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# Spouted Bed and Jet Impingement Fluidization in Food Industry

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## Abstract

Spouted bed and jet impingement fluidization are the cases of classic fluidization modification obtained by proper distribution of the fluid and the construction of device. The condition for the realization of both fluidization types is strict determination of the apparatus dimensions and process parameters. The chapter presents the issues concerning the choice of optimal operating conditions and dimensions of working elements of devices for both spouted bed and jet impingement fluidization, as well as possibilities for analysis of heat transfer in the process. Furthermore, the examples of industrial application of spouted bed fluidization and jet impingement fluidization in food technology are presented.

**Keywords:** spouted bed, jet impingement fluidization, drying, freezing, roasting, food

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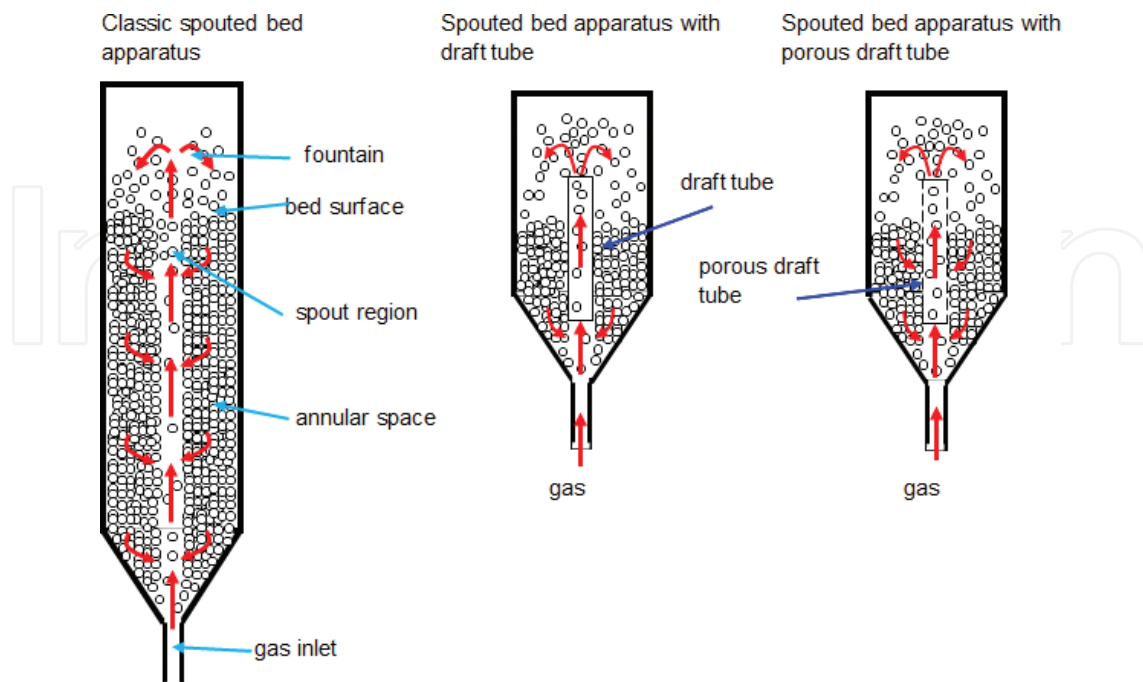
## 1. Introduction

As a result of proper interaction between the gas-solid with the inner construction of fluidized bed apparatus and the mode of gas distribution, modifications of the classic fluidization, such as spouted beds and jet impingement fluidization, are developed. The term spouted bed fluidization was first used by Gishler and Mathur in 1954 who successfully applied this technique for drying of wheat. The first industrial device for drying peas, lentils, and flax was built and installed in Canada in 1962 [1].

## 2. Spouted bed fluidization

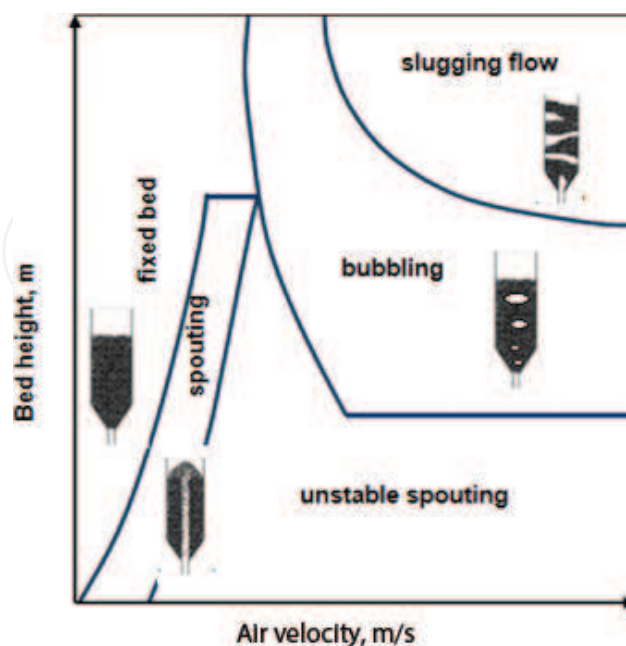
Spouted bed fluidization is the process in which the fluid is vertically introduced upwards at a suitable velocity through the center of the bottom granular material bed. The bed particles are carried up in the central jet by the fluid stream and on reaching the top medium layer they rain back onto the peripheral annular region. Next, they slowly move downward the column. The process cycle is repeated many times causing the bed's circulation (**Figure 1**) [2, 3].

