first consider the equation for discharge of liquid from a cold tank with the following premises: the temperature of the liquid and the vapour phase is the same, temperatures are uniformly distributed in the tank volume, and heat exchange takes place between the tank and the ambient air.

The changes in the masses m_l and m_v of the liquid and the vapour phases with time t are described by the equations

$$\frac{dm_l}{dt} = -G - \frac{dm_g}{dt} \quad (4)$$

$$\frac{dm_g}{dt} = \frac{1}{1 - (\rho_g/\rho_l)} \left(V_g \frac{d\rho_g}{dt} + \frac{\rho_g G}{\rho_l} + \frac{\rho_g m_l}{\rho_l^2} \frac{d\rho_l}{dt} \right) \quad (5)$$

where V_g is the vapour phase volume, and ρ_g , ρ_l are the vapour and liquid phase densities at temperature T .

The value of $d\rho_g/dt$ is described by the equation

$$\frac{d\rho_g}{dt} = - \frac{1}{v_g^2} \frac{dv_g}{dT} \frac{dT}{dt} \quad (6)$$

where v_g is the specific volume of the vapour. The value of dv_g/dT value can be

calculated by means of empirical relationships from the reference literature (see for example [25]). If the process occurs slowly, there is a need to consider the possible preferential boil-off from mixtures (for example, a mixture of propane and butane). In our simple model this effect is not taken into account. Any error arising from this approximation is not high (as will be shown later by a comparison of theory and experiment). A further proposition is that the specific volume of the vapour refers to the value on the saturation line. For liquefied petroleum gases the dependence of v_g on T is described by the empirical formula

$$v_g = v_0 \exp(T_0/T) \tag{7}$$

where v_0 and T_0 are constants for a given substance.

Substituting Eq. (5) and Eq. (6) into Eq. (4) and taking into account that $\rho_g/\rho_l = v_l/v_g$ (where v_l is the specific volume of the liquid phase), we obtain

$$\frac{dm_l}{dt} = -G \left(1 + \frac{Z}{1-Z} \right) + \frac{V_g}{v_g^2} \frac{dv_g}{dT} \frac{dT}{dt} \frac{1}{1-Z} - \frac{Z}{1-Z} \frac{m_l}{\rho_l} \frac{d\rho_l}{dt} \tag{8}$$

where $Z = \rho_g/\rho_l$.

The dependence of V_g on time can be described by an approximate equation

$$\frac{dV_g}{dt} = Gv_l \tag{9}$$

which gives satisfactory accuracy for a cold tank, in which the dependence of v_l on temperature is rather low [25].

In describing the dependence of the temperature T on time we have taken into account that the liquid phase enthalpy changes by the evaporation and by heat exchange with the surrounding air, and we obtain

$$C_{pl} m_l \frac{dT}{dt} = -L_{ev} \frac{dm_g}{dt} + \alpha F_t (T_a - T) \tag{10}$$

where α is the heat exchange coefficient, F_t is the tank surface area, and T_a is the temperature of the surrounding air.

The Z value is usually a small parameter. The value of $d\rho_l/dt$ is also usually small, and therefore the terms with $d\rho_l/dt$ in Eq. (5) and Eq. (8) can be neglected. Substituting Eq. (5) into Eq. (10) and taking into account the above mentioned proposition (including the fact that $Z \ll 1$), we find the equation describing the time dependence of the temperature

$$\frac{dT}{dt} = \frac{1}{C_{pl} m_l - L_{ev} (V_g/v_g^2) (dv_g/dT)} \left[-\frac{v_l}{v_g} G L_{ev} + \alpha F_t (T_a - T) \right] \tag{11}$$

The proposed model was verified by means of experimental data [26]. In [26], experiments were performed in which a mixture of propane and butane was discharged from a vessel of volume 50 dm³ having a perforation in its lower part. The calculated and experimental results are presented in Fig. 2. Also in Fig. 2, for comparison, the results calculated by means of the well known Bernoulli formula are shown. The

