Multitube reactor, especially for catalytic gas-phase reactions

The present invention relates to a multitube reactor and the use of such a reactor for carrying out catalytic gas-phase reactions, in particular for carrying out exothermic and endothermic catalytic gas-phase reactions such as the preparation of phthalic anhydride (PA), acrylic acid (AA), methacrylic acid (MAA), acrolein, maleic anhydride (MA), glyoxal, phosgene, hydrocyanic acid or vinylformamide (VFA).

In the chemical industry, multitube reactors are usually used for carrying out catalytic gas-phase reactions over fixed-bed catalysts.

Multitube reactors usually comprise a reaction tube bundle made up of numerous parallel reaction tubes and arranged within an outer wall. The reaction tubes, which usually contain supported catalysts, have their open ends fixed so as to form a seal to tube plates and each of them opens into a cap connected to the outer wall at the upper or lower end. As well as supported catalysts, the reaction tubes may, alternatively or additionally, contain coated catalysts, unsupported catalysts, ordered packings of catalyst material which have been arranged similarly to a static mixer. It is lastly also possible to coat the inner surface of the reaction tubes with catalyst material. The reaction mixture flowing through the reaction tubes is fed in and discharged via the caps. A heat transfer medium is circulated through the space located between the uppermost and bottommost tube plates and surrounding the reaction tubes, which space may be divided by deflecting plates, in order to introduce or remove heat of reaction. For this purpose, the outer wall of the multitube reactor has means for feeding in and discharging the heat transfer medium, usually suitable annular inlet and outlet channels through which the heat transfer medium is circulated by means of suitable pumps. After leaving the multitube reactor, the heat transfer medium is again brought to a prescribed temperature, for example in an external heat exchanger, before it re-enters the reactor. In the case of exothermic reactions, the temperature may also be controlled using evaporative cooling.

The multitube reactors used in industrial production processes have a diameter of several meters. For economic reasons, as large as possible a number of reaction tubes is used in the reactor. In the case of a reactor having a diameter of several meters, the number of reaction tubes is usually in the range from 10,000 to 50,000, preferably in the range from 10,000 to 30,000 tubes. In

the past, it has been considered important to pack the tubes of industrial multitube reactors as tightly as possible in order to achieve as small as possible a reactor diameter for a maximum number of reaction tubes. The tubes are customarily positioned in a triangular arrangement, usually in an equilateral triangle. A measure used for the compact arrangement of the reaction tubes is the ratio of the tube spacing t to the external diameter d_a of a tube. Here, the tube spacing is the distance from the central internal axes of nearest-neighbor reaction tubes. Known industrial reactors, for instance the reactor described in the examples of DE 44 31 957 A1, have a ratio of tube spacing to external tube diameter of 1.28 or less.

Particularly when carrying out strongly exothermic oxidation reactions, for example the preparation of phthalic anhydride, acrylic acid, methacrylic acid, acrolein, maleic anhydride or glyoxal, precise control of the reaction temperature plays a critical role. In these reactions, a gas mixture is passed through the reaction tubes which contain a fixed bed of a catalytically active multimetal oxide. For example, multitube reactors are used for preparing phthalic anhydride which is an important intermediate for producing synthetic resins, phthalate plasticizers, phthalocyanine dyes and further fine chemicals. The worldwide production of phthalic anhydride is more than 4,000,000 metric tons per year. Most phthalic anhydride is now produced by gas-phase oxidation of o-xylene using air as oxidant. For this purpose, o-xylene is vaporized, mixed with an excess of air and passed at 340-440°C over the catalyst present in the reaction tubes. The catalyst can, for example, comprise a mixture of V_2O_5 and TiO_2 with promoters on ceramic bodies such as porcelain or SiC spheres or rings. Typical dimensions of these ceramic bodies are about 6 mm × 6 mm and 8 mm × 6 mm. In this process, the o-xylene is oxidized to phthalic anhydride with a selectivity of 78-80%. With an enthalpy of reaction of about -1110 kJ/mol, the oxidation is strongly exothermic.

Suitable heat transfer media are, in particular, heat transfer media which are liquid in the preferred reaction temperature range of 250°C to 500°C, preferably 250°C to 380°C. For example, the use of melts of salts such as a melt of a mixture of potassium nitrate, potassium nitrite and sodium nitrite, preferentially used in phthalic anhydride synthesis in particular, is particularly useful.