Areas of spouted bed fluidization application overlap with that of classic fluidization use; however, the mechanisms of fluid flow and bed elements motion differ substantially in both cases. The motion and mixing of particles in the spouted bed are regular and cyclic, imposed by the constant upward-flowing fluid jet, opposed to a more random and complicated particle motion in the conventional fluidized bed [4]. There are at least five solutions of devices for spouted bed fluidization: rectangular, cylindrical flat base, cylindrical cone base, conical, and dilute jet spouted beds [5]. In view of the occurrence of bottom dead zones, the flat-base devices are used infrequently. Whereas equipment with conical beds is applied most often in numerous modifications, yet these are beds of low depth. Dilute jet-type spouted beds were studied extensively in the former Soviet Union over the 1950–1960 period [5]. They show a substantially higher fluid velocity than the minimal velocity of spouted bed fluidization that results in dilute phase occurrence in the entire device [2]. To overcome the constraints of conventional spouted beds (low aeration in the annular region, slow turnover of bed particles), there were developed several modifications concerning the process and devices. The most frequently applied are novel rotating jet spouted bed (RJSB) [6], rotating jet annular spouted bed (RJASB) [7], pulsating spouted bed [8], and draft tube spout-fluid bed (DTSFB) [9]. These



**Figure 1.** Scheme of gas motion in conventional spouted bed and spouted bed with draft tube.

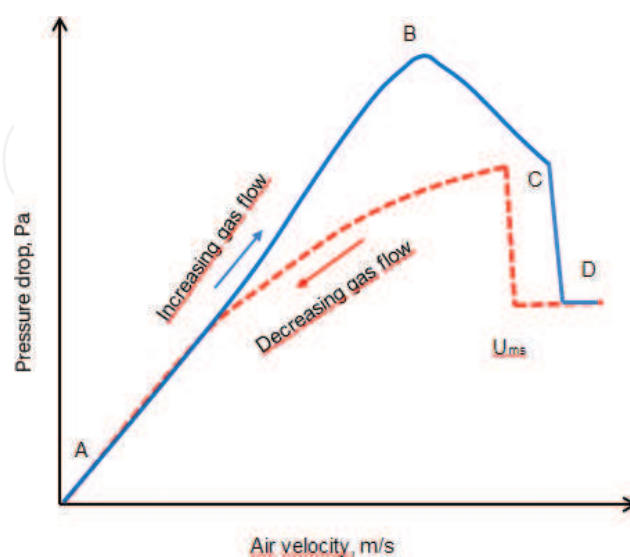
modifications allow to obtain a higher heat transfer coefficient, more uniform distribution of particle residence time inside the device, and lower pressure drops as compared to conventional devices for spouted bed fluidization [10]. One of the most common solutions is using the apparatus with a draft tube where the fountain is formed over the tube top (**Figure 1**). Then neither bed height nor particle size homogeneity is limited and the ease of controlling the bed circulation in order to prevent particles from moving out of the peripheral annular zone to the central core is of importance [11, 12]. However, the application of this solution is restricted to the cases when the medium flowing out of the draft tube is not only the bed particles carrier but plays an active role in a process (drying, cooling, chemical reactions) as well. The reason for this is that the medium from the draft tube cannot percolate to the outer annular zone and therefore, the process is likely to be substantially slowed down through decreased efficiency at lowered heat transfer. To eliminate these shortcomings, a porous draft tube is used. The unique properties of the solution make it applicable, for example, for the thermal disinfestation process of wheat grain where both grain movement and proper conditions of gas-solid contacting with a concurrent slight pressure drop need to be controlled [11]. Another modification of spouted bed fluidization is provided by the Wurster apparatus—a spouted bed device with a draft tube and additional fluidizing air stream (spout-fluid bed). The additional gas stream is injected upwards through the flat or conical porous bottom. The aerodynamic characteristics of the spouted bed system as against conventional fluidization are not only dependent on fluid properties and solid particles making up the bed. The apparatus construction as well as bed height and width are of great importance. Subject to the fluid type and the kind of fluidizing solid particles and apparatus configuration, the types of bed structures/fluidization kinds are best represented by the dependence of bed height on the fluid velocity. **Figure 2** presents the exemplary ranges of the bed structures according to Mathur and Epstein [13].