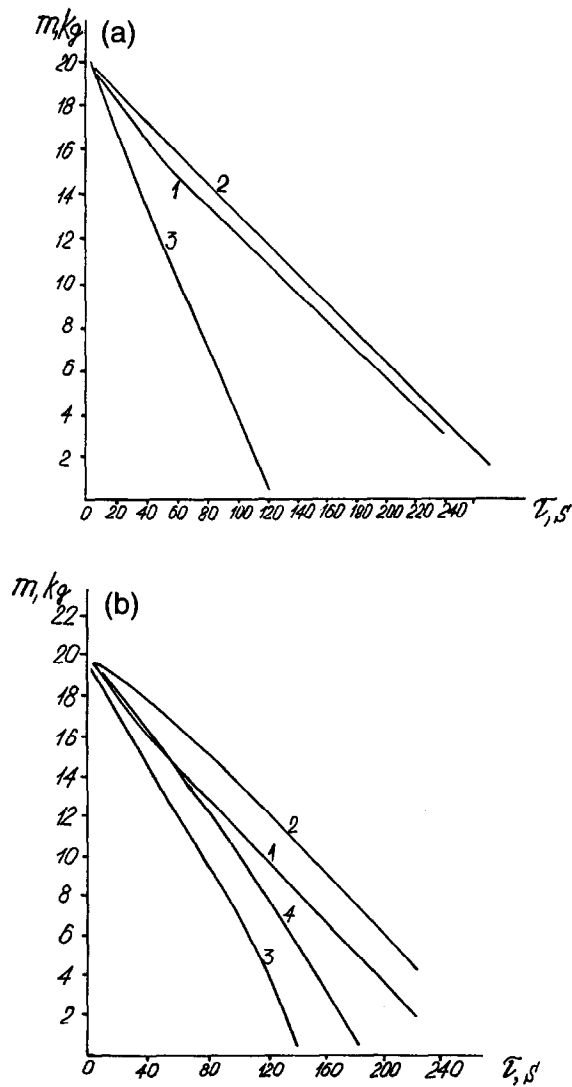


Fig. 2. Dependence of LPG mass m in a tank on time t for LPG compositions: (a) propane 80.7% and butane 19.3% by mass; (b) propane 59.6% and butane 40.4% by mass. 1, experiment [26]; 2, calculation results from this work for perforation diameter $d = 4$ mm; 3, calculations by the Bernoulli formula for $d = 4$ mm; 4, calculation results from this work for $d = 5$ mm.

Bernoulli formula was applied instead of Eq. (2), Eqs. (5), (8) and (11) were solved numerically, proposing that pressure and temperature are connected by the saturation condition. Data for perforation diameter $d = 5$ mm are given for comparison. It can be seen that results calculated by the proposed method are in good agreement with the experimental values, in contrast to those obtained from the Bernoulli formula. It must be

noted that, according to our calculation, the temperature drop of the liquid in the vessel does not exceed 15°C (experimental data for temperature have not been determined).

3.2. Behaviour of a tank with superheated liquid in a fire

The model for the description of the behaviour of a tank containing superheated liquid in a fire was created using the same assumptions as in the previous section. Additionally, account was taken of a thermal flow from the fire to the tank which is uniformly distributed over the tank walls. It was also accepted that the mass liquid evaporation rate is equal to the mass vapour discharge rate through a safety valve, that is, $dm_g/dt \approx 0$. The last assumption is confirmed by literature data (see for example [27]) that, after the operation of a safety valve, the initial discharge process is essentially stationary. For the strict description of the process it is useful to consider a rise in level and liquid droplet carryover, but in our simple model these factors are not taken into account.

The time–temperature dependence is described by the equation

$$\frac{dT}{dt} = \frac{1}{C_{pl}m_l} (QF_t - GL_{ev}) \quad (12)$$

where Q is the effective thermal flow intensity to the substance stored in the tank.

The time dependence of the liquid mass is described by the equation

$$\frac{dm_l}{dt} = -G \quad (13)$$

The dependence of the specific vapour phase volume on temperature is described by Eq. (7), and that on the specific liquid phase volume by the empirical expression

$$v_l = v_{l_0} + k(T - T_{l_0}) \quad (14)$$

where v_{l_0} , k and T_{l_0} are constants for a given liquid and can be taken from the reference literature.