The reaction conditions, in particular reactor temperature control, require particular attention for a number of reasons: the large number of tubes in the reactor makes it necessary for

the gas mixture flowing into all tubes over the entire cross section to be the same and constant over time, so that the reaction proceeds at the same rate in all tubes and does not proceed particularly quickly in a few preferred tubes. However, the high enthalpy of reaction liberated can, in particular, lead to the catalyst sintering or melting or becoming inactive in individual tubes in the case of deviations from the prescribed temperature range. This is associated with considerable risks for the plant. Inhomogeneities in the throughput also lead to different reaction conditions in the tubes. This results in formation of increased amounts of by-products which reduce the yield and have to be separated from the resulting phthalic anhydride in later purification steps and have to be disposed of. In the gas-phase oxidation, the reaction temperature goes through a maximum along a reaction tube in the flow direction and this maximum is referred to as a hot spot. Such a hot spot is desirable in principle but problems occur if the hot spot temperature is too high, since this leads both to a reduced catalyst life and to a decrease in the selectivity of the reaction.

In principle, effective reactor temperature control therefore has the task of reducing temperature inhomogeneities over the cross section of the reactor and to prevent the occurrence of undesirably high hot spots.

In the case of previously known reactors, which generally have a very small ratio of tube spacing to external tube diameter, effective reactor temperature control was possible to only a limited extent. Particularly in the case of cylindrical reactor geometries, the heat transfer medium in the cross section is passed from a region outside the reaction tube bundle to an inner space of the reactor which is free of reaction tubes or vice versa. This leads to a large pressure drop and thus to a restricted flow of heat transfer medium. In the past, one was therefore forced to use high-performance and consequently very expensive pumping facilities for conveying the heat transfer medium.

It is an object of the present invention to provide a reactor which makes possible a more uniform temperature distribution over the radial reactor cross section and also substantially reduces excessively high heat transfer medium hot spots.

We have found that this object is achieved by the multitube reactor having the features of claim 1. According to the present invention, it is proposed that in the case of relatively large

reactors in which a large amount of heat of reaction is generated owing to the numerous reaction tubes and has to be removed, the ratio of tube spacing t to external tube diameter d_a be made dependent on the reactor diameter or on the external tube bundle diameter d_{RBa}. In particular, the present invention proposes providing a ratio of tube spacing t to external tube diameter d_a of at least 1.3. The reaction tubes are preferably arranged so that three adjacent reaction tubes form a preferably equilateral triangle. The tube spacing t is in this case equal to the side length of the triangle.

The present invention accordingly provides a multitube reactor which has a reaction tube bundle comprising numerous parallel reaction tubes arranged within an outer wall and has means for introducing and discharging a heat transfer medium flowing around the reaction tubes, wherein the ratio t/d_a of tube spacing t to external tube diameter d_a of a reaction tube is at least 1.3.

The multitube reactor of the present invention has numerous advantages.

Increasing the ratio of tube spacing to external tube diameter allows higher heat transfer medium flows at a given pump power and thus leads to a more uniform temperature over the cross section and to a reduction in the heat transfer medium hot spot.

Surprisingly, it is found that it is possible to increase the entry temperature of the heat transfer medium without exceeding the maximum permissible exit temperature of the heat transfer medium. This leads to an improved selectivity for the reaction and an attendant increase in the yield. In this way the capacity of the reactor can be increased by up to 2%.

The reduction in the heat transfer medium hot spot leads to greater operational safety since the danger of ignition of the reaction mixture is greatly reduced.

Specifically the use of the tube bundle reactor of the invention in phthalic anhydride synthesis has revealed that a spacing ratio of more than 1.3 allows a significantly higher loading of the inflowing gas mixture with o-xylene.

The tube spacing is only slightly greater than that in the prior art and also leads to only a slight increase in the reactor diameter. Interestingly, it is found that this slight increase in

the dimensions at a given pump power allows an almost doubled heat transfer medium flow.

According to the present invention, it is also proposed that the ratio t/d_a of tube spacing to external diameter of a reaction tube be increased with increasing external diameter d_{RBa} of the reaction tube bundle. This makes it possible to take account of the heat of reaction to be introduced or removed, which becomes very high with a rising external tube bundle diameter and thus a greatly increasing number of tubes.

The spacing ratio of more than 1.3 proposed according to the present invention is particularly advantageously employed in large reactors. In a first preferred embodiment, the reaction tube bundle has an essentially circular cross section having an external diameter d_{RBa} of more than 4 m. Such a reaction tube bundle usually has a tube-free central region by which the heat transfer medium flowing radially past the reaction tubes can flow away in an axial direction.