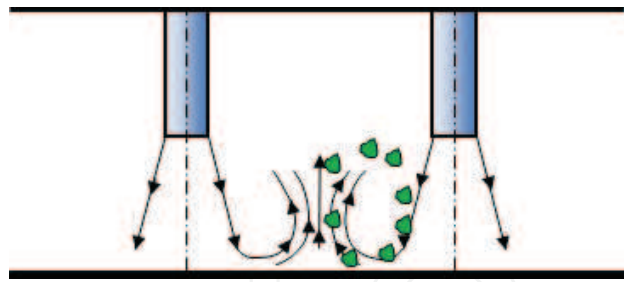
**Figure 2.** Ranges of the bed structures.

The line representing transition between the fixed bed and spouting bed area corresponds to minimum spouted bed velocity, whereas the horizontal line separating spouting from the bubbling area denotes the maximum bed depth. At the height of bed above some value, the transition from the packed bed to spouted bed is not possible. Increasing velocity in this range induces bubble bed formation followed by slugging flow. Working out the technology and spouted bed fluidization devices necessitates the determination of parameters at which the bed is converted into dynamic state as the parameters established for conventional fluidization cannot be used [2]. However, similar to classical fluidization, the dependence between bed pressure drop and fluid flow velocity can be presented (**Figure 3**).

Initially, particles remain stagnant (A-B state) but as the pressure drop increases linearly along with a fluid velocity increase, the so-called dilute zone forming at the bed base is observed to expand. At the B-point, the pressure drop reaches its maximum value while air velocity corresponds to the critical value. The B-C state is characterized by the moment of visible loosening of the bed with bubbling and channeling air flow. At the C-point, fluidization commences and the fluid velocity at the  $U_{ms}$  point is defined as the minimum spouting velocity. The C-D state corresponds to the process of ultimate breaking up of the bed dense phase by the fluid stream. Formation of the characteristic stable fountain of material is observed [14]. The increasing fluid flow rate makes the annular and fountain zones substantially loosened, whereas the further growing velocity results in a uniform bed without division into annular and fountain zones. Thus, the jet-type bed is created [2]. The analysis of the curves run in **Figure 4** shows a pressure drop reduction at increasing and decreasing gas velocity. Minimum spouted bed velocity indicates the onset of the bed stable work. The authors [15, 16] studied a difference in determining the velocity in the cases of increased and decreased gas flow rate. The minimum spouted bed velocity reported at the onset of the observed process is higher than that noted at decreased gas velocity. Some authors [17] established that the most repeatable velocity value of the spouting onset is the value obtained at increasing gas flow rate. Hence, the bottom line serves as the basis for minimum velocity



**Figure 3.** Pressure drop as affected by gas velocity during spouted bed process.



**Figure 4.** Motion of product in jet impingement fluidization.

determination in most papers. On the basis of the relationship between pressure drop and fluid flow rate in the bed, many authors developed the correlations enabling among others, determination of the minimum fluidization velocity [2, 18–20]. For cylindrical devices of up to 0.5 m diameter with/without conical base, the most generalized equation was proposed by Mathur and Gishler [21] to estimate minimum spouting velocity (prediction error within the range  $\pm 15\%$ ) for a wide variety of materials, bed dimensions, inlet diameter, and fluids from air to water (**Table 1**).

More accurate approximations of the minimum fluidization velocity value require knowing a number of process parameters. For instance, Zhong et al. [22] propose the correlation for estimation of minimum fluidization velocity of the bed particle of  $D$  dimension (according to Geldard classification) in the apparatus of cross section of 300 mm  $\times$  30 mm and 2000 mm height (Eq. (2)) and Olazar et al. [23] for materials such as glass or other materials of similar density (Eq. (3)) (**Table 1**). According to Bi [5], bed geometry, which determines the minimum spouting velocity, is sufficiently described by the following three parameters: bed height  $H_o$ ,

Eq. no.	Correlation	Authors
(1)	$U_{ms} = \left(\frac{d_p}{D}\right) \left(\frac{D_o}{D}\right)^{\frac{1}{3}} \sqrt{\frac{2gH_o(\rho_s - \rho_g)}{\rho_g}}$	Mathur and Gishler [21]
(2)	$U_{ms} = 24.5 (2gH_o)^{0.5} \left(\frac{d_p}{D_i}\right)^{0.472} \left(\frac{\lambda_i}{D_i}\right)^{0.183} \left(\frac{H_o}{D_i}\right)^{0.208} \left(1 + \frac{U_f}{U_{mf}}\right)^{-0.284} \left(\frac{\rho_s - \rho_g}{\rho_g}\right)^{0.226}$	Zhong et al. [22]
(3)	$(Re_i)_{ms} = 0.126 \sqrt{Ar} \left(\frac{D_b}{D_o}\right)^{1.68} \left(\tan \frac{\gamma}{2}\right)^{-0.57}$	Olazar et al. [23]
(4)	$(Re_i)_{ms} = 0.202 \sqrt{Ar \left(\frac{D_i}{D_o}\right) \left[\left(\frac{D_i}{D_o}\right)^2 + \left(\frac{D_i}{D_o}\right) + 1\right] / 3}$	Bi [5]