The value of G can be described by the well known expressions

- for a subcritical discharge $\{P_a/P > [2/(\gamma + 1)]^{\gamma/(\gamma-1)}\}$

$$G = \mu F \sqrt{\frac{2\gamma}{\gamma-1} p \rho_g \left[\left(\frac{P_a}{P}\right)^{2/\gamma} - \left(\frac{P_a}{P}\right)^{(\gamma+1)/\gamma} \right]} \quad (15a)$$

- for a critical discharge $\{P_a/P < [2/(\gamma + 1)]^{\gamma/(\gamma-1)}\}$

$$G = \mu F \sqrt{\gamma \left(\frac{2}{\gamma+1}\right)^{(\gamma+1)/(\gamma-1)} p \rho_g} \quad (15b)$$

where μ is the discharge coefficient, F is the perforation area, p is the pressure, and γ is the adiabatic constant.

For verification of the proposed model we used the experimental data [27]. In [27] large-scale experiments were carried out in which the behaviour of propane tanks of

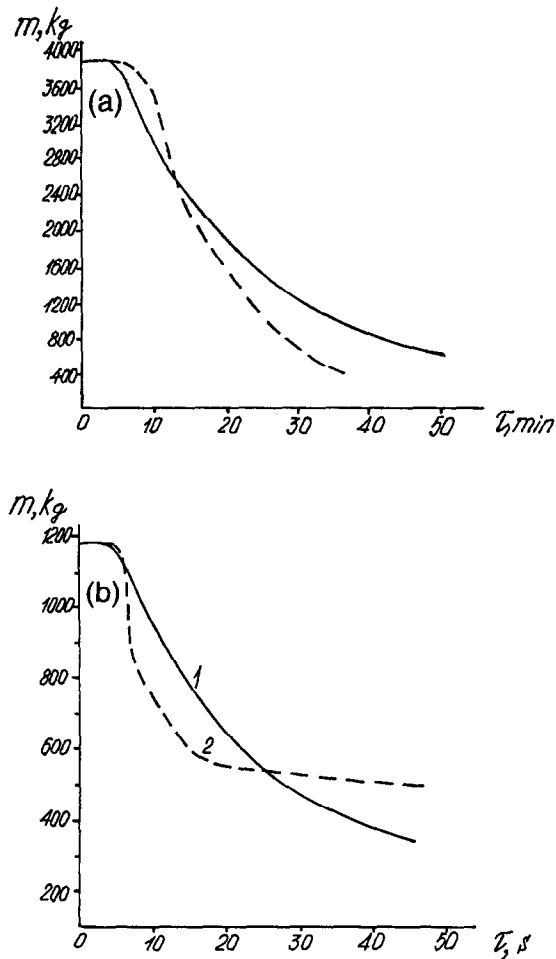


Fig. 3. Dependence of propane mass in a tank m on time t for initial liquid phase volume fractions (a) 72% and (b) 22. 1, calculation results from this work; 2, experiment [27].

volume 10.25 m^3 in a fire was investigated. The operation of a safety valve was taken into account (the safety valve operating pressure was equal to 1.43 MPa).

It was found that Eq. (15a) and Eq. (15b) satisfactorily describe the mass discharge rate at $\mu = 1$. The Q value was accepted as 65 kW m^{-2} . The calculated results and experimental data [27] are presented in Fig. 3. The agreement of the theoretical and experimental results is rather satisfactory.

For the mass discharge rate G of the vapour phase the following empirical correlation from [28] was also used

$$G = F \sqrt{\left(\frac{p_c M}{RT_c} \right) p_c [0.167 p_R^5 + 0.534 p_R^{1.95}]} \quad (16)$$

where p_c is the critical pressure, M is the molar mass, R is the gas constant, T_c is the critical temperature, and $p_R = p/p_c$ is the nondimensional pressure.

The results of calculations with using of Eq. (16) virtually coincide with data obtained with the correlation of Eq. (15a) and Eq. (15b) at $\mu = 1$.