In this embodiment, the ratio t/d_a of tube spacing to external diameter of a reaction tube is particularly preferably in the range from 1.3 to 1.6 for reaction tube bundle diameters of from 4 m to 12 m and is most preferably in the range from 1.3 to 1.5 for reaction tube bundle diameters of from 4 m to 10 m.

Multitube reactors having such diameters generally have from 10,000 to 50,000, preferably from 10,000 to 30,000 reaction tubes.

However, the invention is not restricted to reactors having cylindrical reaction tube bundles. If, for example, reaction tube bundles having a rectangular cross section or a circular cross section with end segments free of reaction tubes are employed, the ratio of tube spacing to external tube diameter is preferably dependent on the depth d_{RBt} of the tube bundle through which the heat transfer medium flows transversely. At a spacing ratio of 1.3 according to the present invention, the reaction tube bundle preferably has an essentially rectangular cross section with a tube bundle depth measured parallel to the flow direction of the heat transfer medium of at least 1.3 m.

The depth d_{RBt} of the reaction tube bundle is advantageously from 1.3 m to 3 m, in which case the ratio t/d_a of tube spacing t to the external diameter d_a of a reaction tube is then in the range from more than 1.3 to 1.6.

The reaction tubes are usually made of ferritic steel and have a typical wall thickness of from 1 to 3 mm. Their internal diameter is generally from 20 to 70 mm, preferably from 20 to 35 mm. The typical length of the reaction tubes and thus the length of the cylindrical region of the reactor is in the range from 1.5 to 7 m.

In the case of reaction tube bundles having an essentially rectangular cross section, the ratio of cross-sectional depth d_{RBt} to cross-sectional length of the bundle is preferably from 1:1 to 1:10, particularly preferably from 1:1.5 to 1:4.

The reactor of the present invention enables flows of the heat transfer medium, for instance a salt melt, of from 10,000 to 20,000 m^3 per hour to be achieved.

A reduction in the reaction-related hot spot in the reaction tube increases the life of the catalyst and also improves the selectivity of the reaction. In particular, the lengthening of the catalyst life represents an important advantage of the reactor of the present invention, since the replacement of the catalyst material required after a certain operating time is associated with high costs and long downtimes of the reactor.

The reaction in the tubes of the reactor is frequently controlled via the catalyst pack. For example, a structured bed of two differently active catalysts can be used to create different reaction conditions in the longitudinal direction of the reaction tube. In a further advantageous embodiment of the tube bundle reactor according to the invention, the reactor interior is subdivided (viewed in the longitudinal direction of the reaction tubes) into at least two zones through which differently temperature-controlled heat transfer medium flows. This measure first creates a further reaction control possibility by thermostating the two catalysts at different temperatures. Secondly, it is even possible with some processes to use only one catalyst to pack the tubes and to effect reaction control exclusively via two or more, preferably up to five, differently temperature-controlled zones. Successive zones preferably realize temperature differences of up to 60°C, advantageously up to 40°C, for the heat transfer medium. The individual zones are preferably separated from each other using tube bottom plates which are disposed substantially horizontally in the reactor and possess recesses for accommodating the reaction tubes.

The present invention also provides for the use of a multitube reactor according to the present invention for carrying out catalytic gas-phase reactions.

In particular, the present invention provides for the use of a multitube reactor according to the present invention for carrying out oxidation reactions, in particular for the preparation of phthalic anhydride, maleic anhydride, acrylic acid, acrolein, methacrylic acid, glyoxal, phosgene, hydrocyanic acid or vinylformamide. Such processes using multitube reactors containing fixed-bed catalysts are known (cf., for example, Ullmann's Encyclopedia of Industrial Chemistry, 5th Edition, Volume B4, Table 7 on page 103, with further references).

In the following, the invention is described in more detail with reference to the preferred illustrated embodiments shown in the attached drawing. The preparation of phthalic anhydride mentioned below is presented purely as an example of the use of the reactor of the present invention in oxidative gas-phase reactions and does not imply any restriction of the invention to this application.