**Table 1.** Empirical equations for predicting minimum spouting velocity.

inlet nozzle diameter  $D_o$ , and diameter on bed surface  $D_t$ . Therefore, the minimum spouting velocity can be estimated from Eq. (4) (for  $D_t/D_o < 1.66$ ). Sutkar and others [24] report that the minimum fluidization velocity increases along with increasing bed height, inlet tube diameter, and operating pressure. Pressure drop in the bed can be defined from the correlation [25]:

$$\frac{\Delta p_m}{H_o p_b g} = 1 + 0.0006 \left( \frac{D_t}{D_o} \right)^{5.04} \left( \frac{d_p}{D_t} - 1 \right)^{-1.92} \left( \tan \frac{\gamma}{2} \right)^{0.7} \quad (5)$$

This equation is valid for  $\gamma = 30\text{--}60^\circ$ ,  $D_o = 0.5$  m,  $H_o = 0.0541\text{--}0.28$  m. Pressure drop in the spouted bed increases along with the increasing bed height, inlet tube diameter, and particle density, while it decreases with elevating particle diameter and gas velocity [24]. Besides the minimum fluidization velocity and pressure drop across the bed, significant factors affecting the operating conditions of spouting beds are geometric parameters of the apparatus [2]. The most critical proved to be a fluid inlet nozzle diameter as well as the nozzle inlet diameter/column diameter ratio. Obviously, a nozzle inlet diameter is dependent on the dimensions of the bed elements. For most cases, the  $D_o/d_p$  ratio should range between 25 and 30 [26], while according to San Jose and others [27] for  $\gamma > 60$  it should be 5:2 at  $2 < D_o/d_p < 60$ , for  $\gamma < 60$  5:2 at  $2 < D_o/d_p < 80$  and for small diameter-solid particles  $d_p = 1$  mm or  $>4$  mm — the ratio should be equal to 2:1. The conical base angle is considered a less critical parameter. For a stable spouting state with mostly solid materials, the ratio should be over  $40^\circ$  [4]. Kmieć and others [2] analyzing the heat and mass transfer in the spouting bed concluded that the problem needs to be addressed separately for the central core zone and fountain zone. Many solutions to this problem are available in the literature, yet universal and generalized dependencies are still lacking. For instance, Rocha and others [25] present the dependence in a dimensionless form:

$$Nu = 0.9892 Re_p^{1.6421} Pr^{0.333} \left( \frac{H_o}{d_p} \right)^{-1.3363} \left( \frac{m_z}{m_g} \right)^{0.71} \left( \tan \frac{\gamma}{2} \right)^{0.1806} \quad (6)$$

which the authors developed studying the granular material drying process in a rectangular-based spout apparatus at the following parameters:  $T = 70\text{--}80^\circ\text{C}$ ,  $75 < Re_p < 844$ ,  $0.1065 < H_o < 0.2456$  m,  $6.49 < d_p < 7.8039$  mm,  $30^\circ < \gamma < 60^\circ$ . Spouted bed fluidization has industrial application in agricultural engineering, chiefly in drying processes. Fountain driers are characterized by high efficiency and performance as well as operational reliability and simple structure. The products dried in such devices tend to maintain their physical properties, sensorial characteristics, and satisfy the microbiological safety requirements [14, 28].