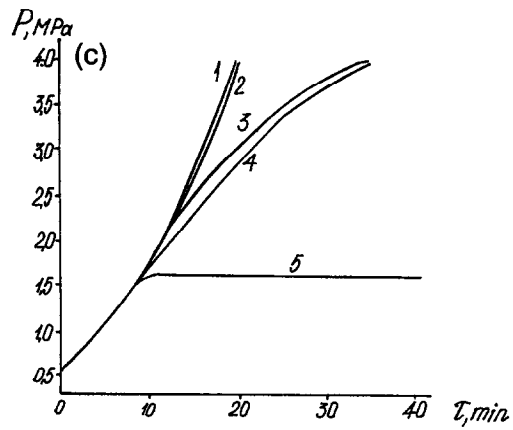
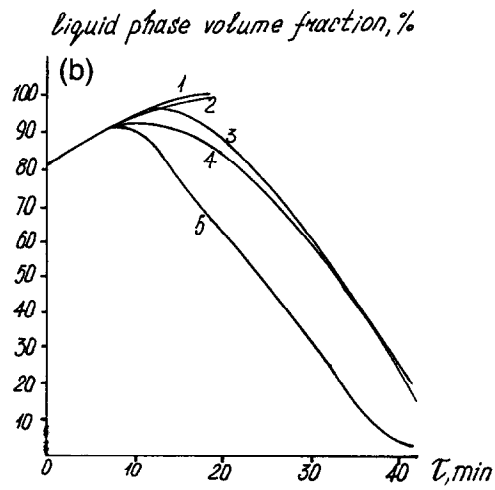
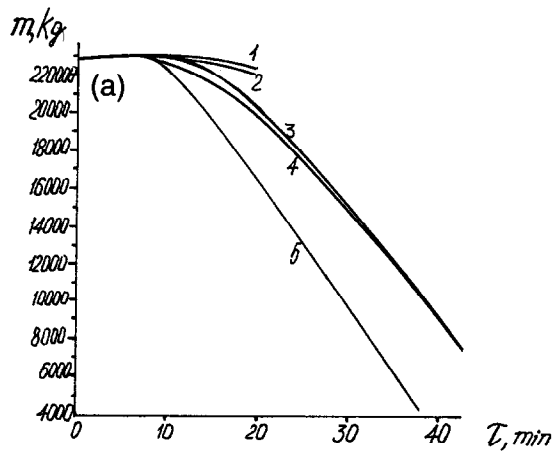
4. Applications of proposed models for analysis of a BLEVE accident (Alma-Ata, 1989) and determination of the required vent devices parameters

The proposed model was used for analysis of the accident involving BLEVE which took place in Alma-Ata, Kazakhstan in 1989, and also for determination of the required cross-sectional areas for vent devices in order to prevent BLEVE. The input data were the following [29]:

- tank volume 54 m³;
- initial liquefied petroleum gas (LPG) mass in the tank 23 t;
- LPG composition by mass propane 45%, butane 55%;
- initial temperature 30°C;
- cross-sectional area of safety valve 77 mm²;
- operating pressure of safety valve 2.26 MPa;
- density of thermal flow on the tank walls 65 kW m⁻².

According to [29], the accident had the following scenario. On the evening of 20 May 1989, because of a serious violation of railway rules, two trains collided, as a result of which one railway tank containing LPG was destroyed. Following the collision, liquid phase flow from a perforation began. Some 3–5 min after the discharge began, ignition of the resulting vapour cloud occurred, and a powerful gaseous torch formed. The fire almost entirely covered an adjacent LPG tank. After a further 15–17 min, a tank explosion (BLEVE) with fireball formation took place, despite the firemen having begun to cool the tank walls by means of water sprays. The fireball during its lifetime caused numerous fires at distances up to 180 m. As a result 200 people were injured and more than 20 people were killed, including nine firemen.