In the drawing:

Figure 1 schematically shows a reactor according to the present invention integrated into a plant for producing phthalic anhydride;

Figure 2 shows a longitudinal section of a first embodiment of the reactor of the present invention;

Figure 3 shows a cross section through the reactor of Fig. 2 along the line III-III;

Figure 4 shows an enlarged view of a section of the reactor of Fig. 3;

Figure 5 shows a longitudinal section of a second embodiment of the reactor of the present invention;

Figure 6 shows a cross section through the reactor of Fig. 5 along the line VI-VI;

Figure 7 shows a cross section through a third embodiment of the reactor of the present invention; and

Figure 8 shows a longitudinal section through a fourth embodiment of the reactor of the present invention.

In Figure 1, it is possible to see a schematic outline of a plant for preparing phthalic anhydride. A detailed description of the process may be found, for example, in Ullmann's Encyclopedia of Industrial Chemistry, 5th Edition, Vol. 20A, p. 181 ff.

o-Xylene or naphthalene is vaporized in the vaporizer 10 and mixed with an excess of air supplied via blower 11 and heated to from about 120 to 300°C in a heater 12. The o-xylene/air mixture goes to the reactor 13 where, in the region of the upper cap 14, it is uniformly distributed over the entire reactor cross section. The upper cap 14 is bounded in the direction of the cylindrical reactor wall 15 by an upper plate 16. The reaction tubes 17 of the tube bundle 18 open into the plate 16. The top ends of the reaction tubes 17 are welded to the plate 16 so as to be free of leaks. The catalyst material (not shown) is located in the reaction tubes 17. The lower ends of the reaction tubes 17 are welded to a bottom plate 19 so as to be free of leaks and open into a bottom cap 20 of the reactor 13. The o-xylene/air mixture flows through the reaction tubes and is mostly oxidized to phthalic anhydride. The hot reaction gas is passed via a line 21 to desublimators or separators 22 where the product is deposited in the form of very fine crystals. The phthalic anhydride is melted off the separators 22 and the pure product is obtained in a subsequent distillation apparatus 23 from the crude phthalic anhydride melted off (in accordance with Ullmann's Encyclopedia of Industrial Chemistry, 5th Edition, Vol. 20A, p. 181 ff.).

The temperature of the reaction tube bundle 18 is controlled by means of a heat exchange circuit which is denoted overall by the reference numeral 24. For this purpose, a salt melt of sodium nitrate, sodium nitrite and potassium nitrite is passed into the cylindrical section of the reactor via openings 25 in the wall and is there conveyed longitudinally, cross-flow transversely, in cocurrent or countercurrent past the reaction tubes 17 of the tube bundle 18 in order to remove the heat of reaction generated in the oxidation of o-xylene.

The heat transfer medium leaves the reactor via openings 26 in the wall and goes to an external heat exchanger 27 which is brought to the desired reaction temperature, usually in a range from 340 to 440°C, by means of a steam generator (not shown) via lines 28, 29. The exact choice of reaction temperature depends,

in particular, on the catalyst material used and should be kept as constant as possible.

Figure 2 shows a first embodiment of the multitube reactor of the present invention in more detail. Figure 3 shows a section through the reactor of Figure 2 along the line III-III. Elements which have a comparable function to elements described above in connection with Figure 1 are denoted by the same reference numerals.

The cylindrical reactor 13 has a vertical reaction tube bundle 18 having a circular cross section and an external diameter d_{RBa} (cf. Fig. 3). The reaction tubes 17 are uniformly distributed over an annulus. The central region 18a of the tube bundle 18 is free of reaction tubes.

The heat transfer medium is introduced into or discharged from the space surrounding the reaction tubes through openings 25, 26 in the wall via annular lines 30, 31 by means of one or more pumps 32. A meandering flow of the heat transfer medium is achieved by means of deflecting plates 33 arranged in the reactor, but radial flow prevails in the region of the reaction tubes 17.

In the cross-sectional view of Figures 3 and 4, the tube spacing t, the external diameter d_a of a reaction tube 17 and the external diameter d_{RBa} of the reaction tube bundle 18 are indicated. It should be noted that the diagram in Figure 3 (likewise those of Figures 6 and 7) is not to scale. In reality, the diameter of the reaction tubes 17 compared to the external tube bundle diameter is significantly smaller.

To enable the essential parameters of the multitube reactor of the present invention to be seen more clearly, a section of the tube bundle 18 shown in Figure 3 is depicted on a larger scale in Figure 4. Especially the illustration in Figure 4 reveals that in each case three adjacent reaction tubes form the corner points of an equilateral triangle.

The embodiment of the reactor of the present invention shown in Figures 5 and 6, and also the variant of Figure 7, each have a reaction tube bundle 18 having an essentially rectangular cross section. Such a geometry gives advantages in terms of the transport of heat transfer medium owing to the lower flow resistances, especially in the case of transverse flow cooling.