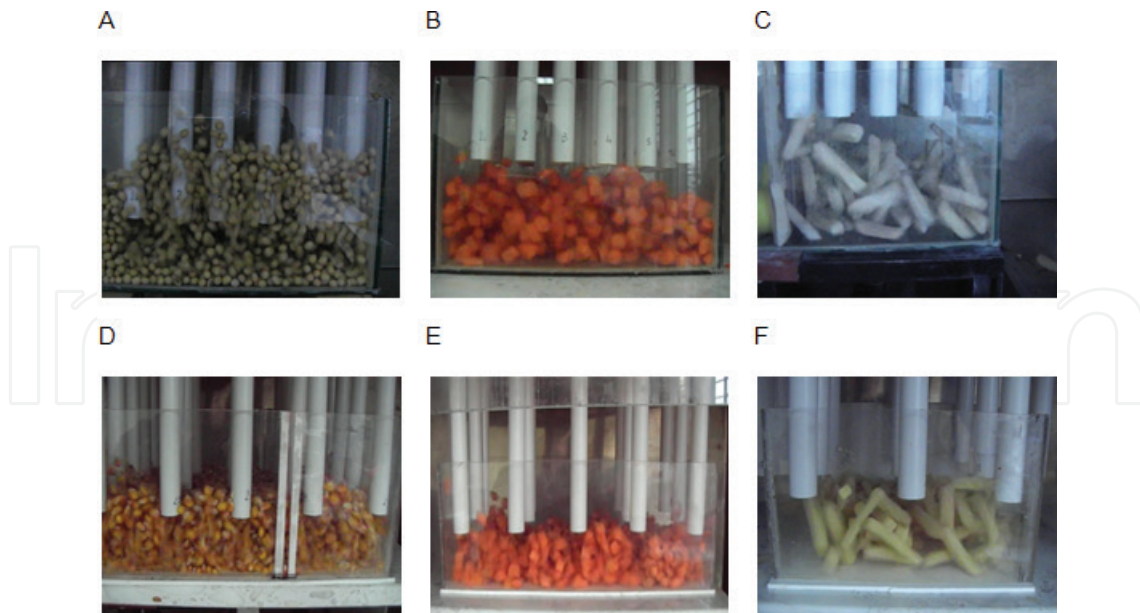
### 3. Jet impingement fluidization

Jet impingement fluidization resembles spouted bed fluidization to some extent. During the jet impingement fluidization process, the bed particles displace in the fluid jet leaving the nozzle system to form characteristic fountains. Explanation of the operating principles of a jet impingement fluidization apparatus needs concerning the formation of a fluid jet flowing out of a single nozzle in the impingement phenomenon. The physical field of fluid jet flow from the nozzle includes three basic regions: free jet flow, stagnation, and horizontal flow or wall jet. The free

jet flow area comprises a few regions whose size depends on the distance between the nozzle and rebound air jet area. This region is divided into two areas, that is, the core and mixing. In the region of the potential core, turbulences do not occur in the jet, whereas they are observed after fluid jet impinging on the stagnation area. Consequently, the flow is disturbed and the jet potential core becomes decentralized. This area is called the mixing region. A level of energy dissipation and length of the fluid jet central core rely mainly on the shape of nozzles and their exit profiling level as well as velocity at the nozzle exit [29, 30]. Vortices in the free jet region along with resulting energy dissipation are the factors of great weight as they contribute to intensive heat transfer in the stagnation area and the stream wall. The stream in the free jet region displays the free flow behavior pattern in the distance equal to 1–1.5 times the nozzle diameter, where upon it starts moving toward the stagnation region. In this region, fluid jet impingement occurs causing the fluid stream bouncing off the apparatus bottom and consequently, it experiences a velocity drop with a considerable increase in static pressure. A negative pressure gradient in the jet wall area induces the initial rapid flow speed increase followed by momentary stagnation and the final decrease [29]. The flow field of the jet leaving the multiple jet arrays is similar to that of a single impinging jet; it also comprises three regions. The differences, however, concern the jet wall area and depend on the nozzles spacing. The interaction mechanisms occurring in arrays of many jets can be observed in three types of cases [31]. First is related to interference between the neighboring jets in the mixing area. Second concerns the appropriate jet-to-jet spacing that promotes the rising fountain phenomenon. The fountain height is determined by, among others, the nozzle length and flow rate of fluid leaving the nozzles. Third pertains to the air outlet arrangement in the plane transversal to the nozzle direction and causes turbulent flow of fluid jet emerging from the working chamber [30]. Jet impingement fluidization is a method that utilizes the impingement phenomenon to induce fluidal bed boiling. The state of fluidal boiling occurs once the minimum fluidization velocity is achieved, that is the buoyancy force exceeds the gravitational force of bed solid particles, which are lifted toward the fountain top. Owing to the inter-particle force of adjacent bed particulates, single-bed elements are entrained to the fountain top to be afterwards pushed under the lower pressure zone or the air jet issuing from the nozzle. Consequently, solid particles fall back toward the bottom of the apparatus working chamber and the process cycle starts anew (**Figure 4**).

Kluza [32], Kluza and Stadnik [33], Góral and Kluza [34–36] thoroughly examined the effect of the apparatus parameters on attainability of fluidal beds of various products. The authors concluded that if the ratio of nozzle symmetry axis spacing to the inner diameters was  $L/D \leq 1.8$ , the bed boiled in a way atypical for jet impingement fluidization. The product tended to displace asymmetrically and was thrown outside the device. Besides, a change in the spacing ( $1.8 \leq L/D \leq 4$ ) resulted in turbulent displacement of the bed solids followed by the occurrence of “bed loosening” phenomenon. Whereas if the  $L/D > 4$  condition was satisfied, dead zones between the particular nozzles were found which were only slightly affected by the air jet emerging from the nozzles. A detailed analysis of the fluidized bed behavior in the jet impingement fluidization technology [33] showed that the fluidal boiling state occurred when the distance between the symmetry axes of nozzles was equal to four times the inner diameter of nozzle ( $L/D = 4$ ). **Figure 5A–C** presents asymmetric displacement of the product in the case of multiple jet array at  $L/D < 4$ .