Results of calculations of LPG parameters in the heated tank are presented in Fig. 4, where the dependences of the mass of LPG in the tank, the volume fraction of the liquid phase and the pressure in the tank are presented up to the moment at which the BLEVE began. The mass of LPG in the tank does not change noticeably, and the pressure reaches a value near 4.0 MPa, which is hazardous for the integrity of the tank even without heating of the tank walls. It can be noted that the liquid phase volume fraction increases; that is, the liquid fills almost the whole tank volume. According to [29], after the whole tank volume is filled by a liquid phase, boiling of liquid near the internal tank walls is made difficult. A regime of heat exchange between the tank walls and the liquid phase in the form of bubble boiling changes to a regime of free convection in the closed vessel. In this case, the heat exchange coefficient decreases, with a corresponding increase of tank wall temperature, and the probability of destruction of the tank increases. We have analysed the influence of vent device parameters (cross-sectional area F , operating pressure p_{op}) on tank behaviour in a fire. A decrease of the safety valve operating pressure p_{op} from 2.26 to 1.6 MPa (the limiting pressure for BLEVE



initiation) at a fixed cross-sectional area F does not cause a noticeable change of LPG characteristics (curves 2 in Fig. 4). However, increasing the F value to 770 mm^2 significantly changes the history of the process. The mass of LPG in the tank after safety valve operation falls noticeably, both at an operating pressure of 1.6 MPa and at 2.26 MPa. The liquid phase volume fraction begins to decrease but the pressure increase in the tank does not stop; that is, a safety valve of cross-sectional area 770 mm^2 does not prevent BLEVE. The safety valve operating pressure also does not have any significant influence on the considered process parameters.

At a safety valve cross-sectional area of 7700 mm^2 , after its operation beginning, the LPG mass and the liquid phase volume fraction decrease rapidly but the pressure remains almost constant. The approximate constancy of pressure after safety valve operation begins, in the case of a sufficient cross-sectional area, was experimentally shown in [5,27]. If, in the accident at Alma-Ata railway station, the LPG tank had been supplied by a safety valve with a large enough cross-sectional area (near 7700 mm^2), BLEVE would not have taken place, and destruction of the tank caused by thermal loading would have been much less severe than in the case of the real incident. In reality, if a flame makes contact with the vapour space, a heated vessel may fail at pressures that can be withstood by a cold vessel. If a failure occurs under these conditions, the fireball is less significant, because explosive evaporation of liquid phase will not take place, and the pressure wave and vessel fragments produced during the incident will be not so hazardous. It must be noted that larger relief valves can give rise to larger jet fires after the relief flow is ignited. But the hazard of such jet fires is significantly smaller because of two factors. First, a jet flame occurring after a safety valve begins to operate has a vertical direction and does not immediately make contact with the vessel walls, because the valve is as a rule placed in the upper part of the vessel. Second, thermal radiation from the jet fire is rather smaller than that from a fire around the vessel.

5. Conclusions

1. The main features of accidents with boiling liquid expanding vapour explosions (BLEVEs) in tanks containing superheated liquids or liquefied petroleum gases (LPG) have been considered. The main methods of BLEVE prevention have been analysed.

2. Simple mathematical models of such processes as discharge of a liquid or vapour phase from a tank holding superheated liquid or LPG and liquid behaviour in a tank exposed to fire are proposed. These models are verified by means of experimental data available in the literature.

3. An analysis of an accident involving BLEVE and fireball formation at a railway tank containing LPG subjected to fire (Alma-Ata, 1989) is presented. It is shown that the

Fig. 4. Dependence of LPG mass in a tank m (a), liquid phase volume fraction (b) and pressure p (c) on time t . 1, safety valve cross-sectional area $F = 77 \text{ mm}^2$, operating pressure $P_{\text{op}} = 2.26 \text{ MPa}$; 2, $F = 77 \text{ mm}^2$, $P_{\text{op}} = 1.6 \text{ MPa}$; 3, $F = 770 \text{ mm}^2$, $P_{\text{op}} = 2.26 \text{ MPa}$; 4, $F = 770 \text{ mm}^2$, $P_{\text{op}} = 1.6 \text{ MPa}$; 5, $F = 7700 \text{ mm}^2$, $P_{\text{op}} = 1.6 \text{ MPa}$.

consequences of this accident could have been prevented or significantly reduced by having a safety valve of cross-sectional area not less than 7700 mm² and operating pressure not exceeding 1.6 MPa.

4. The conclusion can be drawn that the consequences of accidents with BLEVE on tanks containing superheated liquids or LPG in the event of fire can be effectively prevented or significantly reduced by means of vent devices (safety valves or breaking diaphragms) which have the appropriate cross-sectional area and operating pressure.

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