These advantages are further reinforced by the spacing ratio proposed according to the present invention.

Spaces 34 which are free of reaction tubes for distributing or collecting the heat transfer medium are provided at broad side faces of the reactor located opposite one another. In the variant of Figure 6, the reactor wall 15 itself has a rectangular cross section, while the variant of Figure 7 uses a cylindrical reactor. In the latter case, the essentially rectangular cross section of the reaction tube bundle 18 is obtained as a result of the segments 34 which are free of reaction tubes. According to the present invention, the spacing ratio in the case of essentially rectangular reaction tube bundles is chosen as a function of the depth d_{RBt} of the tube bundle 18 through which transverse flow occurs.

Figure 8 lastly illustrates a fourth embodiment of the tube bundle reactor of the present invention, in a schematic longitudinal section. Reactor 35 is in this case a two zone reactor which, viewed in the longitudinal direction of the reaction tubes 17, is subdivided into two differently temperature-controlled zones 36, 37. The zones 36, 37 are supplied by separate heat transfer medium circuits. In the example, a first salt solution is passed into the first zone 36 via the stubs 38, 39 and passes out again via stubs 40, 41. Correspondingly, a second salt solution passes via stubs 42, 43 into the second zone and passes out again via stubs 44, 45. The two zones 36, 37 are separated from each other via a 50 mm thick tube bottom plate 46. The plate has recesses through which the reaction tubes 17 are inserted. After insertion of the tubes, the tubes are hydraulically expanded somewhat, so that good and substantially tight seating of the tubes 17 in the tube bottom plate 46 is achieved. Within a zone, deflection plates 47 are disposed to guide the salt melt radially from the outside into a reaction-tube-free center of the reactor, where the melt is deflected upwardly and led back again radially to the outside. In Figure 8, the large arrows 48 symbolize the direction of flow of the reaction gases and the smaller arrows 49, 50 symbolize the flow of the first and second salt melts respectively.

Examples

The catalysts used in the examples described hereinbelow were prepared as follows.

Catalyst I:
50 kg of steatite (magnesium silicate) rings having an external diameter of 8 mm, a length of 6 mm and a wall thickness of 1.5 mm were heated to 160°C in a coating drum and sprayed with a suspension of 28.6 kg of anatase having a BET surface area of 20 m^2/g, 2.19 kg of vanadyl oxalate, 0.176 kg of cesium sulfate, 44.1 kg of water and 9.14 kg of formamide until the weight of the layer applied corresponded to 10.5% of the total weight of the finished catalyst (following calcination at 450°C). The catalytically active material applied in this way, i.e., the catalyst shell, consisted of 4.0% of vanadium (reckoned as V_2O_5), 0.4% by weight of cesium (reckoned as Cs) and 95.6% by weight of titanium dioxide.

Catalyst II:
50 kg of steatite (magnesium silicate) rings having an external diameter of 8 mm, a length of 6 mm and a wall thickness of 1.5 mm were heated to 160°C in a coating drum and sprayed with a suspension of 28.6 kg of anatase having a BET surface area of 20 m^2/g, 4.11 kg of vanadyl oxalate, 1.03 kg of antimony trioxide, 0.179 kg of ammonium dihydrogenphosphate, 0.045 kg of cesium sulfate, 44.1 kg of water and 9.14 kg of formamide until the weight of the layer applied corresponded to 10.5% of the total weight of the finished catalyst (following calcination at 450°C). The catalytically active material applied in this way, i.e., the catalyst shell, consisted of 0.15% by weight of phosphorus (reckoned as P), 7.5% by weight of vanadium (reckoned as V_2O_5), 3.2% by weight of antimony (reckoned as Sb_2O_3), 0.1% by weight of cesium (reckoned as Cs) and 89.05% by weight of titanium dioxide.

Example 1: Preparation of PA using a reactor according to the present invention.

A tube bundle having an external diameter of $d_{RBa} = 5.435\ m$ was located in the reactor of the present invention. The tube bundle consisted of about 14,000 reaction tubes made of steel which each had a length of 3.5 m and an external diameter $d_a = 29\ mm$. The tube spacing t was 40 mm; the ratio t/d_a was thus 1.3793.
4 standard m^3/h of air having a loading of 98.5% purity by weight o-xylene of 90 g/standard m^3 were passed through the tube from the top downward. The reaction tubes had been packed in such a way that two differently active catalyst zones were formed. To this end, initially catalyst II is packed into every tube until a bed height (measured from the bottom) of 1.3 m was attained. A bed of

catalyst I having a height of 1.7 m was then applied atop the layer of catalyst II.