**Figure 5.** The view of impingement fluidization of different products [33].

There is a lack of visible typical curvilinear closing trajectories determined by the product. In the case of both, soya beans, maize, carrot cubes and French fries, the material was partly thrown outside the apparatus because of too small separation distances between the nozzles and thus, inability of appropriate return of the rebound air jet. **Figure 5D–F** shows typical trajectories of the product determined by the motion of the gas jet leaving the nozzles and rebound off the device bottom. The character of the bed behavior was obtained after setting and application of the optimum dependence between the nozzle diameter and the distance between the nozzles. On the basis of the bed development analysis in jet impingement fluidization, Kluza and Stadnik [33], Góral and Kluza [36] performed the assembly of 11 operating heads (nozzle diameter from 11 to 35 mm) with a nozzle pitch four times the nozzle inner diameter  $L = 4D$ . The heads were fitted with nozzles in a staggered or aligned configuration. To minimize the losses of velocity and pressure of air leaving the nozzles and the character of its flow, the authors assume that the ratio of nozzle length  $S$  to its diameter  $D$  should be higher than 6 ( $S/D > 6$ ) that will ensure a turbulent character of the air flow [37]. According to the assumption, all the heads were equipped with nozzles of 230 mm length. Góral and Góral et al. [38, 39] reported that the maximum velocities of rebound air jets are obtained using the operating heads with nozzles of 23 and 20 mm diameter, while the lowest at 26 mm diameter and staggered configuration. As for the heads with 11, 14, and 16 mm nozzle diameter, the rebound air jet velocity was similar and ranged from 1 up to 5 m s<sup>-1</sup>. The heads fitted with nozzles of 20 and 23 mm diameter enabled to achieve rebound air jet velocity as high as 18 m s<sup>-1</sup> (**Figure 6**).

The objective of working toward the development of optimum processing conditions and dimensional dependencies of the operating elements of a device that provides appropriate conditions for jet impingement fluidization is to obtain proper bed boiling in its whole volume in a treatment process. For the purpose of full realization of the studied process, it is essential to strictly define the characteristic technical dimensions of the apparatus along with

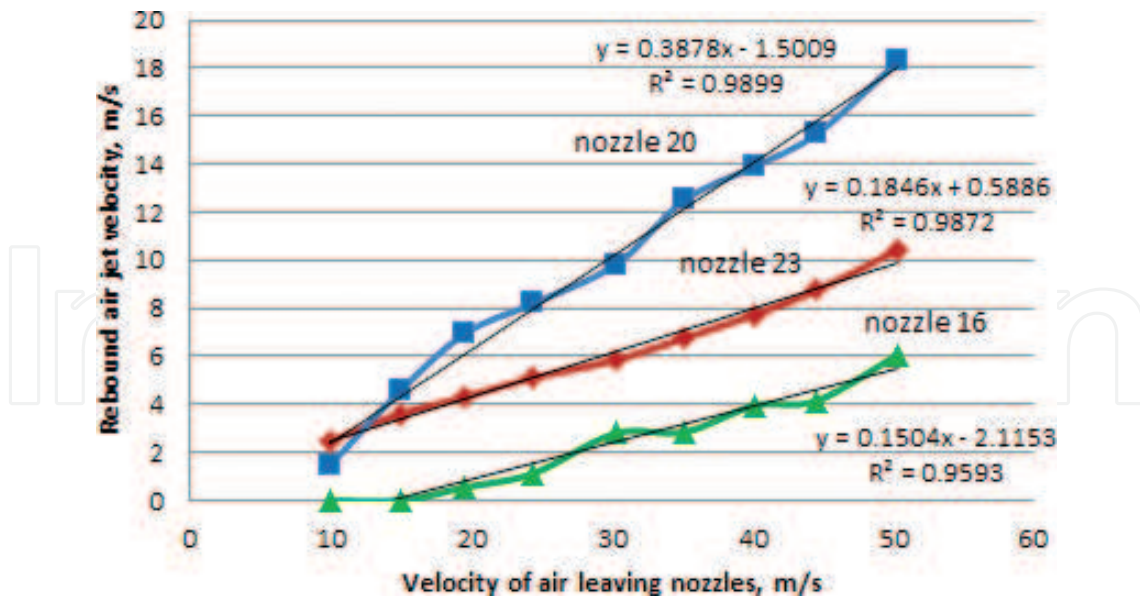


Figure 6. Rebound air jet velocity versus velocity of air leaving nozzles.

process parameters. Notably, for geometrically diversified bed particles, the parameters and dimensional dependencies of the device working space are universal only to some extent. Work on development of optimum conditions for bed fluidal boiling needs to equally consider treatment efficiency, energy saving, and quality of the product. Kluza and Stadnik [33] studied the boiling of carrot cubes beds and found that the greatest heights of beds at rest ensure the subsequent fluidal state with an operating head fitted with nozzles of 20 mm inner diameter in both aligned and staggered pattern (Figure 7).