The heat transfer medium used was a salt melt of KNO_3,$NaNO_2$ and $NaNO_3$ which was passed through the reactor at 348.9°C and a flow rate of 11,000 m^3 per hour. The exit temperature of the melt was 351.1°C. The hot spot temperature of the salt melt was 3.98°C above the salt inflow temperature.

A plurality of temperature sensors were arranged radially over the cross section at various levels. The temperature differences measured over the reactor cross section were not more than about 2.2°C.

The yield of PA was 78.9 mol%.

Example 2: (Comparative example) Preparation of PA using a reactor according to the prior art.

A tube bundle having an external diameter of d_{RBa} = *5.021 m* was located in the reactor of the prior art. The tube bundle consisted of about 14,000 reaction tubes made of steel which each had a length of 3.5 m and an external diameter d_a = *30 mm*. The tube spacing t was 38 mm; the ratio t/d_a was thus *1.267*. Again, the packing of the reaction tubes - as described in Example 1 - created two differently active catalyst zones.

4 standard m^3/h of air having a loading of 98.5% purity by weight o-xylene of 90 g/standard m^3 were again passed through the tube from the top downward. The heat transfer medium used was a salt melt of KNO_3, $NaNO_2$ and $NaNO_3$ as in the example according to the present invention.

The inflow temperature of the melt was 345.9°C; 6200 m^3 of melt per hour were passed through the reactor. The exit temperature of the melt was 349.7°C. The hot spot temperature of the salt melt was 7.2°C above the salt inflow temperature.

The measured temperature differences over the reactor cross section were not more than about 4.2°C.

The yield of PA was 77.8 mol%.

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We claim:

1. A multitube reactor (13) which has a catalyst tube bundle (18) comprising numerous parallel catalyst tubes (17) arranged within an outer wall (15), said catalyst tube bundle (18) having from 10,000 to 50,000 catalyst tubes (17), and having means for introducing and discharging a heat transfer medium flowing around the catalyst tubes (17), wherein the ratio t/d_a of tube spacing t to the external diameter d_a of a catalyst tube is at least 1.3

2. A multitube reactor as claimed in claim 1, wherein the ratio t/d_a of tube spacing t to the external diameter d_a of a catalyst tube (17) rises with increasing transverse dimensions of the catalyst tube bundle (18).

3. A multitube reactor as claimed in claim 1 or 2, wherein the catalyst tube bundle (18) has an essentially circular cross section having an external diameter d_{RBa} of more than 4 m.

4. A multitube reactor as claimed in claim 3, wherein the external diameter d_{RBa} of the catalyst tube bundle (18) is from 4 m to 12 m and the ratio t/d_a of tube spacing t to the external diameter d_a of a catalyst tube (17) is in the range from 1.3 to 1.6.

5. A multitube reactor as claimed in claim 4, wherein the external diameter d_{RBa} of the catalyst tube bundle (18) is from 4 m to 10 m and the ratio t/d_a of tube spacing t to the external diameter d_a of a catalyst tube (17) is in the range from 1.3 to 1.5.

6. A multitube reactor as claimed in claim 1 or 2, wherein the catalyst tube bundle (18) has an essentially rectangular cross section with a tube bundle depth d_{RBt} measured parallel to the flow direction of the heat transfer medium of at least 1.3 m.

7. A multitube reactor as claimed in claim 6, wherein the depth d_{RBt} of the catalyst tube bundle (18) is from 1.3 to 4 m and the ratio t/d_a of tube spacing t to the external diameter d_a of a catalyst tube (17) is in the range from 1.3 to 1.6.

8. A multitube reactor as claimed in any of claims 1 to 7, wherein the reactor is devided, in the longitudinal direction of the catalyst tubes (17), into at least two zones (36,37), with a flow of heat transfer medium of different temperature being provided in each zone.

9. The use of a multitube reactor as claimed in any of claims 1 to 8 for carrying out catalytic gas-phase reactions.

10. The use of a multitube reactor as claimed in any of claims 1 to 8 for carrying out oxidation reactions, in particular for the preparation of phthalic anhydride, maleic anhydride, acrylic acid, acrolein, methacrylic acid, glyoxal, phosgene, hydrocyanic acid or vinyl formamide.