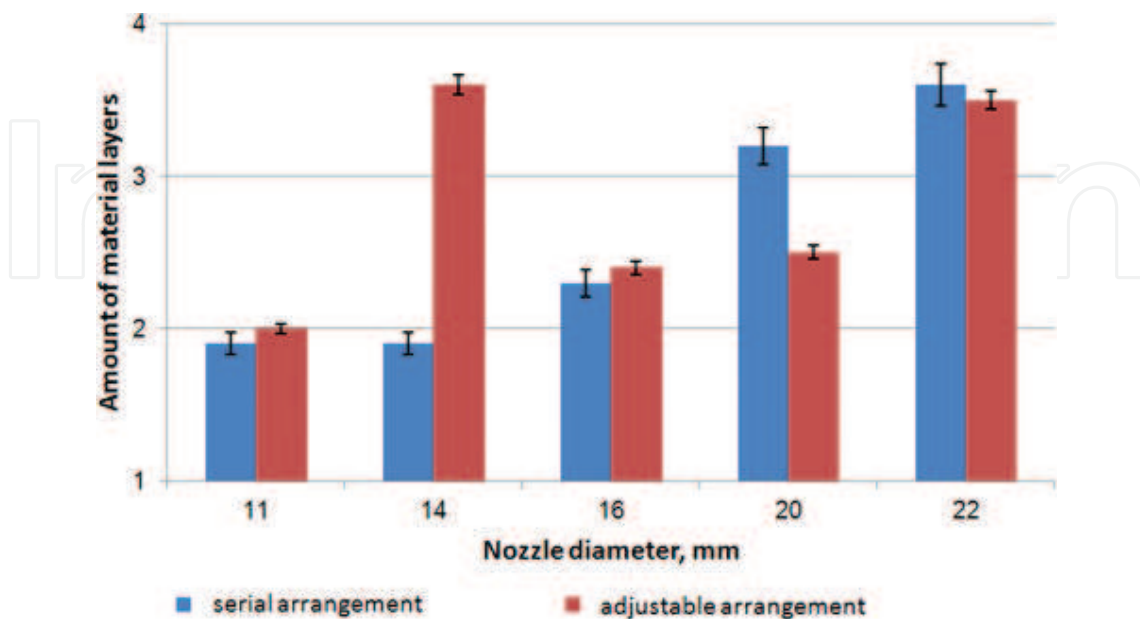


Figure 7. Dependence of bed layers of carrot on inner diameter and arrangement of nozzles.

It was also noted that using heads with nozzles in the in-line arrangement enables fluidization of the beds of the greatest heights. However, for both maize and carrot cubes bed, the best results were obtained with nozzles of 20 mm inner diameter. As for French fries, proper boiling bed phenomenon occurred when the operating heads with nozzles of inner diameter 20 and 35 mm were arranged by aligning. The papers of Góral and Kluza [36] and Góral et al. [39] investigated the problem of heat transfer at vegetable freezing with jet impingement fluidization technique. The authors assumed that the local heat transfer coefficient from the flat plate to air can be estimated from the Nusselt number. In most processing cases, a difference between the value of air temperature at the nozzle exit and of the impinging air jet is so small that the air jet temperature within the bed area can be assumed constant [40]. Prior to the development of a calculation model, Góral and Kluza [36] made the following basic assumptions: air jet flow is turbulent within the velocity range enabling reverse fluidization occurrence, air temperature range varies between  $-50^{\circ}\text{C}$  and  $0^{\circ}\text{C}$ , material shape corresponds to infinite plate and non-stationary character of heat transfer. Assuming a turbulent fluid flow over the plate-shaped product, the estimation of the local Nusselt number characterizing heat transfer commonly involves using the criterial equation [41]:

$$Nu_x = CRe_x^m Pr^n \quad (7)$$

where constants  $C$ ,  $m$ , and  $n$  are determined experimentally. In the investigated process case [36], the values of the Reynolds number were within the 8000–30,000 range and thus, also in the area described as transient flow. Considering the fact that a product moves within the rebound air jet, that is of a changed direction flow, then it can be assumed with high probability that the air flow around the studied product in form of the plate is turbulent. Regarding the scope of air temperature and humidity required in the industrial food-freezing systems, the Prandtl number changes only slightly. Therefore, the Prandtl number from the formula can be included into the constant  $C$  [42], and consequently, the dependence for the forced convection heat transfer can be expressed as:

$$Nu_x = CRe_x^m \quad (8)$$

Besides, as for the exponent  $m$  the value range of  $0.5 < m < 0.8$  should be expected. For the purposes of the model, Góral and Kluza [36] established the  $C$  and  $m$  coefficients empirically during the freezing processes run using the operating head with nozzles of 20 mm inner diameter and the air velocity changing from the determined minimum fluidization velocity to rebound air jet rate equal to  $10.5 \text{ m s}^{-1}$ , essential to maintain a fluidization regime. The obtained Nusselt number values were subjected to regression analysis relative to the corresponding Reynolds number values to get the dependence  $Nu_x = 0.353Re_x^{0.608}$  at the determination coefficient 0.918 [36]. Góral and Kluza [36] compared the experimentally established heat transfer coefficients with those available in the literature. The mean heat transfer coefficients determined varied from 62 up to  $200 \text{ W m}^{-2} \text{ K}^{-1}$  and were lower than  $150\text{--}350 \text{ W m}^{-2} \text{ K}^{-1}$  values [43] obtained at impingement freezing of food, whereas they were considerably higher as against the mean heat transfer coefficients ( $35\text{--}100 \text{ W m}^{-2} \text{ K}^{-1}$ ) determined at the freezing process of the zucchini slices in the fluidized tunnel at  $10 \text{ m s}^{-1}$  air velocity and  $-39^{\circ}\text{C}$  temperature [44]. The paper of Góral and others [45] compares the effect of oval-shaped nozzle exit on uniformity of rebound air velocity fields. The heads fitted with oval nozzles were characterized by non-uniform

distribution of rebound air jet velocity fields. Comparison of the obtained research results and those for the head with round nozzles showed an utterly different character of these fields as they distributed uniformly. However, the values of rebound air jet velocity obtained using all four operating heads were similar and ranged between 2 and 5.6 m s<sup>-1</sup>. The authors assessed the influence of oval cross-sectioned nozzles on attainable boiling bed state studying the French fries freezing process by jet impingement fluidization technique. It was found that fluidal boiling state was achieved using the heads equipped with oval nozzles in the parallel and mixed staggered configuration and at 40 m s<sup>-1</sup> velocity of air jets leaving the nozzles. In the other cases under study, the bed boiling state was not achieved or pneumatic transport of material occurred. Despite several advantages, jet impingement fluidization applicability in food industry is very limited. Industrial equipment based on this phenomenon is mainly used for toasting and puffing of snack pellets, toasting of cereals (corn flakes, bran flakes, rice flakes, etc.), drying of pet foods, toasting and drying of vegetables, toasting of pasta, roasting of coffee beans, peanuts, sunflower seeds, watermelon seeds, sesame seeds, soy beans, corns, grains, drying and baking of bread powder, crouton, and drying of cut meat products [46–48]. Application of the jet impingement fluidization method for freezing and thawing of foods has been still at the laboratory research stage [45].

#### 4. Conclusion

A modification of conventional fluidization through a changed mode of fluid distribution and the construction of apparatus opens new potential application areas for this method. The work and studies on the modification have been carried out in two directions. The first is based on the vertical injection of fluid through the bottom of the device to induce spouted bed fluidization. The other concerns fluid distribution via nozzles as well long ones mounted above the bed. Both modifications have been intensively investigated and implemented. The fluid velocity in both modified methods should be appropriately chosen because of differences in bed particle size and structure and finally, substantial impact on economic conditioning of the realized processes. A too high fluidization velocity is likely to induce considerable worsening of product quality. Therefore, a large body of scientific papers was published and has been still presenting analytical research attempts to determine minimum fluidization velocity. The problem is also related to configuration and dimensioning of operating elements of devices for the realization of both fluidization modifications. These two methods enable better contact between the bed and fluid as compared to other types of fluidization systems. It pertains especially to spouting beds with low resistance of fluid permeability and is associated with enhanced intensity of heat and mass transfer. Introduction of jet fluidization technology for fruit and vegetable freezing allowed product treatment in a very short time. Compared to conventional fluidization, this method enables to obtain far higher heat transfer coefficients reaching 200 W m<sup>-2</sup> K<sup>-1</sup>. As against spouted bed fluidization problem, only few papers address the structure of devices for jet fluidization and they deal with a narrow range of products. Hence, it seems justifiable to state that despite several obvious advantages, applicability of both fluidization modifications still remains studied insufficiently, so working toward apparatus construction development needs further continuation.

## Nomenclature

$Ar$	Archimedes number, $\frac{g d_p^3 (\rho_s - \rho_g) \rho_t}{\mu_g^2}$
$d_p$	equivalent diameter of non-spherical particle calculated as the volume/area surface ratio, m
$D$	diameter of cylindrical column, m/nozzle inner diameters, m
$D_o$	diameter of fluid inlet, m
$D_t$	bed width, m
$g$	gravitational acceleration, $m s^{-2}$
$h$	heat transfer coefficient, $W m^{-2} K^{-1}$
$H_o$	bed depth, m
$L$	nozzle symmetry axis spacing
$Nu$	Nusselt number, $\frac{hx}{k}$
$S$	nozzle length, m
$Re$	Reynolds number, $\frac{d_p U_{ms} \rho_g}{\mu_g}$
$T$	temperature, °C
$U_{ms}$	minimum spouting velocity, $m s^{-1}$
$U_f$	fluidizing gas velocity, $m s^{-1}$
$U_{mf}$	minimum fluidizing velocity, $m s^{-1}$
$x$	characteristic dimension, m
<b>Greek letters</b>	
$\gamma$	cone angle, degree
$\lambda_i$	spout nozzle width, mm
$\mu$	gas viscosity, Pa s
$\rho_s$	density of particles, $kg m^{-3}$
$\rho_g$	density of fluid, $kg m^{-3}$

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